APPENDIX H:
TECHNOLOGIES FOR BIOFUELS PRODUCTION

RENEWABLE FUELS ROADMAP AND
SUSTAINABLE BIOMASS FEEDSTOCK SUPPLY FOR NEW YORK
Final Report

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GLOSSARY OF TERMS

Term	Definition
CHP	Combined heat and power
TCI	Total capital investment
ABE	Acetone, butanol, and ethanol
HDO	Hydrodeoxygenation
$DeCO_2$	Decarboxylation
SMR	Steam methane reforming
NREL	National Renewable Energy Laboratory
ATR	Autothermal reforming
FT or FTL	Fischer-Tropsch reactions, or Fischer Tropsh Liquids, their product
C5	An alcohol or hydrocarbon having five carbon atoms
C6	An alcohol or hydrocarbon having six carbon atoms
oligomer	an intermediate-length polymer that is itself a building block of a larger polymer chain
monomer	the basic molecular building block of a polymer
polymer	a long-chain molecule consisting of many repeating molecular units
LTCHG	low-temperature, catalytic, hydrothermal gasification
SCWG	supercritical water gasification
CO	carbon monoxide
NOx	oxides of nitrogen
PM	particulate matter, a criterion air pollutant

Unit	Definition
gal	U.S. gallons
ton	U.S. short ton
bu	bushel, a measure of volume with implied measures of mass, depending on material
lb	pounds
tonne	metric ton
g	gram
ppmv	parts per million on a volumetric basis

1 BIOREFINERY CONVERSION TECHNOLOGY OVERVIEW

This project investigates the conversion technologies for liquid biofuels that are currently available, and the technologies under development that are far enough along the development path to potentially be available on a commercial basis in the near and mid-term strategic assessment (circa 2020) time frame. It builds on previous work performed for the Western Governors Association in a *Strategic Assessment of Bioenergy Development* (Antares, 2008). This report offers new material on conversion technologies in the pipeline that have made recent gains in terms of progress and investment, as well as an update of previous work. Five new technologies are added to the technology reviews and to the library of conversion models. Each of the previous technology reviews and models are updated, particularly with regard to capital and operating cost data.

Due to the complexity of biofuel conversion technologies, there is no simple equation to describe production costs and process yields. Spreadsheet models were developed to calculate the costs and yields for selected technologies based on key variables: feedstock type, conversion option, and facility size.[1] The technology models used in the assessment were chosen to be *representative of the types of biofuel production processes that the assessment team believes can be commercialized in the mid-term*. They are not endorsements of any specific technology and the assessment team believes that the actual plant configurations built will likely incorporate features that could not be foreseen in this study. The technology choices for detailed analysis and modeling were based on the relative benefits and challenges for each conversion process, status of the technology, and availability of published engineering data. To have a significant presence in future markets, competing technologies will need to have similar or better yields and similar or lower production costs than the representative technologies evaluated in this report.

1.1 TECHNOLOGY ANALYSIS SCOPE AND FRAMEWORK

Fifteen biofuel conversion technologies are evaluated in this report. The sheer number of technologies under development suggests a high level of interest and diversity of approaches being taken to commercialize biofuels. Brief explanations and process descriptions for each of these conversion technologies are given in Table H-1. The size range gives potential facility sizes expected in the 2020 time frame. Reference name abbreviations are also included in this table. For each of the current and mid-term conversion processes, a technical description of the process, a cost and performance analysis and a short discussion of the outlook for the technology are provided.

This report focuses on technologies that convert solid biomass into liquid fuels, as these form the basis for the assessment and are most likely to be available in the circa 2020 time period. Technologies that are already moving beyond the pilot phase into the demonstration stage now are designated as near-term. Those moving into demonstration in the 2015 to 2025 are classed as “mid-term,” and those likely to become available at the far end of the mid-term (2025 or thereafter) will be called “long term.”

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[1] Some of the technologies have different process options that can affect the yield and performance of the conversion. For example, renewable diesel can be generated via stand-alone or co-processing conditions, and lignocellulosic ethanol production has several pretreatment process options.

Table H-1. Current and Advanced Biofuel Conversion Technologies.

Reference Name (Feedstock Product Process)	Acronym	Feedstock	Conversion Technology	Fuel Type	Size Range* (MGY)	Modeled in this Report?	Description
Current Representative Technologies							
Grain to Ethanol - Dry Mill	GEt-DM	Grains / Starches	Enzymatic Fermentation	Ethanol	5 to 100	Yes	Dry milling process - grains are ground into a flour, and the starch is converted into sugar and then fermented to ethanol.
Grain to Ethanol - Wet Mill	GEt-WM	Grains / Starches	Separation and Fermentation	Ethanol	50 to 300	Yes	Wet milling process - grain separated into components and starch is yeast fermented and distilled.
Fatty Acid to Methyl Ester	FAME	Seed Oil / Waste Oils / Animal Fats	Esterification	Methyl Esters	1 to 80	Yes	Vegetable oils and fats are filtered and converted via base catalyzed transesterification, producing biodiesel and glycerin, which must be separated.
Sugar to Ethanol Fermentation	SEt-F	Sugars	Fermentation	Ethanol	5 to 100	No	Sugar crops such as sugar cane are milled and fermented to produce ethanol.
New Technologies Projected to be in Use by 2015 to 2025							
Lignocellulosics to Ethanol - Enzymatic Hydrolysis/Fermentation	LCEt-H/F	Lignocellulosi Biomass	Hydrolysis and Fermentation	Ethanol	20 to 100+	Yes	Cellulose and hemicellulose converted to sugars via hydrolysis. Various options for hemicellulose conversion (pretreatment). Conversion of sugars to alcohol via fermentation.
Lignocellulosics to Ethanol - Gasification/Fermentation	LCEt-G/F	Lignocellulosi Biomass	Gasification and Fermentation	Ethanol	50 to 100+	No	Gasification to produce syngas, which is then conditioned and compressed. The compressed gas is fermented to ethanol.
Lignocellulosics to Middle Distillates - Fischer Tropsch	LCMD-G/FT	Lignocellulosic Biomass	Gasification and Fischer Tropsch Synthesis	Middle Distillates, Gasoline	5 to 100+	Yes	Gasification to produce syngas, which is then cleaned and purified. The clean syngas then undergoes catalytic synthesis (reactor with Co catalyst to maximize diesel fraction). The product is separated and upgraded.
Lignocellulosics to Mixed Alcohol - Gasification/Thermocatalysis	LCMA-G/TC	Lignocellulosic Biomass	Gasification and Thermochemical Conversion	Mixed Alcohols	15 to 100+	Yes	Syngas production via biomass gasification, followed by catalytic conversion to mixed alcohols.
Lignocellulosics to Mixed Alcohol Digestion/Hydrogenation	LCMA-D/H	Lignocellulosi Biomass	Digestion/ Hydrogenation	Mixed Alcohols	30+	No	Anaerobic digestion of biomass with methanogenic inhibition followed by evaporation and fermentation. Produces a mixture of alcohols, carboxylic acids, ketones, and biofuels.
Lignocellulosics to Gasoline - Pyrolysis/Hydrotreating	LCGa-P/H	Lignocellulosi Biomass	Pyrolysis then Coprocessing via Hydrotreatment / Hydrocracking	Bio-oil, Diesel, Gasoline	5 to 100+	Yes	Biomass conversion to bio-oil via fast pyrolysis, co-processing with fossil fuels in petroleum refinery.
Lignocellulosics to Butanol - Hydrolysis/Fermentation	LCBu-H/F	Lignocellulosi Biomass	Thermochemical Conversion or Multistage Fermentation	N-butanol, Iso-butanol	5 to 100	Yes	Pretreated biomass is fermented in two steps by *Clostridia* (n-butanol) or *E. coli* (iso-butanol), or partially fermented and then hydrogenated.
Hemicellulose to Ethanol: Pulp and Paper Application	HCEt-H/F	Hard- and Soft-woods	Hot Water Extraction Hydrolysis and Fermentation	Ethanol	4 to 16	Yes	Extraction and autohydrolysis of hemicellulose from chipped, debarked pulpwood; fermentation to produce ethanol.
High Moisture Biomass: Biorefinery Heat and Power	HMB-CHP	Manures, Food Wastes, Biosludges	Digestion or Gasification	Methane, Syngas	Biorefinery Heat and Power	No	Anaerobic digestion of biomass or hydrothermal gasification followed by purification processes to produce a clean gaseous fuel for biorefinery CHP.
Black Liquor Conversion to Middle Distillates or DME	BLMD-G/FT BLDME-G/TC	Pulp Mill Byproducts	Gasification and Thermocatalysis	FT Liquids, DME	20 to 65	Yes	Gasification of black liquor to form syngas, then either catalytic synthesis to FT Middle distillates or to DME.
Algae to Biodiesel (Methyl Ester)	ARD or AME	Microalgae	Esterification	Methyl Esters or Renewable Diesel	5 to 100	No	Pressing to remove oil, and esterification or hydrotreating to form long-chain fuels.
Fatty Acids to Diesel Fuel - Hydrotreatment (Green Diesel)	FADe-H	Seed Oil / Waste Oils / Animal Fats	Oil Extraction then Coprocessing via Hydrotreatment	Renewable Diesel	5 to 200	Yes	Biomass oils conversion to diesel and other hydrocarbons via hydrotreating methods as in petroleum refinery

There is no guarantee that early demonstrations will be successful or that technologies that appear to be in early phases of development will not make a breakthrough earlier than expected. Further, other conversion technologies and fuels that do not strictly meet the product or timeline constraints of this assessment could potentially be incorporated in the transportation market in the future. For example, there is a potential for the use of compressed biogas as an alternative to natural gas for buses or specialized fleets. Biogas can be generated from manure and wastes via anaerobic digestion, and is also a by-product from landfills that can be harnessed for energy. This technology is described in this report but focusing primarily on how it might provide heat and power services to biorefinery operations to produce liquid biofuels.

1.2 CONVERSION TECHNOLOGY MODELS

Conversion technology models were developed to provide the analytical means to project to the mid-term the cost and performance of the selected technologies for the feedstock types available and the facility size. The models are used to optimize the location and technology choices in a GIS-based analysis for siting biorefineries. It is part of a larger model of the biorefining industry being developed for DOE called the National Biorefinery Siting Model (NBSM). These technology models are based on data available from published studies. An overview of the major assumptions and methodology that apply to the models is given here for reference. Some of the limitations of the models are also discussed.

In some cases, in consultation with the overall assessment team, some technologies were not provided with cost performance models. The basic standard for selection of representative technology models was that the engineering analysis for the commercial scale technology was reported in detail, publicly available and peer reviewed.[2] Modeling based on this standard provides reasonable confidence in the projected cost and performance for modeling Bioenergy industry development in the U.S. Studies that relied on goals for estimating process efficiency, rather than engineering process data analysis based on lab and better pilot plant operational experience, were considered to be more speculative and not as useful in gauging what might be realistically accomplished by the biofuels industry in the mid-term timeframe.

In general, the key input variables for each model are the feedstock input type and quantity. Key outputs are biofuel yield, capital and O&M costs, and by-product quantity and value. Levelized non-feedstock production costs are also calculated from the key outputs and economic factors described below.

Some of the overall standards for reporting are as follows:

- All reported cost values are given in U.S. 2008 dollars unless otherwise noted. Price conversions from other years were performed using the Chemical Engineering Plant Cost Index (CEPCI) and standard relations.

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[2] It must be noted that the lignocellulosic butanol model is an exception to this rule, due to its many similarities with LCEt-HF. Here, data from engineering studies of a proven corn butanol production plant were combined with detailed engineering studies of a lignocellulosic feed pretreatment for ethanol fermentation. The data comes from two engineering analysis sources, which made it possible to build a hybrid model.

- Models for current commercial grain ethanol and FAME biodiesel production technologies are based on current cost and performance data from operating facilities. These technologies are already well developed. No projections of future incremental improvements were made for this report.

- Advanced technologies models are based on detailed engineering cost and performance projections available in the literature. The projected costs are typically estimates for the n^{th} plant, with accuracy in the range of +/- 25% to 30%.

- Facility sizes/capacities available within each model are limited in consideration of: the amount and range of data used to project cost for larger and smaller facilities, any technological size limitations, and the relative capital and operating costs of competing technology types. These facility size limits will be noted individually for each model within the pertinent chapters later in this Appendix.

- All cost and performance models for the advanced technologies include projections for the 2015 - 2025 ("mid-term") timeframe. Models of other time periods were also included for technologies that had additional projections available.[3]

- The production cost uses a Fixed Charge Rate (FCR)[4] of 12.3%, calculated based on an economic lifetime of 25 years and a weighted cost of capital of 9.7%.

- All biorefinery co-products with significant energy contents and relatively limited markets, such as lignin co-produced with cellulosic ethanol, are assumed to be used within the refinery in order to supply heat and power to the conversion processes. The remainder of the technology's fuel requirements is assumed to be met with fossil fuel sources like natural gas. The type of fossil fuel used was kept consistent with the engineering design study from which the model was constructed, ultimately for consistency with existing lifecycle analyses.

Although efforts were made to perform a comparable analysis of each technology, it was not possible to make the same assumptions for each model, since these models are based on data from published studies with varying levels of detail. This can be particularly significant for items that have highly fluctuating market values. For example, electricity costs and fuel prices have increased sharply over the past few years. Where possible, these increased costs have been accounted for by directly substituting current values into calculation of these operational costs. However, in some studies there is not sufficient detail on process power consumption to make this substitution. In these cases the utility costs may be underestimated. In order to make the assumptions for each analysis as transparent as possible, as much detail as practical is included in the report regarding the estimations and calculations for each model.

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[3] For example, lignocellulosic ethanol production is modeled for 3 time periods – near term (2010), mid term (2015), and long term (2025+).

[4] FCR is defined as the annual amortized cost of an item divided by the total cost of that item over its lifetime. It is a measure of the aggressiveness of a payback schedule. 12.3% was chosen as a conservative estimate of this parameter.

1.3 BIOFUEL CONVERSION TECHNOLOGY DEVELOPMENT PERSPECTIVES

While this report carefully avoids making declarations about which technology is most likely to succeed, we have characterized the state of development and investment for each technology in the body of the report. In this section, we compare performance and cost attributes at a high level as a perspective on where technology development is headed. In the U.S., the “current” motor fuel of choice in terms of total demand is gasoline. New York State has recognized this and invested significantly in conversion technologies to supply this market. The state’s efforts with developers Mascoma Corporation and Catalyst Renewables Corporation are described in more detail in the “Lignocellulosic Ethanol via Hydrolysis and Fermentation” section later in this document.

Data are presented in Table H-2, comparing some of the yield characteristics of the technologies evaluated in this report that produce gasoline substitutes. Grain ethanol facilities have improved yields over the years. On an energy throughput basis they rank at the top. That is largely because nature has done the first-stage processing, concentrating starches in the grain kernels. Those feedstocks usually cost more ($/ton delivered) and in certain conditions could compete with food demands. Lignocellulosic biorefineries (LCB) range in conversion efficiency (expressed as a % on an HHV basis) from the low 30s to the mid 40s. The higher end can be achieved by either the advanced hydrolysis and fermentation technologies or the gasification and alcohol synthesis technologies, so that both have good promise for energy conversion. However, the thermochemical route as modeled in this work produces ethanol and other alcohol fuels. Those other alcohol fuels must be counted as energy carriers to attain the high efficiencies projected. Obtaining market outlets for the co-products is essential to economic success for the thermochemical plant. Obviously, higher catalyst specificity for ethanol production can also enhance biofuel output if overall yields are not sacrificed. This is clearly an area of research that will benefit those technologies.

Despite the higher heat content of the fuel product, the conversion efficiency to biobutanol ranks near the bottom for LCBs as modeled in this study. Increasing yields for biobutanol will be essential research to improve the competitive position of this biofuel.

The conversion efficiency for hemicellulose extraction at a pulping facility is very high but the process uses a smaller fraction of the incoming feedstock matter for fuel production and therefore the basis for calculating yields (hemicellulose input only) is not the same as the other technologies. The cellulose and lignin fractions of the feedstock are co-products produced in comparable quantities and the success of that technology will depend on markets for all the products created or internal use of the lignin for energy production.

Table H-2. Yield Comparison for Selected Technologies.

Technology	Feedstock	Yield Ethanol (gal/dry ton)	Yield n-Butanol (gal/dry ton)	Yield Mixed Alcohols (gal/dry ton)	Conversion to Fuel % HHV
Technologies - Gasoline					
Grain to Ethanol - Dry Mill	Corn-Wet Mill	89.3			56.8%
	Corn-Dry Mill	100			63.6%
Lignocellulosics to Ethanol - Enzymatic Hydrolysis/Fermentation (*Near term facility with dilute acid pretreatment*)	Corn Stover	70.0			37.2%
	Wheat Straw	76.9			34.2%
	Switchgrass	80.9			35.0%
	Poplar	76.8			35.9%
	Pine	69.3			37.2%
Lignocellulosics to Ethanol - Enzymatic Hydrolysis/Fermentation (*Mid term facility with dilute acid pretreatment*)	Corn Stover	80.6			41.3%
	Wheat Straw	76.8			37.9%
	Switchgrass	77.4			38.7%
	Poplar	85.9			40.1%
	Pine	90.2			41.5%
Lignocellulosics to Ethanol - Enzymatic Hydrolysis/Fermentation (*Long term facility with LHW pretreatment*)	Corn Stover	86.0			44.4%
	Wheat Straw	82.0			40.5%
	Switchgrass	82.6			41.3%
	Poplar	91.9			42.9%
	Pine	96.4			44.3%
Lignocellulosics to Ethanol - Gasification/Fermentation	Corn Stover	75.8		12.8	46.5%
	Wheat Straw	78.7		13.3	46.5%
	Switchgrass	77.8		13.1	46.5%
	Poplar	83.4		14.1	46.5%
	Pine	84.6		14.3	46.5%
Hemicellulose to Ethanol: Pulp and Paper Application	Spruce	135			65.9%
	Silver Birch	133			65.9%
	Eucalyptus	125			65.9%
	Scots Pine	110			65.9%
Lignocellulosics to Butanol - Hydrolysis/Fermentation	Switchgrass	1.18	48.5		30.8%
	Hybrid Poplar	1.24	51.9		30.7%
	Pine	1.32	55.0		32.1%
	Mixed Paper	1.22	51.3		33.3%
	Wheat Straw	1.16	56.5		35.4%

From a capital investment perspective (Table H-3), grain ethanol production also has an advantage of lower capital investment per unit capacity output ($/gal/year). Again, grain ethanol facilities benefit from a feedstock that is readily converted to fermentable sugars. For the LCB technologies, the initial step of breaking down cellulose and hemicellulose to fermentable sugars along with other process requirements can double or even triple the plant capital cost compared to grain ethanol for a comparable production capacity. Combining this high cost with the risk of introducing new technology, it is easy to see why these plants are only being built in the U.S. with government cost share and only at demonstration scales. The projected investment required for the advanced lignocellulosic technology does match the cost for a grain wet mill (the more complex process for grain ethanol) but remains well above the investment cost for a dry mill. Dry mills represent a much lower investment risk from this perspective.

Table H-3. Cost Attributes for Conversion Technologies.

Technology	Feedstock	Plant Feedstock Demand Dry tons/yr	Plant Capacity MGY	Non-Feedstock Production Cost ($/gal)	Non-Feedstock Production Cost ($/MMBtu)	Non-Feedstock O&M Cost ($/gal)	Total Capital Investment ($/(gal/yr))	Total Capital Investment (1) ($/ton feed)
Biomass Conversion Technologies								
Lignocellulosics to Butanol - Hydrolysis/Fermentation	Switchgrass	1,000,000	48.5	$ 1.40	$ 13.36	$ 0.37	$ 6.34	$ 308
	Hybrid Poplar	1,000,000	51.9	$ 1.23	$ 11.74	$ 0.37	$ 6.09	$ 316
	Pine	1,000,000	55.0	$ 1.15	$ 10.98	$ 0.37	$ 5.85	$ 322
	Mixed Paper	1,000,000	51.3	$ 1.23	$ 11.74	$ 0.37	$ 6.14	$ 315
	Wheat Straw	1,000,000	56.5	$ 1.14	$ 10.88	$ 0.37	$ 5.73	$ 324
Lignocellulosics to Ethanol - Enzymatic Hydrolysis/Fermentation (Mid term facility with dilute acid pretreatment)	Corn Stover	1,000,000	80.6	$ 0.65	$ 7.70	$ 0.19	$ 3.78	$ 305
	Wheat Straw	1,000,000	76.8	$ 0.66	$ 7.86	$ 0.19	$ 3.82	$ 293
	Switchgrass	1,000,000	77.4	$ 0.66	$ 7.86	$ 0.19	$ 3.81	$ 295
	Poplar	1,000,000	85.9	$ 0.65	$ 7.70	$ 0.19	$ 3.74	$ 321
	Pine	1,000,000	90.2	$ 0.64	$ 7.65	$ 0.19	$ 3.70	$ 334
Lignocellulosics to Ethanol - Enzymatic Hydrolysis/Fermentation (Long term facility with LHW pretreatment)	Corn Stover	1,000,000	81.6	$ 0.61	$ 7.22	$ 0.27	$ 2.80	$ 241
	Wheat Straw	1,000,000	82.0	$ 0.61	$ 7.26	$ 0.27	$ 2.82	$ 232
	Switchgrass	1,000,000	82.6	$ 0.61	$ 7.26	$ 0.27	$ 2.82	$ 233
	Poplar	1,000,000	91.9	$ 0.61	$ 7.22	$ 0.27	$ 2.76	$ 254
	Pine	1,000,000	96.4	$ 0.60	$ 7.18	$ 0.27	$ 2.73	$ 264
Lignocellulosics to Ethanol - Gasification/Fermentation	Corn Stover	1,000,000	75.8	$ 1.05	$ 12.47	$ 0.57	$ 3.92	$ 297
	Wheat Straw	1,000,000	78.7	$ 1.02	$ 12.14	$ 0.56	$ 3.77	$ 297
	Switchgrass	1,000,000	77.8	$ 1.03	$ 12.24	$ 0.56	$ 3.82	$ 297
	Poplar	1,000,000	93.4	$ 0.98	$ 11.68	$ 0.54	$ 3.56	$ 297
	Pine	1,000,000	84.6	$ 0.97	$ 11.57	$ 0.54	$ 3.51	$ 297
Hemicellulose to Ethanol: Pulp and Paper Application	Spruce	1,000,000	5.9	$ 3.74	$ 44.56	$ 3.06	$ 5.56	$ 768
	Silver Birch	1,000,000	6.3	$ 3.64	$ 43.32	$ 2.97	$ 5.45	$ 749
	Eucalyptus	1,000,000	3.8	$ 4.93	$ 58.71	$ 4.93	$ 6.34	$ 839
	Scots Pine	1,000,000	5.6	$ 3.92	$ 46.65	$ 3.23	$ 5.65	$ 651
Grain Ethanol	Corn-Wet Mill	1,000,000	89.3	$ 0.25	$ 2.95	$ (0.07)	$ 2.62	$ 234
	Corn-Dry Mill	1,000,000	100.0	$ 0.36	$ 4.24	$ 0.11	$ 1.98	$ 198

1) In the hemicellulose extraction case, total capital investment is stated per ton of hemicellulose feedstock, which is a derivative of the primary feedstock, whole wood. The "Plant Feedstock Demand" total is stated as tons of wood. The "Total Capital Investment" total is stated per ton hemicellulose, as this is the incremental feedstock to the retrofitted biofuels facility.

The potentially attractive feature of the hemicellulose extraction and conversion process applied to an existing pulp mill is that, while the production capacity may be smaller, the benefit of building on existing infrastructure should also lower capital costs. However, based on the data used in the analysis for this project, the lower cost expectation is not realized. Further, the benefit of using existing infrastructure must be weighed against the trend in the U.S.

toward reduced capacity for basic paper products.[5] The unexpected result for the analysis of this technology is the very high O&M costs projected for this technology based on the engineering studies available. We believe the O&M costs reported in the engineering studies used for this report are on the high side, but clearly engineering development to reduce those costs will go a long way to improving the technology's prospects for implementation.

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[5] Specialty tissues and paper products are still competitively produced in the U.S.

2 PRODUCTION OF BIOFUELS FOR THE GASOLINE FUELS MARKET

2.1 CURRENT TECHNOLOGY: GRAIN AND SUGAR CONVERSION TO ETHANOL

Ethanol is currently produced in significant quantities using commercial conversion technologies. These technologies are based on enzymatic fermentation of sugars and starch (grains) to produce ethanol. Ethanol (ethyl alcohol) production is primarily a fermentation process that converts glucose into alcohol.[6] The main difference between the conversion technologies currently commercially available is the pretreatment of the feedstock before fermentation (Antares 2008).[7]

2.2 GRAIN ETHANOL PRODUCTION

According to the Renewable Fuels Association, as of May 7, 2009, U.S. biorefineries had a capacity of 10.6 billion gallons per year for operating refineries, with 1.9 billion gallons per year under construction at new or expanding refineries. These figures are out of a total of more than 200 facilities (Energy Supply Logistics). More than 98% of the current U.S. ethanol production capacity comes from corn. Of the corn ethanol plants in operation approximately two-thirds are dry-mills. These facilities have high overall ethanol conversion rates and can be built at a scale that is affordable for large agricultural cooperatives and companies focused on ethanol production. About 80% of the U.S. ethanol production comes from dry-mill facilities, indicating proportionally more capacity in dry mills than in wet mills (Solomon, Barnes and Halvorsen 2007).

2.2.1 Technology Description

Dry Mill Ethanol Facilities

Figure H-1 illustrates the basic steps of the dry-milling or enzymatic fermentation process.[8] In this process, the corn or grain feedstock is initially ground into a flour or fine meal to release the starch. This material is then mixed with water to produce a mash, which is processed in a high temperature cooker with enzymes to convert the starch to sugar and reduce bacterial contamination before fermentation. Ethanol is produced during fermentation, which usually takes 40 to 50 hours (RFA 2005). The ethanol is purified and separated from the stillage during distillation. Further distillation and dehydration (as with molecular sieves) must be done to purify the ethanol for use as a fuel and increase the alcohol concentration.

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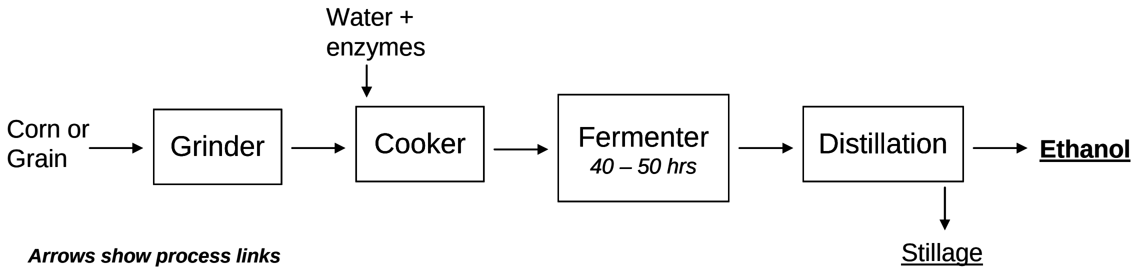
[6] Fermentation generally refers to the conversion of sugars to alcohol using *Saccharomyces* (yeast, either baker's or brewer's) under anaerobic conditions. A more general interpretation of "fermentation" is as the biological or biochemical conversion of biomass compounds like starch and sugars into alcohols or acids.

[7] The text in this and the following sections on grain- and sugar-based ethanol production was revised from material originally presented by Antares in 2008 in a *Strategic Assessment of Bioenergy Development.*

[8] A more detailed schematic diagram is given in Exhibit D.

Figure H-1. Schematic Diagram of Dry Milling Ethanol Production Process.



The whole stillage (WS) by-product of the dry-milling process, also known as distillers' grain, includes the fiber, oil and protein components of the grain, as well as the non-fermented starch. It is currently sold as an animal feed in a variety of forms. As a result of its high moisture content, stillage is readily perishable and only has a shelf life of 6 to 10 days. To increase durability of the product, thin stillage can be separated and dried to viscous syrup which is mixed back with the solids to create a feed product known as wet distillers' grains with solubles (WDGS). WDGS, containing 65% moisture, can be used directly as an animal feed product. However, the shelf life of WDGS is only slightly longer than WS (one to two weeks), so end-uses such as feedlots must be nearby (typically within about 50-100 miles of the mill) to avoid spoiling. To further increase shelf life and lower transportation costs, WDGS is usually dried to 10-12% moisture to produce a product known as dried distillers' grain with solubles (DDGS). However, the drying process is extremely energy-intensive, and can consume about one-third of the energy requirements of the entire dry mill facility.

Wet Mill Ethanol Facilities

In the wet milling process, the grain is separated into components (germ, gluten, fiber, and starch) before fermentation, yielding a number of valuable by-products. Wet mill facilities are good examples of the term "biorefinery."[9] Figure H-2 shows the basic steps of the separation process that is used to produce a fermentable starch.[10] The grain is first steeped in a mixture of water and dilute sulfurous acid for 24 to 48 hours to facilitate the separation into components (RFA 2005). The germ slurry is then separated, and goes through a process of grinding and screening to separate the fibrous material from the germ. The germ can be sold to crushers, or further processed on-site into an oil product. The gluten is separated from the starch, producing a gluten meal by-product that can be used for animal feed. The separated starch can then be hydrolyzed, fermented and distilled to produce ethanol, using the same steps as those described above for dry milling.

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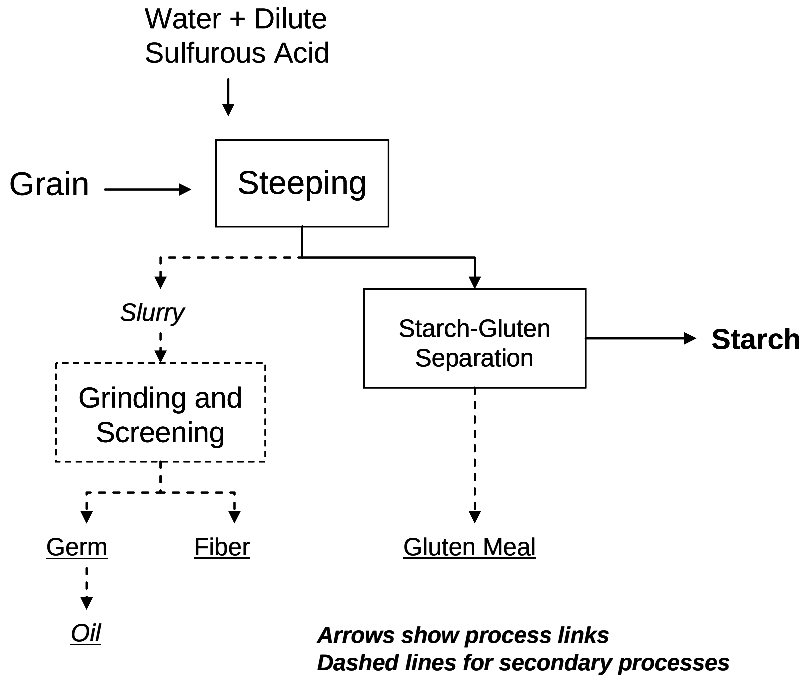
[9] There is some debate as to how many products a facility must produce before it can be termed a "biorefinery." For example, a corn ethanol dry mill facility may have three separate, diverse product streams: carbon dioxide, DDGS or WDGS, and ethanol. Under a general "biorefinery" definition that requires only a diverse product slate, dry mill facilities could thus be considered biorefineries. A definition requiring both high-value, low-volume and low-volume high-value products would not permit this designation.

[10] Exhibit D has a more detailed schematic diagram of this process.

The gluten meal produced from the wet-milling process has high moisture content and is readily perishable. Thus the same problems occur as with the whole stillage by-products of dry milling – the gluten meal must either be used quickly or dried.

Figure H-2. Schematic Diagram of Wet-Milling Process for Conversion of Grain to Starch.



2.2.2 Performance and Cost Analysis

Dry Mill Ethanol Facilities

Ethanol yield is directly related to amount of fermentable starch in corn kernels, plant efficiency, plant age, types of equipment, and plant management. Shapouri, Gallagher and Graboski (2002) reported that the average ethanol yield was 2.68 gal/bu for small facilities (<40MGY), and 2.65 gal/bu for large plants.[11] For newer facilities, ethanol yield has increased to 2.8 gallons per bushel.

There have been several technology improvements since the 1970's that have reduced the cost of ethanol production, including a 20% increase in yield of ethanol production from corn due to improved biochemical processing, higher starch quantities in the kernels, and a 35% reduction in heat energy required for conversion by re-use of process heat (i.e. use of "waste" heat) (Gallagher, Brubaker and Shapouri 2005). New dry-mill plants use 30,000 Btu of thermal energy and less than 1kWh of electricity to produce 1 gallon of ethanol and by-products

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[11] The authors report no rationale for the small difference in ethanol yield between the size categories. This difference is not statistically significant, given an overall sample size of 21 dry mill facilities.

(Shapouri and Gallagher 2002). Furthermore, new ethanol plants have minimal discharge of wastewater, and typically do not require wastewater treatment facilities (Shapouri and Gallagher 2002).

There is a clear relationship between capital cost and plant capacity for dry mill facilities, as investigated in Gallagher, Brubaker and Shapouri (2005). This study analyzed several dry mill facilities of varying sizes constructed over a 25 year period, and found that capital costs decrease on a $/gallon basis with increasing size up to a capacity plateau of about 55-74 MGY (with a minimum at 65 MGY), beyond which costs increase with size. The capital cost at the minimum is $1.14 million per MGY ethanol production capacity (Gallagher et al. 2005).

In 2004, there were about 40 large dry-mill ethanol plants (40-100 MGY) in operation or under construction (Shapouri and Gallagher 2005). The capital costs for new plants at this time ranged from $1.05 to $3.00 million per MGY of capacity. (The minimum capital cost presented in this study is significantly lower than the $1.14 per gpy value presented by Gallagher et al (2005) above.) As expected, the cost for expansion of existing plants was much lower, from $0.20 to $1.00 per gallon capacity (with an average of $0.50/gal).

According to Shapouri, Gallagher and Graboski (2002) DDGS made up 70% of all distiller's grain products from dry mill facilities, while WDGS and MDGS (modified distillers grains) accounted for 21% and 9%, respectively (all on a dry basis). Furthermore, 17 of 21 ethanol facilities surveyed indicated that they sold wet and modified distiller grains.

Wet Mill Ethanol Facilities

Since wet mills are more complex than dry mills, they have a different plant-size to capital cost relationship (Gallagher, Brubaker and Shapouri 2005). Wet mills also have higher capital and O&M costs, leading fewer investors to choose wet mills over dry mills for new construction. However, energy expenses tend to be lower in wet mills than dry mills as they typically include cogeneration of steam and electricity (Shapouri, Gallagher and Graboski 2002).

Wet mills have a number of valuable by-products, including corn gluten feed, corn gluten meal, and corn oil. The separated starch component can either be fermented to produce ethanol, or it can be further processed into corn syrup, corn starch and sugar. In this analysis we assume that all of the starch is fermented in an ethanol producing wet mill facility. However, it is important to note that the flexibility of product distribution from wet mills can insulate these facilities from large price fluctuations for one of more of the potential products.

Outline for Model Analyses

The key values and assumptions used to calculate the current technology analysis and economic performance model for wet and dry mills are described below. An example of the detailed analysis is shown in Appendix H-F.

- Feedstock input quantity is a variable, and should be given on an as-received basis (typically about 15% moisture). Dry mills can use corn or sorghum, while wet mills use only corn.

- Applicable facility size ranges are 5 MGY to 100 MGY for dry mills, and 50 MGY to 300 MGY for wet mills, based on data from Gallagher et al. (2005).

- Ethanol yield is calculated based on the average of values reported in the literature. New dry mills typically produce 2.8 gallons per bushel of corn (100 gal/ton), and wet mills produce about 2.5 gal/bu (89.3 gal/ton) (Shapouri and Gallagher, 2005; McAloon et al., 2000, Butzen and Hobbs, 2002).

- The conversion efficiency of corn to fuel is estimated to be 64% (on an energy basis) for a dry mill and 57% for a wet mill. This is calculated using a typical HHV for ethanol of 84,000 Btu/gal, and a HHV for corn of 6,600 Btu/lb (at 15% moisture content).

- It is assumed that all distillers' grains from dry mill facilities are converted to DDGS. It is estimated that dry mill facilities produce 6.7 lb DDG per gallon of ethanol produced (McAloon et al. 2000).

- The wet mills by-products include corn gluten feed, corn gluten meal, and corn oil, of which 11.4 lb, 3 lb, and 1.6 lb are produced from 1 bushel of corn, respectively (Butzen and Hobbs, 2002).[12]

- The amount of CO_2 production from fermentation[13] for a dry mill facility is estimated to be 17.5 lb per bushel of corn (Antares Group Inc., 2005a; McAloon et al., 2000). The CO_2 production for wet mills is based on the same value, but is scaled by the ethanol conversion rate.

- The average water requirement for a dry mill facility is 4.7 gallons[14] of water per gallon of ethanol produced (Shapouri & Gallagher 2005). Based on data from Shapouri, Gallagher and Graboski (2002), wet mills use 5.2 times more water than dry mills.

- Capital cost data for dry mill facilities is based on the relationship derived from Shapouri and Gallagher (2005). The minimum capital cost on a $/gal basis occurs at 65 MGY. For larger facilities the cost increases with capacity.

- Wet mill capital costs are based on data from Whims (2002), using a scaling factor of 0.6.[15] The capital for a wet mill facility is higher than for a dry mill facility due to additional process equipment requirements.

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[12] 1 bushel of corn is equivalent to 56 lbs.

[13] This refers to only that CO_2 emitted during the fermentation process, and does not refer to any CO_2 emissions incurred by process fuel combustion, product fuel combustion, feedstock cultivation, or feedstock transportation. Unless othewise noted, CO_2 emissions reported in this Appendix for all technologies will refer only to fermentation-related emissions, which are often purified and sold as a coproduct for use in soft drinks and other compressed carbon dioxide applications. Lifecycle emissions are discussed in Appendix G.

[14] While the total water embodied in a gallon of ethanol has decreased in recent years, this embodiment is largely driven by irrigation water use, rather than conversion facility process water use (Suh et al, 2009). According to Suh et al (2009), average process water consumption for corn ethanol facilities (both wet and dry mills) in the U.S. is 3.3-4 gallons per gallon ethanol. This is a net, rather than absolute value and does not distinguish between dry and wet mills, so the most recent value available that did differentiate mill types (Shapouri and Gallagher, 2005) was used. This may present an overestimate due to progress made in water consumption reductions over the past four years.

- Average wholesale values of the co-products are based on current data from United States Department of Agriculture (USDA 2007). The value of DDGS is $90 - $115 per ton (similar to data from McAloon et al. 2000; Shapouri and Gallagher 2005). Corn gluten feed is $40-60/ton, corn gluten meal is $335-350/ton, and corn oil is $0.31-0.32/lb. These current wet mill co-product values from USDA are comparable to those reported for 2003 in Shapouri and Gallagher (2005).

- Consumables for both processes include enzymes, yeast, chemicals, and denaturant. The consumables cost for small dry mills (< 40 MGY) is $0.13/gal, while for large dry mills (40 to 100 MGY) the cost is $0.12/gal (Shapouri and Gallagher, 2005). The cost of consumables for wet mills is $0.19/gal (Shapouri, Gallagher and Graboski, 2002).

- Utilities include electricity, fuels, water and waste management, and does not reflect the use of combined heat and power (CHP) beyond basic heat integration. The use of CHP greatly affects the consumption of fuel to supply the process heat and electricity demands quantified by Shapouri et al. (2002), and should be recommended for all dry mills due to their significant use of natural gas.[16] This analysis assumes that heat is supplied in a standard natural gas-fired boiler with no CHP, and electricity is purchased from the grid. The cost for small dry mills is $0.27/gal and for large dry mills the cost is $0.19/gal (Shapouri and Gallagher 2005). Average electricity use is 1.19 kWh/gallon of ethanol produced, and the average heat use is 34,800 Btu per gallon of ethanol. The cost of electricity for the dry mill facilities has been updated to 2007 average wholesale value for the industrial sector of 7.9¢/kWh. New dry mill facilities have minimal wastewater discharge. Wet mill utility costs are $0.19/gal (Shapouri et al. 2002).[17] Wet mill energy usage is lower than dry mills as they typically employ cogeneration of steam and electricity.

- Annual labor costs for small dry mill facilities are $0.077/gal, and $0.060/gal for large facilities (Shapouri and Gallagher 2005). Labor costs for wet mills are $0.096/gal, based on data from Shapouri, Gallagher and Graboski (2002).

- No economy of scale is included for the wet mill operating costs—no distinction is made between large and small facilities. The 2002 Cost of Production Survey aggregated wet mill data over all facility sizes, rather than over several size ranges.

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[15] This scaling factor was derived from the Whims (2002) data. The term "scaling factor" refers to the exponent of a capital cost vs. capacity or size relationship. For a linear relationship in which there is no economy of scale, this exponent is one. If economy of scale exists, this exponent will be smaller than one.

[16] Neither of New York State's dry mill ethanol facilities currently use CHP. As of 2008, CHP use was planned or already implemented in 24 of 206 of the nation's planned or constructed dry mill facilities (EPA CHP 2008)

[17] Electricity prices have not been updated to reflect current market value as there was not sufficient detail in the Shapouri et al. (2002) to perform the calculations.

- Other fixed O&M costs include maintenance, administrative costs, and other costs. For dry mills, other fixed costs are equivalent to \$0.090/gal for small facilities and \$0.106/gal for large facilities (Shapouri and Gallagher 2005). For wet mills other fixed costs are \$0.108/gal (Shapouri, Gallagher and Graboski 2002).

- The levelized non-feedstock production cost of ethanol is calculated based on the annual operating expenses and capital cost payment, co-product credit, and quantity of fuel produced.

Model Results

Some of the key cost and performance results for current ethanol production processes are shown in Table H-4 and Table H-5. These tables show results for the range of applicable facility sizes based on selected input quantities, and illustrate the high and low end costs. In general, the analysis shows that ethanol production from wet mill facilities has lower costs than the dry mills as a result of the co-product value.[18] However, it is important to note that the value for these co-products will fluctuate as the market changes. For similar sized facilities the wet mills have higher capital and O&M costs than dry mills, as the process is more complex. It should also be noted that both tables show results only for the non-feedstock portion of the production costs.

Table H-4. Example Cost and Performance Outputs for Dry Mill Facility.

Sample Model Results - Grain Ethanol (Dry Mill)		
Feedstock Input (as rec'd ton/yr)	50,000	1,000,000
Ethanol Yield (MGY)	5	100
DDG Yield (ton/yr)	16,750	335,000
CO2 Stream (ton/yr)	15,625	312,500
Water Consumption (1000 gal/yr)	23,500	470,000
Capital Cost (Million \$)	\$ 15.0	\$ 198.1
Annual O&M Costs (Million \$/yr)	\$ 3.39	\$ 58.0
By-Product Credit (Million \$/yr)	\$ 2.34	\$ 46.7
Non-feedstock Production Cost (\$/gal)	***\$ 0.58***	***\$ 0.36***

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[18] This invites the question of why most new corn ethanol facilities, and most existing facilities, are dry mill plants. It is important to note the significantly larger capital outlay and higher operating costs for wet mill facilities, which may discourage developers from choosing this option. Moreover, this result raises an issue of co-product accounting: should co-product revenue streams be counted toward the overall production cost of the ethanol, or should they be held economically separate? As this report focuses strictly on biofuels and not the entire varied product slate of biorefineries, in this and subsequent analyses, the biofuel will be considered the primary product and all other products co-products. To this end, the facility's economics will be stated as a non-feedstock production cost per unit of fuel. The costs of production are difficult to attribute between products and co-products (the energy to cool corn ethanol fermenters, for example, applies to both the ethanol produced and the DDGS, but would not be strictly necessary if the object of the process were to produce feed), but all of the costs and revenues of the co-product streams together contribute to the success or failure of a biorefinery. Cost and energy input attribution can be done artificially by many respected methods, including assigning these inputs on a mass or heating value basis, but these do not reflect the fact that co-product values can and likely will enable ethanol production in a low-priced fuel market where margins are slim, particularly in the case of corn ethanol. A key result of this report is that obtaining higher co-product value is key to the financial viability of not only corn ethanol, but also many biofuel conversion technologies. Given this fact, it was decided that exclusion or artificial separation of co-products from the plant financial analysis would be inconsistent with the scope of the report and potentially misleading.

Table H-5. Example Cost and Performance Outputs for Wet Mill Facility.

Sample Model Results - Grain Ethanol (Wet Mill)		
Feedstock Input (as rec'd ton/yr)	560,000	3,360,000
Ethanol Yield (MGY)	50	300
Corn Gluten Feed Yield (ton/yr)	114,000	684,000
Corn Gluten Meal Yield (ton/yr)	30,000	180,000
Corn Oil Yield (ton/yr)	16,000	96,000
CO2 Stream (ton/yr)	156,250	937,500
Water Consumption (1000 gal/yr)	1,222,000	7,332,000
Capital Cost (Million $)	$ 163.9	$ 480.4
Annual O&M Costs (Million $/yr)	$ 34.7	$ 208.4
By-Product Credit (Million $/yr)	$ 38.6	$ 231.6
Non-feedstock Production Cost ($/gal)	***$ 0.33***	***$ 0.12***

2.2.3 Outlook

According to USDA Economic Research Service, the average price for corn in 2007 through 2008 and again from 2008 through early 2009 was around $4.20/bu ($150/ton) (including an estimated delivery cost of $0.30/bu ($11/ton)).[19] This compares to $3.04 ($109/ton) from 2006 through 2007. This converts to a feedstock fuel equivalent cost for ethanol production of $1.35/gallon for dry mills and $1.51/gallon for wet mills, based on the conversion yields used in the model. The resulting ethanol production costs are about $1.71 to $1.92 per gallon. For comparison, current FOB ethanol prices are around $1.75-$2.15/gal (USDA AMS 2007). The USDA results are very similar to the results of the ethanol production models (combining feedstock and non-feedstock costs) reported here.

Projected prices for corn at the farm gate from FAPRI's 2007 and 2008 U.S. and World Agricultural Outlook are shown in Figure H-3. Since the 2008 data was not available at the time of analysis, the 2007 data is used in the report. Note that although the more recent data shows higher corn price projections, both data sets have relatively flat price escalation curves.[20] The higher corn prices from the 2008 data likely reflect increasing costs of energy.

According to the 2007 data, corn is projected to cost about $3.33/bu ($119/ton) in 2015 including the same $11/ton transportation cost as above. This leads to ethanol production costs of $1.55 to $1.74 per gallon for dry mills and $1.57 to $1.74 per gallon for wet mills.

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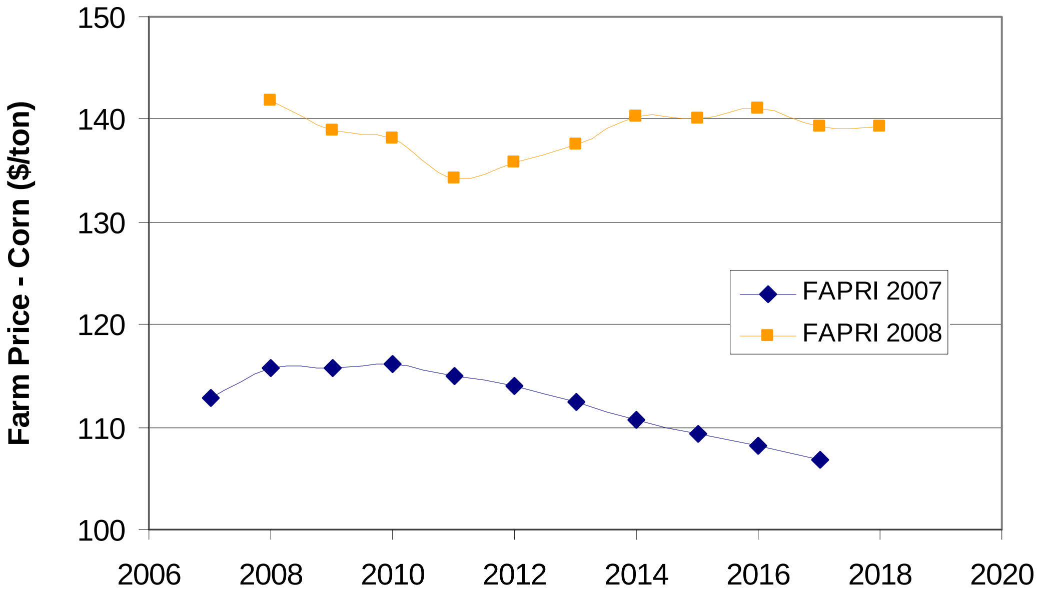
[19] Corn price based on data from USDA AMS 2007. Estimated delivery cost based on Antares experience, using a typical cost of $0.20/ton-mile for truck transportation and a distance of 50 miles.

[20] The projected corn price escalation from 2008 to 2015 is -5.6% with the 2007 FAPRI data, and -1.3% with the 2008 FAPRI data. This relatively small change can be considered indicative of a flat market.

Figure H-3. Projected Corn Prices.



Data from FAPRI 2007 and FAPRI 2008

Dry-mill produced DDGS competes with corn, soybean meal, and dicalcium phosphate in animal feed markets as a source of protein and phosphorus. DDGS has a very long shelf life, and is easily transported across the U.S. and all over the world. Domestic markets account for about 75% of U.S. production, while the remaining 25% is produced to satisfy the demand of international markets (Antares Group Inc. 2005a).

As more dry-mill ethanol facilities come on line, the market can become saturated with these types of animal feed products and alternate end-uses or markets will need to be explored in order to keep a high value for this by-product. In general, the overall animal feed market is far larger than existing DDGS production capacities, thus theoretically there is room for growth in that area. However, DDGS producers will need to consider expanding into non-traditional feed markets, beyond feed for ruminating animals like dairy beef and cattle. An additional option is increasing exports to other countries, but this can be stifled by high freight costs.

Despite the potential for market expansion for growth in the corn and other grain ethanol markets, a combination of political and economic pressures have narrowed the margins of corn and other grain ethanol producers. Politically, the Obama administration's emphasis on advanced biofuels and the federal Renewable Fuel Standards—which grant 2.5 as many renewable fuel units to a gallon of lignocellulosic ethanol than to a gallon of corn ethanol—have created a hierarchy of ethanol producers. A great deal of research and pilot funding is being funneled into commercializing lignocellulosic ethanol, while corn ethanol producers are seeing fewer yield increases at their production facility and escalating commodity prices. Corn, as noted above, has reached such a high price that ethanol producers are barely meeting operating expenses, while petroleum prices are low and ethanol sales prices correspondingly low.

This has led to a record number of corn ethanol producers filing for bankruptcy. VeraSun and Aventine Renewable Energy Holdings, two of the nation's largest corn ethanol producers, last year announced Chapter 11 bankruptcy, as did 14 other corn ethanol plants, including NorthEast Biofuels in New York.[21] The total number of companies going out of business since the start of 2008 is expected to total forty by mid-2009.

2.3 SUGAR FERMENTATION TO ETHANOL

The conversion of sugars (like sugarcane) to ethanol is simpler than the conversion of starches as it requires fewer process steps. This process is very similar to the dry milling process described above except that no cooking is required. The feedstock is milled and pressed, and then fermented to generate ethanol. The resulting product is purified during distillation, and can be dehydrated to increase the alcohol concentration.

Ethanol production directly from sugars is generally only performed in locations where there is a large quantity of available feedstocks, such as sugarcane crop in Brazil and sugar beets in parts of Europe. According to the USDA National Agriculture Statistics Service (NASS), the entire U.S. produced about 34 million tons of sugar beets and 30 million tons of sugarcane in 2006 (USDA NASS 2007).[22] For comparison, the U.S. soybean crop from 2006 was nearly 90 million tons and the corn crop was 300 million tons (USDA NASS 2007). As a potential U.S. biofuels feedstock, sugar crops are of limited availability, are generally expensive to produce or import, and are most valuable to the food market rather than the lower-cost biofuels market.

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[21] This facility was purchased by Sunoco for $8.5 million, in May, 2009. The facility is planned to continue as a corn ethanol producer, expected to reopen in early 2010. (*Biofuels Digest*, 2009)

[22] The top states with sugarcane crops are Florida, Louisiana, Hawaii and Texas, while the top sugar beet growing states are Minnesota, North Dakota, Idaho, Michigan and California (USDA NASS 2007).

2.4 TECHNOLOGIES READY FOR COMMERCIALIZATION IN THE MIDTERM (2015 – 2025)

There are a number of advanced biofuel technologies being developed that may be available in the 2010 to 2025 time frame. One of the key differences between these and the current commercial biofuel technologies is the types of biomass feedstocks used for production. The current technologies typically use grain and oil seed crops (which are relatively expensive and compete with food uses). In contrast, most of the advanced technologies use lignocellulosic biomass such as wood, grasses and agricultural residues. Although these biomass feedstocks are generally more difficult to convert to biofuels, they are not a human food source and can be less expensive than grain and seed crops.

Cost and performance models for several of the advanced technologies have been developed. Since the advanced technologies do not currently have commercial plants in operation, these models are based on detailed engineering analysis for commercial scale plants (typically the projected cost and performance projections for the nth plant) available in the literature. The projected costs are typically accurate in the range of +/- 25% to 30%. All cost and performance models for the advanced technologies include projections for the 2020 ("mid-term") time period. Models of other time periods were also included for technologies that had additional projections available in published studies.

2.5 LIGNOCELLULOSICS TO ETHANOL: HYDROLYSIS AND FERMENTATION

Lignocellulosic biomass consists mainly of cellulose, hemicellulose and lignin. This includes woody biomass, grasses, and agricultural residues. Lignocellulosic biomass typically contains 40-60 wt% cellulose (dry weight) and 20-40 wt% hemicellulose, both which can be hydrolyzed to sugars and fermented to produce ethanol. Most of the remaining fraction (10-25 wt%) is lignin, a complex polymer which is resistant to biological degradation and cannot be fermented. However, the lignin fraction can be thermochemically treated for conversion to biofuels or used directly to produce heat and/or power.[23] There are also small amounts of proteins, lipids, and ash (Antares 2008).[24]

The chemical composition of the biomass (i.e. the lignin/hemicellulose/cellulose ratio) is a major factor in the ethanol yield. Materials with high sugar content (cellulose and hemicellulose) have the potential to produce the greatest amount of ethanol per ton of feedstock consumed. Table H-6 shows typical biochemical compositions for various biomass feedstocks. Most woody biomass has about 27% lignin, while grasses such as switchgrass have about 18% (Hamelinck, van Hooijdonk and Faaij 2005). Switchgrass has a large fraction of ash and other materials which cannot be converted to ethanol or used for heat and power production.

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[23] The higher heating value (HHV) of lignin is 10,495 +/- 515 Btu/lb (dry), whereas the (hemi)cellulose and other components have a much lower HHV around 7,310 Btu/lb (dry) (Hamelinck, van Hooijdonk and Faaij, 2005).

[24] The text in this section was revised from material originally presented by Antares in 2008 in a *Strategic Assessment of Bioenergy Development.*

Table H-6. Typical Biochemical Compositions of Various Feedstocks.[25]

	Feedstock Composition (wt%)			
Feedstock	**Hemicellulose**	**Cellulose**	**Lignin**	**Other**
Agricultural Resources	23-26%	33-40%	17-24%	10-27%
Switchgrass	25%	32%	18%	25%
Hardwoods (i.e. Poplar)	13-19%	42-50%	26-28%	3-19%
Pine (Softwood)	22%	45%	28%	6%

Sources: U.S. DOE EERE Biomass Program Jan 25, 2006; Hamelinck, van Hooijdonk and Faaij 2005

2.5.1 Technology Description

Production of ethanol from lignocellulosic materials is similar to production from grains, except that it is more difficult to break down the feedstock into fermentable sugars. In other words, hydrolysis of lignocellulose is more difficult than hydrolysis of starch.[26] The hydrolysis or saccharification of cellulose produces glucose (a sugar containing 6 carbon atoms), and hydrolysis of hemicellulose produces various 5 carbon and 6 carbon sugars.

Figure H-4 shows the basic process steps for the conversion process. The biomass is first pretreated to clean and size the materials and alter the cell structure so that it is more accessible to conversion. Hemicellulose hydrolysis is part of the pretreatment, and can be performed by a variety of methods (see below). After pretreatment, the solids (cellulose and lignin) are filtered and pressed and undergo cellulose hydrolysis, while the sugar-containing liquids resulting from the hemicellulose hydrolysis are diverted to fermentation. Cellulose hydrolysis is the main hydrolysis step, in which the cellulose is converted to sugars. The primary methods for this reaction are described below. After hydrolysis, the sugars are converted to ethanol via fermentation, and the ethanol is recovered in a distillation column and concentrated. The lignin and other solid residuals can be thermochemically treated (gasified) or used directly to produce heat/power via combustion. Unfermented sugars in liquid effluent can also be dried and fired in a boiler. An alternate use of the lignin is to produce high-value co-products such as high-octane hydrocarbon fuel additives, but these conversions are complex and expensive. Note that although each step is shown below as a separate process, integration of two or more process steps is a potential way to reduce costs and increase overall efficiencies. In the descriptions below, each major process step is discussed separately in order to clarify the different aspects of the process.

It is important to mention another potential method of ethanol production from lignocellulosic materials; hemicellulose conversion of biomass in a forest product biorefinery. In this process sugars are extracted from the hemicellulose component of the biomass and fermented to ethanol. The rest of the biomass is then used to produce high value pulp and paper products. Production of lignocellulosic ethanol as a by-product of the forest products industry is very appealing, as the required biomass infrastructure is already well developed. There is significant

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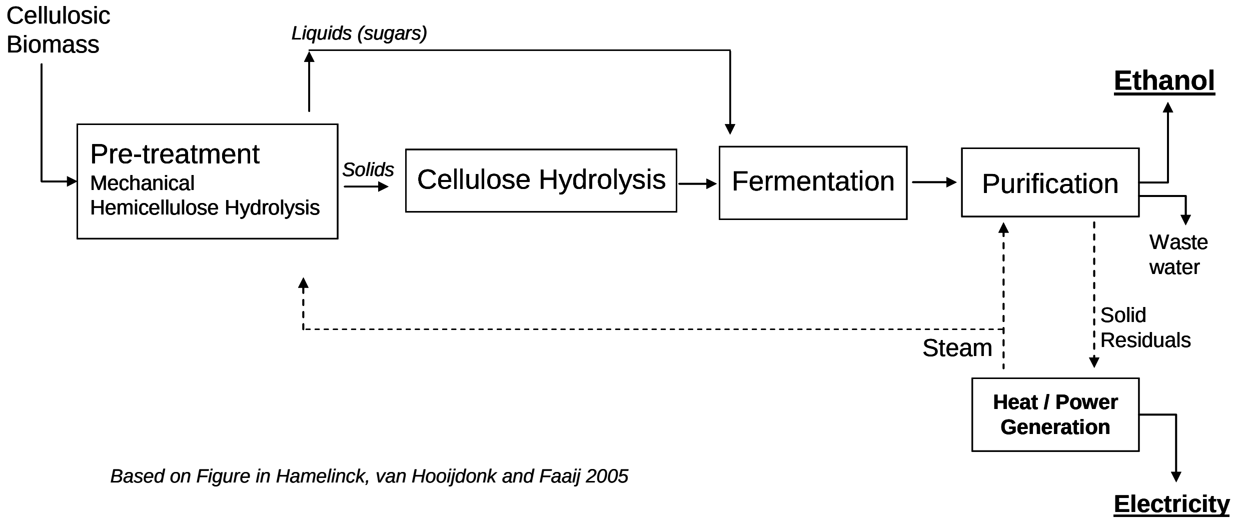
[25] "Other" includes ash, acids, and extractives (low molecular weight organic materials like aromatics and alcohols).

[26] Hydrolysis is the breakdown of organic materials through the use of water.

interest in this type of biorefinery, and sufficient data publicly available for a detailed characterization of this process for this analysis. This process is detailed in the "Pulp and Paper Biorefining" section later in this report.

Figure H-4. Schematic Diagram of LCEt Hydrolysis and Fermentation Process.



Based on Figure in Hamelinck, van Hooijdonk and Faaij 2005

Pretreatment and Conditioning

Pretreatment is an important part of the conversion process because it has a very significant impact on the ethanol yield. It is also a very energy intensive step – sizing the biomass can take up to one third of the total energy requirement for the process. Hemicellulose saccharification is also included in the pretreatment step, as it frees the cellulose for conversion to sugars. In fact, cellulose hydrolysis without the hemicellulose hydrolysis pretreatment step typically has yields below 20%, while after this pretreatment step the cellulose hydrolysis can have yields greater than 90%. Some of the primary hemicellulose hydrolysis methods are described below.

- ***Chemical pretreatment processes, such as dilute acid***: The acid catalyzed hydrolysis process uses dilute sulfuric, hydrochloric, or nitric acid.[27] This method is currently available and gives reasonably high sugar yields (75-90% for a concentrated acid process; 50-75% for a dilute acid process). However, the acid and some inhibitors that are products of the pretreatment (furfural, for example) must be recovered or eliminated from the pretreated material in order to reduce acid consumption, and speed fermentation. This pH-reduction step takes place immediately prior to enzymatic hydrolysis and fermentation, and is done by separating remaining solids from the liquid, followed by washing and pressing. For concentrated acid processes, the acids are then removed or recovered from the liquid fraction via membrane separation (about 80% effective recovery) or by continuous

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[27] Note that there is also a concentrated acid hydrolysis process, but it is currently very expensive and will not be discussed in detail here. This is a pretreatment process intended to hydrolyze hemicellulose and to make cellulose more vulnerable to subsequent hydrolysis. Thus, both dilute and concentrated acid processes should be considered distinct from acid hydrolysis of cellulose, which is a subsequent process intended to attack cellulose, rather than lignin or hemicellulose as in pretreatment.

ion exchange (97% recovery with 2% of sugars lost). What acid is not recovered must be neutralized before fermentation—low acid concentrations can be neutralized by adding lime, causing a conversion to gypsum, which can then be filtered out for disposal.[28] Conditioning is also used to remove inhibitors produced during pretreatment, as they reduce ethanol yields.

A further complication of acid pretreatment is the very small biomass particle size it requires, typically about 1 to 3 mm, as smaller particles have larger percentages of surface area available (Hamelinck, van Hooijdonk and Faaij, 2005). This means that the energy that must be devoted to grinding and milling the biomass feedstock is greater than in other pretreatment options.

- ***Physical pretreatment processes such as steam explosion and liquid hot water (LHW):*** In steam explosion, biomass is heated using high pressure steam for several minutes, and the reaction is stopped by sudden decompression to atmospheric pressure. This process results in sugar yields around 45-65% (Hamelinck, van Hooijdonk and Faaij 2005). LHW uses compressed hot liquid water to hydrolyze the hemicellulose, resulting in high yields of 88-98%. These high yields and the fact that no acid or chemical catalyst is required make the process extremely attractive. However, LHW is currently in the demonstration stage of development, and is not expected to be available for large scale commercial projects for another 5 years. Biomass feedstock size required for these processes are around 19mm, much larger than the dilute acid hydrolysis step (Hamelinck, van Hooijdonk and Faaij 2005). This significantly lessens the pretreatment energy requirement.

- ***Biological pretreatments (Fungi):*** The process is characterized by low energy use and mild environmental conditions. However, it also has low conversion yields and long reaction times. This method tends to be more effective in combination with chemical treatments (Hamelinck, van Hooijdonk and Faaij 2005).

Cellulose Hydrolysis

Cellulose hydrolysis follows one of the pretreatment options noted above, which has made the cellulose more vulnerable to chemical attack, and has hydrolyzed hemicellulose. Cellulose hydrolysis can be performed in two ways: ***enzymatic hydrolysis*** and ***acid hydrolysis***.

The most common of these options is enzymatic hydrolysis. In this process, cellulase enzymes convert the cellulose to sugars under mild process conditions.[29] This method results in relatively high yields (75-85%) and has lower maintenance costs than acid hydrolysis (described below). Cellulase is currently a very expensive part of this ethanol production system, although there are research activities that focus on lower costs and increasing yields.[30]

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[28] There is typically 0.02 kg gypsum produced per kg feedstock after recovery. With acid catalyzed hydrolysis with no acid recovery this can be as high as 0.6 - 0.9 kg of gypsum per kg of feedstock.

[29] Cellulase enzymes are actually a complex mixture of enzymes, produced by organisms that live on cellulosic material (Hamelinck, van Hooijdonk and Faaij 2005).

[30] Key enzyme producers including Genencor International and Novozymes Biotech have both been successful in these developments in recent DOE-supported research.

Yields are expected to increase to 85-95% by 2020. Recovery and reuse of the enzymes also helps to minimize costs, although they do degrade over time.

Acid hydrolysis of cellulose is the only alternative currently to enzymatic hydrolysis. Acid hydrolysis can only be used in conjunction with an acid hydrolysis pretreatment (2-stage acid process). *Dilute acid hydrolysis,* the oldest cellulosic conversion to ethanol process (the first commercial plant was operating in 1898), is not currently competitive with the enzymatic hydrolysis process. However, *concentrated acid hydrolysis* has significantly higher conversion rates than dilute acid hydrolysis, with sugar yields around 90%. Since acid hydrolysis has high operating costs, minimal acid use and maximal recovery are needed.

Fermentation:

In the fermentation process, microorganisms (bacteria, yeast, or fungi) convert carbohydrates to ethanol in an anaerobic environment. The chemical reactions for conversion of 5-carbon and 6-carbon sugars to ethanol are shown below.

$$3C_5H_{10}O_5 \rightarrow 5C_2H_5OH + 5CO_2 \tag{1}$$

$$C_6H_{12}O_6 \rightarrow 2C_2H_5OH + 2CO_2 \tag{2}$$

Genetic engineering has been used to generate bacteria and yeast that can coferment both types of sugars (Hamelinck, van Hooijdonk and Faaij 2005). Continued research to provide microorganisms with higher conversion efficiency and resistance to ethanol poisoning and harsh process conditions would further improve process economics. According to McAloon et al. (2000), the hydrolysis and fermentation steps to produce ethanol from lignocellulosic biomass may take up to seven days. Other research suggests that this time may be reduced to as few as two days.[31] For comparison, the existing mill technology for converting starch to ethanol (not including pretreatment) only takes two days.

McAloon et al. (2000) estimate that 39 lb lignocellulosic residue per gallon of fuel ethanol is produced using a dilute acid pretreatment process with corn stover feedstock. The residue typically has a moisture content around 60%. Potential co-products from ethanol production include electricity, cell matter, furfural, and acetic acid.[32]

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[31] Please note that fermentation time is a function of desired conversion (which follows a curve of reducing time benefit), fermentation type (batch, semibatch, or continuous), and fermentation organism chosen. Also, long-term projections include reduced time for fermentation. For example, Wooley et al. (1999) estimate a decreasing residence time requirement as technology and microorganisms are further developed, such that fermentation only requires two days in the long-term scenario. As there is not a “standard” industry organism at this stage of development, it is misleading to state a single fermentation residence time value for the entire industry. Seven days is used here only as a conservative time estimate in order to generate a conservative cost estimate; this will be detailed in the next subsection.

[32] As detailed studies have not been made on the development of these co-products, they will not be included in this analysis. Electricity is the only by-product from the process evaluated here.

2.5.2 Performance and Cost Analysis

Projected cost and performance for ethanol production from the LCEt-HF process has been developed by looking for different pretreatment options for three distinct time periods. The time periods in this analysis include near term (2015), mid term (2015-2025), and long term (2025+). The near term model only includes dilute acid pretreatment and early commercial sizes (25 to 60 MGY). The mid term model uses dilute acid pretreatment for medium size facilities (60 to 100 MGY) and steam explosion for large facilities (>100 MGY). The long term model uses dilute acid pretreatment for medium size facilities (60 to 100 MGY) and liquid hot water (LHW) for large facilities (>100 MGY). (The latter technology is also discussed in the Pulp and Paper Biorefining section of this report.) The pretreatment technology and size ranges for each time period were selected based on the projections from NREL, Utrecht University, and other institutions (details given below).[33]

This LCEt-Hydrolysis Fermentation production model has been derived based on available data from published studies.[34] All of the studies used to develop the analysis use enzymatic hydrolysis for cellulose conversion, and a variety of pretreatments for hemicellulose hydrolysis (dilute acid, steam explosion, and LHW). The references use either poplar or corn stover feedstocks, and all except one use a boiler for power and steam production (there is one that uses biomass integrated gasification combined cycle - BIGCC).

Figure H-5 shows the values reported in various studies for ethanol conversion efficiency (bars) and yield (points) for various time periods.[35] The conversion efficiency and yield are directly related. The conversion efficiency is calculated as the ratio between the heating value (HHV) of the ethanol produced and the HHV of the feedstock input. The reported values include a variety of pretreatment technologies over a range of projected time periods. There is a clear correlation between ethanol conversion efficiency (yield) and time period, as the long term processes are projected to approach theoretical conversion efficiencies. This development of production performance will be an important factor in the economic favorability of these technologies.

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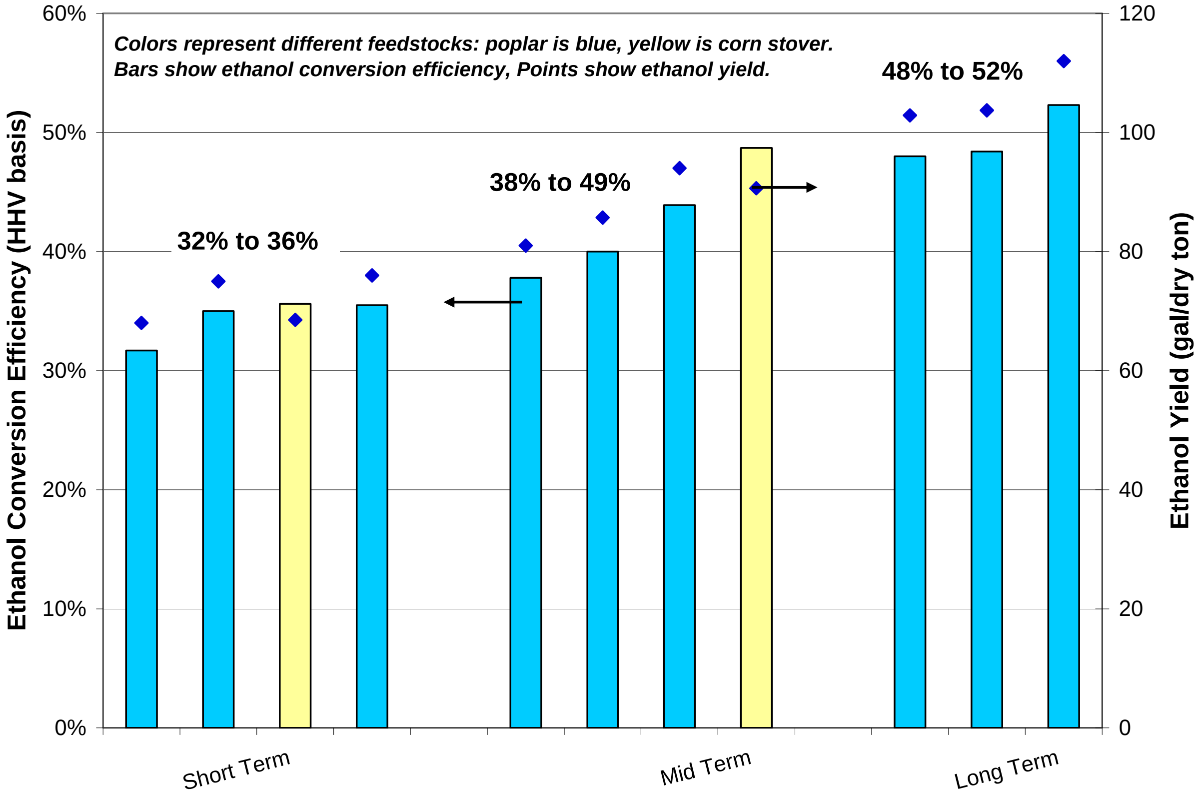
[33] The combinations of facility size and pretreatment technology chosen for each time period reflect both the practicality of using a particular pretreatment process at small or large scales, and the expected degree of development of each type of pretreatment process. Dilute acid hydrolysis, for example, is technically viable in the near term, while steam explosion will not be a commercially demonstrated technology until at least the mid-term, and will be better suited for larger facilities due to its lack of membrane recovery processes.

[34] Sources: Hamelinck, van Hooijdonk and Faaij 2005; Wooley et al. 1999; Lynd et al. 2005; Aden et al. 2002; and McAloon et al. 2000.

[35] Ibid

Figure H-5. Ethanol Production Efficiency and Yield from Published Data for Various Time Periods.



Note: Arrows are used on the figure to reference which axis pertains to each series. The bars represent ethanol conversion efficiency (left axis), while the points show ethanol yield (right axis).

A near term conversion process based on dilute acid hydrolysis with a 35% conversion to ethanol has an overall process efficiency of about 60% including electricity production from lignin and other residuals in a boiler (Hamelinck, van Hooijdonk and Faaij 2005). Improvements in the pretreatment methods and other processes could lead to a 48% ethanol conversion efficiency (or greater), and overall process efficiencies of 68% (Hamelinck, van Hooijdonk and Faaij 2005).

Figure H-6 shows the total capital investment in relation to the ethanol production capacity and time period.[36] As with the ethanol production performance data discussed above, these values are from a number of studies and include several pretreatment technologies. As there are no commercial lignocellulosic ethanol production facilities in operation,[37] these values have been derived using factored estimation analysis[38] or detailed process design. The

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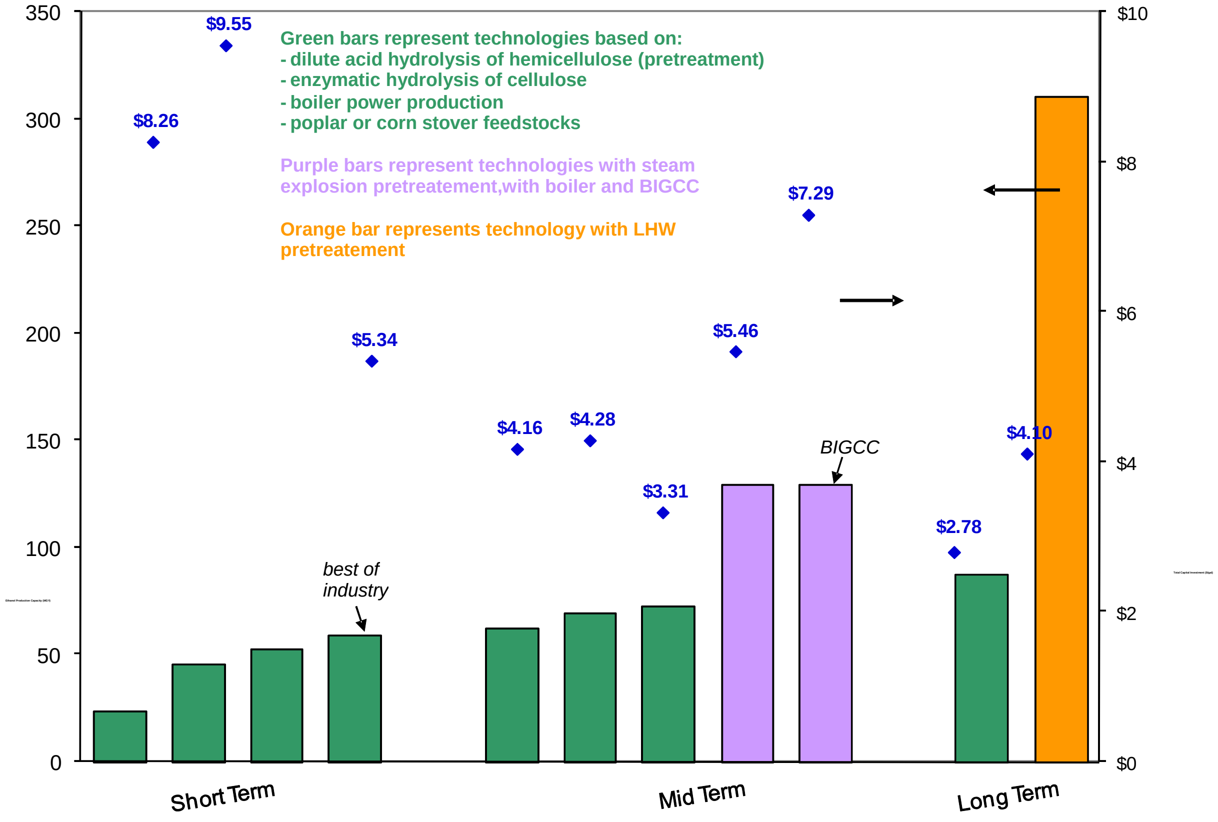
[36] Ibid (except Lynd et al. 2005).

[37] Iogen currently has an operational demonstration facility in Canada. A number of companies have demonstration and commercial facilities planned for start-up in the near-term (Solomon, Barnes and Halvorsen 2007).

[38] Factored estimation is a cost estimation method in which a detailed major equipment cost is calculated, and other cost components, such as working capital, contingency, and insurance are estimated as certain percentages of this major equipment cost. This is commonly used in construction estimation.

Total Capital Investment (TCI) is generally +/- 30% (or better), and includes the equipment costs, installation, contingency and other indirect costs.[39]

Figure H-6. Ethanol Production Capacity and Total Capital Investment for Various Time Periods.



As expected there is a clear trend towards increased plant sizes with time. The average short term plant size is 45 MGY, while the long term facilities are projected to be about 100 to 300 MGY. Furthermore, the capital cost decreases with time, which corresponds with the economies of scale as plant sizes increase, as well as the availability of lower cost equipment.

In the short term, the highest capital cost is that of the "best of industry" (labeled in graph), which has a TCI of $5.34/gallon. This reflects advanced equipment, and more efficient processes. The largest capital cost in the mid term is the facility that includes gasification for power production instead of a boiler (labeled "BIGCC"). The TCI for this facility is $7.29/gal, significantly higher than the equivalent facility that uses a boiler ($5.46/gal), reflecting the current infancy of gasification as a commercial technology. As both of these examples have significantly

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[39] All costs are reported in 2008 US$, converted using the Chemical Engineering Plant Cost Index (CEPCI) where necessary.

different technologies from the other modeled facilities, their cost information is not included in the average used for the ethanol production cost model.

The relative costs of the major equipment areas from several studies for near term LCEt Hydrolysis and Fermentation facilities using dilute acid pretreatment technology are shown in Table H-7. Note that the first two facilities include cellulase production, while the 69 MGY plant does not.[40] Note that at least for the near term, it is more likely that enzymes would be purchased rather than produced.

The power and steam production area is the most expensive, followed by pretreatment. This suggests that lowering the costs of these process steps could be most beneficial in reducing overall capital costs.

Table H-7. Capital Cost Break-Out for LCEt Hydrolysis and Fermentation with Dilute Acid Pretreatment.

Plant Size (MGY)	**25**	**52.5**	**69**
Major Equipment Area		***% of total cost***	
Feedstock Handling	4.0%	3.6%	6.6%
Pretreatment	21.9%	19.5%	23.6%
Saccharification / Fermentation	10.6%	9.9%	8.3%
Cellulase Production	13.3%	11.5%	0.0%
Distillation / Separation	10.5%	9.6%	19.2%
Power / Steam Production	27.6%	33.0%	33.7%
Other*	12.2%	12.9%	8.8%
Total	100.0%	100.0%	100.0%
Source	*McAloon et al. 2000*	*Wooley et al. 1999*	*Aden et al. 2002*

** Other includes wastewater treatment, storage, and utilities*

<u>Outline for model analysis</u>

Due to the complexity of the LCEt Hydrolysis and Fermentation process, it was necessary to use a number of assumptions to generate a simple and useable model. The cost and performance model was developed as follows:

- Hemicellulose hydrolysis (pretreatment) varies based on facility size and time period
    - The only available technology for short term (2010) is dilute acid pretreatment, in the size range from 25 to 60 MGY.

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[40] While some of the LCE Fermentation/Hydrolysis facilities from the published studies used in the analysis include on-site production of cellulase for enzymatic hydrolysis, others assume the cellulase is purchased for use, depending on which option was more economically favorable. The cost of purchased cellulase is included in the consumables portion of the Variable O&M costs. While Consolidated Bioprocessing (CBP) is a future possibility, insufficient publicly available cost data exists to include this as part of the model.

  - Mid term (2015-2020) pretreatment technologies are dilute acid for medium size facilities (60 to 100 MGY),[41] and steam explosion for large facilities (> 100 MGY).
  - Long term (2025+) pretreatment technologies include dilute acid for medium size facilities, and Liquid Hot Water (LHW) for large facilities.
  - It is assumed that small facilities (< 60 MGY) are not constructed after the short term time period.

- Cellulose hydrolysis step is performed using enzymatic hydrolysis (with cellulase) for all modeled processes.

- Cellulase is produced on-site for most scenarios used in this analysis, although in some cases it is purchased. The selection for on-site production versus purchase of cellulase follows the choices reported in the literature for each technology option. On-site cellulase production will affect O&M costs, capital costs, and overall ethanol yields, as a fraction of the sugar stream is diverted to feed the microorganisms. Purchased cellulase only affects O&M costs, and is generally more expensive overall. Consolidated Bioprocessing (CBP) is a third possibility in which cellulase is produced by the fermentation organism itself and cellulase need not be purchased or produced independently. Insufficient publicly available cost data exists to include this option as part of the model. A CBP scenario would have both a different capital cost breakdown and different O&M costs.

- Electricity and steam are produced with a boiler and turbogenerator from residual solids (lignin, unconverted cellulose, cell mass), and unconverted sugars that are partially dried to a syrup. This was assumed for all time periods, but it must be recognized that in the near term, on-site heat and power generation may not be adopted by every facility.

- Electricity is the only co-product produced from the process, and is valued at 7.97¢/kWh based on the average industrial electricity cost from Energy Information Administration (EIA 2009).[42] Note that EIA's projections for industrial electricity for the 2010 to 2030 time period is 5.8¢/kWh, which is significantly lower than the value used (EIA AEO 2007).

Selection of feedstock type is a critical component of the model, as the composition of the biomass affects the ethanol production. In particular, the hemicellulose, cellulose, and lignin content of the feedstock affect yield (gallons of ethanol produced per dry ton of feedstock). The pretreatment technology also has a significant effect on the ethanol yield, as each technology has a different conversion factor. The feedstock composition and yield in turn affect the material available for power production – a portion of the unfermented sugars are dried and the resulting

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[41] The size ranges used were chosen for consistency with the NREL research from which their cost parameters were in part derived (Hamelinck, van Hooijdonk and Faaij 2005).

[42] After modeling had been completed for all scenarios, it was determined that this price was incorrectly determined and a value of 7.01¢/kWh should have been used instead. This change will be integrated into spreadsheet technology models before any further modeling is performed.

syrup is fired in the boiler along with the lignin and cell mass. In addition, different feedstocks will have different costs and availability.

The conversion rates of hemicellulose and cellulose into sugars and the fermentation of sugars into ethanol are technology dependent. These rates also increase with time, such that for each pretreatment process the conversion rates in the long term are significantly higher than in the short term rates. Conversion rates for saccharification of cellulose and hemicellulose and fermentation of the glucose and other sugars are expressed as a percent of theoretical conversion for each technology available in each time period, as shown in Table H-8.

Table H-8. Conversion Rates for LCEt Hydrolysis and Fermentation.

	Conversion (% of theoretical) (1)				
	Short Term	***Mid Term***		***Long Term***	
Pretreatment Technology	*Dilute Acid*	*Dilute Acid*	*Steam Explosion*	*Dilute Acid*	*LHW*
Saccharification					
Hemicellulose	83%	85%	55%	85%	93%
Cellulose	75%	85%	93%	90%	98%
Fermentation					
Xylose & other sugars (2)	86%	90%	85%	95%	94%
Glucose (2)	93%	94%	93%	95%	94%
Source	*Hamelinck et al. 2005, Wooley et al. 1999*	*Wooley et al. 1999*	*Hamelinck et al. 2005*	*Wooley et al. 1999*	*Hamelinck et al. 2005*

1) Average conversion estimates based on poplar feedstock and the reported pretreatment technologies. The advanced technologies for mid and long term pretreatments may not be suitable for all feedstocks.

2) Cellulose is converted to glucose during saccharification, while hemicellulose is converted to xylose and other sugars

The theoretical ethanol yield from biomass feedstocks is derived based on the dry weight of fraction of C-5 and C-6 sugars. Theoretical yield is equal to the total weight percent of C6 sugars times 172.82, plus the total weight percent of C5 sugars times 176.87 (U.S. DOE EERE Biomass Program 2006).[43] The calculation for actual ethanol yield is very similar, except that the percent of theoretical conversion for each step is taken into account.

Total electricity production depends on the amount of material available for boiler fuel and the steam use. These are both related to the facility type. The dilute acid process and LHW pretreatment have high steam requirements, which implies lower electricity production. Facilities with higher conversion rates will have fewer unconverted sugars to use as boiler fuel.

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[43] These factors are derived from the following conversions: 1.11 pounds of C6 sugar per pound of C6 polymeric sugar, 1.36 pounds of C5 sugar per pound of C5 polymeric sugar, each pound of sugars can produce a maximum of 0.51 pounds of ethanol, and there are 6.55 pounds of ethanol per gallon.

Operation and Maintenance (O&M) costs are separated into fixed and variable components. The fixed O&M includes annual costs that are a function of the plant's capacity rather than its actual annual ethanol production. These costs include labor, maintenance, taxes and insurance. Variable O&M costs include raw materials (such as sulfuric acid and lime for dilute acid pretreatment, ammonia and corn steep liquor for cellulase production, or purchased cellulase if it is not made in the process) and solids disposal (i.e. ash and gypsum). The O&M values used in the model are derived from published studies for each technology type.

The key values and assumptions used to calculate the mid-term (2015) technology analysis and economic performance model for LCEt Hydrolysis and Fermentation are described below. An example of the detailed analysis for all periods is shown in Appendix H-F.

- Feedstock type and input quantity are variables. Applicable feedstocks include corn stover, wheat straw, switchgrass, poplar (hardwood) and pine (softwood), and portions of the MSW stream (see Appendix H-G for details on biomass from MSW). These feedstocks were selected based on available composition data, and can be used to represent a range of wood and agricultural residues.

- The applicable size range for facilities in this time period is 60 to 150 MGY, based on facility sizes from sources (Hamelinck, van Hooijdonk and Faaij 2005; Wooley et al. 1999; Aden et al. 2002). Facilities with capacities from 60 to 100 MGY are modeled to use dilute acid pretreatment, while larger facilities (> 100 MGY) use steam explosion pretreatment.

- Ethanol yield is based on the feedstock composition and pretreatment technology. Biochemical compositions (hemicellulose, cellulose, and lignin percentages) for the feedstocks are based on data from U.S. DOE EERE Biomass Program (2006) and Hamelinck, van Hooijdonk and Faaij (2005), as shown in Table H-6 above. The actual yields are calculated based on projected conversion of hemicellulose and cellulose to sugars during saccharification, and the fermentation of these sugars to ethanol. The projected conversion rates for each technology are given in Table H-8 above. Yield data for various feedstocks are shown in Table H-9.

- The total conversion efficiency (HHV) is 38-41% for dilute acid and 36-40% for steam explosion, depending on the feedstock type. Corn stover and pine have the highest conversion rates of the selected feedstocks.

- Capital cost data is based on average of results from (Hamelinck, van Hooijdonk and Faaij 2005; Wooley et al. 1999; Aden et al. 2002; and McAloon et al. 2000).[44] Each pretreatment technology has a different base capital cost, which is scaled by facility size using a typical scaling factor for biofuels of 0.8. The base cost for a mid term dilute acid facility is \$3.92/gal for a 68 MGY facility, and the cost for a 130 MGY steam explosion facility is \$5.46/gal.

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[44] Note that the BIGCC case and best of industry are excluded from capital cost estimates, as they are not representative of an average facility.

- Net electricity production is based on heat and power production with partially dried solid residuals (lignin, cell mass) and unfermented sugars (dried to a syrup), fired in a boiler. The amount of electricity generated is dependent on the pretreatment technology. For the mid-term period, electricity production for the dilute acid process is about 2 kWh per gallon of ethanol produced, following Wooley et al. (1999), while steam explosion pretreatment gives about 6.3 kWh/gal based on data from Hamelinck, van Hooijdonk and Faaij (2005). Note that the net electricity production for the steam explosion pretreatment is much higher than other technologies as this method has lower steam requirements. Furthermore, cellulase is not made on site in the steam explosion case, which means a larger quantity of non-fermented sugars and residuals are available for power generation.

- Annual water consumption is calculated based on the annual feedstock input, using data from Wooley et al. (1999) for near-term dilute acid process. Potential changes in water consumption for different time periods or for different pretreatment processes are not included in this analysis.

- The CO_2 production stream is estimated by assuming that the amount of CO_2 produced is approximately 96% of the weight of ethanol produced, since the theoretical max yield is 51% ethanol and 49% CO_2 (by weight) from sugars.

- Annual Fixed and Variable O&M costs are calculated based on data in published studies. Mid-term dilute acid O&M costs are based on Wooley et al. (2005), and steam explosion data is from Hamelinck, van Hooijdonk and Faaij (2005). Variable O&M includes raw materials (sulfuric acid & lime for dilute acid, ammonia & CSL for cellulase production, other chemicals) and solids disposal (i.e. ash and gypsum). Fixed O&M includes labor, maintenance supplies, insurance and property taxes.

- The co-product credit for electricity is based on the net electricity available for export and a current electricity value of $0.0797/kWh, based on the 2008 average wholesale price of electricity for industrial sector from EIA (2009) (see footnote 35).

- The levelized non-feedstock production cost for ethanol is calculated based on the annual operating expenses and capital cost payment, co-product credit, and quantity of fuel produced.

Table H-9. Feedstock Composition and Yield Data.

	Yield	Actual Yield (gal/ton) (2)				
Feedstock	**Theoretical Max Yield (1) (gal/dry ton)**	**Short Term (Dilute Acid)**	**Mid Term (Dilute Acid)**	**Mid Term (Steam Explosion)**	**Long Term (Dilute Acid)**	**Long Term (LHW)**
Agricultural Resources						
Corn Stover	107.6	72.6	80.6	77.3	86.0	82.0
Wheat Straw	98.8	69.3	76.8	73.6	82.0	77.9
Grass Resources						
Switchgrass	99.7	70.0	77.4	74.4	82.6	78.7
Woody Resources						
Hybrid Poplar	109.9	76.9	85.9	81.2	91.9	98.6
Pine	115.1	80.9	90.2	83.7	96.4	103.5
Other Resources						
Yard Trimmings / Green Waste	108.0	63.9	70.0	68.6	74.4	73.4
Mixed Paper	116.2	76.8	86.0	83.5	92.1	98.8

1) Maximum yield values based on sugar content, calculated from EERE's Theoretical Ethanol Yield Calculator (http://www1.eere.energy.gov/biomass/ethanol_yield_calculator.html)

2) Actual yields calculated based on average conversion values for all feedstocks in short term, and wood for all terms. Agricultural and grass feedstock yields for advanced mid and long term pretreatment technologies based on average increase of wood feedstock yields from short to mid term (4.4% of theoretical), as the conversion technology information does not apply to these materials, but other comparable advanced technologies may apply.

Model Results:

Some of the key cost and performance results for projected mid-term LCEt Fermentation/ Hydrolysis processes are shown in Table H-10 and Table H-11. These tables show results for the range of applicable facility sizes based on selected input quantities, illustrating the high and low end costs for each pretreatment technology. The analysis shows that in general the non-feedstock production cost of ethanol for dilute acid is significantly lower than for steam explosion. Although the steam explosion process generates significantly more electricity, it also has much higher capital and O&M costs than dilute acid. However, steam explosion has environmental and other benefits that are not accounted for in this simple economic analysis, which may encourage development of these facilities.

Table H-10. Example Cost and Performance - Mid Term Dilute Acid LCEt Hydrolysis and Fermentation.

Sample Model Results - LCEt Fermentation/Hydrolysis w/ Poplar (dilute acid)		
Feedstock Input (dry ton/yr)	700,000	1,160,000
Ethanol Yield (MGY)	60	100
Electricity Production (MWh/yr)	120,800	200,200
CO2 Stream (ton/yr)	190,400	315,500
Water Consumption (1000 gal/yr)	373,900	619,600
Capital Cost (Million $)	$ 241.1	$ 361.2
Annual O&M Costs (Million $/yr)	$ 21.0	$ 34.8
By-Product Credit (Million $/yr)	$ 9.6	$ 16.0
Non-feedstock Production Cost ($/gal)	***$ 0.68***	***$ 0.63***

Table H-11. Example Cost and Performance - Mid Term Steam Explosion LCEt Hydrolysis and Fermentation.

Sample Model Results - LCEt Fermentation/Hydrolysis w/ Poplar (steam explosion)		
Feedstock Input (dry ton/yr)	1,250,000	1,850,000
Ethanol Yield (MGY)	101	150
Electricity Production (MWh/yr)	642,160	950,390
CO2 Stream (ton/yr)	321,500	475,800
Water Consumption (1000 gal/yr)	667,700	988,200
Capital Cost (Million $)	$ 578.8	$ 791.9
Annual O&M Costs (Million $/yr)	$ 46.7	$ 69.0
By-Product Credit (Million $/yr)	$ 51.2	$ 75.7
Non-feedstock Production Cost ($/gal)	***$ 0.89***	***$ 0.61***

Hamelinck, van Hooijdonk and Faaij (2005) estimate the mid term ethanol production cost from poplar using steam explosion pretreatment is about $1.67 per gallon for a facility with 600,000 dry ton per year input. Using similar feedstock cost ($67.16 per dry ton) and facility size gives a comparable ethanol production cost from the model, about $1.50/gal. For further comparison, Aden et al. (2002) estimate an ethanol production cost of $1.59/gal for a 69 MGY dilute acid facility using corn stover feedstock.[45] The feedstock cost from the study is about $44.77/dry ton. This gives an ethanol production cost from the LCEt model of $1.23/gal for this configuration, slightly lower than the projection from Aden et al (2002). Different economic analysis methods may contribute to this variation.

2.5.3 Outlook

LCEt-HF has garnered a significant amount of attention in recent months. Federal Renewable Fuel Standards have mandated at least 7.5 billion gallons of renewable biofuels in the nations fuel supply by 2012. More importantly, the terminology used in the RPS creates an important distinction between grain ethanol and lignocellulosic ethanol. As an "advanced biofuel," cellulosic ethanol counts over twice as much as corn ethanol toward this goal, at 2.5 instead of one compliance unit. This, combined with recent federal emphasis on advanced, non-grain biofuels, has meant that much R&D and many grants and subsidies are being funneled into LCEt-HF.

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[45] Costs have been converted to 2008$ using CEPCI.

New York State

In New York State, one such grant of $25.8 million was issued by the Pataki administration in 2006 to two cellulosic ethanol startups attempting to build demonstration facilities. Mascoma Corporation was awarded $14.8 million for the construction of a 500,000 gpy demonstration facility in Rome, NY, and the remainder was earmarked for Catalyst Renewables Corporation for a 130,000 gpy pilot plant in Lyonsdale, NY. Mascoma's facility began producing ethanol in January, 2009 and the company plans to build a commercial-scale facility in Kinross, MI in mid-2010. Mascoma recently entered into an agreement with Chevron Technology Ventures, in which it will supply the lignin byproduct from its pilot plant to Chevron so that the properties and possible value-adding products of the lignin can be explored. The Catalyst Renewables project is currently in a research partnership with the SUNY College of Environmental Science and Forestry. A construction date has not yet been set.

National Efforts

On a national scale, the mandates of the EISA (discussed previously) have put cellulosic ethanol into the spotlight as a commercializable "advanced biofuel." Many companies have begun developing processes and test facilities to study cellulosic ethanol production via hydrolysis and fermentation. Some notable efforts include:

Abengoa Bioenergy (The Netherlands). In 2003, Abengoa Bioenergy was awarded $35.5 million by the DOE in order to research and develop an integrated biorefinery process for cellulosic ethanol production based on distiller's grains and corn stover. By 2007, the company had begun operating a 1.4 ton per day pilot biorefinery. It now also has a 70 ton per day demonstration facility and plans a commercial-scale integrated biorefinery in Hugoton, KS with DOE assistance. This facility is to use both LCEt-HF and thermochemical (discussed in the next section) methods (Abengoa 2009).

BlueFire Ethanol (Irvine, CA). Another awardee of DOE funds for commercialization of biorefining, BlueFire Ethanol uses concentrated acid hydrolysis on a variety of feedstocks, including rice and wheat straw, agricultural residues, and sorted municipal solid waste. It has undergone permitting and is awaiting final financing for a 3.9 MGY facility in Lancaster, CA to use sorted municipal solid waste, and is planning another such facility in Fulton, MS (BlueFire 2009).

DuPont Danisco (Itasca, IL). This joint venture between DuPont and Genencor was formed to focus on the production of bulk biocatalysts and packaged biorefineries using corn cobs and bagasse as feedstocks for ethanol production. DuPont Danisco has since worked with the University of Tennessee to plan and construct a 250,000 gpy, $40 million demonstration facility in Monroe County, TN. This plant is to use corn cobs, switchgrass, and fibrous biomass as feedstocks (DuPont 2009).

POET (Souix Falls, SD). This established corn ethanol produced began researching cellulosic ethanol production via hydrolysis and fermentation and developed the BPX process, which it claims reduces cooking requirements during pretreatment. The company's 20,000 gpy demonstration plant in Scotland, SD uses corn cobs and corn stover as feedstocks. POET plans a commercial-scale, 125 MGY plant using this technology as an add-on to its Emmetsburg, Iowa grain ethanol facility. This is to begin operating in late 2011 (POET 2009).

This influx of both venture capital and research breakthroughs, combined with the environmental/sustainability benefits of LCEt-HF (see the chapter later in this report on emissions) compared to grain ethanol, its near-term commercialization potential, and its reduced threat of direct food replacement will make this technology and others like it a common choice for large scale deployment of biofuels.

2.6 LIGNOCELLULOSICS TO ETHANOL: THERMOCHEMICAL CONVERSION AND FERMENTATION

In the Lignocellulosics to Ethanol via Thermochemical conversion and Fermentation process, gasification is used to convert biomass feedstocks into a syngas rich in CO and H_2, which is then fermented to produce ethanol. As this technology is based on gasification, it has a wide range of feedstock flexibility. Furthermore, unlike the LCEt Hydrolysis and Fermentation process, this process uses the lignin part of the biomass directly to contribute to syngas production. Gasification requires dried and sized material, so biomass feedstocks must be pretreated before use (Antares 2008).[46]

2.6.1 Technology Description

Gasification Overview

Gasification is the process of converting a solid fuel into a gaseous fuel. Gasification is appealing for biomass because a wide variety of materials, including complex biomass structures like lignin, can be gasified to create a uniform syngas feedstock. There is a lot of flexibility in the end use of the gas, provided it is cleaned of contaminants. Syngas can be used to make biobased liquid fuels and chemicals, or it can be used in high-efficiency power generation systems such as a gas turbine-combined cycle and fuel cells.

The two main reactions occurring during gasification are as follows:

The Boudouard Reaction: $C + CO_2 \Leftrightarrow 2\,CO$ (3)

The Water-Gas Reaction: $C + H_2O \Leftrightarrow CO + H_2$ (4)

These reactions require heat input. In simple gasifiers the heat is drawn from the combustion of the pyrolysis gases formed during gasification, while in more complex gasifiers it is provided externally. These reactions produce more syngas from the biomass and typically leave little unreacted carbon in the ash.

The main challenges facing biomass gasification are as follows:

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[46] The text in this section was revised from material originally presented by Antares in 2008 in a *Strategic Assessment of Bioenergy Development.*

- The high moisture content of un-dried woody biomass;
- Commercial availability of gas clean-up technologies that are sufficiently robust and effective to allow the product gas to be used in a catalytic reactor, fermentation process, turbine or engine; and
- The relatively high cost of gasification equipment, which is currently built on a custom basis.

Furthermore, gasification for power generation is currently a challenge as there is limited industry experience in using biomass gasifiers in advanced power generation cycles.

In general, the carbon conversion in a gasifier increases as the temperature increases. Furthermore, a recent study has shown that the feedstock is an important factor in the conversion behavior (Moilanen 2006). For example, pine wood has high carbon conversion at relatively low temperatures, whereas pine bark has relatively low conversion efficiency even at higher temperatures. Additionally, straw has high conversion efficiencies above 850ºC, but ash sintering causes operational problems.[47] This study also demonstrated that the reactivity of the biomass (gasification rate) is dependent on the biomass fuel properties, the gasifying agent, and the temperature and pressure conditions during reaction (Moilanen 2006). Some of these effects seem to be related to the behavior of the ash-forming substances during gasification, which are difficult to characterize.

There are currently two principal types of gasifiers: fixed bed and fluidized bed. The former is typically simpler, less expensive and produces a low BTU syngas. The latter is more complicated, more expensive and produces a syngas with a higher heating value.

In fluidized bed systems, steam and/or air are passed through a particle bed and cause the bed to levitate (i.e., become "fluidized"). Fuel is added to the bed, which is often composed of inert heat transfer media such as sand or dolomite. Other materials such as alumina may be used to further enhance the heat transfer. Fluidized bed devices are noted for their high energy output (per area of bed) and fuel flexibility.

Syngas Fermentation to Ethanol

Figure H-7 shows the general arrangement for the LCEt Fermentation/TC process. After gasification, the syngas is conditioned and compressed for fermentation. In the fermentor, anaerobic micro-organisms ferment the syngas to produce ethanol. Although these organisms generally prefer CO to H_2, the syngas can contain any mixture of these gases for conversion. The chemical reactions for ethanol production for this process are as follows:[48]

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[47] Note that these experimental results can be used to describe the characteristics of the fuel, but conversion in a gasifier will be affected by many additional factors, such as the partial pressure of the gasification agent and the size distribution of feedstock particles.

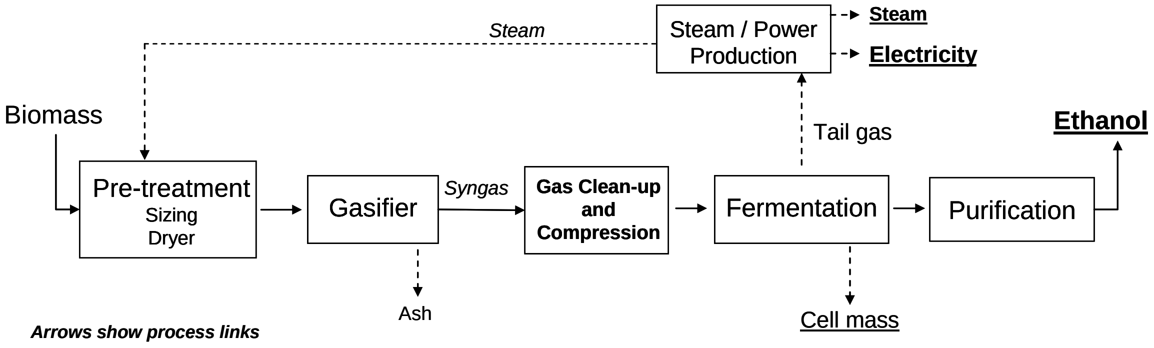
[48] Source: (Spath and Dayton 2003)

$6CO + 3H_2O \rightarrow CH_3CH_2OH + 4CO_2$ (5)

$2CO_2 + 6H_2 \rightarrow CH_3CH_2OH + 3H_2O$ (6)

Typical CO conversions reported in literature for laboratory scale fermentation are ~90%, while conversions for H_2 only about 70% (Spath and Dayton, 2003).

Figure H-7. Schematic Diagram Showing LCEt Fermentation/TC Process.



Source: Diagram based on figure in (Spath and Dayton 2003)

Unlike the by-products of corn ethanol production, the cell mass produced during fermentation of syngas is not currently approved for animal feed[49] (Spath and Dayton 2003). An alternate use for the material is to recycle it to the gasifier for re-processing. The exhaust gas produced from fermentation can be used to produce steam and power. The steam can be used to dry the biomass feedstock during pretreatment.

2.6.2 Performance Analysis

While a very high percentage of the chemical energy in biomass can be converted to syngas, not all of the syngas will be converted to ethanol. Estimates of the ethanol yield produced from biomass with this process can be estimated using the following equation, assuming all of the syngas produced by the gasifier (after conditioning) goes to fermentation:[50]

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[49] Cell mass is also a byproduct of LCEt Hydrolysis and Fermentation, but this has not been widely considered for animal feed due to the fact that it is mixed with lignin and other non-fermented substances that cannot be digested by ruminants. Such a substance would be difficult to sell as animal feed, but relatively easy to gasify or combust for heat and power within the conversion facility.

[50] Source: (Spath and Dayton 2003)

$$Ethanol(MGY) = \frac{Feed_{in}(ton/day) \times HHV(Btu/lb) \times \eta_{gasification} \times C}{1.5 \times 10^{5}}; \tag{7}$$

Where $Feed_{in}$ is the rate of feed input to the gasifier in dry tons per day;

HHV is the higher heating value of the feedstock;

$\eta_{gasification}$ is the cold gas efficiency of the gasifier and gas conditioning; and

C is the conversion rate of CO and H_2 to ethanol, as a fraction of theoretical

Projected yields for LCEt Fermentation/TC process are 70-105 gallons of ethanol per dry ton of feedstock (Spath and Dayton 2003). (Yields from LCEt-HF processes are similar at between 70 and 103 gallons per ton feedstock.) These are lower than the yields from current grain ethanol production technologies, which are around 100 gal/ton. Spath and Dayton (2003) note that this is a result of losses from gasification, gas conditioning, and fermentation steps that partially counteract the benefits of direct use of the lignin component of the biomass.

2.6.3 Outlook

At present, there are numerous small, fixed-bed gasifiers in use around the world. These units typically convert wood and agricultural residues into a low-BTU gas for use in process heating or district heating. These applications have developed as alternatives to traditional boilers. There has been a significant effort to develop large-scale (>10 MW_e) gasifiers for use in a biomass integrated gasification combined cycle (BIGCC). These efforts have not yet resulted in a sustainable commercial operation.

University of Arkansas and Bioengineering Resources Inc (BRI) have been working on LCEt Fermentation/TC technology since the 1980s. BRI has a patented fermentation bacterium for the process (Bioengineering Resources, 2007), and is part of a team that was recently selected to build a demonstration facility with funding from DOE (U.S. DOE, 2007). Several other universities have also been involved in research efforts. A notable private player is highly publicized biofuels startup Coskata, which has received funding from General Motors and Khosla Ventures. With a 40,000 gpy demonstration plant under construction and plans to build a 50-100 million gpy facility in an undisclosed location, the company seems to be the farthest along in commercializing an LCEt-Fermentation/TC process. As yet this technology appears to be a long way from commercialization, therefore there is not sufficient information about the cost and performance for a detailed model.

2.7 LIGNOCELLULOSICS TO MIXED ALCOHOLS: THERMOCHEMICAL CONVERSION

2.7.1 Technology Description

The conversion of Lignocellulosics to Mixed Alcohols via Gasification and Thermocatalysis combines the process described above for syngas production with a catalyst conversion step.[51] It is similar to the LCMD – G/FT process that will be discussed later in the context of diesel substitutes, although the end products are alcohols, not distillates. Figure H-8 shows a block diagram for a typical LCMA Synthesis/TC process. Higher Alcohol Synthesis (HAS) is done in reactors similar to those used for FT and methanol synthesis. The most effective types of catalysts include modified methanol synthesis catalysts, modified FT catalysts, and alkali-doped molybdenum catalysts (Nexant Inc., 2006). The modified methanol synthesis and Mo catalysts show higher alcohol yields than modified FT catalysts. The Mo catalysts also have the best selectivity for higher alcohols, a high tolerance for CO_2, and sulfur in the syngas (Nexant Inc., 2006). However, although this higher sulfur tolerance requires less clean-up of the syngas before conversion, it may require sulfur removal downstream in the mixed alcohol fuel (Antares 2008).[52]

Figure H-8. Schematic Diagram of Thermochemical Ethanol Production Process.

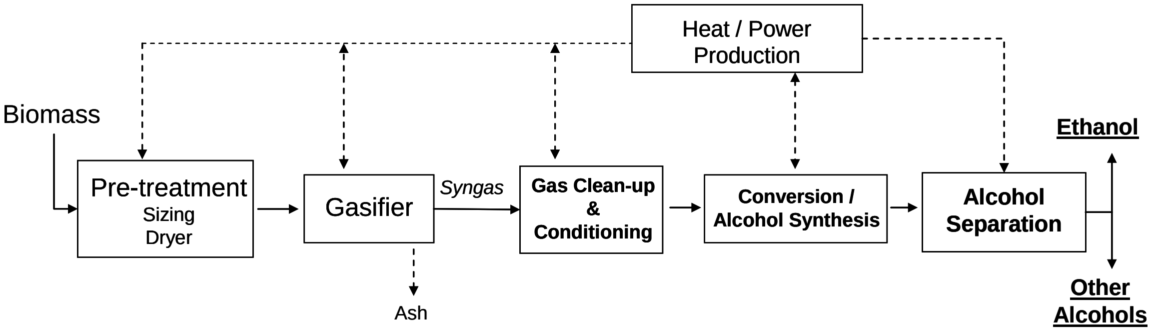

Diagram based on Figure in Phillips et al. (2007)

LCMA Synthesis/TC involves a complex set of reactions that produce a variety of products, depending on the catalyst used and process conditions. The major reactions include methanol synthesis, FT reactions, higher alcohol synthesis, and water-gas shift. The process is optimized at syngas compositions with ratios of $H_2/CO \sim 1$ (Spath and Dayton 2003). Regardless of the type of catalyst used, typically 40 to 90% of the product stream needs to be recycled to maximize mixed alcohol production (Nexant Inc. 2006). The main by-products of the process are CO_2 and water. Large quantities of methane are also often produced.

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[51] See previous section “Lignocellulosics to Ethanol: Thermochemical Conversion and Fermentation” for further details.

[52] The text in this section was revised from material originally presented by Antares in 2008 in a *Strategic Assessment of Bioenergy Development.*

2.7.2 Performance and Cost Analysis

The performance and cost model for LCMA Synthesis/TC has been developed primarily based on recent work by NREL (Phillips et al. 2007). In this report, NREL developed a detailed process model and economic analysis for a projected thermochemical ethanol facility. The analysis is based on the DOE estimates for synthesis yields and selectivity for a 2,205 dry ton/day facility (equivalent to 772,000 ton/yr for a plant with 96% availability).

The NREL model configuration uses an indirect steam gasifier and a conventional steam power cycle. The syngas clean-up and conditioning step includes tar reforming, water scrubbing (for cooling and quench), and acid gas removal. Clean syngas is converted to alcohols in a fixed bed reactor, using a MoS_2 catalyst with very high ethanol selectivity.[53] The alcohol separation section includes dehydration and separation of alcohols. Methanol is recovered and recycled through the alcohol synthesis section to increase yield of ethanol and higher alcohols.

A portion of the unconditioned syngas is diverted to generate electricity and heat. Although this model produces exactly the amount of energy required to sustain the process (consuming 28% of the syngas), an actual plant could vary the energy production depending on favorable market conditions to buy or sell electricity from the grid.

If all the syngas was converted to alcohols, the ethanol yield would be 110.9 gallons per dry ton of biomass, and the total alcohol production would be 130.0 gal/ton (Phillips et al. 2007). With 28% of the syngas used for energy production, the ethanol yield is 80.1 gal/ton.

Outline for model analysis

The key values and assumptions used to calculate the advanced technology analysis and economic performance model for thermochemical ethanol production via higher alcohol synthesis are described below. An example of the detailed analysis is shown in Appendix H-F. All aspects of the model analysis are based on data from Phillips et al. (2007), unless otherwise noted.

- Feedstock type and input quantity are variables. Applicable feedstocks include woody and agricultural residues. The moisture content of these feedstocks must be less than 50% for optimal operation.
- This model is applicable for facility sizes with greater than 185,000 dry ton biomass input per year, following the size range for Fischer Tropsch synthesis, which has similar technology characteristics.
- As different feedstocks have distinct heating values, the feedstock choice affects the syngas produced in the gasifier and thus overall yields of and composition of the alcohol products. Yield of ethanol and other alcohols are calculated based on the HHV conversion efficiency. Ethanol conversion is equivalent to 38.9% of the input

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[53] This is a modified Fischer Tropsch catalyst based on the former Dow/UCC catalyst, with conversion performance modeled based on target results. In addition to higher total CO conversion and higher alcohol selectivity, the projected distribution of ethanol and methanol used in the NREL model differ from current results. The Phillips et al. (2007) model assumes that 71% of the alcohol production is ethanol, and 5% is methanol. In contrast, the Dow distribution was 30-70% methanol and 34.5% ethanol.

energy from biomass, converted to gallons using a typical ethanol HHV of 84,000 Btu/gal. The yield of other alcohols is based on an HHV efficiency of 7.6%. Nearly this entire stream consists of propanol, which has an HHV of about 97,300 Btu/gal.

- Annual water consumption is estimated to be 1.94 gallons per gallon of ethanol produced. This water use includes cooling tower make-up, process water, and other system losses.

- Capital cost data is based on values from Phillips et al. (2007) for a 2,205 dry ton/day plant and a scaling factor of 0.8, which is typical for biofuel conversion facilities.

- Annual Fixed O&M costs include a labor costs and other fixed O&M. Labor costs are calculated following Phillips et al. 2007, with a base cost of $4,720,000/yr for a 772,000 ton/yr facility. The labor costs are scaled by facility size using a scaling factor of 0.25, following the Fischer Tropsch analysis (see above for details). The other fixed O&M costs (maintenance, insurance, and taxes) are calculated as a percentage (4%) of the total capital investment.

- Annual Variable O&M costs include consumables (such as catalysts, gasifier bed material, and other raw materials) and waste disposal. These are calculated on a feedstock input basis. No electricity purchase is required, as a portion of the syngas is diverted for heat and power production.

- The co-product credit for mixed alcohols is calculated following the methodology of Phillips et al. (2007). The value of the alcohols is estimated to be $1.87/gal, based on the current wholesale gasoline prices from EIA of $2.39/gal (average from January 2008 through February 2009), a 78% discount for lower energy density, and a 10 ¢/gal reduction for potential additional processing requirements before use.[54]

- The levelized non-feedstock production cost for the ethanol is calculated based on the annual operating expenses and capital cost payment, co-product credit, and quantity of fuel produced.

Model Results

Some of the key cost and performance results projected for LCMA Synthesis/TC are shown in Table H-12. This table shows results for the range of applicable facility sizes based on selected input quantities, illustrating the high and low end costs. The low end feedstock input shows the minimum applicable facility size, and the high end was selected to show a reasonable range of sizes and costs. In general, the analysis shows that the non-feedstock production cost of ethanol from thermochemical conversion of biomass to mixed alcohols is relatively low, although it is slightly higher than the non-feedstock production cost of ethanol from the LCEt Hydrolysis and Fermentation process.

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[54] The 10 ¢/gal discount is applied to account for the fact that no special effort is made to ready the mixed alcohols for blending or sale.

Table H-12. Example Cost and Performance for Ethanol from LCMA Synthesis/TC.

Sample Model Results - LCE Fermentation/TC w/ Poplar		
Feedstock Input (dry ton/yr)	185,000	1,500,000
Ethanol Yield (MGY)	15.4	125
Other Alcohols Yield (MGY)	2.6	21
Water Consumption (1000 gal/yr)	29,920	242,570
Capital Cost (Million $)	$ 77.0	$ 410.6
Annual O&M Costs (Million $/yr)	$ 11.4	$ 24.1
By-Product Credit (Million $/yr)	$ 4.9	$ 39.4
Non-feedstock Production Cost ($/gal)	***$ 1.04***	***$ 0.28***

Phillips et al. (2007) estimate the minimum selling price of thermochemical ethanol is $1.27/gal for a 772,000 dry ton per year facility.[55] The ethanol production cost from the model for a similar sized facility using a delivered feedstock cost of $41.24/dry ton for poplar is $1.48/gal. Since the Phillips et al. (2007) study is the basis for this analysis, it is likely that different economic factors and analysis methods contribute to this variation.

The LCMA Synthesis/TC process is very similar to LCMD – FT, such that one would expect similar capital cost projections and biofuel yields for both processes. However, a comparison of the example results from the technology characterization models (Table H-21 and Table H-12) portrays LCMA Synthesis/TC as having much lower capital costs and higher yields than the FT process. It is possible that this is due to more optimistic technology development projections for LCMA Synthesis/TC. There do not seem to be any published studies showing experimental support for the targets given in Phillips et al. (2007).

2.7.3 <u>Outlook</u>

Although the process to generate higher alcohols from syngas has been known since the early 1900s, the technology has not yet reached commercialization (Spath and Dayton 2003). As of April 2005, there were no commercial plants that solely produce mixed alcohols in the C_2 to C_6 range (Nexant Inc. 2006). However, Range Fuels, Inc. is planning a demonstration facility to generate ethanol and other alcohols via thermochemical conversion (see details below). The main technical hurdles for higher alcohol synthesis include poor selectivity to higher alcohols and low yields. Typical conversion rate for single pass processes is about 10% production of alcohols, which is mostly methanol (Spath and Dayton 2003). The methanol can be recycled back through the process to generate higher alcohols.

Research and development efforts for HAS have been performed by several companies since the early 1980s. Some of the most advanced processes were developed by DOW, IFP and Snamprogetti (Spath and Dayton 2003). However, none of these companies are currently active in this area of research (Nexant Inc. 2006). Recent efforts for commercialization of the process have been spurred by new catalyst developments, new project developers, and

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[55] Production cost converted from 2005$ to 2006$ using CEPCI.

the interest in alternative fuels (Nexant Inc., 2006). Some of the current commercialization efforts are described below based on information from Nexant Inc. (2006):

- Pearson Technologies has a 30 ton/day biomass gasification with syngas conversion to alcohols in Aberdeen, Mississippi. Pearson is also trying to develop a demonstration plant in Hawaii.
- As mentioned above, Range Fuels Inc. of Colorado is developing a 1,200 ton/day (wood chip input) demonstration facility to generate ethanol and methanol. This project is supported by DOE and is scheduled for completion in 2011 (see U.S. DOE 2007).
- Power Energy Fuels is continuing to work on its proprietary Ecalene^{TM} process. Although this process is currently only developed at the bench-scale, there are two or three pilot plants under consideration that would produce mixed alcohols from biomass sources (wood chips, RDF, and tires). One of these pilot facilities is a 2,000 gallon per day plant that would be located at Wabash River Coal Gasification facility (with ConocoPhillips).
- Standard Alcohol Company of America is continuing to work on their Envirolene^{TM} process. This is only a bench scale process currently, but the company is interested in developing a pilot.

Despite this recent surge of development, there are still a number of technical and economic hurdles that must be overcome for the commercialization of the LCMA Synthesis/TC process. According to Nexant Inc. (2006, p. 3-2), the primary technical barriers include: "the overall process feasibility to produce the desired product slate, the ability to scale-up the process to a commercial level, the appropriate process conditions both in the reactor and upstream units, performance of various catalysts at commercial conditions, catalyst sensitivities, and appropriate syngas compositions." Additionally, large-scale mixed alcohol synthesis will require detailed consideration of the market interest and production costs relative to other fuels.

2.8 LIGNOCELLULOSIC-BASED BUTANOL BIOFUELS

Several options for converting lignocellulosic biomass to biofuels have already been discussed, including fermentative (biochemical) conversion and thermochemical conversion to produce ethanol. A lesser-known liquid fuel that can be produced from these same feedstocks is n-butanol, or more rarely its chemical relative, iso-butanol.[56] As will be discussed in more detail, n-butanol and iso-butanol both have key advantages, in their chemical properties, over ethanol as a fuel additive or potentially as a stand-alone fuel. This makes butanol

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[56] It is important to note that there are four isomers of butanol ($C_4H_{10}O$): *tert*-butanol, *sec*-butanol, n-butanol, and iso-butanol. While the nonspecific term "butanol" can refer to any of these, in the context of biofuel production by fermentation, n-butanol is the implied isomer. ABE fermentations, to be discussed later in the report as one of the most likely ways to make biobutanol, produce n-butanol. While metabolic pathways may be developed to produce other isomers, such as the higher blending octane iso-butanol, this report will focus on n-butanol. In this report, "butanol" will be used to refer to n-butanol, unless otherwise indicated.

conversion, though still under development and lacking distribution infrastructure, a tempting technology for biofuels producers to pursue.

2.8.1 Technology Description

One route to biofuels that is similar to lignocellulosic-based ethanol production via hydrolysis and fermentation, is to convert lignocellulosic biomass into a longer-chain alcohol fuel, n-butanol, to produce a high-octane gasoline additive. N-butanol is a four-carbon alcohol in widespread use as an industrial solvent, with a U.S. market size of some 370 million gallons per year at a bulk sales price of about $3.75 per gallon (Hence the annual market sales represent approximately $1.4 billion.). Following the introduction of oxygenates in gasoline, butanol was recognized as a viable oxygen-containing gasoline booster, provided that it could be cost-competitively produced.

Originally produced by fermentation, starting nearly 90 years ago (using *Clostridium acetobutylicum*), butanol gradually became a petro-chemically-derived product in the 1950s as the price of petro-chemicals dropped below that of starch and sugar substrates like corn and molasses. An additional contributor to this shift was the fact that the U.S. lost access to low-priced sugar cane from Cuba in 1954. Virtually all of the butanol in use today is produced petro-chemically (Green Car Congress 2005)

Biobutanol, purified to the same standards as petro-butanol, can be used in blends with gasoline. BP and DuPont have stated that blends containing up to 16% biobutanol are safe for use in conventional passenger cars (BP/DuPont Fact Sheet 2008) Biobutanol advocates have claimed that up to 100% butanol can be used with no engine modifications, but debate continues (ButylFuel 2006; Green Car Congress 2005).

From a purely technological perspective (policy and economic factors aside), conversion of biomass to biobutanol can happen in two ways:

1. A thermochemical conversion that relies on heat and pressure, often combined with biochemical processes
2. A biochemical route that relies on microorganisms

Thermochemical Route

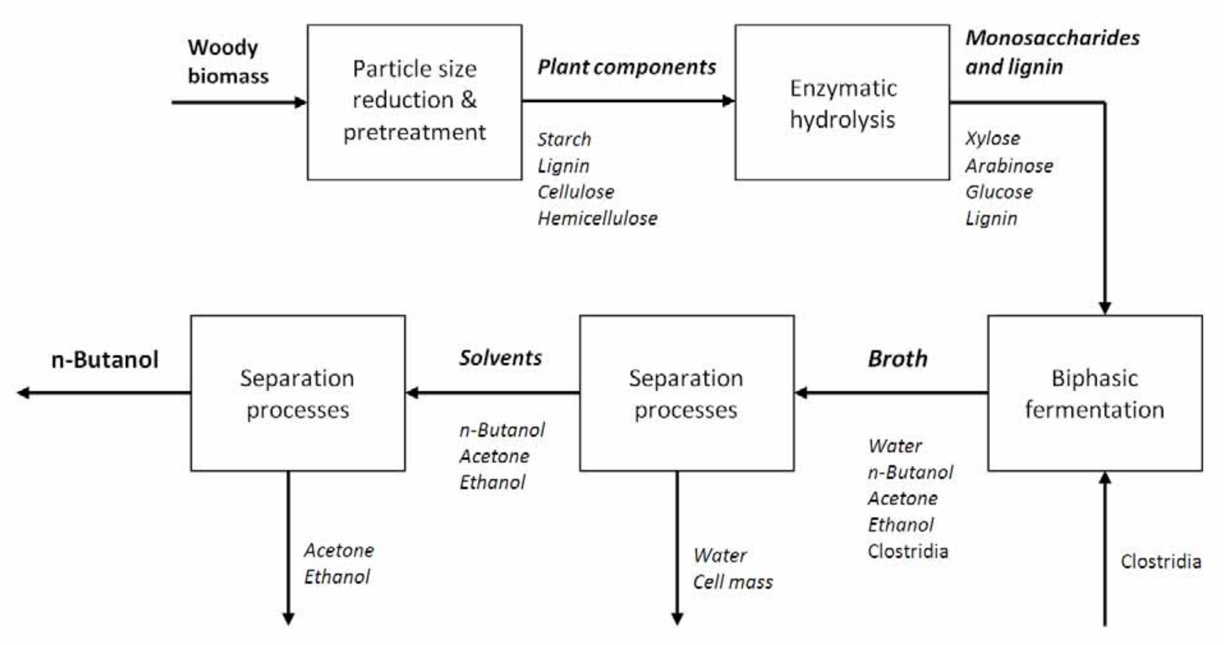
Since 1994, Wilhelm Keim and his research group at the University of Aachen have performed R&D under a DOE-sponsored, Air Products-led program using syngas (a mixture of carbon monoxide and hydrogen created thermochemically, as described in the previous section) as a feedstock for butanol production (Verkerk, 1999). The syngas can theoretically be made into butanol in a number of ways. It can be converted to methane and subsequently fermented by methane-consuming organisms, or fermented directly in the presence of a biocatalyst. It can also be reacted in the presence of a chemical catalyst in order to form a range of different alcohols. Because there are not any truly selective catalysts for these reactions, a diverse mixture of chemical products is formed that incurs significant separation costs. The catalysts that exist appear to have high production costs and are not currently produced at a commercial scale.

Due to the relatively far-term commercialization potential of this thermochemical route, this report will focus on the nearer-term technology of the biochemical pathway to produce butanol (described below).

Biochemical Route

Biochemical production of n-butanol from lignocellulose begins with particle size reduction and pretreatment of the biomass feedstock in order to expose long-chain cellulose molecules to enzymes designed to break them into simpler component sugars as in the LCEt-HF route. *Clostridia*[57] (traditionally) or other types of bacteria like *E. coli* (in some newly developed technologies) then anaerobically process the sugars into acids and the acids into a mixture of acetone, ethanol, and the desired product, n-butanol. This mixture is known as "ABE" and such biological processing is thus called "ABE fermentation." Subsequent separation units remove pure n-butanol from the dilute fermentation broth.

Figure H-9. Basic Steps in the Production of Biobutanol via Batch Fermentation.



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[57] Many different strains of the bacterium *Clostridium* (most commonly *C. beijerincki* and *C. acetobutylicum)* process xylose to form a mixture of solvent chemicals. Clostridium is not, however, the only type of bacteria that can be used. As will be discussed later, *E. coli* and other bacteria have been engineered to produce ABE mixtures.

Considering the process in more detail, feed preparation begins with particle size reduction in order to increase the surface area of the biomass and thus its vulnerability to chemical attack. This is a common step in either ethanol or butanol production. Either hot water or ammonia is then used to expand the cellulose fibers and make the biomass more accessible to cellulase and amylase enzymes. Enzymatic hydrolysis reduces specific long-chain molecules—hemicellulose and cellulose, or amylose and amylopectin—into monomer sugar components. These simple sugars can include both six-carbon (e.g. glucose, mannose) and five-carbon (e.g. xylose, arabinose) molecules.

After feed preparation, there are a number of possible fermentations that can be used to transform the simple sugars into n-butanol. The most traditional method, depicted in Figure H-9, is ABE fermentation, in which the sugars are fermented in a two-step fermentation process with *Clostridia*, usually *C. acetobutylicum* or *C. beijerincki*. *Clostridia* process five- and six-carbon sugars at nearly equal rates (Ezeji 2008), making them ideal for wood and other heavily lignocellulosic materials.

In the first step of fermentation, acidogenesis, the bacteria convert sugars to acids that gradually lower the pH of the solution. At some critical pH, the bacteria begin a solventogenic step in which the acids are converted to a mixture of ketones and alcohols of which acetone, butanol and ethanol are the most important. In traditional ABE processes with *C. acetobutylicum*, these products are produced in a ratio of about 3 : 6 : 1 (acetone : butanol : ethanol) by mass. More recently studied strains of *Clostridia*, such as *C. beijerinckii P260*, have demonstrated better selectivity for butanol production, with product ratios of about 3 : 16 : 1 (acetone : butanol : ethanol) by mass (Qureshi 2008). Hydrogen and carbon dioxide are co-produced in this fermentative conversion process.

As the butanol concentration increases during fermentation, it begins to inhibit cell membrane activity, including the uptake of sugars that keep the bacteria alive and the uptake of butyric acid, butanol's precursor. Fermentation ceases at a limiting concentration of butanol where the bacteria can no longer process such precursor acids. (Lee, 2008) Bioengineering and gene splicing have improved this butanol tolerance from 20 g butanol/L broth for traditional *C. acetobutylicum* to nearly 40 g/L for modified strains of *E. coli* (Van Noorden 2008). Selectivity has also been improved so that butanol far exceeds acetone and ethanol in the product mixture (Lammers 2008; Qureshi 2008).

Another possible fermentation that has sparked renewed interest in biobutanol has been developed by researchers at Ohio State University and promoted by David Ramey of ButylFuel, LLC. In this pathway, fermentation with *C. tyrobutyricum* replaces the first phase of the biphasic fermentation described above. *C. tyrobutyricum* processes sugars into butyric acid (the precursor acid for n-butanol) and hydrogen with almost complete selectivity, meaning that there are only very small amounts of acetone- and ethanol-precursor acids in the product mixture. This butyric acid solution is then pumped into a fibrous bed reactor where *C. acetobutylicum* converts butyric acid into butanol, exactly as in the second stage of traditional ABE fermentation. The main difference is that without a feed containing significant levels of the precursor acids for ethanol and acetone, *C. acetobutylicum* can produce almost exclusively n-butanol (Yang 2006).

This latter pathway is still subject to the same bacterial butanol poisoning constraints as the ABE process, but offers better selectivity. As the process has only recently been patented, and appears to have been tested only in individual phases and only at a bench scale, costs, productivity, and yield potential remain unclear. (Yang 2006). The technology's developers are currently also considering a "hybrid" pathway in which *C. tyrobutyricum* is used to convert corn hydrolyzate into butyric acid as above, but with the second fermentation step replaced by a thermochemical hydrogenation step. This utilizes the hydrogen co-product from the initial fermentation step (Ohio State University 2008). Again, this technology appears to be in its infancy, with little data available.

A final fermentation pathway was recently developed by researcher James Liao at UCLA. This method, licensed by Gevo, uses a modified strain of *E. coli* to process glucose into not n-butanol, but rather iso-butanol. The highly engineered bacterium can process six-carbon sugars, but not five-carbon sugars. Its butanol tolerance is relatively low, at 2% butanol by volume, but the isomer it produces has several advantages[58] over the usual fermentation product, n-butanol. These advantages may allow this conversion pathway to compete with those of higher yield and more versatile feedstocks (Van Noorden 2008).

After any type of fermentation, a series of separations processes are used to remove the desired product, pure butanol, from the mixture of dead bacteria, water, and (in some cases) acetone, and ethanol. Traditionally, this has entailed an energy-intensive series of distillations, but many innovative separations processes have been proposed and/or implemented at a pilot scale. Pervaporation[59] can be used to remove n-butanol *in situ* from the fermenting broth and limit butanol poisoning of the *Clostridia*. The pervaporated mixture of water and solvents, primarily butanol, must still be distilled to remove butanol, but the radically decreased water content decreases the energy requirements of the overall process. As energy costs are often a key deciding factor in the techno-economic viability of a conversion process, this type of *in situ* butanol removal could greatly contribute to the overall viability of n-butanol production for fuels. Liquid-liquid extraction and argon-gas stripping are also more efficient contenders for butanol removal. Reverse osmosis has been proposed as an economical solution, but at a bench scale it is crippled by membrane fouling (Lee 2008).

The question of overcoming the limits imposed by butanol poisoning of *Clostridia* has led to modifications to the traditional batch fermentation. A fed batch process would rapidly poison the bacteria with butanol. Because the fermentation occurs in two steps, a purely continuous process would produce only acids rather than an ABE product. (A single-stage continuous process would require that both steps occur simultaneously, rather than in series, in order

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[58] Iso-butanol offers several advantages over its isomer n-butanol, with which the rest of this report is concerned. It is even less miscible with water than n-butanol, making it attractive from a fuel distribution standpoint. Most attractive, however, is the higher octane number of iso-butanol (Motor Octane Number (MON) of 89 for iso-butanol versus 78 for n-butanol, 102 for ethanol and 81-89 for gasoline). This allows refiners that produce gasoline for blending with a biofuel to reduce the severity of their reforming processes, leading to higher gasoline yields or product revenue per barrel of crude oil. Higher MON also reduces engine knocking. While neither isomer of butanol can compete with the MON of ethanol, iso-butanol has a blending advantage over n-butanol. Both butanol isomers are at least comparable to gasoline in MON.

[59] Pervaporation is the separation of two liquids using a membrane through which one liquid preferentially evaporates. The vapor is thus concentrated in a particular component, and can be condensed and used as is, or can be further concentrated with further pervaporation stages.

to reach significant amounts of ABE product.) The most efficient fermentation appears to be a multistage continuous process, or a fed batch process with *in situ* butanol recovery (Lee, 2008). The former would dedicate separate reactors for each fermentation step, so one colony of bacteria would operate under the correct conditions for acidogenesis and the other under the correct conditions for solventogenesis. The latter, *in situ* removal in fed batches, would mean that sugars could be replenished during fermentation to optimize the fermentation rate, without butanol accumulating beyond the bacterial tolerance.

Apart from fermentation process structure, other optimizations have been made by tweaking the chemistry of the fermentation broth. Buffers can be added to increase the amount of acid produced before the critical pH is reached, thereby increasing the amount of butanol precursor available. Stimulants like furfural and hydroxymethyl furfural increase the activity of *Clostridia* and the rate of fermentation (Lee 2008).

The extensive bioengineering of *Clostridia* has worked toward fermentations that selectively produce butanol rather than ethanol and acetone (These products will still occur in small amounts, however, and could be marketable as solvents). Another research target has been to increase the hydrogen yield of the fermentation process, while boosting the butanol productivity. In a natural fermentative process, some of the hydrogen produced by *Clostridia* would be used (inter-species transfer) by methane-producing bacteria (methanogens) in the inoculum. Reducing or eliminating the methanogens is thus one approach to increasing the ultimate yield of hydrogen. Researchers have found that heat treatment is one of the effective techniques for accomplishing that. A *Clostridium* bacterium will form a bacterial spore in the presence of heat, and thereby survive. The methanogens are non-spore-forming; the heat kills them. The application of heat thus effectively selects for the *Clostridia* population and so for production of hydrogen, while eliminating the competing process of methanogenesis. An additional benefit of using heat rather than some other selection technology is the possibility of recycling low level heat from the fermentation and distillation processes.

2.8.2 Performance and Cost Analysis

Researchers at a bench scale have used multistage continuous or immobilized cell processes to attain 112-139 gallons butanol per ton of glucose feed (Lee 2008). Values for a lignocellulosic feed are apparently unpublished, but the similar reactivity of *Clostridia* with hexose and pentose would indicate similar yields of butanol per gram of hydrolyzed lignocelluloses. Yields calculated on the basis of research by Qureshi et al. (2008) for *C. beijerinckii* range between 50 and 55 gallons per dry ton lignocellulosic feedstock. For comparison, mid-term lignocellulosic ethanol technology could be expected to yield 70 to 80 gallons of ethanol from the same amount of feedstock (Ethanol has a lower energy density, however, so this comparison is not the entire story. This will be addressed below).

Few biomass-based n-butanol plants are active today, even fewer using lignocellulosic pathways (Those plants will be discussed below, as well as companies and research institutions with notable butanol development efforts). Little plant cost and productivity data has been reported, so any extensive biobutanol production analysis must be based on detailed engineering process simulations.

One currently operating example of a pilot plant, albeit not one producing the commonest n-butanol isomer, is Gevo's 20,000 gpy lignocellulosic iso-butanol pilot facility, a recommissioned ethanol plant in Colorado. While the technology driving this facility was licensed from researcher James Liao and probably centers around *E. coli*, the overall production process is proprietary and its details remain unclear. Gevo's scientific advisory board claims that such a retrofit incurs capital costs of $0.25 to $0.30 per gpy installed capacity. Gevo announced that the next step is to partner with ethanol plant designer ICM Inc. to build a larger, one million gpy facility at ICM's biofuels research center in St. Joseph, Mo (Lammers 2008). This facility is scheduled to come on line during Summer 2009 (Gevo 2009). The company is currently developing a customer base for commercial-scale iso-butanol production (Gevo 2009).

ButylFuel, LLC (ButylFuel LLC 2009) has provided some preliminary production cost data, which it hopes to confirm by completing tests on its Bu-100 (100 gallons n-butanol per week or 5000 GPY) demonstration plant and its Bu-1,000 pilot plant. The company's cost estimates claim that it can produce biobutanol from corn for about $1.20 per gallon, not including a credit for the hydrogen produced. The company compares this with ethanol production costs of about $1.28 per gallon (Both of these calculations use a feedstock price of corn at $2.50 per bushel). As a further point of reference, ButylFuel LLC suggests that butanol produced from petroleum costs about $1.35 per gallon to manufacture. The economics of the ButylFuel, LLC process are claimed to be even more attractive when cellulosic waste material is used as feedstock instead of corn: the price to produce a gallon would drop to $0.85. The cost assumed for this "waste" feedstock is unknown. In the case of such a waste use, the cost to grow and prepare corn or another dedicated feedstock for fermentation, by far among the major cost items, would be eliminated.

A simple calculation puts this feedstock cost in perspective. Based on University of Missouri Agricultural Publication G4020 (Murphy 2001), one bushel of shelled corn weighs 56 lb. Using the aforementioned total production cost data, one can determine the feedstock component of the production cost of biobutanol. Interpreting Lee's (2008) data of 120 gallons butanol per ton of glucose feed as 74.3 gallons per ton of corn and using ButylFuel LLC's cost data (both production cost of $1.28 per gallon and feedstock cost of $2.50 per bushel corn cost) leads to a feedstock element of $0.82/gallon of biobutanol and $0.38/gallon for capital and operating expenses. This confirms the high cost of feedstock in the production of biobutanol from corn.

Gapes (2000) presents detailed estimates of capital and annual operating cost for two n-butanol facilities the author designed in the 1980s. These facilities used traditional ABE fermentations, one a batch process and the other a continuous process. The batch facility would produce 4,500 metric tons (1.14 million gpy) of ABE annually. Capital investment (and all related figures updated to $2008) would be 1.6 million EUR ($3.31 million), or about $2.91 per gpy production capacity. Production costs would be 0.8 million EUR ($1.66 million), or about $1.42 per gallon.

In the case of a 7.58 million gpy continuous process, capital costs would be $9.34 million, or $1.37 per gpy. Production costs would be $3.62 million per year, or about $1.24 per gallon (Gapes 2009). This would suggest that continuous processes are both less expensive to operate and cheaper to build. However, because the two plants have different production scales, their per-gallon costs can only be used as an order-of-magnitude estimate, and cannot be directly compared.

In the general case of lignocellulosic feedstocks, capital and materials handling costs can be expected to increase, while the feedstock cost can be expected to decrease. The authors' calculations presented in the "Lignocellulosic Ethanol" section of this report indicate that capital costs for lignocellulosic ethanol plants can be as much as six times as large as for corn-fed plants. This radical increase is due to equipment for additional processing steps, longer pretreatment and fermentation residence times, and lower yields per ton of feedstock. Operating costs appear to almost double between corn and lignocellulosic feedstocks. If a similar ratio holds for n-butanol plants, this may indicate that decreased feedstock costs for lignocellulosic materials may not compensate for increased capital and operating expenses.

Outline for Model Analysis

The important values and assumptions used to construct the biobutanol production model are described below. A detailed presentation of the model is shown in Appendix H-F.

- Key model variables are:
    - feedstock type (switchgrass, hybrid poplar, pine, mixed paper, or wheat straw)
    - yearly plant feedstock consumption (on a dry basis)
    - technology time frame (near, mid and long)
- The different feedstocks have different yields of butanol per ton, so capital cost was scaled differently for each feedstock to account for the fact that two plants of the same biobutanol output capacity that use two different feedstocks will have to handle different amounts of the raw feedstocks. Thus, the size of the front end equipment (like feed handling and pretreatment tanks) will have a different relationship to the back end equipment (like secondary distillation columns and fuel storage tanks) for different feedstocks with different yields. The differential scaling to compensate for this was done based on plant sections (e.g. "feedstock handling and storage" or "fermentation") rather than an itemized major equipment roster.
- Feedstocks were assumed to contain 30% moisture as received. This is intended to be a conservative estimate, so that feed handling equipment will be sized to have slightly excess rather than inadequate capacity. Most feedstocks should have less than 30% moisture. Corn stover generally has 15% moisture (Aden 2002). Wood chips are highest in moisture as received with as much as 50% water (Wooley 2000), but air dried wood is generally 15-20% moist (Scurlock 2003).

- Capital cost data was drawn from three detailed economic assessments of butanol and ethanol production. Front-end major equipment costs were individually scaled from a study by Aden et al. (2000) of a lignocellulosic ethanol plant. Because feed handling and pretreatment are comparable for lignocellulosic ethanol and lignocellulosic butanol (the same sugars are made accessible for fermentation in the same amounts, despite the different organisms used), the front end of each type of plant should be very similar, despite different subsequent process steps. Equipment for saccharification and all subsequent process steps was modeled after a corn butanol cost study done by Wu et al. (2007). Fermentation volume requirements were determined to be approximately the same for both types of feedstock (lignocellulosics and corn), since residence time and butanol tolerance were similar for the two cases. When no more detailed capital scaling factor was available, 0.6 was used as per McAloon et al. (1999). All costs were inflated to 2008$ using the CEPCI index as described in Appendix H-C. A Lang factor[60] of 3.15 was used, based on the implied Lang factor used by Aden et al. (2002), who used installation factors and contingency factors to generate a total installed cost. This was in good agreement with the effective Lang factor generated from the major equipment roster by the individualized Lang factor cost estimation method of Peters and Timmerhaus (1980), which was 3.08. The larger result was used in order to give a more conservative estimate.

- Yields were calculated based on the relative amounts of hemicellulose and cellulose in each feedstock, as fermented with *C. beijerinckii P260*. Qureshi et al. (2008) performed laboratory studies of batch-fedsimultaneous saccharification and fermentation (SSF) of wheat straw using this organism. Their mass balance is the source of the yield used in the model, 0.42 g ABE/g hydrolyzate. Hydrolysis efficiencies were based on earlier work by Antares (2007): 90% (cellulose) and 95% (glucose) hydrolysis efficiencies. Though a different fermentation organism is used, these values are very similar to the yield and efficiencies reported in the corn butanol case with *C. beijerinckii BA101*: 0.465 g/g hydrolysate and 93.2% hydrolysis efficiency. The relative distribution of the ABE products (the mass ratio between acetone, ethanol, and butanol) is assumed to be the same as in the Qureshi (2008) study and the same for all feedstocks.

- Conversion efficiency is based on the HHV data presented by Qureshi et al. (2000) on a dry mass basis. Only the desired liquid fuel products--butanol and ethanol--are counted toward the output energy value. Acetone is considered a byproduct for the purposes of this study.
    - N-butanol HHV: 37.5 MJ/kg
    - Ethanol HHV: 29.2 MJ/kg

- Operating costs were scaled from a detailed techno-economic assessment of a corn butanol plant by Wu et al. (2007). As with capital costs, O&M costs were scaled individually either by feedstock or product volume as was most applicable. Unfortunately, the operating cost estimates given by Wu (2007) and

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[60] A Lang factor is a multiplier used to estimate a total installed plant cost based on the plant's total major equipment cost. This is used for preliminary cost estimation, when the details of piping, electrical work, and siting are not yet known accurately.

McAloon (1999) did not itemize utilities or materials requirements between pretreatment and subsequent process areas to the extent necessary for operating costs to be scaled separately as per the lignocellulosic ethanol case. Thus, pretreatment operating costs, with the exception of enzyme costs, are assumed to have the same costs as corn pretreatment. Because enzyme costs are the most significant of the pretreatment costs (McAloon, 1999), this use of well-characterized lignocellulosic enzyme costs may minimize the inaccuracy of the overall operating cost value. This analysis assumes that the same materials (lignocellulosic biomass) are being hydrolyzed into the same desired products (six-carbon sugars) in both ethanol and butanol processes, and thus enzyme costs for lignocellulosic butanol should be similar to those of lignocellulosic ethanol. The differences in process parameters only occur later, when identical materials are fermented by different organisms.

- The model assumes an enzyme cost of $12.08/ton cellulose and $4.37/ton starch, based on average enzyme costs of $0.18 and $0.0206 respectively per gallon lignocellulosic ethanol. (Aden et al. 2002).

Model Results

Sample results from the biobutanol production and cost analysis model are presented in Table H-13 below. Although the theoretical nature of the engineering studies used to derive this model mean that its results should be considered only preliminary, it indicates costs of production that are significantly higher per gallon than those of ethanol produced via LCEt – HF. Whether real-world data will verify this higher per gallon (and per-BTU) production cost will become apparent as operating data from demonstration plants becomes available. Clearly, improved butanol yields could help close this economic gap.

Table H-13. Biobutanol Cost and Performance Model Example Results.

Sample Model Results		
Feedstock Type	Wheat Straw	Wheat Straw
Feedstock Input (ton/yr)	442,478	1,327,434
Yield Biobutanol (MGY)	25	75
Conversion Efficiency (HHV)	35.4%	35.4%
Consumables and By-Products		
Acetone Yield (99.5%; tons/yr)	37,911	113,733
Ethanol Yield (99%; tons/yr)	2,044	6,133
Biobutanol Production Costs		
Fixed O&M ($/yr)	$ 48,402,835	$ 101,890,985
Variable O&M ($/yr; does not include feedstock cost)	$ 9,132,354	$ 27,397,062
Byproduct Credit ($/yr)	$ 35,361,775	$ 106,085,326
Annual Operating Cost ($/yr)	***$ 22,173,413***	***$ 23,202,720***
Total Capital Investment ($)	***$ 197,424,154***	***$ 384,642,391***
Fixed Charge Rate	12%	12%
Non-Feedstock Production Cost ($/gal)	$ 1.86	$ 0.94

2.8.3 Outlook

At the global scale, 3.5 million metric tons of n-butanol are produced annually via petrochemical pathways and sold as a solvent and chemical reagent at three times the price of ethanol (Van Noorden, 2008). Though installation of more production capacity could be expected to reduce this price, n-butanol today is a much higher-value product than ethanol. This lure of a high-priced solvent market could bring biobutanol producers online and increase production volumes until fuel butanol becomes cost-competitive with fuel ethanol. If equal production costs are achieved, the advantages of n-butanol over ethanol (explained below) would make biobutanol the preferred fuel. Biobutanol produced from traditional ABE processes had a much higher production cost than petro-butanol, leading to the decline of biobutanol in the 1940s and 1950s. More productive, more butanol-tolerant microorganisms, more energy-efficient separations processes, and different feedstocks are needed to reduce this margin to the point of feasibility.

The strengths of biobutanol include: its technical and materials overlap with bioethanol; little necessary retrofitting of existing infrastructure; and its behavior with water. Its disadvantages center on comparatively low yields, its toxicity to humans and other mammals, odor and (possibly) blending octane number. N-butanol's specific advantages include:

- Biobutanol could take advantage of many aspects of the ethanol boom of the last decade. Butanol can be produced from the same feedstocks—both cellulosic and non-cellulosic—with very similar equipment. Ethanol plants can be recommissioned or converted to butanol plants with little modification and at low cost. Pilot butanol plants can be constructed as additions to existing ethanol plants in order to share equipment and reduce capital cost.

- N-butanol is non-hydroscopic and non-corrosive, and does not dissolve and transport contaminants common to petroleum piping and tanks.[61] This would make pipeline transport of gasoline/butanol blends, unlike gasoline-ethanol blends, possible. Gasoline-ethanol blends are currently prepared by transporting the components individually and blending them in depots and distribution centers.

- For the refiner, butanol has several possible advantages. Due to its relatively low vapor pressure (the Reid Vapor Pressures for n-butanol and ethanol are 0.33 and 2.0 respectively), n-butanol could be used at higher concentrations than ethanol in gasoline blends without altering cars' existing VOC capture systems (Cascone 2008). Alternatively, for the same commercial Reid Vapor pressure specification, a higher amount of butanol than ethanol can be blended in commercial gasolines.

- N-butanol has a higher energy-to-mass ratio than ethanol (88%, rather than 70% the energy density of gasoline), leading to a higher fuel efficiency blend (Cascone 2008).

One important consideration when discussing the energy density of n-butanol is the overall energy conversion efficiency of the biobutanol process. While n-butanol does have a higher energy density than ethanol, it also generally has a lower yield per ton of feedstock. Based on data presented in the biochemical lignocellulosic ethanol chapter of this report, one ton of switchgrass can be estimated to yield 5.3 MMBTU (LHV basis) in ethanol fuel if processed with dilute acid pretreatment. Using n-butanol yield data from Qureshi et al. (2008) for a lignocellulosic feedstock, the same amount of switchgrass could be expected to yield 4.8 MMBTU (LHV basis) of butanol and its ethanol co-product. Despite n-butanol's high energy density, the overall energy yield of the n-butanol fermentation is less in this case. A similar calculation for several other lignocellulosic feedstocks leads to varying levels of disparity between lignocellulosic ethanol and n-butanol. In the case of wheat straw, energy yields are nearly equal. However, such simple calculations do not consider the entire life cycle of the fuel. Lower energy inputs for product separation could provide n-butanol with a better overall energy balance than ethanol.

Another important consideration is engine compatibility of butanol. Engines appear to tolerate butanol well: BP has claimed that blends containing up to 16% butanol are safe for passenger cars (Fact Sheet 2008).[62] Butanol producers

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[61] The isomer iso-butanol is even less miscible with water than n-butanol, increasing its long-term storage potential and its pipeline transportability.

[62] Due to a scarcity of biobutanol, no demonstration cars appear to be on the road. The question of how much biobutanol maximizes performance, or what maximum level of biobutanol is non-damaging to engines has been rarely studied in literature. All data appears to have come out of BP. Early BP press releases endorse use of biobutanol at 10 vol% concentration (GCC 2008). At a German energy conference in 2007, BP's Ian Dobson presented data supporting their endorsement of 16% blending (GCC 2008). The non-BP data that is available pertains to butanol blending with diesel rather than gasoline. Researchers used branched-chain iso-butanol rather than straight-chain n-butanol, and found

claim that n-butanol/gasoline mixtures are fully compatible with conventional gasoline engines, damaging neither the fuel systems nor the engines of modern cars (ButylFuel, LLC 2009). No detailed research has been published to support this claim. BP states that it has tested blends of butanol and gasoline in conventional car motors, but it is unclear what the extent of this testing has been. It has not released any data apart from its 16% blending claim (Fact Sheet 2008). This lack of a consensus between motor manufacturers and fuel refiners - as to if, and how much, butanol is safe for engines - may lead consumers and fuel developers to avoid the risk of an unendorsed, unstandardized fuel technology.

In considering the possible effects of butanol on conventional fuel systems, one must note that butyric acid is the precursor for the n-butanol in the biobutanol process. Complete removal of the butyric acid from the fuel component seems to be a minimum requirement to prevent pitting and corrosion of metal engine parts. Engines require many non-metal parts, which also must be checked for compatibility with butanol. Viton ® elastomers, a common fuel system material, have been shown to swell and soften less in pure n-butanol than in gasoline blends (DuPont 2007). Butanol fuel proponents suggest that this allows the Viton ® components a longer service time and improves durability when used with butanol rather than gasoline. In actual practice, however, the reduced swelling (i.e. shrinking of already swollen material) of the engine packing materials may introduce engine leaks if not resized to accommodate a lower degree of swelling.

Butanol proponents claim that from July 14 to August 15, 2005 a '92 Buick Park Avenue was successfully driven across the U.S. on pure butanol (ButylFuel, LLC 2009). Others put this in perspective by pointing out that the butanol used was commercial, petrochemical-grade butanol purchased from Ashland (Green Car Congress 2005). Proponents note a beneficial environmental performance during these tests in ten states: they claim emission reductions of hydrocarbons by 95%, carbon monoxide to 0.01% and nitrogen-oxides by 37% for the pure butanol fuel. High altitude performance is not discussed (ButylFuel, LLC 2009).

Industry Perspectives

Overall, butanol appears promising for its compatibility with existing vehicle technology. Another important question to be examined is butanol's compatibility with existing fuel distribution infrastructure. Logistically, the best advantage of n-butanol is its behavior with water. Because butanol is only slightly miscible with water, butanol/gasoline blends show no separation in the presence of water (BP, DuPont 2006). While good for fuel processing and transport, this also reduces but does not completely eliminate the risk of groundwater contamination in the event of a butanol spill or leak. Due to the low density of butanol, it is likely to form a slick on the surface of a body of water rather than an underlying layer or an emulsion, and quickly evaporate (Butanol MSDS). While there appear to be no studies examining butanol contamination in groundwater, MSDS data indicates that butanol

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that engine performance reached a peak between 30 and 40 vol% butanol (Al Hasan 2008). This could suggest the use of biobutanol as an additive to diesel fuel, as well as to gasoline.

biodegrades quickly, especially upon exposure to air. Groundwater contamination is more likely in cases of a release into soil than into water or air (Butanol MSDS).

The most significant drawback to n-butanol seems to be its acute toxicity to humans and other mammals.[63] Animal testing indicates inhalation toxicity 1.6 times greater than that of ethanol, ingestion toxicity ten times as great as ethanol, and skin toxicity four to five times that of ethanol (Butanol MSDS, Ethanol MSDS). Gasoline is reportedly less acutely toxic than n-butanol or ethanol by ingestion, but possibly more toxic than both upon inhalation or skin exposure. Because of gasoline additives like benzene, gasoline's most significant health concern is chronic exposure (Gasoline MSDS). Though n-butanol's chronic effects appear to be limited, its acute effects are cause for health and safety concerns for both end product users and production employees (Cascone 2008).

Finally, consumer acceptance may be limited due to the offensive cod liver-like smell of butyric acid, even when present in only trace amounts.

Research and Development

BP announced that it will be working with the University of California, Berkeley, on a $500 million, 10-year program, part of which will be devoted to research on improving biofuels such as butanol. And last year BP announced a partnership with DuPont to develop new technology for making butanol. DuPont will provide expertise in biotechnology.

BP and DuPont have taken advantage of the "synergy" of biobutanol and bioethanol production by constructing a $400 million bioethanol facility in the UK. The first phase of the DuPont-BP venture will consist of using existing technology to convert sugar beets into 30,000 tons, or 9 million gallons, of biobutanol annually at British Sugar's facility in Wissington, England, east of Cambridge (Associated Press 2006). The companies plan to eventually either add a biobutanol facility to the site or convert the plant to biobutanol production (BP, DuPont 2006).

A smaller company, Gevo, has made headlines for buying exclusive rights to the technology developed by James Liao (now on Gevo's advisory council) of UCLA, described earlier in this chapter. Gevo has garnered a remarkable amount of venture capital funding from sources including Virgin Green Fund, Khosla Venture Capital, and the Malaysian Life Sciences Development Fund. Gevo, founded in 2005 by researchers at CalTech, has recently partnered with ICM, Cargill (which licensed Gevo a fully or partially fermentative butanols production technology), and Total Oil and Gas. As noted above, it has both a working pilot butanol plant and a one million gpy iso-butanol facility scheduled to begin production in 2009.

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[63] In the case of iso-butanol, toxicity is less of a concern. Iso-butanol is not absorbed dermally, and has a lower ingestion toxicity than does n-butanol. As with n-butanol, iso-butanol is readily biodegradable (*iso-butanol*, 2009).

2.9 PULP AND PAPER BIOREFINING – HEMICELLULOSE EXTRACTION

2.9.1 Technology Description

Pulp mills may already be classified as biorefineries. However, enhanced pre-processing and treatment of byproducts may allow these facilities to expand their product slate. By adding either finished biofuels or important intermediaries, these established mills can increase the value of the incoming feedstock and leverage substantial processing and handling infrastructure.[64]

Paper can be made in different ways: from completely fresh woody material, from recycled paper material or a combination thereof. Since the second path is purely reprocessing of material, the following discussion will concentrate on the production of paper from fresh woody material. This material comes to the paper mill in the form of whole logs. Both bark and wood consist of three basic long-chain compounds: hemicellulose, lignin and cellulose. Of these three components, it is the cellulose that forms the largest portion of paper. The paper making process can be properly viewed as a pathway for stripping cellulose material from lower-value hemicelluloses and lignin, while using these byproduct materials in ways that extract their maximum value.

Traditionally, paper mills fall into one of two major types: Kraft mills (over 80% of the paper mills in the U.S.) or mechanical mills (EPA 1990). A Kraft mill applies "white liquor," a hot solution of concentrated sodium hydroxide and sodium sulfide, to wood that has been stripped of bark and chipped. This chemical digestion breaks down hemicellulose and lignin while leaving cellulose fibers largely intact. The process typically lasts one to three hours. Mechanical liquids/solids separation with screens or sweeps is then used to remove "black liquor," the depolymerized hemicellulose and lignin dissolved in a caustic solution of sodium sulfate and sodium carbonate. This is concentrated through evaporation and then burned to provide heat to the energy-intensive digesters. Regenerated sodium hydroxide and sodium sulfide are recycled to the digesters (Amidon 2008).

The digested wood chips, now "brown stock," are forced with high pressure into a tank of water, where the cellulose fibers expand and are rinsed of contaminant black liquor. This "brown stock' is the base material for cardboard-like products. For the production of pure white paper a bleaching step will follow. Bleaching can reduce stock mass by as much as 5% and reduce cellulose fiber strength by reducing the amount of lignin bound to cellulose (PTF, 1995).

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[64] Though New York State has paper mills, it does not have a significant pulping industry, with only one pulp mill currently in operation (International Paper's Kraft mill in Ticonderoga, NY) at the time of writing (NYSDEC 2009). The lack of a hardy pulping industry in New York State means that some of the biofuels technologies considered—hemicellulose extraction and black liquor gasification, which will be discussed later as a value adding strategy for pulp mills—are not immediately applicable to New York State. While hemicellulose extraction may be applied to other facilities that are common in the region, including wood chipping and fuel pellet operations, black liquor gasification will be used in New York State only if additional pulp industry develops. It is important to note, however, that integration of biofuel manufacturing into pulp and paper production may improve the economics of a facility such that it may be more attractive to build such a pulp mill in New York. The current lack of pulp industry may not indicate a future lack of industry.

Either white or brown stock is made into paper and cardboard respectively by extrusion onto wire mesh and subsequent drying and pressing.

Mechanical mills, which do not attempt to remove the lignin from the cellulose fraction, process the wood fibers by grinding wood chips with refiner plates, usually after a steam treatment step to expand the fibers and make them less resistant to tearing. This is extremely energy intensive, and produces more physical cuts in the cellulose fibers, which decrease the strength of the pulp when used in paper. Modern mills consume about 2600 kWh per ton of pulp product (Paprican 2008).

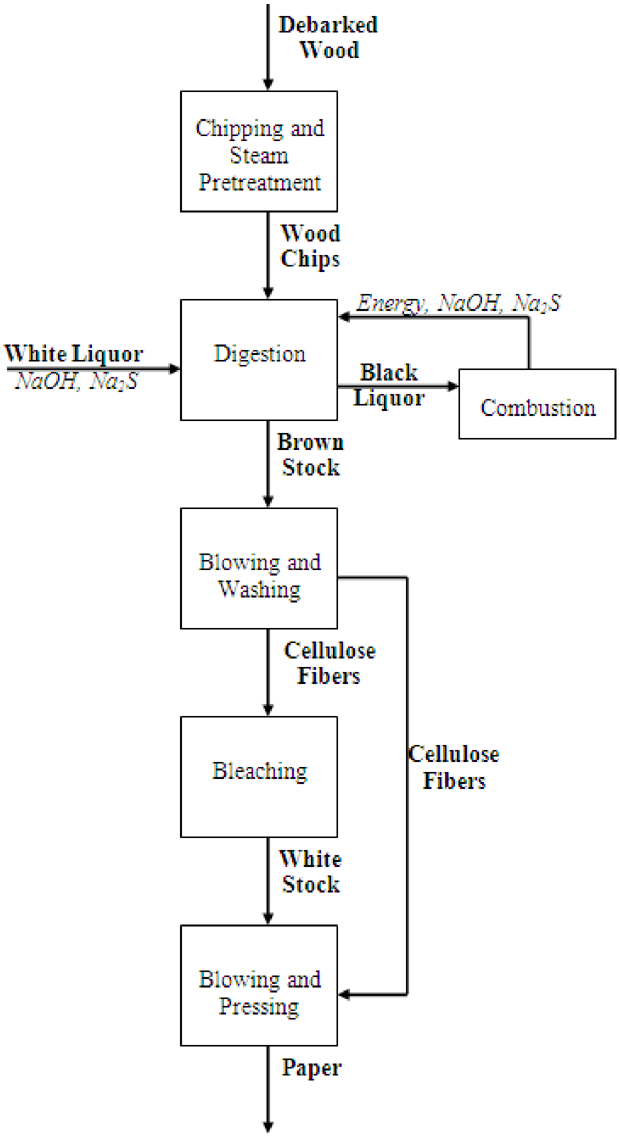

Figure H-10. Conventional Kraft Mill Process Design.

Although several types of biorefining have been proposed, each one exploits the same idea: that hemicellulose and lignin are worth more as co-products than for their simple heating value. Hemicellulose comprises between 19 and 32% of the mass of de-barked wood chips, lignin 22-31%, and cellulose 40-45% (Ragauskas, 2005). This means that less than half the input mass can be converted into pulp products—nearly as much as is currently burned for its fuel value. There are a large variety of technology routes to better utilize hemicelluloses and lignin. Hot Water Extraction (HWE) is an option of particular interest for both the pulp mill application and the LCEt-HF technology described earlier.

Hot Water Extraction was described in a previous section as a pretreatment process in the context of LCEt-HF. Water is used to soak a woody feedstock, dissolving (and to some extent hydrolyzing) the hemicellulose components and thereby making the cellulose components of the feedstock more accessible to digestion. Hemicellulose, a polymer comprised of five- and six-carbon monomers, can be dissolved and extracted as oligomers (shorter-chain polymer subcomponents) using hot water. Research has shown that near-complete extraction can be accomplished with treatment times of around two hours.[65] After this pretreatment is performed, the hemicellulose-rich water extract and the woody components of the feedstock are moved together through process steps that degrade cellulose and allow fermentation of its component sugars.

In the case of pulp and paper biorefining, the hemicellulose-rich water does not remain with the cellulose-rich woody solids, but rather is separated from them. The solids are then processed into paper products rather than ethanol. Only the hemicellulose solution is processed into ethanol, rather than all components. Lignin, which is irreducible to simple sugars and thus not fermentable, is used for fuel/heat, while the cellulose remains available for paper production. By virtue of this separation, the hemicellulose is used as a liquid fuel precursor rather than for its simple heating value—its ultimate fate in traditional pulping. The process as proposed by Amidon et al. (2008) is shown below in Figure H-11.

After extraction, the hemicellulose (resident in the extraction liquor) is subjected either to autohydrolysis or enzymes that partially reduce it to a solution of simple sugars: xylose and arabinose. These sugars can be fermented with microorganisms to form ethanol and potentially butanol (This is discussed subsequently in the section on biobutanol). The wood chips, now about 25% deficient of hemicellulose, enter the digestion phase, where lignin and the remaining hemicellulose are extracted as black liquor and burned. Frederick et al. (2006) proposed the same basic process for pine pulpwood, but considered the chemical effects of the amount of hemicellulose removal. While nearly all hemicelluloses can be extracted, this removal comes at the expense of cellulose mass. The long residence times required to extract nearly all the hemicellulose results in a portion of the cellulose being extracted as well. In context of value, this has the effect of transferring mass from a primary product (paper) into a blower-value biofuel co-product. At lower extraction efficiencies, less cellulose is co-extracted, but less hemicellulose is extracted for liquid fuel. The ideal operating point will depend on product and co-product values. For pine, one

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[65] Amidon et al (2008), using hardwood as feedstock.

study suggests that the point will be near 5% total mass removal, where little cellulose is leached, but around 15% of available hemicellulose is extracted (Frederick 2006; Finland 2006).

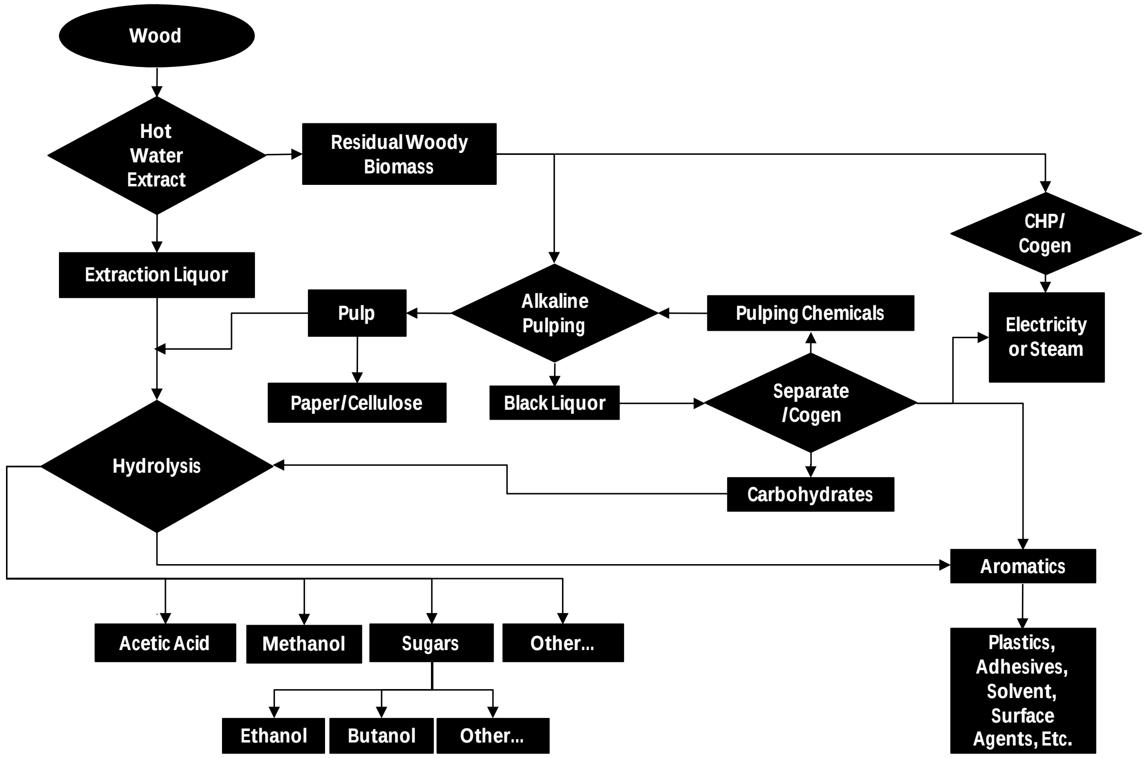

Figure H-11. Process Schematic Proposed by Amidon et al (2008).

2.9.2 Performance and Cost Analysis

There are tradeoffs to be considered in this approach. Hemicellulose extraction with hot water leads to complicated water supply issues. At a commercial-scale throughput, a two-hour residence time for extraction implies a large amount of water and tankage, probably equal to those of the existing digestion reactors. Thus, non-bleaching water requirements and reactor space may be approximately double. There is also an additional reactor heat requirement—which researchers have deemed “small” but not quantified (Frederick 2006; AIChE 2006)—and less heat is generated as result of less hemicellulose in the black liquor available for combustion.

Extracting hemicellulose before digestion also means that the digester need not process it. In theory, this reduces the amount of caustic and sodium sulfide required to process the chips into brown stock, as well as the residence time of digestion and thus digester volume. Since hemicellulose is a far easier molecule than lignin to attack, these reductions may be small, but the increased porosity of the extracted wood chips might eliminate the need for sodium sulfide altogether (Amidon et al. 2008). Moreover, lignin has a far higher heating value than hemicellulose, meaning that there is a relatively smaller energy loss than fuel mass loss upon burning.

A final complication may be reduced strength of paper products with near-complete removal of hemicellulose. Researchers have shown that cornstalk-based paper[66] tears most easily and binds least effectively when hemicellulose is greatly reduced (Ahmed 2006). Removing 17% of the total mass of cornstalk in the form of hemicellulose and trace cellulose (62% of the total amount of hemicellulose present) led to poor pulp binding and low tear energy (Ahmed 2006). Conventional wood-based pulp generally contains 19-32% hemicellulose by mass, but hot water extraction's two-stage removal of hemicellulose and lignin could reduce hemicellulose levels in wood-based pulp by 95% or more (depending on the chosen extraction time) leading to pulp hemicellulose contents of only1-2% by mass.[67] How this will affect paper quality remains undetermined. However, it is possible that this presents a boundary condition for the technology, since mills must carefully preserve the quality of their product stream (their pulp and/or paper).

Based on data from a recent HWE analysis paper (Amidon 2008), it appears that a mature application of HWE might result in a hemicellulosic ethanol yield of 136 gallons per ton hemicellulose feedstock. As the process is still early in development, capital cost and mill energy consumption changes are based on engineering simulations rather than plant construction data, and hemicellulose extraction appears to have only been tested at pilot scale. The largest extraction reactor for which data has been published measured 65 ft^3 and was run as a batch reactor rather than continuously (Amidon 2008). Commercial scale wood chip processing would require a continuous throughput. While there is published data on the fermentation efficiency of simple sugars, xylose and arabinose with microorganisms to form either butanol or ethanol, the effect of hot water extraction-specific by-products (particularly sulfur) on the fermentation remains unclear.

Frederick et al. (AIChE 2006) estimated the ethanol-related portion of the total annual production cost of pulp and ethanol from extracted hemicellulose at a thermochemical pulp mill to be $1.47 per gallon in $2008. The researchers' earlier estimates of a breakeven price ranged from $2.42-$3.11 ($2008) for facilities producing 6-16 million gallons per year (Frederick et al. 2006; Finland 2006). They compared capital costs of lignocellulosic ethanol facilities to make either ethanol and pulp fibers or ethanol alone, and found that integrated hemicellulose extraction required less capital, but had twice the break-even price. They concluded that it would be more economical to produce ethanol alone, rather than fiber and ethanol.

Mao (2008) simulated a hemicellulose HWE process for a small (2.7 MGY ethanol), a medium (3.8 MGY) and a large (5.5 MGY) Kraft pulp mill. These simulations indicated capital costs between $9.70 and $11.05 per gpy installed capacity. Ethanol-related annual operating costs ranged between $14.4 million (small) and $27 million (large).

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[66] Cornstalk paper has been investigated as a way to increase useof agricultural byproducts, particularly in Midwest states with relatively low forest inventories and large agricultural volumes. While still at only a bench scale, cornstalk paper appears to have been the only type of paper whose properties have been examined as a function of hemicellulose content.

[67] Pulp hemicellulose content for Kraft mill from Roberts, 1996. The range presented is for hard- and softwood feedstocks.

Frederick and Amidon both considered hemicellulose extraction at thermochemical Kraft mills, rather than thermomechanical pulping plants. Thermomechanical mills could also be candidates for hemicellulose extraction because they already handle large volumes of process water containing hemicellulose. Such streams, however, are typically very dilute (0.5 g/L) (Persson 2008), so the economic feasibility of using this hemicellulose depends on concentrating it so that fermenters and reactors can be of reasonable size.

Persson et al. (2008) estimated the economic feasibility of hemicellulose extraction in a thermomechanical pulp mill based on a computer simulated scale-up of laboratory data. Using ultra- and micro-filtration to separate hemicellulose from process water taken from a thermomechanical Swedish pulp mill, they estimated a cost of $88/ton (originally €670/ton) hemicellulose extracted and concentrated to 30 g/L. Assuming Amidon's ethanol yield estimates, this translates to $0.56 spent on extraction and concentration alone per gallon ethanol produced-fermentation, purification, and chemical analysis would raise this value dramatically. While the final concentration is excessive for fermentation purposes and the cost thus overestimated, Persson's cost analysis suggests that hemicellulose extraction at thermomechanical pulp mills may be infeasible due to the high dilution of their hemicellulose-containing streams.

Outline for Model Analysis

The important values and assumptions used to construct the hemicellulose extraction/ethanol production model are described below. A detailed presentation of the model is shown in Appendix H-F.

- Model inputs are:
    - feedstock type (Southern pine, eucalyptus, silver birch, and blue spruce; chosen due to widely available composition data and the fact that they represent both hard- and softwoods)
    - yearly plant feedstock consumption (on a dry basis)
- The facility was modeled as a retrofit to an existing pulp mill. The model assumes no land purchase, and all costs *are incremental* to the costs of the existing mill.
- Only part of the available hemicellulose (15 % by mass) was estimated to be extracted. This value was chosen based on data presented by Frederick et al. (2006), in which hemicellulose removal and cellulose removal were tracked as a function of time. Cellulose (destined for pulp production) remains almost completely undissolved by hot water treatment until a time corresponding to 15% removal of hemicellulose. After this point, the slope of the cellulose solubility line increases dramatically, and larger amounts of cellulose are lost from the pulp product to the ethanol product.
- Feedstocks were assumed to contain 30 wt% moisture as received. Air dry wood contains 15-20 wt% moisture (Scurlock, 2008), but green wood inputs can contain as much as 50 wt% moisture. Thirty percent was chosen as an intermediate value so as not to grossly under- or over-size processing equipment.

- Suggested feedstock costs (which are not included in operating or production costs presented, but rather only for reference) are conservative estimates of feedstock cost, drawn from the North Carolina Timber Report and previous studies conducted by Antares.

- Capital cost data was drawn from two detailed economic assessments of ethanol production. Mao et al (2008) constructed a cost estimate for retrofitting a hemicellulose extraction facility onto an existing Kraft mill, based on detailed process simulations in ASPEN. This study determined a scaling factor of 0.699 for capital cost and a Lang factor of 3. It did not take into account the extra tankage required for the hot-water extraction, so additional tanks were added to the capital cost presented here. Frederick et al. (2006) also performed an economic assessment, which agreed well with Mao's results. All costs were inflated to 2008$ using the CEPCI index.

- Energy balance data from Mao et al. (2008) was used to estimate utility demands for the facility. Non-labor, non-feedstock variable operating cost assumptions are based on previous Antares technology models (2008), which scaled these data from McAloon et al. (1999). Electricity and natural gas were priced as the 2008 industrial sector average, based on EIA data (2009). Labor was allocated as in Brown (1999) and priced by the authors' experience.

- Reduced pulp yield due to diversion of hemicellulose mass to ethanol production is added to the overall cost of production. U.S. 2008 market prices for bleached pulp (NBSK, or long-fiber northern bleached softwood Kraft pulp) ranged between $700 and $800/ton based on FOEX PIX as delivered prices. $750 per ton was chosen as a pulp value and discounted 20% to account for the lack of a transportation cost in the value of the pulp to the mill. Reduction in pulp yield was calculated based on a digestion hemicellulose removal efficiency of 80%, based on average Kraft pulp hemicellulose content of 12.5%.

- Yields were calculated based on the relative amounts of hemicellulose and cellulose in each feedstock. As per Amidon et al. (2006), who present typical values for fermentation and hydrolysis efficiencies, the model uses 90% fermentation efficiency and 90% hydrolysis efficiency.

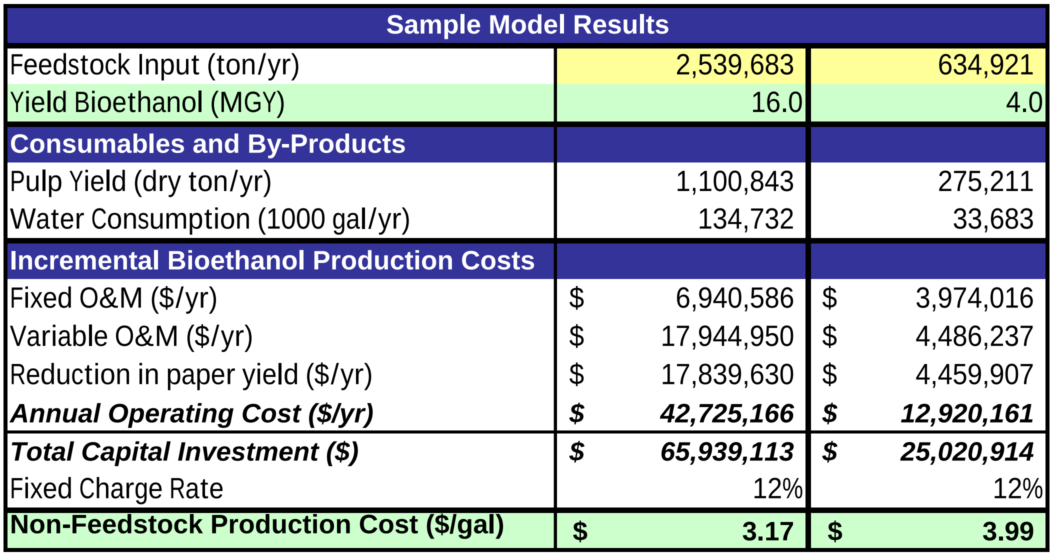
- Conversion efficiency is based on the HHV data from the EERE feedstock database and is presented on a dry mass basis. Only the liquid fuel (by) product-- ethanol--is counted toward the output energy value. The total hemicellulose initially present in the feedstock is counted as input energy.

- The ethanol production volume is presented as ethanol denatured to 95% with gasoline.

Model Results

Sample results from the hemicellulose extraction production and cost analysis model are presented in Table H-14 below. It is important to note that though non-feedstock production costs are high compared to grain and lignocellulosic ethanol technologies, the feedstock cost is likely to be very low, depending on how the cost of a mill's pulpwood is debited against each product line: pulp and ethanol. This is a function of the small mass fraction of a wood feedstock that would be used to produce the fuel co-product. Masses of feedstock in the examples below are stated as whole wood, rather than as tons of hemicellulose extracted from that wood.

Table H-14. Example Cost and Performance from HCEt Model.



Sample Model Results		
Feedstock Input (ton/yr)	2,539,683	634,921
Yield Bioethanol (MGY)	16.0	4.0
Consumables and By-Products		
Pulp Yield (dry ton/yr)	1,100,843	275,211
Water Consumption (1000 gal/yr)	134,732	33,683
Incremental Bioethanol Production Costs		
Fixed O&M ($/yr)	$ 6,940,586	$ 3,974,016
Variable O&M ($/yr)	$ 17,944,950	$ 4,486,237
Reduction in paper yield ($/yr)	$ 17,839,630	$ 4,459,907
Annual Operating Cost ($/yr)	***$ 42,725,166***	***$ 12,920,161***
Total Capital Investment ($)	***$ 65,939,113***	***$ 25,020,914***
Fixed Charge Rate	12%	12%
Non-Feedstock Production Cost ($/gal)	**$ 3.17**	**$ 3.99**

2.9.3 Outlook

The deployment timeline and commercial outlook for HWE in biorefining is complex, more so than for HWE in LCEt-HF. HWE as a part of a pulp mill involves the integration of new and existing facilities in an industry that is short on capital and profits. Stand-alone ethanol plants, on the other hand, must consider some kind of pretreatment for lignocellulosic biomass. The concern in this latter case is which technology to choose, rather than whether to produce biofuels at all.

Research and pilot scale testing suggests that the technology has merit, but because it relies on integration with existing mills, protective measures and compromises will be made to preserve the quality of a mill's primary product, pulp. Additionally, there seems to be some skepticism in the literature as to whether the introduction of the technology will actually improve a mill's bottom line. However, since the technology is relatively early in its development and a large demonstration project is not available for testing, there is probably much that remains to be seen with respect to challenges and benefits for full integration of the technology. That said, it is clear that if the technology can improve the economic and environmental performance of a pulp mill, it could play an important part in preserving a U.S. industry that is finding it ever more difficult to compete in a global economy.

2.10 LIGNOCELLULOSICS TO GASOLINE: UPGRADING PYROLYSIS OIL

Lignocellulosics to Gasoline via Pyrolysis and Upgrading (LCGa Upgrading/Pyrolysis) involves the conversion of biomass to bio-oil via fast pyrolysis, and then upgrading via hydrotreatment followed by hydrocracking at a petroleum refinery to produce a hydrocarbon fuel. In this process the bio-oil is mixed with petroleum products during some stage in the refining process, which enables usage of advanced refinery technologies and economies of scale. In the following discussion the general method of biomass pyrolysis will first be reviewed, followed by an analysis of the development of bio-oil co-processing in refineries (Antares 2008).[68]

2.10.1 Technology Description

Biomass Pyrolysis

Pyrolysis of biomass is a rapid thermal process conducted in an environment without oxygen to prevent combustion. By controlling temperatures and reaction times, pyrolysis can be used to convert solid biomass materials into a liquid oil comprised of low molecular weight fragments of the lignin, cellulose and hemicellulose. This “bio-oil” can be used directly as a substitute for petroleum fuel oil in boilers and gas turbines. Alternately, bio-oil can be further treated to produce refined fuels or natural chemicals such as adhesives, resins, polymers and flavorings. A variety of biomass feedstocks have been used for pyrolysis, including wood, bark, paper, bagasse, corn fibers and other agricultural residues. The bio-oil yield primarily depends on the processing conditions with the rate of heat transfer and product vapor residence time being the key parameters. Feedstock composition, principally the ash component, can also be a factor but plays a lesser role. Under optimum process conditions liquid yields of 60-80% by weight can be realized (Ensyn Group Inc. 2001).

To prepare the biomass for fast pyrolysis, the feedstock is dried to less than 10% moisture content and sized to small relatively homogeneous particles. This is important to ensure a rapid heat transfer. In commercial pyrolysis processes, the reaction occurs in a fluidized bed reactor using an inert material such as sand to transfer heat to the incoming biomass particles. This process occurs so fast that the biomass is flash vaporized, becoming a mixture of gas, vapor, aerosols and solid char. Fast pyrolysis produces higher yields of the liquid fraction in bio-oil than conventional pyrolysis, with a typical composition of 75% liquid, 12% char, and 13% gas (Ringer, Putsche and Scahill 2006).[69]

The pyrolysis step typically occurs at temperatures around 450-500ºC, at atmospheric pressure, with a reaction time from less than one second to several seconds. After the char is separated out (using a cyclone), the vapor product is rapidly cooled and condenses into liquid bio-oil. The remaining non-condensable gas is recycled within the process and used as a fluidizing gas for pyrolysis and as a fuel combined with the solid char to provide process heat.

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[68] The text in this section was revised from material originally presented by Antares in 2008 in a *Strategic Assessment of Bioenergy Development.*

[69] Conventional pyrolysis involves lower reaction temperatures and long residence times, and typically results in nearly equal fractions of liquid, gas, and char produced (Bridgwater 2007).

The bio-oil product is an oxygenated fuel that consists primarily of carbon, hydrogen and oxygen and small amount of nitrogen and sulfur. The exact composition of the product depends on the feedstock composition and the reaction conditions.

Co-processing bio-oil in petroleum refineries

Crude pyrolysis oil can be used directly to produce heat and power. However, upgrading bio-oil to a hydrocarbon fuel via hydrotreatment and hydrocracking can significantly increase the value of the product by producing fungible gasoline and diesel fuels. Hydrotreating and hydrocracking are well-developed processes currently used for rejecting nitrogen, oxygen and other heteroatoms[70] from crude petroleum oils. This potential upgrading approach is the focus of efforts to co-refine pyrolysis oils with petroleum.

One of the main issues with crude bio-oil is the lack of stability over time. According to Ringer, Putsche and Scahill (2006), bio-oil viscosity increases with time at much faster rate than petroleum products. Such highly viscous oil cannot be used as motor fuel. Furthermore, the separation between aqueous and organic phases that accompanies the increase in viscosity is also a problem for usage. Char fines in the bio-oil seem to be a critical factor causing this lack of stability. Advanced technologies that can successfully remove fine particles can significantly increase the shelf life of the product and are currently under development.[71]

Other potential issues with bio-oil include re-vaporization from liquid state and potential environmental and health effects. Bio-oil has a very complex chemical composition, and more than 300 specific compounds have been identified so far (Ringer, Putsche and Scahill 2006). Some of the compounds present are known carcinogens, and experimental tests have shown mixed results with respect to health risks. Further, although bio-oil biodegrades in soil more easily than petroleum-based hydrocarbon fuels, its acidity can be harmful to water environments in the event of large spills (Ringer, Putsche and Scahill 2006).

Additionally, incorporation of new fuels into a mature and accepted framework can be difficult. Using bio-oil directly as a transportation fuel would likely require engine modifications and new infrastructure. However, "if bio-oils could be upgraded chemically to produce a product that looked more like petroleum hydrocarbons, then the end use device would require little to no modification. This would be the fastest way to gain acceptance of biomass-based fuels into the existing infrastructure." (Ringer, Putsche and Scahill 2006, p. 27)

The bio-oil upgrading process primarily involves oxygen removal by hydrotreatment followed by hydrocracking. Catalytic hydrotreating and cracking deoxygenate the bio-oil and reform the remaining carbon and hydrogen into hydrocarbon compounds in the gasoline and diesel boiling point range. Since the carbon oxides and water are removed, these processes reduce the yield of the liquid product of bio-oil (by weight), although hydrotreating gives

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[70] "Heteroatom" is an organic chemistry term used to describe any atom that is not carbon or hydrogen.

[71] A 6-month shelf (equivalent to that of petroleum products) life can be achieved with the addition of alcohol or other solvents and removing the char fines. The fine particle removal is difficult and somewhat analogous to gas-clean up from gasification. Additional research is needed in order to find an easy, effective, low-cost solution.

higher yields than catalytic cracking (Ringer, Putsche and Scahill, 2006). The hydrocarbon product from catalytic cracking processes is also dominated by aromatic compounds which limit the amount that can be blended into motor fuels (Ringer, Putsche and Scahill 2006).

Pyrolysis oil contains a mixture of products including water-soluble oxygenated compounds (derived from the hemicellulose), and insoluble pyrolytic lignin (UOP 2005). The pyrolytic lignin is a lower molecular weight version of lignin, which results from the thermal conversion of biomass feedstocks. This component of the pyrolysis oil is lower in oxygen and has a higher energy content than the water soluble portion. It can be separated out via gravity separation or by adding water to the pyrolysis oil to precipitate out the insoluble pyrolytic lignin portion. Hydrotreatment of the pyrolytic lignin requires less hydrogen than the pyrolysis oil because of the lower oxygen content, and can also be done at mild reaction conditions.

High oxygen and acid content are two challenges for processing bio-oils. Typical refinery feedstocks like crude oil have little or no oxygen, but pyrolysis oils have oxygen contents on the order of 35-45% (UOP 2005). The pyrolytic lignin has less oxygen than the water-soluble carbohydrate portion of the bio-oil, and therefore requires less hydrogen consumption during the upgrading process. However, the carbohydrate portion of bio-oil has been shown to be a promising feedstock for reforming to hydrogen, which could provide a synergy with co-processing of both components of the bio-oil. The water soluble portion can also be used as a fuel to generate heat and power for the upgrading process.

Pyrolysis oils are much more acidic than other refined fuels. The acidity of refinery processed feedstocks is measured by total acid neutralization (TAN) number.[72] The TAN for crude is <1, while a typical pyrolysis oil has TAN of 78 (UOP 2005). This level of acidity requires processing in stainless steel vessels, which are expensive and not standard in refineries. Pretreatment of the pyrolysis oils can be done in stainless steel vessels to reduce the acidity, making co-processing in conventional downstream vessels possible.

Figure H-12 shows a potential arrangement for producing pyrolysis oil and co-processing in a petroleum refinery. The biomass feedstock is first sized and dried for pyrolysis. After the reaction, the pyrolysis oil is separated into components - the oil fraction and pyrolytic lignin are co-processed to produce conventional fuels while the water-soluble components are converted into oxygenated products. Process residues are used for energy generation.

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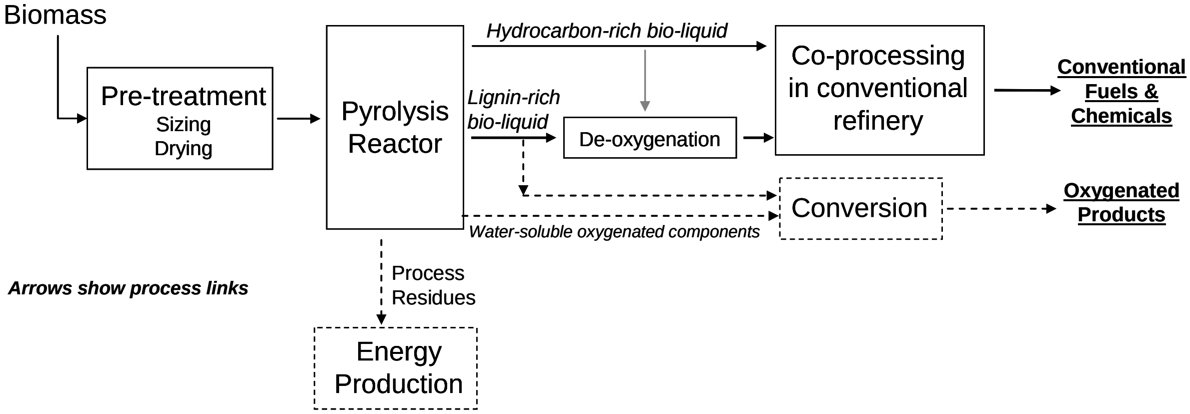
[72] Total acid neutralization number is defined as the number of milligrams of potassium hydroxide to neutralize one gram of feedstock (UOP 2005).

Figure H-12. Schematic Diagram of the LCGa Upgrading/Pyrolysis Process.



Based on figure in Solantausta (2006)

2.10.2 Performance and Cost Analysis

Bio-oil yield is affected by a number of factors, including feedstock composition, reaction temperature and heat transfer rate, and amount of inert gas in the reaction environment. Furthermore, the mineral matter in biomass acts as a catalyst for cracking and polymerization during pyrolysis, which affects the composition of the bio-oil product (Ringer, Putsche and Scahill 2006). However, despite the importance of many factors in bio-oil production, the heating value of the product is relatively constant (on a dry basis) at about 7,500-8,000 Btu/lb (Ringer, Putsche and Scahill 2006). This is only a slight reduction from the heating values of the biomass feedstocks used for pyrolysis, which are generally around 8,500 Btu/lb.

The water content in the biomass feedstock is also an important factor in bio-oil production. In fact, even pyrolysis of bone-dry biomass produces bio-oil with 12-15 wt% water (Ringer, Putsche and Scahill 2006). Any moisture in the input biomass will contribute to an even higher percentage of water in the bio-oil. As water is an unfavorable component in bio-oil, and also acts as a heat sink, it is advantageous to remove as much water as possible from the incoming biomass before pyrolysis.

Upgrading bio-oil is important for a number of reasons. Hydrotreating and hydrocracking bio-oil converts it to a hydrocarbon fuel that can be blended and transported with petroleum fuels. This provides access to a wide infrastructure already in place. Upgrading can also significantly increase the value of the bio-oil.

The ratio of pyrolytic lignin to water soluble pyrolysis oils is such that in the near term, a substantial electricity/steam plant should accompany development of any fuel project based on this technology to ensure maximum use of the incoming resource. As complete and efficient use of incoming feedstock is critical to any project, the energy/liquid fuel product ratio will be an important consideration in siting. A potential alternative for the long-term is to reform the water-soluble portion of the bio-oil to generate hydrogen for use in the upgrading process. Although the reforming technology is currently too expensive to support a stand-alone pyrolysis oil upgrading facility, it is a possible future development.

The evolution of pyrolysis oil production and upgrading is assumed to be as follows:

- *Short term*: no pyrolysis oil upgrading technologies available (still in research phase)
- *Mid term*: hydrotreating / hydrocracking of pyrolytic lignin and oil fraction in a separate (skid-mounted) process co-located at a conventional refinery; resulting products blended with petroleum products for storage and distribution.
- *Long term*: no analysis of the technology for this period. However, hydrotreating / hydrocracking the pyrolytic lignin and oil fraction via co-processing in a stand-alone refinery, then reforming the water-soluble portion of the bio-oil for hydrogen production, is a potential long-term scenario.

Outline for model analysis

The key values and assumptions used to calculate the mid-term technology analysis and economic performance model for LCGa Upgrading/Pyrolysis are described below. An example of the detailed analysis is shown in Appendix H-F.

- Feedstock type and quantity are variables for input. Types of materials that can be used for pyrolysis include woody and agricultural feedstocks. The applicable size range for the model is biomass input of 30,000 - 800,000 dry tons per year, based on reported facility sizes in UOP (2005); Ringer, Putsche and Scahill (2006); and Polagye, Hodgson and Malte (2007).
- The facility is assumed to be operational at full load 330 days per year, which is equivalent to a capacity factor of 90%.
- Pyrolysis yield depends on feedstock type, and the values used are based on published data from Ensyn and Dynamotive (Ensyn Group Inc. 2001; Dynamotive Energy Systems Corporation 2000).
- The conversion efficiency of biomass to pyrolysis oil is estimated to be 66-83% (on an energy basis) depending on the feedstock, with bark giving the lowest yield and bagasse and mixed paper giving the highest. This is calculated using the HHV of the feedstocks and typical pyrolysis oil properties.
- Pyrolytic lignin (including the oil fraction) is the only portion of the bio-oil that undergoes hydrotreatment / hydrocracking for upgrading to hydrocarbon fuels, and the hydrotreatment and hydrocracking units are sized based on this feed stream.

- The pyrolytic lignin content (weight percentage) for all herbaceous feedstocks based on values for bagasse, and percentage for woody feedstocks is based on data for softwood (from Dynamotive Energy Systems Corporation, 2000).

- The only significant by-products of pyrolysis oil production are solid char and non-condensable gases. It is assumed both of these products are recycled into the process to use as a heat source and fluidizing gas for the reaction. There are several products from upgrading, including gasoline and diesel, light hydrocarbons and the un-treated water-soluble bio oil. Gasoline is the primary product, accounting for 30% of the converted pyrolytic lignin (by weight).

- The diesel and light hydrocarbon by-products are valued at $1.76/gal and $0.93/gal respectively, based on the average 2005 wholesale values from EIA AER (2005).[73]

- The water-soluble bio-oil can be cofired with natural gas (or refinery off-gas) to provide heat and/or electricity. This material needs to be co-fired because of its relatively high water content (20-30%) and low heating value (5,000 Btu/lb on a moisture-free basis). If the water-soluble bio-oil was not combusted it would become a disposal liability. To simplify the model, it is assumed that the value of the heat/electricity produced from co-firing the water-soluble bio-oil offsets the cost of natural gas consumed for combustion. As such, there is no net value or cost associated with the water-soluble component of the bio-oil.

- The yield of the concentrated CO_2 generated from hydrotreatment is derived from UOP (2005). No value is assigned to this stream in the analysis.

- Water consumption is estimated based on data from Ringer et al. (2006). According to Ringer et al. (2006), a 200,750 dry ton per year facility uses about 160 thousand gal/hr water, mostly for cooling. It is assumed that only 2% of this water is lost in blowdown, so that the annual consumption is about 28,000 thousand gal/year. This value is scaled based on feedstock input. No additional water consumption is added for hydrotreatment.

- Capital costs for the pyrolysis process and hydrotreatment are calculated separately, as they have different size and economies of scale.

- Total capital investment values for pyrolysis oil production for a range of sizes were available from various sources.[74] These values were used to determine a relationship between facility size and capital cost, shown in Figure H-13. A scaling factor of 0.8 is used for capital cost calculations.

- The capital cost for hydrotreatment is based on a value of $30 million for a unit that processes 2,250 barrels per day, and a standard scaling factor of 0.6 (UOP 2005).

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[73] For comparison, the projected values of diesel and Liquefied Petroleum Gas in 2015 are $2.04/gal and $1.34/gal ($633/ton), respectively (EIA AEO 2007).

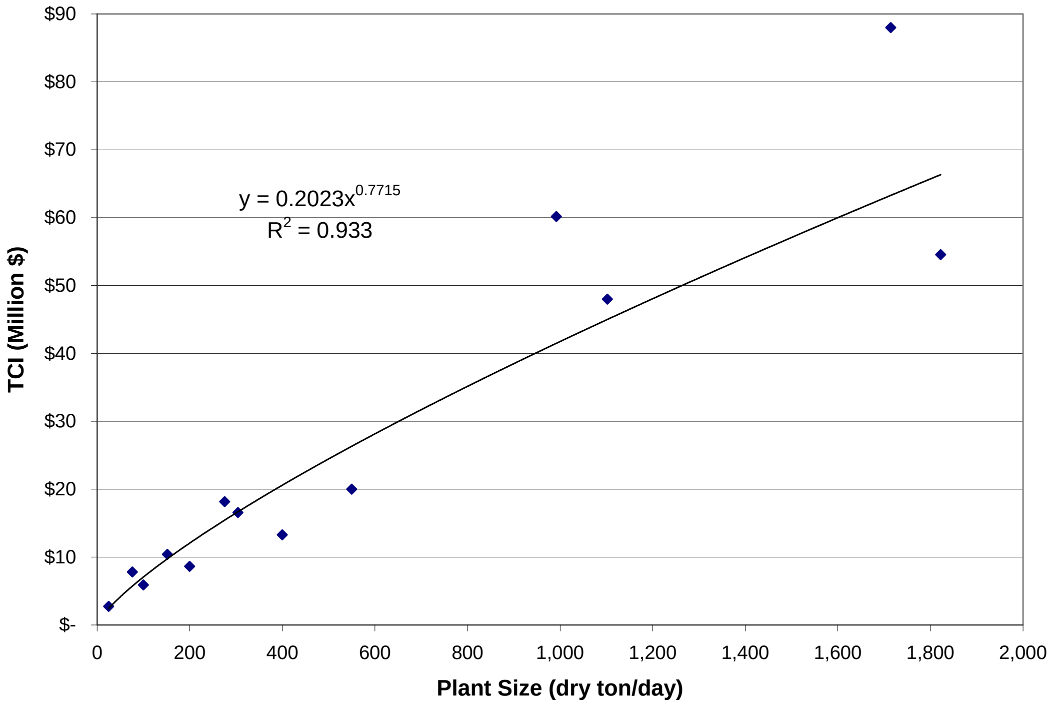
[74] Sources: Ringer, Putsche and Scahill 2006; Polague et al. 2007; UOP 2005; and Cole Hill Associates 2004.

- Variable O&M costs for the process include electricity and other utilities, hydrogen gas for hydrotreatment, and waste disposal. These annual costs are estimated individually, based on plant size. Electricity cost for pyrolysis oil production has been updated to 5.7 cents/kWh based on the 2005 average cost for electricity in the industrial sector from EIA (2006). The electricity cost for hydrotreatment has not been updated as there was not sufficient detail in the published studies, but the difference is expected be minimal and would not have significant effect on the production cost.

- Annual fixed O&M costs include labor, maintenance, insurance, overhead, and other costs. Except for the labor component, these costs are estimated as a percentage of total capital cost. The labor portion only applies to the pyrolysis oil production process, and is scaled based on project size using a relation derived from the available data from Polagye et al. (2007) and Ringer, Putsche and Scahill (2006). This relationship is shown in Figure H-14.

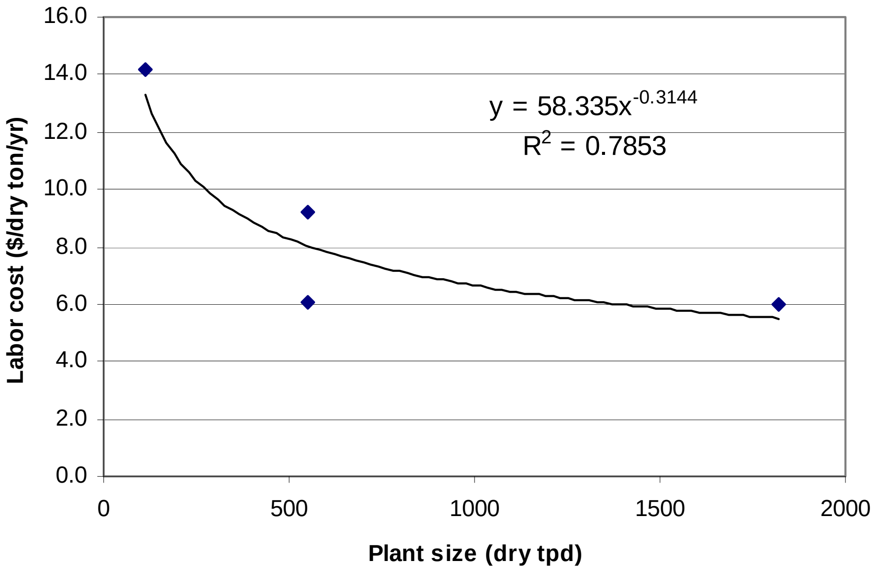
- The levelized non-feedstock production costs for gasoline are calculated based on the annual operating expenses and capital cost payment, co-product credit, and quantity of fuel produced.

Figure H-13. Capital Cost and Plant Size Relation for Fast Pyrolysis.



Based on data from Cole Hill Associates (2004), Ringer et al. (2006), Polagye et al. (2007), UOP (2005).

Figure H-14. Labor Cost Scaling by Facility Size for Pyrolysis Oil Production.



Based on data from Polagye et al. (2007); and Ringer, Putsche and Scahill (2006).

Model Results

Some of the key cost and performance results projected for LCGa Upgrading/Pyrolysis are shown in Table H-15. This table shows results for the range of applicable facility sizes based on selected input quantities, illustrating the high and low end costs. In general, the analysis shows that the non-feedstock production cost of gasoline with this method is quite expensive. This suggests that further development is needed to make pyrolysis oil with upgrading competitive with other biofuel technologies. For example, co-processing in a refinery could have significant impact on the production cost if the processing challenges can be overcome.

Table H-15. Example Cost and Performance for LCGa Upgrading/Pyrolysis.

Sample Model Results - LCGa Upgrading/Pyrolysis w/ Wood		
Feedstock Input (dry ton/yr)	30,000	800,000
Crude Pyrolysis Oil Yield (MGY)	4.58	122
Gasoline Yield (MGY)	0.507	13.5
Diesel Yield (MGY)	0.126	3.30
Light Hydrocarbon Yield (ton/yr)	839	22,380
CO2 Stream (ton/yr)	7,600	202,620
Water Consumption (1000 gal/yr)	4,180	111,420
Capital Cost - pyrolysis (Million $)	$ 7.78	$ 98.0
Capital Cost - hydrotreatment (Million $)	$ 4.86	$ 34.8
Annual O&M Costs (Million $/yr)	$ 2.41	$ 44.7
By-Product Credit (Million $/yr)	$ 0.82	$ 21.8
Non-feedstock Production Cost ($/gal)	***$ 6.20***	***$ 2.90***

2.10.3 Outlook

Until recently there were only two very active companies that developed commercial pyrolysis oil technologies - Ensyn Corporation and DynaMotive Energy Systems Corporation. Each company has a patented pyrolysis technique and a wide range of experience producing bio-oil from different feedstocks. Their products have been tested and used in various applications. By 2005 Ensyn had 7 commercial RTP^{TM} biomass plants, and the largest plant can process 160 green tons of wood per day (Ensyn Corporation n.d.). DynaMotive Energy Systems Corporation reached commercialization in 2004 with a 110 ton/day biomass pyrolysis BioTherm^{TM} plant in Ontario. DynaMotive also completed their first modular 220 ton/day biomass pyrolysis plant in December 2006 (Dynamotive Energy Systems Corporation 2006).

UOP, a specialist in refining process technologies, became active in renewable fuel technology, forming a separate Renewable Energy & Chemicals business in late 2006. Since then, UOP has commercialized the UOP/Eni Ecofining™ process to produce green diesel fuel from biological feedstocks and has also developed process technology to produce renewable jet fuel under a contract from the U.S. Defense Advanced Research Projects Agency (DARPA). UOP has ongoing research efforts in biofuels, with specific focus on second-generation feedstocks working with organizations such as the U.S. Department of Energy (DOE), the DOE’s National Renewable Energy Lab, Pacific Northwest National Lab, and the U.S. Department of Agriculture (USDA) Agriculture Research Service Eastern Regional Research Center (ARS-ERRC). In 2009 UOP launched Envergent Technologies, LLC, a joint venture with Ensyn Corp., to offer technology and equipment to convert lignocellulosic biomass like forest and agricultural residuals to pyrolysis oil for use in power and heating applications. The joint venture will accelerate development efforts to commercialize the technology for fast pyrolysis and pyrolysis oil upgrading into transport fuels such as green gasoline, green diesel and green jet fuel.

BIOCOUP is a European consortium of research partners that have formed to evaluate the development of co-processing bio-liquids in refineries. According to a presentation given at the 2006 European Conference on Biorefinery Research (Solantausta 2006), the consortium’s objectives for pyrolysis oil co-processing include: reducing bio-oil production costs, develop and scale de-oxygenation technology for upgrading bio-oil, assess the viability of co-processing upgraded bio-liquids, optimize recovery and fractionation strategies for conversion to chemicals, and develop life cycle analysis and scenario analysis.

One of the key barriers for development of the co-processing technology is that refineries are unwilling to experiment with processing for small quantities of pyrolysis oil that may detrimentally affect their catalysts or end product (UOP 2005). A typical U.S. petroleum refinery generates 150,000 barrels per day (bpd), while the total production of pyrolysis oil is only a small fraction of that amount (UOP 2005).

3 PRODUCTION OF BIOFUELS FOR THE DIESEL FUELS MARKET

Biodiesel is currently produced in significant quantities using commercial conversion technologies. It is also produced in very small operations by "do it yourself-ers," but these rarely achieve the efficiency and solvent recycling levels of commercial facilities and will not be discussed here. The technology is based on transesterification of oils and greases to generate biodiesel, technically known as fatty acid methyl ester (FAME). There are three basic conversion routes for FAME production: base catalyzed transesterification of oil with alcohol; direct acid catalyzed transesterification of oil with methanol; or conversion of oil to fatty acids, followed by acid catalysis to alkyl esters (EUBIA 2006; NBB n.d.).

3.1 CURRENT TECHNOLOGY: FAME

The base-catalyzed transesterification option tends to be the most economic for virgin feedstocks and as such is most commonly used to produce esters on a commercial scale.[75] There are several reasons that favorably affect the process economics, including low temperature and pressure processing requirements, high conversion rates (around 98%), lower reaction times, and direct conversion to methyl esters without producing any intermediate compounds (EUBIA 2006; Antares 2008).[76]

3.2 FATTY ACID TO METHYL ESTERS: TRANSESTERIFICATION (BIODIESEL)

Acid catalyzed transesterification[77] is expected to be the preferred method for conversion of waste oils, since it is less sensitive to free fatty acids in the feedstock. This conversion method seems to be more economical than base catalyzed transesterification of waste oils, which requires an extra pretreatment step to remove impurities (Zhang et al., 2003b). Although acid catalyzed transesterification has not yet been optimized for commercial scale production, it is expected that this reaction method could be available by 2015.

3.2.1 Technology Description

Figure H-15 illustrates the main process steps for transesterification of vegetable oil. Before transesterification, the vegetable oils and fats are filtered to remove water and contaminants. The oil is then mixed with an alcohol (often methanol, because of its low cost) and a catalyst (usually sodium hydroxide or potassium hydroxide for base catalyzed reaction and sulfuric acid for the acid catalyzed process) in a simple closed reactor system at low temperature and pressure. For base catalyzed process, the molar ratio of methanol to oil is about 6:1, while for an acid catalyzed process the ratio is about 50:1 (Zhang et al. 2003a). The mixture is left to settle in the reaction vessel for 1 to 8 hours, after which the glycerin and crude biodiesel (methyl esters) are separated. The majority of the

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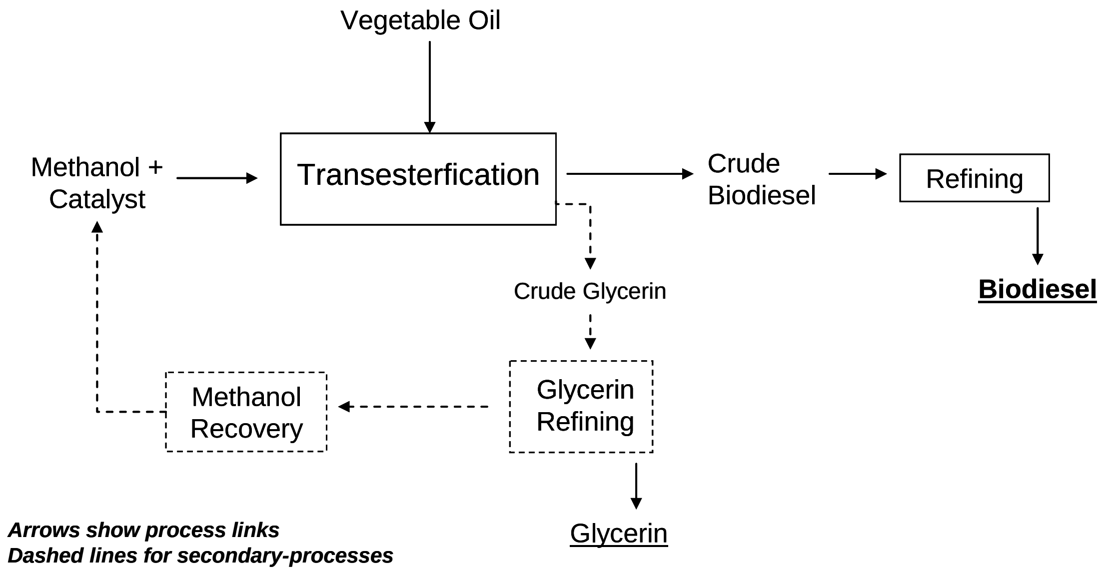
[75] Virgin feedstocks include vegetable (seed) oils and animal fats.

[76] The text in this and the following section was revised from material originally presented by Antares in 2008 in a *Strategic Assessment of Bioenergy Development.*

[77] Acid-catalyzed transesterification should be considered distinct from acid-catalyzed esterification, which is a commonly used pretreatment step in base-catalyzed transesterification. (Cardoso 2008)

alcohol is recovered from the glycerin by-product and recycled back into the system. The biodiesel is purified, usually by a washing process, to remove residual catalyst and soaps.

Figure H-15. Schematic Diagram of Transesterification Process.



Arrows show process links
Dashed lines for secondary-processes

The purity and yield of the methyl esters are affected by the molar ratio of glycerides to alcohol, type of catalyst, reaction time, reaction temperature, and the free fatty acids and water present (Vicente, Martinez and Aracil 2007; Ma and Hanna 1999).[78] The presence of water causes soap formation during the reaction, and free fatty acids significantly reduce ester yields (particularly for base catalyzed reaction). The formation of soap is an undesirable by-product as it reduces the biodiesel yield and makes glycerol separation more difficult (Vicente, Martinez and Aracil 2007).

Experiments involving the transesterification of sunflower oil feedstock showed that the initial catalyst concentration was the most important factor for biodiesel purity and yield. The catalyst concentration was found to positively affect the biodiesel purity, and negatively affect the yield (Vicente, Martinez and Aracil 2007). In other words, a higher initial catalyst concentration produced more pure biodiesel in smaller quantities than a lower catalyst concentration. Vicente, Martinez and Aracil (2007) also found that higher temperatures negatively affected biodiesel yield and positively affected the purity.

Although under similar conditions the acid catalyzed transesterification process has a slower reaction rate than the base catalyzed process, this can be overcome by a number of methods, including longer reaction times, using a

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[78] Fatty acids can be bonded to another molecule (such as glycerol in triglycerides). Free fatty acids are not attached to another molecule, and are a type of impurity in oils used for biodiesel production. Ma and Hanna (1999) note that if oils are not dry and free fatty acids minimal (<0.5%), ester yields may be significantly reduced.

larger quantity or increased concentration of acid catalyst, or a higher reaction temperature (Zhang et al., 2003a). However, the acid catalyzed process has not yet been done on a commercial scale.

3.2.2 Performance and Cost Analysis

Biodiesel has been of interest as a substitute for petro-diesel both for reasons of biodegradability and for its physical characteristics. When burned, biodiesel has been shown to have reduced particulate emissions and possibly lower NO_x emissions. Studies of NO_x production have shown a stronger link between engine type and NO_x emissions than between fuel type and NO_x emissions, but biodiesel appears to be at least comparable to, if not an improvement over, petro-diesel in nitrogen oxide production.[79] Biodiesel has also proven to be more compatible than petro-diesel with NO_x control technologies, due to its lack of sulfur contamination. Sulfur is a potent catalyst poison that quickly disables NO_x control systems that use Selective Catalytic Reduction (SCR). Further, biodiesel has been shown to be more compatible than petro-diesel with diesel particulate filters (DPFs) that control soot emissions (Williams 2006).

Potential drawbacks to biodiesel as a fuel substitute center on its cold-weather behaviors, which may include gelling and increased viscosity. Reluctance among motor manufacturers to certify engines for more than 5% biodiesel (B5) in diesel blends has also inhibited its growth as a diesel substitute. Chrysler and GM have approved blends up to 20% (B20) for some trucks, but Ford, Volkswagen, and Mercedes remain the only auto manufacturers to certify even B5 for most of their diesel models (HybridCars 2008).

A study by Haas et al. (2006) investigated the economics of a 10 MGY acid catalyzed FAME facility using virgin oil feedstock. They found that the largest portion of the capital cost was storage tanks to hold a 25 day supply of oil feedstock, which made up about 1/3 of equipment costs. An analysis by Zhang et al. (2003b) showed that the transesterification reactor vessels and distillation columns were the largest portions of the equipment costs, particularly for the acid catalyzed process (that has a much larger methanol requirement and requires two reactors and stainless steel vessels). However, their analysis did not include the cost of feedstock storage vessels.

For a transesterification processes using virgin oil, the feedstock is the bulk of the production costs. For example, in the analysis by Haas et al. (2006), soybean oil made up 88% of the calculated production costs. Furthermore, production cost was found to vary linearly with feedstock costs, and inversely with market value of glycerol. In their analysis, Zhang et al. (2003a) showed that the base catalyzed process with virgin oil was the most simple and had the least process equipment, but also had very high feedstock costs. Zhang et al. (2003b) showed that the economics of acid catalyzed process using waste oil feedstocks can be competitive with base catalyzed process using virgin feedstocks. Although the acid catalyzed process has higher capital and O&M costs, the feedstock is much less

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[79] Several studies have examined this question. NREL, in its high altitude engine test laboratory, has performed stationary emission tests on various biodiesels (Graboski, M.; McCormick, R., 1994). These indicated a marked decrease in hydrocarbon emissions, but a significant increase in NO_x emissions upon switching from petroleum- to plant-derived diesels. Later studies by McCormick (2005) found that in engine-alone tests, NO_x emissions increase, particularly in older engines, but that actual vehicle testing shows a modest reduction, rather than increase in NO_x emissions. There does not appear to be a consensus among researchers whether biodiesel in practice elevates or decreases these emissions rates.

expensive. In their analysis, Zhang et al. (2003b) showed that plant capacity and feedstock oil price had the largest effect on production costs of biodiesel for both types of processes.

Outline for model analysis

The key values and assumptions used to calculate the current technology analysis and economic performance model for transesterification are described below. An example of the detailed analysis is shown in Appendxix H-F.

- Feedstock type and input quantity are variables. The feedstock options are virgin oil, animal fats, and yellow grease, and the applicable size range is 1 to 80 MGY.
- Biodiesel yield is calculated based on the average of values reported in the literature for each feedstock type (summarized in Table H-16). In general, the virgin feedstocks have higher yields than yellow grease. Reported yields from virgin oil are from 246.2 to 270 gal/ton (90% - 99% conversion on a mass basis), and for yellow grease the yields are 234.9 to 263.2 gal/ton (86% - 97% conversion).[80]
- The conversion efficiency of oil/fat to fuel on an energy basis is estimated to be 88-94%, depending on the feedstock. This is calculated using an average HHV of oil feedstocks of 16,500 Btu/lb, and a typical biodiesel heating value of 117,000 Btu/gal.
- Glycerin is the only significant by-product of transesterification. The average production rate is 0.8 lb glycerin per gallon of biodiesel produced from virgin feedstocks (Bender, 1999, Haas et al., 2006). These studies refer to the production of a partly purified glycerol product, which contains about 80% glycerol by mass. Although waste oils may have slightly higher glycerin yields, this has not been included in the analysis. The value of the purified glycerol is estimated to be $0.05/lb, based on current market price data from Nilles (2006) and personal communication with Leland Tong (National Biodiesel Board contributor). Note that this is a variable market and it is difficult to predict what the product value will be in the future, particularly as more biodiesel manufacturers come on-line and glycerin becomes even more plentiful.
- Annual water consumption is estimated to be 0.25 lb per gallon biodiesel produced, based on data from Haas et al. (2006).
- Capital cost for base catalyzed transesterification is based on data for a 10 MGY facility from Haas et al. (2006). The acid catalyzed facility cost is based on equipment cost from Zhang et al. (2003b) for a 2.4 MGY facility, with added cost for storage facilities from Haas et al. (2006). Additionally, the multiplier from Haas et al. (2006) was used to calculate the total capital investment for this facility. Capital cost is scaled using a scaling factor of 0.6, based on data from Bender (1999).

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[80] Sources: Haas et al. 2006, Sheehan et al. 1998, Zhang et al. 2003a, Canakci and van Gerpen 2001.

- In general, the capital costs of acid catalyzed transesterification are estimated to be higher than base catalyzed transesterification because the process uses a greater quantity of methanol, which requires more and larger transesterification reactors and a larger distillation column (Zhang et al. 2003b). Additionally, the acid catalyzed process is much more corrosive and requires expensive stainless steel process equipment. The increased methanol requirement and feedstock pretreatment also leads to increased O&M costs for the acid catalyzed process.

- Labor costs are calculated based on data for a virgin oil facility from Haas et al. (2006), but the operator salary was increased to \$20/hr from \$12.5/hr based on the authors' experience. Values in Zhang et al. (2003b) suggest that acid catalyzed facilities have higher labor requirements, so the authors' estimate was that there would be one additional operator per shift for a 10 MGY facility.[81] The total labor cost is scaled by facility size, with a scaling exponent based on data available for similar types of facilities.

- Annual non-labor fixed O&M costs are calculated as a percentage (2.1%) of the capital cost, following Haas et al. (2006).

- Annual Variable O&M costs include consumables (methanol, sodium methoxide, hydrochloric acid, sodium hydroxide, and water) and utilities (natural gas, WWT, waste disposal, and electricity). The annual costs for utilities are calculated on a \$/gallon basis of biodiesel produced, based on data from Haas et al. (2006) for the base catalyzed process and Zhang et al. (2003b) for the acid catalyzed process.[82] The annual cost of consumables for both processes is based on data from Haas et al. (2006). However, the quantity and price of the methanol and sulfuric acid for the acid catalyzed process are substituted using data from Zhang et al. (2003b).

- The levelized non-feedstock production costs for biodiesel are calculated based on the annual operating expenses and capital cost payment, co-product credit, and quantity of fuel produced.

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[81] For comparison the base catalyzed facility had two operators per shift.

[82] Electricity prices have not been updated to reflect current market value as there was not sufficient detail in the published studies to perform the calculations.

Table H-16. Biodiesel Yields for Various Feedstocks.

Feedstock (1)	Yield range (gal/ton)	Average Yield (gal/ton)	*Sources*
Virgin Oil (2)	246.4-270.0	258.2	*Haas et. al 2006, Sheehan et al. 1998, Zhang et al. 2003a, Canakci and van Gerpen 2001*
Animal Fats	266.3	266.3	*Bender 1999*
Yellow Grease (3)	234.9 - 263.2	249.1	*Zhang et al. 2003a, Canakci and van Gerpen 2001*

1) All data based on feedstocks with negligible water content.

2) Based on data for degummed soybean oil. High end of the yield range from Haas et al 2006, which assumes a negligible free fatty acid content in the feedstock.

3) Conversion yield for waste greases based on acid catalyzed reaction. For comparison, the transesterification reaction for virgin feedstocks use base catalysts.

Model Results:

Some of the key cost and performance results for current FAME production processes are shown in Table H-17 and Table H-18. These tables show results for the range of applicable facility sizes based on selected input quantities, and illustrate the high and low end costs. In general, the analysis shows that the levelized non-feedstock production cost of biodiesel for the base catalyzed transesterification process is significantly lower than the acid catalyzed process, as expected. However, the feedstock cost will be much more expensive for the virgin oil transesterification facility (typically waste grease is about half as expensive as virgin oil).

Table H-17. Example Cost and Performance for Base Catalyzed Transesterification Facility.

Sample Model Results - FAME Biodiesel w/ Virgin Oil		
Feedstock Input (ton/yr)	5,000	310,000
Biodiesel Yield (MGY)	1.3	80.0
Crude Glycerin Yield (ton/yr)	516	32,020
Water Consumption (1000 gal/yr)	39	2,400
Capital Cost (Million $)	$ 3.9	$ 46.6
Annual O&M Costs (Million $/yr)	$ 0.59	$ 23.7
By-Product Credit (Million $/yr)	$ 0.08	$ 5.1
Non-feedstock Production Cost ($/gal)	***$ 0.78***	***$ 0.31***

Table H-18. Example Cost and Performance for Acid Catalyzed Transesterification Facility.[83]

Sample Model Results - FAME Biodiesel w/ Waste Grease				
Feedstock Input (ton/yr)		5,000		310,000
Biodiesel Yield (MGY)		1.25		77
Crude Glycerin Yield (ton/yr)		498		30,880
Water Consumption (1000 gal/yr)		37		2,310
Capital Cost (Million $)	$	6.12	$	72.9
Annual O&M Costs (Million $/yr)	$	1.16	$	54.3
By-Product Credit (Million $/yr)	$	0.0797	$	4.94
Non-feedstock Production Cost ($/gal)	***$***	***1.47***	***$***	***0.76***

3.2.3 Outlook

According to USDA AMS (2009), the current cost for crude soybean oil is about 36.24 cents per pound ($2.53/gallon), and the cost for biodiesel is around $3.08/gallon FOB production point. This implies a non-feedstock production cost of $0.55 for other chemical inputs and processing costs. This figure is similar to the projections from the model for small base catalyzed transesterification facilities. The DOE Clean Cities Alternative Fuels Price reports estimated the January 2009 retail price of B100 at $3.47/gallon. Thus, roughly 40¢/gallon can be attributed to distribution and retailers' profit. Haas et al. (2006) estimate a biodiesel production cost of $2.00/gallon for a 10 MGY facility using crude soybean oil at $1.81 per gallon (23.5¢/lb). For a similar size facility using the same feedstock price, the FAME biodiesel model gives a production cost of $2.20/gal.

The Food and Agricultural Policy Research Institute (FAPRI) 2007 U.S. and World Agricultural Outlook, projects 2015 soybean oil prices will be about $681.40/ton (34.1¢/lb). Including $10/ton for delivery, this gives production costs of $2.98 - $3.36 per gallon for biodiesel.[84] Note that this does not include the cost of distribution, which would be included in the retail fuel price. Refer to Appendix H-E for details on the conversion of oilseed crops to oil.

3.3 TECHNOLOGIES READY FOR COMMERCIALIZATION IN THE MIDTERM (2015 – 2025)

There are a number of advanced biofuel technologies being developed to produce diesel substitutes, which may be available in the 2010 to 2025 time frame. In many ways the variety of approaches mirrors the work on gasoline substitutes, except for research developing algal biomass as an industrially grown source of feedstocks for biodiesel production. Algae, as with the lignocellulosic feedstocks, can be an important industrially grown energy source. In contrast with the seed oils, it is more difficult to convert the bio-oil in algae to biodiesel as a commercial scale process for removing the oil from algae has not been optimized. Once removed from the cell matter, this oil can be as easily converted to biodiesel as any conventional vegetable oil.

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[83] Please note that the large case presented here likely exceeds the amount of waste grease feedstock that would be aggregated in a single place. The smaller scenario is likely more representative of future facilities. Please see the feedstock assessment portions of the main body of this report.

[84] Delivery cost estimated for truck transportation, using $0.20/ton-mile and a 50 mile delivery radius. Including delivery, the soybean oil price is $2.66/gallon.

Cost and performance models for several of the advanced technologies have been developed. Since the advanced technologies do not currently have commercial plants in operation, these models are based on detailed engineering analysis for commercial scale plants (typically the projected cost and performance projections for the nth plant) available in the literature. The projected costs are typically accurate in the range of +/- 25% to 30%. All cost and performance models for the advanced technologies include projections for the 2020 ("mid-term") time period. Models of other time periods were also included for technologies that had additional projections available in published studies.

3.4 FATTY ACIDS TO RENEWABLE DIESEL: EXTRACTION AND HYDROTREATMENT

Green diesel and renewable diesel are some of the terms used to describe a hydrocarbon fuel produced from biomass oil feedstocks, to differentiate it from biodiesel generated via transesterification. These technologies will be termed here "Fatty Acids to hydrocarbon renewable Diesel" (FADe). Although renewable diesel is produced from vegetable oils and waste greases, unlike fatty acid methyl ester (FAME) biodiesel, this product does not have oxygen in its molecular structure and can thus be considered a true hydrocarbon (Antares 2008).[85]

Renewable diesel presents a number of benefits that may lure biofuels developers into its commercial production. It can utilize oil feedstocks with higher amounts of free fatty acids than the FAME transesterification process. Renewable diesel, like FAME, has superior properties (including low particulate emissions) to petro-diesel, but renewable diesel is chemically more compatible (e.g. more miscible) than FAME biodiesel with petroleum-based fuels, as it is a hydrocarbon fuel (UOP 2005). For example, green diesel typically has less than 10 ppm sulfur, has cetane numbers as high as 100, and has cloud point temperatures as low as -30° C (which minimizes cold weather gel issues). And in contrast to FAME production, there is no glycerin co-product produced, which could become a waste disposal issue for that industry. While the body of literature on emissions from biodiesel is far larger than that on renewable diesel, the work that does exist hints at a better emissions profile than for both conventional petroleum based diesel and FAME biodiesel. Particulates, hydrocarbons, and CO are all reduced (Climate Change 2006). Renewable diesel can be produced in petroleum refineries to take advantage of economies of scale and similar process requirements, which helps minimize the cost of production.[86] This has not yet been proven in practice, as will be discussed later in this section.

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[85] The text in this section was revised from material originally presented by Antares in 2008 in a *Strategic Assessment of Bioenergy Development.*

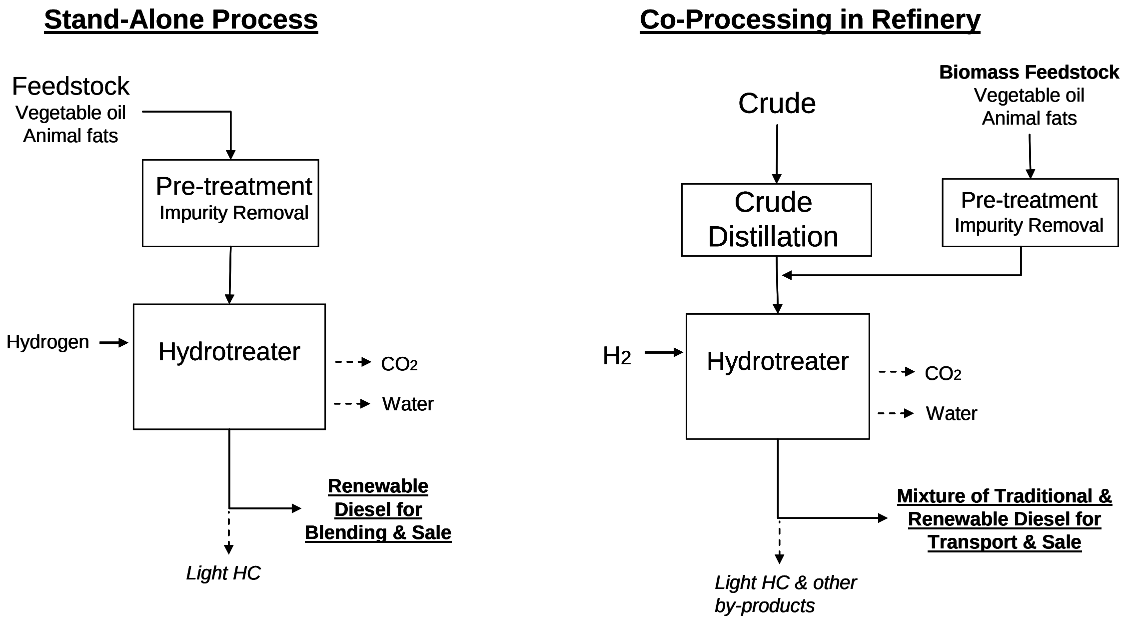
[86] This concept of integrating renewable diesel into existing refineries is hampered by the fact that, without expanding the size of the existing refinery, replacing a portion of the crude petro-diesel feedstock with crude renewable diesel only reduces the potential output of the refinery. Without economic incentives to produce renewable diesel, or high crude oil prices to disadvantage petroleum as a feedstock, existing refineries are unlikely to choose such a capacity reduction that will also require additional pretreatment equipment. As will be discussed later in this section, only a low-cost or "waste stream" feedstock like animal fats may currently have enough of a price advantage to encourage such a decision.

3.4.1 Technology Description

The Fatty Acids to hydrocarbon renewable diesel – Hydrotreatment (FADe - H) production process for renewable diesel in petroleum refineries is very similar to the process for upgrading pyrolysis oils described above, except that it does not typically involve hydrocracking. However, the Super Cetane Technology developed by the CANMET Energy Technology Center (CETC) at Natural Resources Canada does involve a hydrocracking step to tailor the molecular weight of some of the hydrocarbon compounds. They have reported cetane numbers exceeding 100 with their technology (Natural Resources Canada, n.d.). As with the hydrotreatment / hydrocracking of pyrolysis oil, upgrading biomass oil feedstocks can be done in a stand alone skid-mounted unit or pretreated and co-processed with petroleum feedstocks. Schematic diagrams of these configurations are shown in Figure H-16.

In the FADe - H process, vegetable oil or animal fat is upgraded using hydrogen gas, producing diesel and propane or other light hydrocarbons. Some water and CO_2 are also released in the process. A heterogeneous catalyst is used for hydrotreatment, which often may be less expensive than the homogeneous catalyst required for FAME biodiesel (UOP 2005).[87] The product distribution is typically 83-86% diesel and 2-5% light hydrocarbons by weight (in terms of incoming biomass) (UOP 2005). By comparison, the FAME process results in approximately 96% diesel production. This higher biodiesel yield from FAME is due to the retention of oxygen in the product. In hydrotreatment the oxygen is rejected in the form of CO_2 or H_2O.

Figure H-16. Schematic Diagram of Renewable Diesel Production in Petroleum Refinery.



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[87] The overall expense of the catalyst is not a function only of its per-gram purchase price, but also of the quantity required for a given reactor and of the rate of catalyst poisoning. Thus, the relative expense of heterogeneous versus homogeneous catalysts is a complex question that is very facility-dependent.

Oils and greases typically have an oxygen content of 11-12%, which is much lower than pyrolysis oils but still significantly higher than petroleum (UOP 2005). As with pyrolysis oil hydrotreatment / hydrocracking, the acidity of the oils generally means that they must be processed in stainless steel vessels. This is particularly true for waste greases which tend to have higher acidity than virgin feedstocks (UOP 2005).

UOP analyzed co-processing in a distillate hydrotreater for ultra-low sulfur diesel, and found that for brown grease feedstocks "standard refinery distillate hydrotreating units do not appear to be suitable for green diesel production in a co-processing scheme" (UOP 2005, p. 18). In addition to the potential effect on reactor and distillation column metallurgy requirements, co-processing can detrimentally affect catalyst lifetime. This is partly due to the fact that hydroprocessing conditions for standard hydrotreating for ultra-low sulfur diesel in refinery favors hydrodeoxygenation, which produces water from biomass feedstocks and can have a negative effect on catalyst activity. UOP (2005) further noted that although co-processing with brown grease had a positive effect on a catalyst in terms of sulfur, higher heat release causes increased deactivation rate, which lowers the catalyst lifetime.

Stand alone, skid-mounted processing of biomass oil and grease feedstocks seems to be an effective alternative, as it avoids some of the issues caused by co-processing but still has some benefits from refinery integration. In this configuration, biomass feedstocks are first pre-processed to remove contaminants, typically using some combination of the following: hydrocyclones,[88] dewatering, acid washing, ion exchange, desalting, and fixed guard bed catalyst.[89] The feedstocks can then be hydrotreated with a skid mounted system. A benefit of a skid-mounted system is the ability to optimize the reaction for biomass feedstocks, so that the hydrotreatment conditions favor decarboxylation ($DeCO_2$) instead of hydrodeoxygenation. The $DeCO_2$ process removes oxygen from the feedstock as CO_2 instead of water, which can be captured relatively easily. $DeCO_2$ also reduces hydrogen consumption, as it only requires 0.8 wt% H_2 whereas hydrodeoxygenation requires 3%.

Both skid-mounting and co-processing are potential treatment options in refineries. Despite the current processing issues, it is likely that with further technology developments and pretreatment methods, co-processing of waste greases could be an economically favorable process in the long-term. Furthermore, several of the companies developing renewable diesel refining use co-processing techniques with vegetable oils and animal fats, which suggests this process may be feasible for virgin feedstocks.

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[88] The term "hydrocyclone" refers to a cyclonic solids separation device that uses centrifugal force to remove particulates from a liquid stream.

[89] A guard bed of catalyst, often referred to as a "sacrificial bed," is nearly-spent or low-cost catalyst that is used to collect contaminants from a material stream before it flows over the main catalyst bed. In effect, it is low-value catalyst that is deliberately poisoned in order to prevent the active catalyst bed from being poisoned.

3.4.2 Performance and Cost Analysis

Outline for model analysis

The key values and assumptions used to calculate the technology analysis and economic performance model for stand-alone and co-processing FADe - H processes are described below. An example of the detailed analysis is shown in Appendix H-F.

- Feedstock input quantity is a variable.

- Applicable feedstocks are virgin oil, animal fats, and yellow grease. However, feedstock choice does not affect the product yield in this model. The typical heating values of the feedstocks are very similar, generally about 16,500 to 17,000 Btu/lb HHV (US EPA 2001). Although processing of yellow grease feedstocks may lead to higher capital costs as a result of additional pre-processing (such as desalting) and different metallurgy requirements due to higher acidity, the potential added cost has not been included as it is expected to be minimal (on the order of a few percent according to discussions with industry contacts).

- The applicable size range for the model is 15 to 200 MGY of renewable diesel, based on current and planned facility sizes for ConocoPhillips, Neste, and Petrobras. The smallest facility reported by these companies can produce around 1,000 barrels per day of renewable diesel, and the largest (planned) facility will be able to generate about 12,000 barrels per day.

- Stand-alone FADe renewable diesel yield is 86 wt% based on data from UOP (2005). Co-processing yield is 81 wt% based on data for Petrobras H-BIO system (Petrobras, n.d.). The renewable diesel yields are converted to volumetric quantities using a typical density of 6.5 lb/gal.

- The conversion efficiency of oil/fat to fuel (on an energy basis) is estimated to be 98% for skid-mounted and 96% for co-processing. This is calculated using an average LHV of oil feedstocks of 16,000 Btu/lb, and a typical renewable diesel (lower) heating value of 123,200 Btu/gal.

- Hydrogen consumption is calculated as a percentage of the feedstock input (by weight). The stand-alone system favors $DeCO_2$, which requires less hydrogen than the hydrodeoxygenation reaction favored in co-processing. According to UOP (2005), typical hydrogen consumption ranges from 1.5 - 3.8 wt% of feedstock input. The low end value is used for the stand-alone hydrotreatment in the model, while the high end value is used for co-processing.

- Light hydrocarbon production is estimated to be 3.5 wt% for stand alone process and 4.4 wt% for co-processing, based on input feedstock (UOP 2005; Petrobras n.d.)

- Water and CO_2 are also generated during hydrotreatment. The processing conditions for the stand alone process favor CO_2 production, while conditions for co-processing typically produce water. The total amount of CO_2

and water produced is calculated as the difference between the inputs and other outputs that give results similar to UOP (2005).[90]

- Water consumption data is not included for hydrotreatment processes, as it is assumed to be minimal.
- Near term capital costs are based on reported costs of facilities under development from Neste (stand alone) and Petrobras (co-processing), and a scaling factor of 0.6.[91]
- Mid term capital costs are decreased based on next generation facility development and a technology (i.e. cost per gallon capacity) improvement rate of 20% (McDonald and Schrattenholzer 2002). This rate is similar to that found for ethanol production facilities and retail petrol processing.
- Variable O&M costs include utilities and hydrogen. The utility cost is based on data for pyrolysis oil hydrotreatment from UOP (2005). Annual hydrogen cost is based on hydrogen consumption and cost data from UOP (2005).
- Fixed O&M costs include maintenance, insurance, and overhead. These costs are estimated to be 5.5% of near term capital cost, based on pyrolysis oil hydrotreatment O&M. No labor cost is added, as renewable diesel production is part of larger refinery operation and will not require additional operators.
- The propane/light hydrocarbon co-product value is calculated using the average 2005 wholesale price of \$0.93/gal (EIA AER 2005).[92] The estimated added cost for compressing propane is negligible and has not been included.[93]
- The levelized non-feedstock production costs for FADe - Hydrotreatment are calculated based on the annual operating expenses and capital cost payment, co-product credit, and quantity of fuel produced.

Model Results

Some of the key cost and performance results projected for the FADe - H processes are shown in Table H-19 and Table H-20. These tables show results for the range of applicable facility sizes based on selected input quantities, illustrating the high and low end costs. In general, the analysis shows that the non-feedstock production cost of FADe - H renewable diesel via co-processing in a refinery is cheaper than a stand-alone process, as expected. However, it seems that either method could be used to produce biofuel economically. The projected non-feedstock costs on a \$/gallon basis are competitive with FAME biodiesel, and the resulting fuel has the blending advantages over FAME that were noted above. Most importantly, FADe can also be readily integrated into existing infrastructure.

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[90] UOP (2005) estimates that the total water and CO_2 output is equivalent to 12-16 wt% of feedstock input.

[91] This scaling factor is equivalent to the value used for pyrolysis hydrotreatment, following UOP (2005).

[92] For comparison, the projected value of Liquefied Petroleum Gas is projected to be \$1.34/gal (\$633/ton) in 2015 (EIA AEO 2007).

[93] Antares estimates that compressing 17,000 tons per year of propane would cost about \$100K/yr, using an electricity price of 9¢/kWh.

Table H-19. Example Cost and Performance for FADe - Hydrotreatment Stand-Alone Process.

Sample Model Results - FADe Hydrotreatment (skid-mounted) - mid term		
Feedstock Input (ton/yr)	20,000	785,000
Renewable Diesel Yield (MGY)	5.1	200
Light Hydrocarbon Yield (ton/yr)	700	27,475
Capital Cost (Million $)	$ 30.4	$ 274.9
Annual O&M Costs (Million $/yr)	$ 2.5	$ 36.6
By-Product Credit (Million $/yr)	$ 0.4	$ 14.4
Non-feedstock Production Cost ($/gal)	***$ 1.17***	***$ 0.28***

Table H-20. Example Cost and Performance for FADe- Hydrotreatment Co-Processing.

Sample Model Results - FADe Hydrotreatment (co processing) - mid term		
Feedstock Input (ton/yr)	20,000	800,000
Renewable Diesel Yield (MGY)	5.0	200
Light Hydrocarbon Yield (ton/yr)	880	35,200
Capital Cost (Million $)	$ 4.8	$ 43.9
Annual O&M Costs (Million $/yr)	$ 1.5	$ 48.9
By-Product Credit (Million $/yr)	$ 0.5	$ 18.5
Non-feedstock Production Cost ($/gal)	***$ 0.32***	***$ 0.18***

3.4.3 Outlook

The projected 2015 price for crude soybean oil is about $681.40/ton (FAPRI 2007). Including an estimated delivery cost of $10/ton,[94] the feedstock cost for soy-based renewable diesel production is $2.71/gal for a skid-mounted process and $2.77/gal for co-processing. This gives a total production cost of $2.99 to $3.88 per gallon for stand-alone and $2.95 to $3.05 per gallon for co-processing.[95] These costs are very similar to the projected production costs for FAME biodiesel from virgin oil in 2015, which range from $2.98 to $3.36 per gallon for facilities from 1 to 80 MGY (see previous section regarding FAME for details).[96] For comparison, the national average wholesale price for diesel over the past year is $2.03 per gallon (EIA 2007). Also note that the current retail biodiesel price is about $3.20/gal (USDA AMS 2007).

ConocoPhillips has completed a commercial demonstration of FADe renewable diesel production at Whitegate refinery in Cork, Ireland, which co-processes vegetable oil with crude oil. The facility was able to produce 42,000 gallons per day of renewable diesel product (Green Car Congress 2006). ConocoPhillips has also partnered with Tyson to develop a facility to process animal fats in the US. This facility was planned for start up in late 2007, and

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[94] Based on a truck transportation cost of $0.20/ton-mile and a 50-mile delivery radius.

[95] The production cost does not include distribution costs or profit, which are both factored into retail fuel prices.

[96] This would indicate that FAME biodiesel will be competitive in the mid-term with renewable diesel. The key factors that may lead one fuel to succeed over the other would be: familiarity to motor manufacturers and consumers, who may not have dealt with renewable diesel, and the physical fuel characteristics noted above.

was to be able to produce up to 500,000 gallons per day of renewable diesel (ConocoPhillips 2007). The facility was able to begin producing renewable diesel, but the partnership took the plant off-line in May, 2009, citing the removal of federal subsidies as making the process uneconomical (Domestic Fuel 2009).

Neste also has developed a process to generate renewable diesel, using stand-alone hydrotreatment of vegetable oils and animal fats. Their NExBTL (NExt Biomass To Liquid) process was demonstrated at the Porvoo oil refinery in Finland, with a 60 MGY renewable diesel facility (Neste Oil n.d.). In late 2006, Neste Oil announced plans to build a second 60 MGY plant adjacent to the first plant to meet growing domestic demand. In addition, Neste Oil has another FADe renewable diesel production facility under development with OMV in Austria. Although the NExBTL process is stand-alone, it takes advantage of existing refinery infrastructure for energy, blending, logistics, and laboratories (Koskinen, Nurminen and Sourander 2005). For example, a refinery-based NExBTL plant has readily available access to hydrogen and utilities. Additionally, the renewable diesel can be blended, stored, and transported with petroleum based diesel, which improves process economics and logistics.[97]

The Petrobras H-BIO process uses co-processing of vegetable oils to generate FADe renewable diesel. Petrobras has tested a range of oils and process conditions, and the process has a conversion rate of 95% by volume or higher (Petrobras n.d.). Petrobras planned to have H-BIO process operations in at least three refineries by the end of 2007, with a total capacity to process more than 250,000 tons of vegetable oil annually. Further implementation in two more refineries was planned for 2008 (Petrobras n.d.). The company installed H-BIO facilities in four such refineries, but in 2009, in response to increasing vegetable oil prices, the company reduced renewable diesel production to a minimum (GreenMomentum 2009).

Other companies that are currently involved in FADe renewable diesel production include Nippon Oil in Japan, which expects to have a commercial process within three years, and BP, which in 2007 integrated a 80,000 gallon per day demonstration facility into its Bulwer refining facility in Australia (ConocoPhillips 2007).

One factor that may inhibit these commercialization efforts is the lack of an ASTM standard for renewable diesel blending. The D975 and D396 standards that govern petroleum diesel and heating oil have only been extended to include FAME biodiesel, and not yet FADe renewable diesel. FAME biodiesel has also been assigned an ASTM standard (ASTM D 6751), where FADe has not. Another possible drawback is the improved compatibility of FAME biodiesel over petro-diesel with particulate emissions controls (Williams 2006). Because FADe is chemically much more similar to petro-diesel than to FAME, it would likely demonstrate worse compatibility than FAME with such control technologies.

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[97] NExBTL diesel is compatible with existing fuel logistics and vehicles, and meets diesel fuel standards in all properties except density, which is about 780 kg/m^3 (Neste Oil, n.d). As such, the density value is the only blending constraint and limits blends of NExBTL product to 65% by volume (Koskinen, Nurminen and Sourander, 2005).

The chemical similarity of FADe diesel to petro-diesel also means that it is likely similar in toxicity. As petro-diesel is more toxic by ingestion and skin exposure than is FAME biodiesel, this could indicate that FAME is less acutely toxic than renewable diesel.

3.5 LIGNOCELLULOSICS TO MIDDLE DISTILLATES: FISCHER TROPSCH

Fischer Tropsch (FT) synthesis is a method of generating hydrocarbons from gaseous mixtures of CO and H_2 using a metal catalyst. This process has been known since the early 1900s, and the first commercial plant was built by Sasol in South Africa in 1955 (using coal syngas). FT synthesis produces a range of hydrocarbon products, and a large amount of heat from the highly exothermic reaction. The distribution of products generated from FT synthesis is affected by reaction temperature and pressure, input syngas composition, and catalyst type and composition. Among many other products, FT can be used to generate diesel and gasoline fuels. Although the FT process is relatively well characterized, there has not been a lot of analysis or testing with biomass feedstocks (Antares 2008).[98]

3.5.1 Technology Description

The general steps for the LignoCellulosics to Middle Distillates – Fischer Tropsch (LCMD – FT) process are shown in Figure H-17. As typical for gasification processes, the biomass must first be sized and dried for use (see detailed gasification description in the discussion of thermochemical synthesis of ethanol, above). After gasification, impurities such as particulates, sulfur, and tar must be removed from the raw syngas. The amount of gas clean-up required will depend on the type of process and economics. However, this is a very important step in the LCMD - G/FT process, as the FT catalysts (Iron (Fe) or Cobalt (Co)) are extremely sensitive to chemical poisoning by sulfur, as well as by halides and nitrogen.

After clean-up, the syngas is then sent to *either* a Low Temperature (200-240ºC) *or* a High Temperature (300-350ºC) FT reactor (Spath and Dayton 2003). Reactor pressures are generally 145-580 psi (Spath and Dayton, 2003). Low temperature synthesis typically uses Co catalysts, and is used to maximize the diesel fraction of the product, while high temperature synthesis uses Fe catalysts and maximizes gasoline production.

There are four types of reactors that have been used commercially, including:

- ***Multi-tubular fixed bed***, also known as the ARGE reactor. This has conversion efficiencies around 70%, and the products are about 50% wax.

- ***Circulating fluidized bed***, such as the original Synthol reactors that produce high fractions of gasoline and light olefins. These reactors are complex, and the circulating bed material causes considerable erosion which leads to high O&M costs. They have reached design limit for capacity at 7,500 bpd.

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[98] The text in this section was revised from material originally presented by Antares in 2008 in a *Strategic Assessment of Bioenergy Development.*

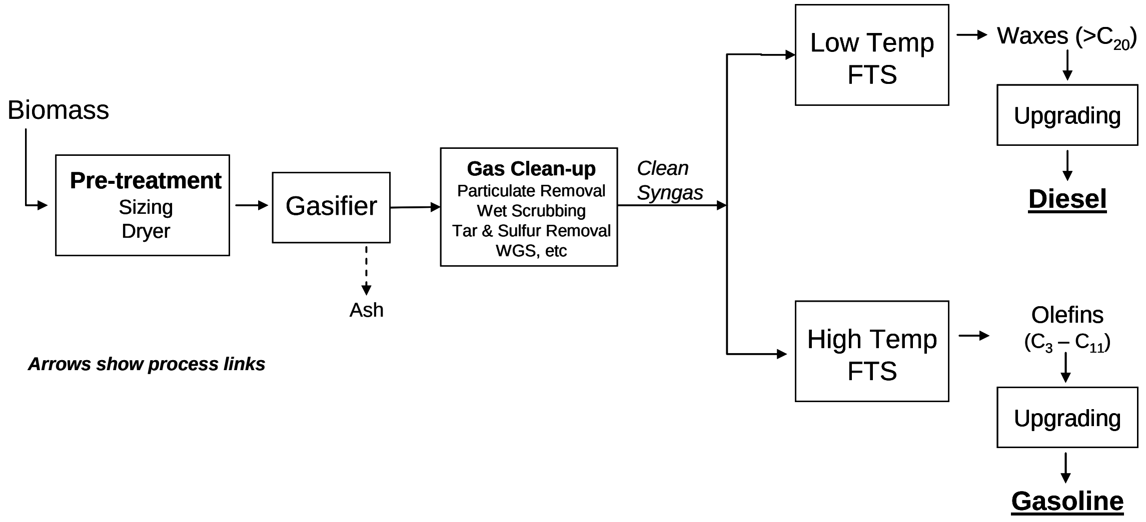
- ***Fixed fluidized bed***, or Sasol Advanced Synthol reactor, which has replaced the older Synthol reactor. It is half the size and cost for same capacity as the Synthol reactors, and has lower operating costs. Size limit is 20,000 bpd.

- ***Fixed slurry bed*** reactors, which are low temperature reactors optimized for wax production. They cost less than tubular reactors, but have only recently been used for commercial FTS. Separating the wax from the catalyst can be a challenge.

Temperature control is critical in the reactor, as it affects the product distribution and can harm the catalyst. In addition to the chemical poisoning mentioned above, catalysts lose activity due to sintering, carbon deposition, and oxidation. Furthermore, reactor temperatures above 400ºC encourage methane formation, an unwanted gaseous product (Spath and Dayton 2003).

The off-gas produced from the FT reactor can be diverted and recycled back through the reactor to generate additional hydrocarbon products, or it can be used to generate power/steam (not shown in the figure). Product liquids generated from FT synthesis are upgraded using a combination of hydrotreating, hydrocracking, hydroisomerization, and then separated.

Figure H-17. Schematic Diagram Showing General Process Steps for FTS.



Based on figure in (Spath and Dayton 2003)

Fischer Tropsch synthesis is characterized as a polymerization reaction with stepwise chain growth. The primary reaction that occurs during synthesis is given below.

$$\text{FT Synthesis Reaction: } CO + 2H_2 \rightarrow \text{--}CH_2\text{--} + H_2O \quad (8)$$

The $–CH_2$-- is a building block for hydrocarbon chains. A secondary reaction occurs when Fe catalysts are used, Water-Gas Shift (WGS).

WGS Reaction: $CO + H_2O \rightarrow H_2 + CO_2$ (9)

The combination of these reactions gives the net overall reaction for Fe catalyzed FT synthesis:

$2CO + H_2 \rightarrow --CH_2-- + CO_2$ (10)

The syngas entering the reactor needs to have a H_2/CO ratio of 2.15 for a cobalt catalyst, and 1.7 for an iron catalyst (lower because of the WGS reaction). The composition of the syngas can be altered by changing the gasification conditions, or by processing the syngas after clean-up via WGS, CO_2 removal, or methane reforming (which converts methane with steam into CO and H_2).

It is clear that FT catalysts are a very important part of the synthesis. Co catalysts tend to last longer than Fe, and they have better carbon conversion to products, since there is no major impact from the secondary WGS reaction. Additionally, Co does not require pressures as high as Fe, which equates to lower operating costs. However, Fe is attractive because it is much less expensive than Co (approximately 200 times cheaper).

The product selectivity from FT synthesis is based on the catalyst's ability to encourage chain growth versus chain termination. Regardless of operating conditions, the Fischer Tropsch reaction always produces a range of olefins, paraffins, and oxygenated compounds like alcohols, aldehydes, acids, and ketones. Only methane and paraffin wax can be produced with very high selectivity (Spath and Dayton 2003). The maximum gasoline product fraction selectivity is 48%, while for diesel the maximum is about 40% (Spath and Dayton 2003).

3.5.2 <u>Performance and Cost Analysis</u>

The LCMD - G/FT performance and cost model has been developed primarily based on recent work performed for US DOE and NREL (Antares 2003a; Antares 2003b; and Antares 2003c). In this analysis, process models, capital costs, and O&M costs were developed for several FT systems with varied size and technology types. One of the commercial scale systems modeled previously was selected for use as the basis for mid-term (2015) LCMD - G/FT technology in this analysis.

The defining characteristics of the configuration are the gasifier, FT reactor, and steam reforming system. The selected configuration uses an air-blown, indirect, atmospheric gasifier, represented by the BCL/FERCO gasifier (abbreviated as BCL). After cleaning and conditioning, the compressed syngas is processed in a steam methane reformer (SMR) to convert much of the hydrocarbons to carbon monoxide and hydrogen.[99] The syngas then undergoes a shift reaction to produce an H_2 to CO ratio of about 2:1. This configuration uses two (low temperature) fixed bed FT reactors with Co catalysts. Each reactor has a CO conversion of about 42%, so that the overall conversion is 66%. Two reactors are used in this process to increase the yield of FT-liquids, as recycling is not an effective option for air-blown gasifiers. The unconverted syngas is used for electricity and heat production in a combined cycle power plant. The FT-liquids produced in the reactor are hydrocracked and undergo primary distillation before they are sent to an off-site refinery. The final products are distillate and naphtha. A diagram showing the major components and process flows for this system is shown in Figure H-18.

In general, the total capital investment for cases involving SMR is greater than that for the cases involving an alternative reforming process, autothermal reforming (ATR). However, it is important to realize that the reformer choice affects O&M costs. Relative to ATR, the use of SMR increases FT liquids production and boosts exportable power revenues, thus resulting in lower total O&M costs on a $/bbl basis.

The overall process efficiency of this system including electricity and steam production is 58% on LHV basis (Antares 2003a). For this configuration of the LCMD - G/FT process, about 73% of the energy produced is embodied in the FT liquids (C5 and larger hydrocarbons), and the remainder is exportable power.[100]

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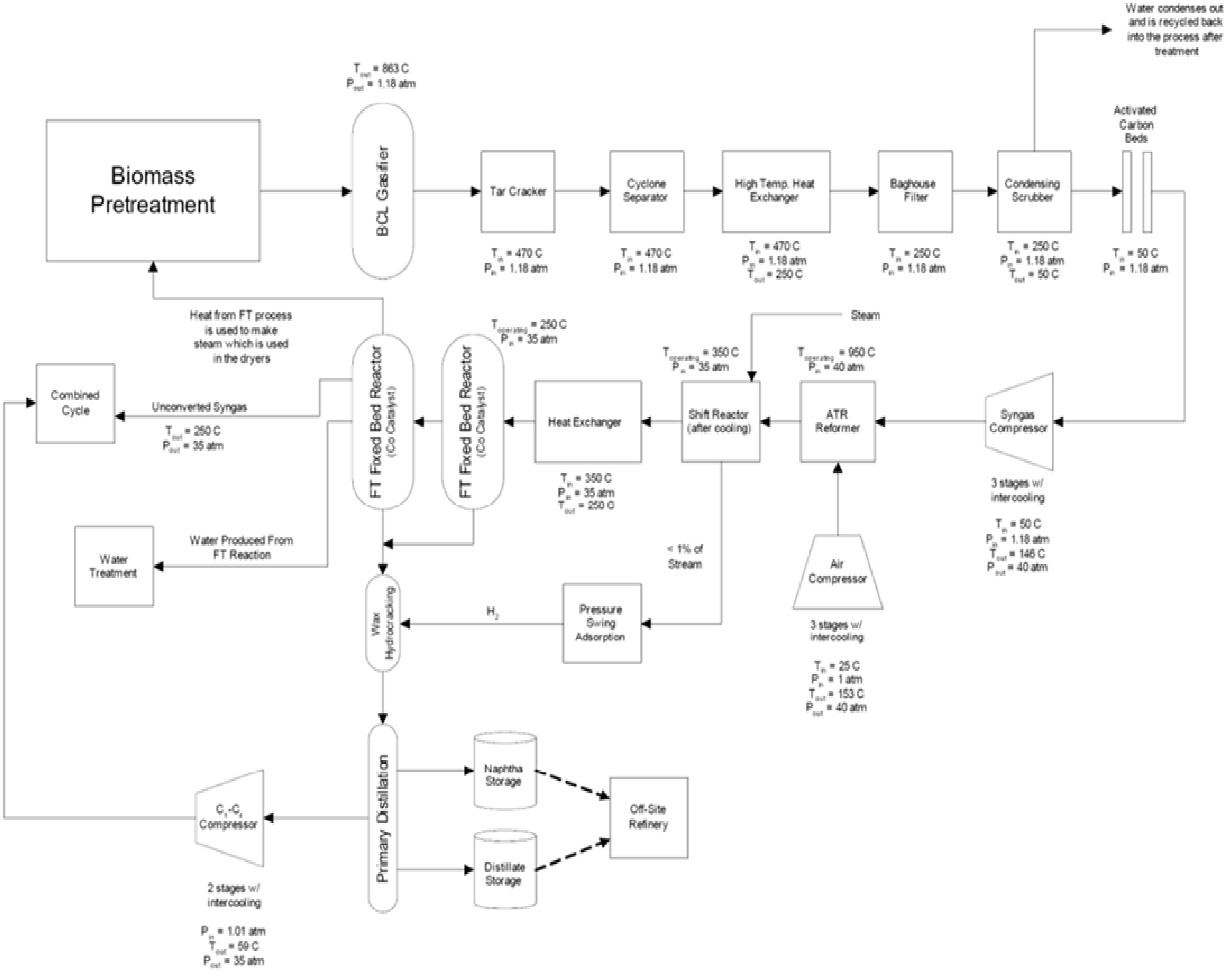
[99] Note that although SMR has higher capital costs than autothermal reforming (ATR), it is also more efficient. ATR is currently the only practical reformer to use for biomass-derived syngas, but it is expected that STR reforming could be available for use by the mid-term period. An optimistic view of SMR is used in the model.

[100] BIG-FT stands for Biomass Integrated Gasification – Fischer Tropsch.

Figure H-18. Basis for Mid-Term LCMD - G/FT Technology Characterization.



Outline for model analysis

The key values and assumptions used to calculate the advanced technology analysis and economic performance model for Fischer Tropsch synthesis are described below. An example of the detailed analysis is shown in Appendix H-F.

- Feedstock type and input quantity are variables. Applicable feedstocks include woody and agricultural residues.

- This model is applicable for facility sizes with greater than 185,000 dry ton biomass input per year, based on Tijmensen, Faaij and van Hardeveld (2002).

- Yield of FT liquids are calculated based on the LHV conversion efficiency. As different feedstocks have distinct heating values, the feedstock choice affects the amount of FT liquids produced. The conversion efficiency for distillates and naphtha has been calculated in Antares (2003a). For the selected system, LHV efficiencies for the products are 30.4% for distillates and 12% for naphtha.[101] This leads to a total energy conversion efficiency of 42% for the process.

- The FT liquids produced from this configuration have estimated heating values (LHV) of 125,684 Btu/gal for the distillates and 115,263 Btu/gal for the naphtha.

- The amount of electricity generated is calculated to be about 560 kWh per barrel of FT liquids produced, following Antares (2003a).

- Annual water consumption is equivalent to about 33,500 thousand gallons for a facility with a feedstock input of 700,000 dry ton per year, based on data from Antares (2003a).

- Capital cost data is based on values for a commercial facility with a biomass input of 80 dry ton per hour (about 700,000 ton/yr) from Antares (2003c). The scaling factors are 0.74 for a biomass input from 185,000 – 750,000 dry ton/yr, and 0.91 for a biomass input greater than 750,000 dry ton/yr, following Tijmensen, Faaij and van Hardeveld (2002).[102]

- Annual non-labor fixed O&M costs are calculated as a percentage (2.3%) of the capital cost, following Antares (2003a). Maintenance and repair accounts for 2% of the cost based on capital investment, and the operating supplies are estimated to be 15% of the total maintenance cost.

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[101] These yields were calculated using the Anderson-Schulz-Flory (ASF) theory to model the FT product distribution (Antares 2003a). For the cobalt catalyst the chain growth probability is assumed to be 0.93, which corresponds to a C5+ selectivity of 94.9%.

[102] "When scale is between 100 and 400 MWth the overall scaling factor for the entire plant (with respect to overall total investment costs) is approximately 0.74. When capacities go beyond 400 MWth, the average scaling factor increases to 0.91" (Tijmensen, Faaij and van Hardeveld 2002, p. 143)

- Labor costs are calculated based on the number of people required to operate the facility, with an assumed current labor rate of $82,000 per year per employee (including fringe and benefits). The base labor requirement for a facility with 700,000 ton/yr biomass input is $5,660,000/yr. This cost is scaled by facility size using a scaling factor of 0.25, as in Tijmensen, Faaij and van Hardeveld (2002).

- Annual Variable O&M costs include steam, water, wastewater treatment, ash and catalyst disposal, consumables, and FT catalysts. These are calculated on a $/gallon basis of the FT liquid products, following Antares (2003a).

- The co-product credit for electricity is based on the net electricity available for export and a current electricity value of $0.0797/kWh (based on the 2008 average wholesale price of electricity for industrial sector from EIA (2009)). The co-product credit for naphtha is estimated to be $76/bbl, based on average market data over 2008 from Plastemarte (2008).[103]

- The levelized non-feedstock production cost for the FT distillates are calculated based on the annual operating expenses and capital cost payment, co-product credit, and quantity of fuel produced.

Model Results

Some of the key cost and performance results projected for LCMD - G/FT are shown in Table H-21. This table shows results for the range of applicable facility sizes based on selected input quantities, and illustrates the high and low end costs. The low end feedstock input shows the minimum applicable facility size, and the high end was selected to show a reasonable range of sizes and costs. In general, the analysis shows that the non-feedstock production cost of distillates from FT synthesis is fairly expensive. However, the co-product credit (for naphtha—a co-distillate—and electricity generated from off-gas) is a critical factor and can have a significant effect on the production costs.

Table H-21. Example Cost and Performance for Fischer Tropsch Facility.

Sample Model Results - Fischer Tropsch Synthesis w/ Poplar		
Feedstock Input (dry ton/yr)	185,000	1,500,000
FT Distillate Yield (MGY)	7.5	61
Naphtha Yield (MGY)	3.2	26
Electricity Production (MWh/yr)	142,940	1,158,980
Water Consumption (1000 gal/yr)	8,860	71,800
Capital Cost (Million $)	$ 296.5	$ 1,604.0
Annual O&M Costs (Million $/yr)	$ 11.7	$ 68.4
By-Product Credit (Million $/yr)	$ 17.2	$ 139.6
Non-feedstock Production Cost ($/gal)	***$ 4.12***	***$ 2.07***

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[103] The market value of naphtha averaged $720 per tonne in 2008 (Plastemarte 2008).

Tijmensen, Faaij and van Hardeveld (2002) project the cost of FT-liquids is about \$3.18/gal for an 80 tph facility (about 700,000 dry ton per year).[104] The production of FT-distillates from the model for a similar sized facility using a feedstock cost of \$44.66/dry ton for poplar is \$3.65/gal. Although this result is slightly higher than the cost suggested by Tijmensen et al. (2002), these results cannot be directly compared as the FT-liquids from the model are separated into components. In the conversion technology model, the naphtha is sold as a lower value by-product, at a value of about \$1.77/gallon.

3.5.3 Outlook

There has been a lot of research and development of FT synthesis technologies. Sasol has several commercial FT plants in South Africa that use coal and crude oil feedstock. Sasol's first FT plant was built in 1955 and because of its success, two more were built in the 1980s. These facilities are currently using Sasol Advanced Synthol reactors, which have much lower O&M costs than the original Synthol reactors.

Two other plants became operational in the 1990s, both of which use natural gas feedstock. The Mossgas plant is a Fischer Tropsch facility in South Africa that produces 1.1 million tons/yr of FT products (Spath and Dayton 2003). Shell commissioned a plant in Malaysia in 1993 which produces 0.6 million tons per year of FT products.

3.6 BLACK LIQUOR TO BIOFUELS: GASIFICATION AND THERMOCATALYSIS

3.6.1 Technology Description

Pulp mills may already be classified as biorefineries. However, enhanced pre-processing and treatment of byproducts may allow these facilities to expand their product slate. By adding either finished biofuels or important intermediaries, these established mills can increase the value of the incoming feedstock and leverage substantial processing and handling infrastructure.

Paper can be made in different ways: from completely fresh woody material, from recycled paper material, or a combination thereof. Since the second path is purely reprocessing of material, the following discussion will concentrate on the production of paper from fresh woody material. This material comes to the paper mill in the form of whole logs. Both bark and wood consist of three basic long-chain compounds: hemicellulose, lignin and cellulose. Of these three components, it is the cellulose that forms the largest portion of paper. The paper making process can be properly viewed as a pathway for stripping cellulose material from lower-value hemicelluloses and lignin, while using these byproduct materials in ways to extract from their maximum value.

Traditionally, paper mills fall into one of two major types: Kraft mills (over 80% of the paper mills in the U.S.) or mechanical mills (EPA 1990). A Kraft mill applies "white liquor," a hot solution of concentrated sodium hydroxide and sodium sulfide, to wood that has been stripped of bark and chipped. This chemical digestion breaks down hemicellulose and lignin while leaving cellulose fibers largely intact. The process typically lasts one to three hours.

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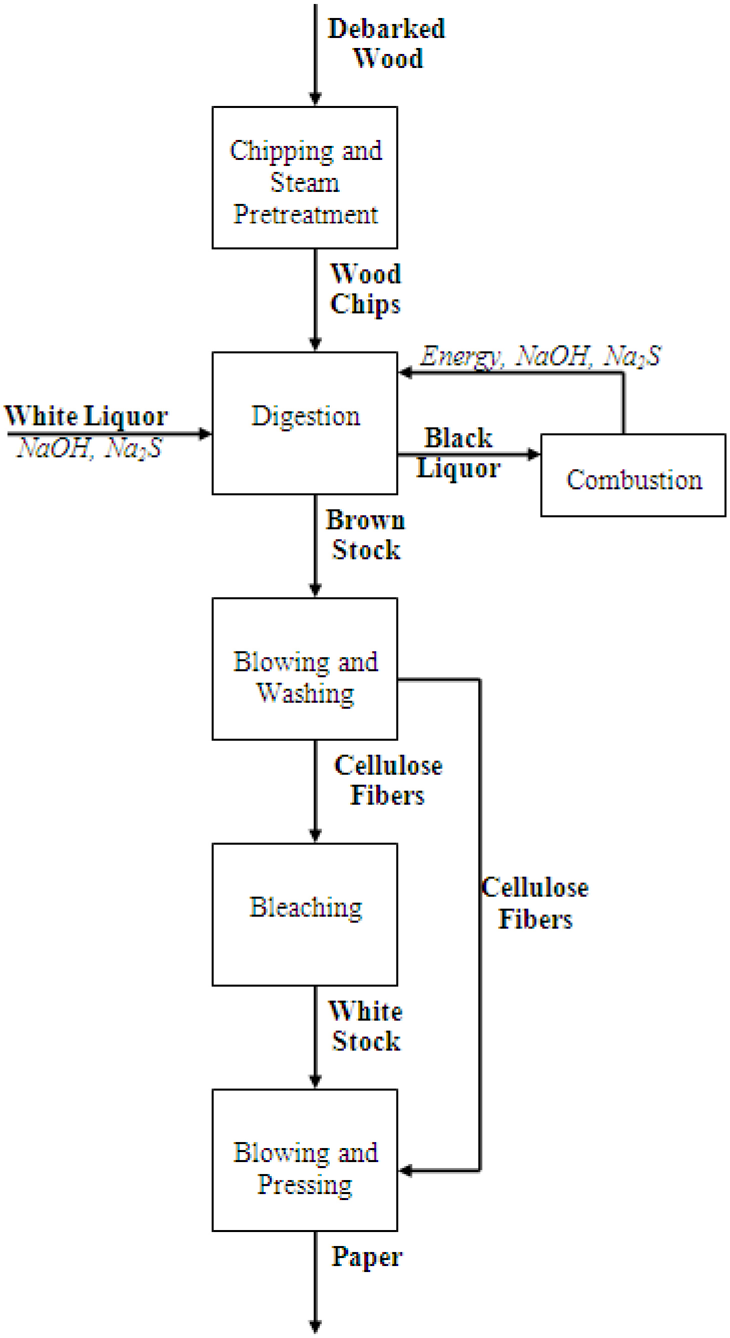
[104] Production cost converted to 2008\$ using CEPCI.

Mechanical liquids/solids separation with screens or sweeps is then used to remove “black liquor,” the depolymerized hemicellulose and lignin dissolved in a caustic solution of sodium sulfate and sodium carbonate. This is concentrated through evaporation and then burned to provide heat to the energy-intensive digesters. Regenerated sodium hydroxide and sodium sulfide are recycled to the digesters. (Amidon 2008)

The digested wood chips, now “brown stock,” are forced with high pressure into a tank of water, where the cellulose fibers expand and are rinsed of contaminant black liquor. This “brown stock’ is the base material for cardboard-like products. For the production of pure white paper a bleaching step will follow. Bleaching can reduce stock mass by as much as 5% and reduce cellulose fiber strength by reducing the amount of lignin bound to cellulose (PTF 1995). Either white or brown stock is made into paper and cardboard respectively by blown extrusion onto wire mesh and subsequent drying and pressing.

Mechanical mills, which do not attempt to remove the lignin from the cellulose fraction, process the wood fibers by grinding wood chips with refiner plates, usually after a steam treatment step to expand the fibers and make them less resistant to tearing. This is extremely energy intensive, and produces more physical cuts in the cellulose fibers, which decrease the strength of the pulp when used in paper. Modern mills consume about 2600 kWh per ton of pulp product (Paprican 2008).

Figure H-19. Conventional Kraft Mill Process Design.



Although several types of biorefining have been proposed, each one exploits the same idea: that hemicellulose and lignin are worth more as co-products than for their simple heating value. Hemicellulose comprises between 19 and 32% of the mass of de-barked wood chips, lignin 22-31%, and cellulose 40-45% (Ragauskas 2005). This means that less than half the input mass can be converted into pulp products—nearly as much is currently burned for its fuel value. There are a large variety of technology routes to better utilize hemicelluloses and lignin, but one representative technology that has been considered in detail is highlighted below.

A technology route examined by Larson et al. (2007) built upon prior work based on black liquor gasification.[105] In their prior work, black liquor gasification was examined as a commercial route to syngas.[106] The product gas would then be used to power a gas turbine to provide heat and power to a pulp integrated pulp and paper mill (black liquor gasification combined cycle or BLGCC). Considerable thermodynamic and economic modeling was completed.

Leveraging that work, Larson et al. (2007) have proposed that the syngas generated from the black liquor gasification could instead be used to create liquid fuels such as compressed dimethyl ether (which is a gas at standard conditions) or FT liquids. A simplified process diagram of possible configuration is provided below. It is worth noting that the proposed arrangements still employ a combined cycle heat/power arrangement. In the cycle, black liquor (for liquid fuels) and additional outside biomass (for heat and power) are utilized. For reference, the overall cycle is designated either Black Liquor to Middle Distillates via Gasification and Fisher Tropsch Synthesis or Black Liquor to Dimethyl Ether (DME) via Gasification and Direct DME Synthesis.

Again, based on prior work, Larson et al. selected a Chemrec gasifier for the conversion black liquor conversion system. The Chemrec gasifier is modeled to operate at 950-1000º C and high pressure converts the black liquor into a 50:50 gas/condensed-phase mixture. Sodium migrates to the condensed phase and can be recycled, while the partial oxidation of lignin and hemicellulose enriches the gas phase in carbon monoxide and hydrogen. This syngas can then be used either in a Fischer-Tropsch reaction to produce long-chain liquid hydrocarbons like diesel fuel, in a liquid-phase (i.e. high pressure) DME reactor, or through a reactor for mixed alcohol synthesis. In addition to different capital and operating costs, the systems vary with respect to syngas recycle rates, the amount of non-black liquor biomass feedstock needed, and clearly the product fuels.

As noted above, other biomass (such as bark, corn stover or wheat straw) can be added to a separate gasifier dedicated to heat and power. In this way, the overall energy balance of the plant can be maintained while maximizing the added value generated by the BLG/MF process.

A schematic of the BLG process is provided in Figure H-20 and a representative process schematic for the entire DME cycle is provided in Figure H-21.

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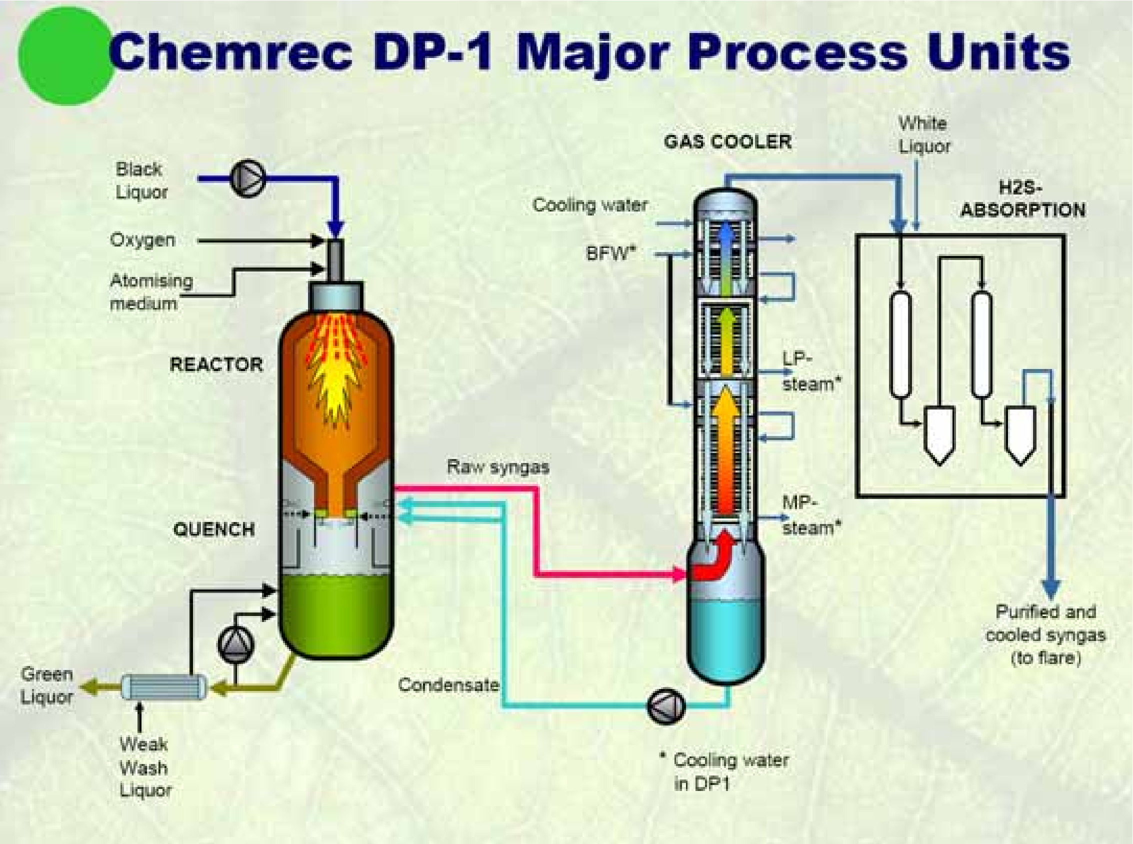
[105] It is important to note that the technology proposed by Larson et al (2007) is a particularly advanced form of black liquor gasification, with sophisticated heat integration and combined cycle technology. As such, it may represent a more energy efficient, but more capital-intensive and possibly less cost-effective form of black liquor gasification. The Larson material is examined preferentially here because of the detailed design and simulation that it presents, and the thorough nature of its analysis. Less sophisticated designs do not appear to have been considered with the same degree of detail.

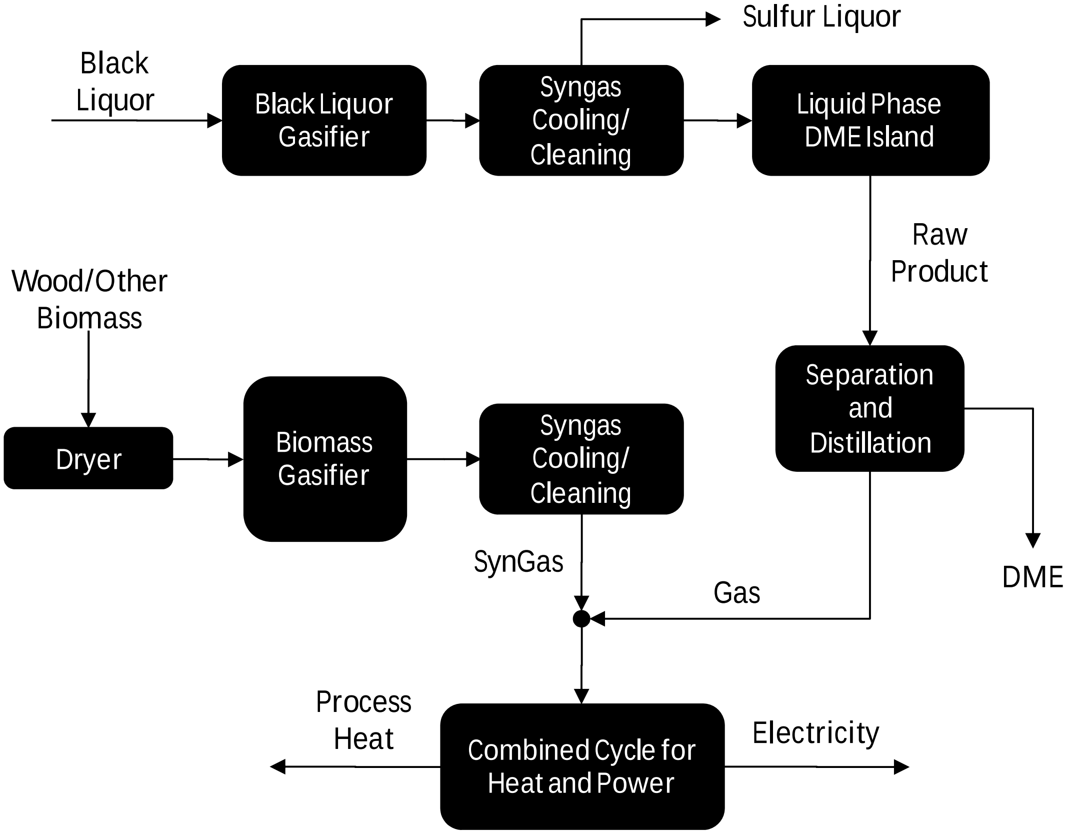
[106] This was not a novel idea. At this time, two black liquor gasification demonstration plants were under construction or in operation. These will be discussed in the "Cost and Performance" subsection later in this section.



Chemrec DP-1 Major Process Units

Figure H-20. Diagram of Sample Chemrec Gasification System (Landaly 2006).

Figure H-21. BLGCC / DME Process Proposed by Larson et al (2007).



3.6.2 Performance and Cost Analysis

Larson et al. (2007) estimated yield and economic data for a number of processes, all applied to the same type of feedstock, with the same annual wood input. The results, in terms of gallons of gasoline equivalent per dry ton of biomass input are provided below in Table H-22. The biomass input is comprised of all the wood entering the mill, including hog fuel and in some cases forest residuals purchased solely for gasification fuel.

Table H-22. Biofuel Yields by Plant and Fuel Type (Larson et al 2006).

Plant Configuration (1)	Average Yield (gal/ton wood feedstock) (2)	Average Yield (gal/ton pulp coproduct)
DME (with syngas recycle, no gas turbine)	50.2	126.4
DME (with syngas recycle and gas turbine)	42.7	126.4
DME (with gas turbine, no syngas recycle)	22.3	55.8
FTL (medium gas turbine, syngas split to power, fuel feedstock)	19.8	51.6
FTL (large gas turbine, syngas split to power,fuel feedstock)	14.3	51.6
FTL (medium gas turbine, all syngas used as fuel feedstock)	40.6	158.6

1) Assumes 65:35 distribution of hardwood to softwood feedstock by mass. Calculated from Larson et al, 2006.

2) Calculated based on the total amount of feedstock input, including forest residuals purchased for fuel.

Estimated capital costs for these systems are shown below (Plant types are abbreviated for simplicity, but correspond to Table H-22 in their placement). These are overall capital costs associated with building a complete mill, not merely a biofuels addition to an existing facility. In comparison, Larson (2006) estimated in the same study that a new, traditional, Tomlinson-type mill with no biofuels production but approximately the same annual wood consumption would cost about $171 million in $2008. The capital costs for the biofuels-producing mills range from $4.57 per gpy capacity to $22.30 per gpy, depending on plant type. Due to the relatively advanced design of the Larson plants, these may be higher than for less sophisticated gasification systems.

Table H-23. Capital and non-Feedstock Operating Costs Based on Simulation (Larson et al 2006).[107]

Plant Configuration	Capacity (MMGPY)	Capital Investment (millions of $)	Capital Investment ($ per gpy)	Annual Operating Cost (millions of $)	Annual Operating Cost ($ per gallon)
Adjusted for $2008					
DMEa	69.29	$ 317	$ 8.81	$ 12	$ 0.18
DMEb	69.29	$ 523	$ 14.49	$ 20	$ 0.29
DMEc	30.61	$ 407	$ 21.76	$ 16	$ 0.51
FTa	28.3	$ 415	$ 24.00	$ 16	$ 0.56
FTb	28.3	$ 631	$ 36.49	$ 24	$ 0.86
FTc	86.93	$ 584	$ 11.00	$ 22	$ 0.26

Operating costs are also included in Table H-23. These vary far more dramatically between plant types than do capital investments, peaking at $0.86 ($2008) and dropping as low as $0.18. It is important to note that these are total operating costs associated with the entire pulp mill. The portion related specifically to biofuels production ranges from $3 million to $6 million less than this total cost.

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[107] "DMEa" refers to the first entry in Table H-22, "DMEb" the second, "DMEc" the third, etc.

Based on these data, Larson et al. (2007) estimated an internal rate of return of 14-18% for the installation of a black liquor refining facility at an existing mid-size pulp mill. This estimate assumes significant tax credits for production of all biofuel types, and assumes capital investment incentives that are not available in the U.S. Under current economic conditions and with present government incentives, this rate of return estimate would be significantly lower.

Net energy consumption reductions and carbon footprint reductions are likely to be small for BLGCC/MF facilities. Larson et al. (2006) estimated wheel-to-well net carbon dioxide reduction of between 0.5 and 1.5 million tons per year for use of Fischer Tropsch gasification rather than modern conventional pulping technology in a mid-size pulp and paper mill. Net power consumption reductions were small—only in one of eight cases studied could a biorefinery sell power back to the grid. (As before, this may be a function of the highly sophisticated plant configurations chosen, and may not reflect the results of a simpler gasification system).

This scenario is different when DME is produced instead of FT liquids, as DME production follows a production route via methanol. Methanol can be produced from both carbon monoxide and carbon dioxide. Such use of multiple carbon-containing components significantly increases the carbon efficiency of the system. However, it also increases the capital costs, as DME has similar volatility as propane and thus needs to be stored under pressure.

Outline for Model Analysis

The important values and assumptions used to construct the BLGCC/FT production model are described below. A detailed presentation of the model appears in Appendix H-F.

- Model inputs are:
    - Feedstock type (percentage of hardwood vs. softwood in the feed).
    - Yearly plant feedstock consumption (on a dry basis). This feedstock input includes fiber for pulp feedstock, hog fuel generated in debarking process and wood residuals purchased solely for boiler fuel.
- The facility was modeled as both a Greenfield construction and as a retrofit to an existing pulp mill. The Greenfield model assumes no land purchase. All retrofit costs are incremental to the costs of the existing mill.
- Due to the lack of any demonstration plants in operation, both yield and cost data is based on a detailed series of AspenPlus simulations completed by Larson et al. (2006). It examines two types of end products—DME and Fischer Tropsch Liquids (FTL)—each achieved in three different plant configurations.
    - DME (with no gas turbine, no syngas recycle)

    - DME (with syngas recycle and gas turbine)
    - DME (with gas turbine, no syngas recycle)
    - FTL (medium gas turbine, syngas split between use for power and as a feedstock)
    - FTL (large gas turbine, syngas split between use for power and as a fuel feedstock)
    - FTL (medium gas turbine, all syngas used as fuel feedstock)
- The two scenarios found to be most economically viable were the first and last of the preceding list, so the authors chose these as most representative of potentially commercializeable facilities.
- The biofuel product yields are stated to be undenatured and without additives. The FTL case assumes no on-site refining apart from removal of light gases. The cost of refining and transportation to an oil refinery are not included in this model.
- Pulp yield outputs assume 48.75 wt% (softwood) and 49.75 wt% (hardwood) digester yields, as per Larson et al. (2006).
- Capital cost was scaled on the basis of dry tons of total feedstock input into the mill using a scaling factor of 0.6 as established by Aden et al. (2002). All costs were inflated to 2008$ using the CEPCI index.
- Feedstocks were assumed to contain 50% moisture as received with slight differences in hardwood and softwood energy content. A hardwood HHV of 7817 BTU/lb and a softwood HHV of 8127 BTU/ lb were calculated from Yin (2006), for a 50% moisture content.
- Suggested feedstock costs are conservative estimates of feedstock cost, drawn from Larson et al. (2006).
- The average volumetric yields tabulated by plant type assume the same feed wood type distribution as in the study by Larson et al. (2006): 65 wt% hardwood and 35 wt% softwood. These will vary depending on the ratio of soft- to hardwoods, but were assumed for simplicity to be constant in this model regardless of wood feed composition.
- Total annual operating cost applies to the entire Kraft mill, not fuel production exclusively. This was estimated to be 4% of the total installed cost as per Larson et al. (2006) but was inflated from Larson's original 2005$ estimate with 2.67% annual inflation as per the authors' original study. The incremental operating cost compares this integrated facility to a modern Tomlinson boiler-driven Kraft mill of the same pulp capacity, with no biofuel production.

- Conversion efficiency is calculated based on the total amount of feedstock input, including forest residuals purchased for fuel. Total annual fuel energy production for each plant configuration was scaled from Larson et al. (2006).

- The allocation of costs between pulp production and biofuels production was handled as simply as possible. A biofuels-producing mill was compared to a conventional Tomlinson mill of the same yearly pulp output, and the annual O&M cost and capital charge differences between the two cases were considered to be the biofuels production cost. In a conventional biofuel product/feed co-product biorefinery case like corn ethanol plants, costs are often attributed on a mass basis because the product and co-product processes are not inherently incremental. Because there is some question as to whether the biofuel in the BLGMF case is a product or a co-product and its implementation is most likely as a retrofit to an existing plant, the authors considered the most informative presentation of biofuels production cost to be simply the incremental cost of biofuels production over conventional mill operations.

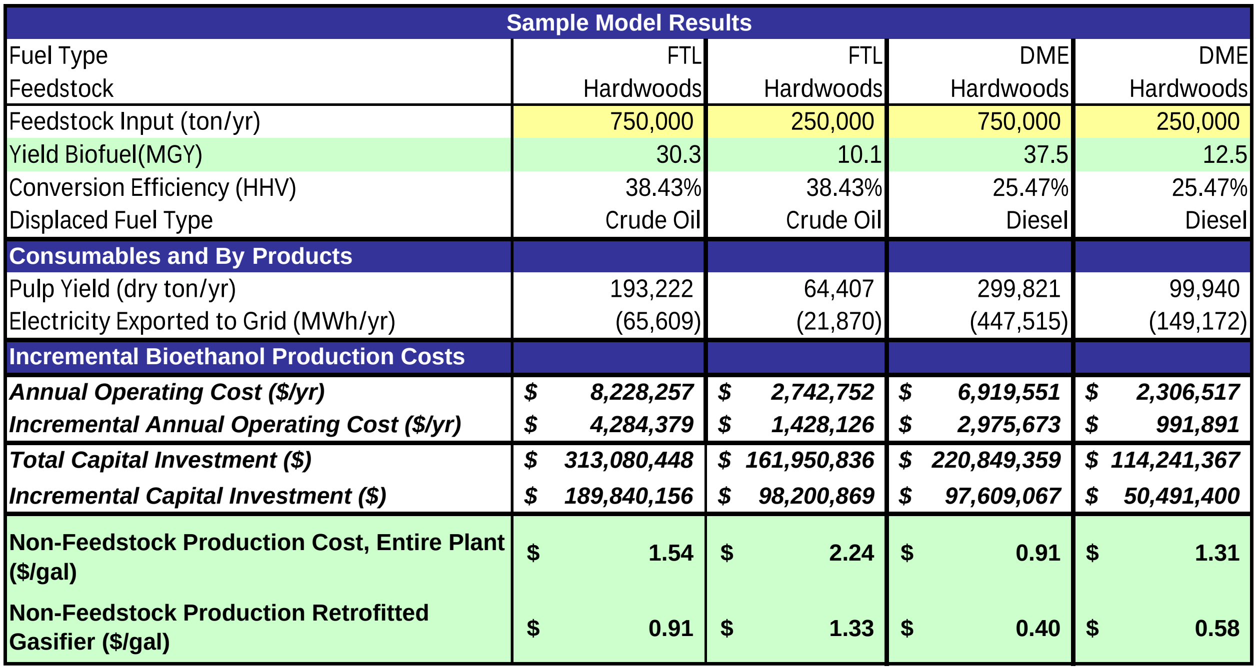
- The volumes of biofuels presented assume densities of 668 kg/m^3 and 780 kg/m^3 for DME and FTL respectively. (ConocoPhillips 2006; CRC Handbook 2006).

- Production efficiencies and yields are all assumed to remain constant regardless of scale. Similarly, parasitic utility demands are assumed to vary linearly with total feedstock input. As this is an important liability to the model, a relatively narrow plant size range has been chosen in order to minimize error.

Model Results

Sample results for the BLGCC/MF model for both main fuel types are noted below in Table H-24. While DME production seems to have far more favorable economics than Fischer-Tropsch liquids production, it is important to note that DME is a less thoroughly evaluated fuel with important storage and handling complications, as noted above. The two fuel types also serve different market segments, as DME is used as a diesel fuel replacement, and FTL requires refining before use as either gasoline or diesel substitutes.

Table H-24. Example Cost and Performance for Black Liquor Gasification Model.



Sample Model Results				
Fuel Type	FTL	FTL	DME	DME
Feedstock	Hardwoods	Hardwoods	Hardwoods	Hardwoods
Feedstock Input (ton/yr)	750,000	250,000	750,000	250,000
Yield Biofuel(MGY)	30.3	10.1	37.5	12.5
Conversion Efficiency (HHV)	38.43%	38.43%	25.47%	25.47%
Displaced Fuel Type	Crude Oil	Crude Oil	Diesel	Diesel
Consumables and By Products				
Pulp Yield (dry ton/yr)	193,222	64,407	299,821	99,940
Electricity Exported to Grid (MWh/yr)	(65,609)	(21,870)	(447,515)	(149,172)
Incremental Bioethanol Production Costs				
Annual Operating Cost ($/yr)	***$ 8,228,257***	***$ 2,742,752***	***$ 6,919,551***	***$ 2,306,517***
Incremental Annual Operating Cost ($/yr)	***$ 4,284,379***	***$ 1,428,126***	***$ 2,975,673***	***$ 991,891***
Total Capital Investment ($)	***$ 313,080,448***	***$ 161,950,836***	***$ 220,849,359***	***$ 114,241,367***
Incremental Capital Investment ($)	***$ 189,840,156***	***$ 98,200,869***	***$ 97,609,067***	***$ 50,491,400***
Non-Feedstock Production Cost, Entire Plant ($/gal)	**$ 1.54**	**$ 2.24**	**$ 0.91**	**$ 1.31**
Non-Feedstock Production Retrofitted Gasifier ($/gal)	**$ 0.91**	**$ 1.33**	**$ 0.40**	**$ 0.58**

3.6.3 Outlook

Although a greatly simplified block diagram has been presented to explain the fundamentals of this concept, a black liquor refinery facility would be a capital-intensive and complex addition to a mill. Clearly, pulp mills already employ extensive chemical processing to refine wood to their existing product slate, but black liquor refining would add completely new unit processes, substantial new heat exchange equipment and liquid fuel handling and storage. By all standards, existing sites would be dramatically transformed. This will undoubtedly present many early operational, maintenance and management challenges, and, as in the petroleum refining industry, there will be reluctance to add a new process that might adversely affect the primary product and/or plant availability.

The Flambeau Rivers BioFuels biorefinery under development in Park Falls, WI is one example of how integration of biorefining into a paper mill may occur. Flambeau Rivers was awarded a federal grant to co-locate a gasification-based liquid biofuels facility with a paper mill, using both black liquor and woody biomass as feedstocks and supplying heat to the adjacent paper mill (Reuters 2008). This parallel biorefinery will not directly affect mill processes, but will supply the mill with a valuable commodity—heat—while allowing the biofuels facility an energy-rich feedstock. This may be a low-risk way to introduce black liquor gasification that allows a decoupling of biofuels production from paper production.

Additionally, the pulp and paper industry, at least in the U.S. has been undergoing steady decline in recent years. In fact, according to a recent publication by Price Waterhouse Coopers (2008), nine of the largest public U.S.-based forest and paper companies posted earnings of $808 million in the third quarter of 2008. This figure was down $149 million from earnings in the same period of 2007 (Price Waterhouse Coopers 2008). This economic reality will mean that mills may be less and less capable of making the investment and taking the associated development risk of implementing these types of projects, especially as early adopters.

Additional challenges to a fast deployment include limited operating experience with an integrated concept. Although black liquor gasification has been demonstrated in a limited fashion (Chemrec's New Bern and DP-1 pilot plants, for example), only a single black liquor gasifier is currently running at a commercial scale. Norampac's Trenton, Ontario TPI gasifier, operating since October, 2006, supplies the paper mill's entire steam load, offsetting natural gas purchases (TRI, 2007). With only a single, small plant in operation, BLG cannot be considered a commercial technology. Likewise, experience for using BLG- or biomass-derived syngas for FT or DME is also limited. Although in 2006 UPM-Kymmene (Finland) announced plans to construct a commercial-scale, gasification-based FT liquids plant alongside a Finnish paper mill to process waste biomass, the company later canceled construction due to financial strain, and the plant was never built. It did later plan a pilot-scale production line, which is currently under construction.[108]

Despite these obstacles, biomass-derived DME has gained some momentum in Europe. *Biofuels Business* reported in September 2009 that Volvo and Ford Europe had performed engine testing on DME. In Volvo's case, this DME was derived from black liquor gasification and showed a high cetane number and favorable emissions. The same report noted that Chemrec had broken ground on a black liquor gasification demonstration plant to produce DME in Pitea, Sweden. It was expected to begin producing DME and methanol in 2010, and appears to be slated to be the first such facility in operation. (*Biofuels Business* 2009) Stora Enso and Neste Oil commissioned a 12 MW gasification demonstration system in June, 2009 at a mill in Varkaus, Finland. The companies plan to perform product testing and process optimization at the plant (Stora Enso 2009).

The use of DME, whether biomass- or petroleum-derived, as a fuel has garnered interest and research funding, particularly in Asia and developing countries. China, faced with growing demand for diesel substitutes and rapidly deteriorating air quality, announced construction of a one million ton per year plant that co-produces DME and methanol via coal gasification. Japan DME and NKK Corp. have also made and are planning significant investments in DME production facilities to supply DME as a diesel substitute. (Olah 2006) While these facilities use non-biomass feedstocks, the accompanying development of infrastructure for and motor testing of DME may lead to greater interest in biomass gasification for DME.

In principle, a biorefining addition to existing facilities should be less capital intensive than if built as a stand-alone biofuels facility (Larson 2007). Additionally, much of the chemistry involved is well-understood and presumably considerable operating expertise and experience can be brought to bear to overcoming challenges. Still, this technology is a long- rather than short-term option for commercial-scale biofuel production.

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[108] Personal communication with Benjamin Thorp, 8/10/09.

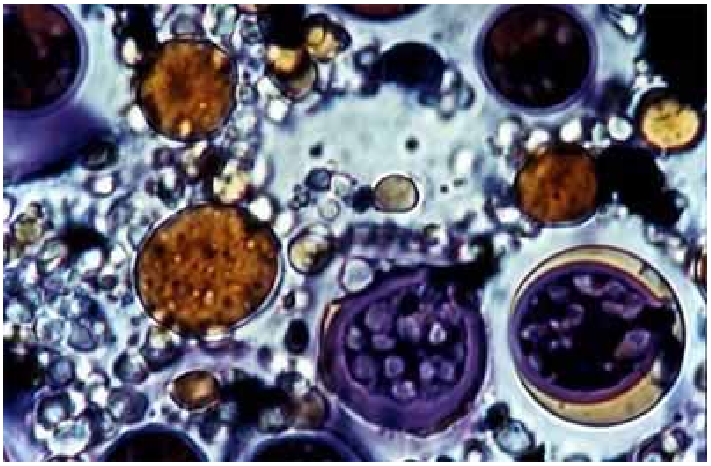
3.7 ALGAE PRODUCTION AND CONVERSION TO BIOFUELS

Algae-based biofuels have recently garnered both significant attention from the press and extensive venture capital funding. These microscopic organisms have potentially very high biofuel yields per acre of land used because of their high cell oil contents. This oil can either be extracted and converted to diesel fuel substitutes (same as with other plant-derived oils like soy oil and rapeseed oil), or the algae can be directly thermochemically converted to diesel substitutes without an oil extraction step.

3.7.1 Technology Description

The term "algae" applies to a large number of aquatic, photosynthetic organisms that fall into several different biological classes, including both protists (single-celled organisms) and eukaryotes (so-called "higher" plants). Microscopic forms of algae include plankton that float or flagellate in seawater. Larger algae like kelp and pond scum resemble land-based plants and can form large masses of growth. Of these myriad species, the simplest organisms, microalgae, may offer the highest yields as a potential biofuel feedstock. They, like all plants, capture energy from the sun, sequester carbon from the atmosphere, and release oxygen into the water and air. An important co-product of this metabolic process is oil formed within the algae cells. Microalgae in particular can reach high levels of oil content, leading to a potentially high biofuel yield if this oil is converted to biodiesel (Please refer to previous sections on biodiesel production for more information about lipids-to-biodiesel conversion processes).

Figure H 22. Cyanobacteria (U.S. DOE 2006).



Different types of microalgae include:

- Green algae. These are generally freshwater species with filament-like structures. These include pond scum, potentially high-oil yield algae called *Gamophyta*.

- Blue-green algae. These are often referred to as cyanobacteria due to their close relationship to bacteria in biochemistry and structure. They play an important role in fixing nitrogen from the atmosphere.

- Brown and red algae. These are generally saltwater algae, and relatively complex in structure. (Columbia 2009)

Microalgae are potentially attractive as a biofuels feedstock for many reasons:

- Under the right growing conditions, they can produce high lipid, or oil, content (by mass) which can be extracted and processed into biodiesel.

- Microalgae can be harvested daily because they have a rapid growth rate and are generally unicellular. In a closed system and the right climate conditions microalgae can be grown and harvested year-round.

- They could be grown on non-arable land because they do not use a soil medium. This would imply a low incidence of land conversion events, and if this rate is lower than for other feedstocks, an increase in overall photosynthetic carbon sequestration.[109]

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[109] It can be argued that conventional biodiesel feedstocks are oil byproducts of food production, meaning that they do not in themselves cause land conversion events. However, if such feedstocks are derived from dedicated energy crop production, the resulting land conversion events would put conventional feedstocks like soy and canola at a carbon sequestration disadvantage.

- Algae could be integrated with power or combined heat and power plants. Researchers are currently experimenting with using the waste heat and scrubbed, carbon-rich flue gas to grow algae and thereby sequester carbon. Algae could also be incorporated into wastewater treatment as a means to enhance aerobic digestion and lower the nutrient load of wastewater.

- Biodiesel is likely to be the dominant algal biofuel product, and current biodiesel ASTM standards would allow biodiesel to be produced from algae oils. However, algae also contain cellulose and protein that could be processed into alcohols (such as ethanol) or other biofuels to recover more value from the algae feedstock.

- There are thousands of species of algae. This vast genetic diversity gives researchers many options when choosing species with desirable traits.

- Algae or the byproducts of oil extraction are well suited to hydrothermal gasification (i.e. conversion to syngas, a mixture of simple carbon and hydrogen compounds) because of their high moisture contents.[110] As long as the algae are grown in a low-sulfur environment, the gasification products, water and syngas, could be respectively recycled for further cultivation and converted to a number of possible biofuel products.

- Algae, when deprived of sulfur, can produce hydrogen. Sulfur reduction depletes cellular oxygen flow, leading to the activation of a hydrogen-producing metabolism (Melis 2001).

Feedstock Production and History

Feedstock production for algae is not described in the resource portion in this assessment project and therefore is covered briefly here. Historically, interest in algae has been closely tied to culture. In Japan and China, algae have been both harvested from nature and cultivated by farmers for nearly 4,000 years, largely as a food additive. The Japanese have attempted to use algae for biological CO_2 fixation. In the 1990s the Ministry of International Trade and Industry spent US$117 million in a microalgal CO_2 utilization program at the Research Institute of Innovative Technology for the Earth (Huntley et al. 2006). No technologies were ever commercialized as a result of this program.

In the U.S. and Europe, in contrast, the only significant interest in algae has been within the last 50 years. In the 1950s and ‘60s algae was of interest for its nutritional value to humans and livestock, but researchers were never able to achieve the theoretical yields needed for use of algae as a commodity food ingredient.

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[110] Hydrothermal gasification, unlike conventional gasification, uses water as an oxidant, making high moisture contents desirable, as long as sulfur and other catalyst poisons are removed from the material before catalysis. Hydrothermal gasification is still performed at only a pilot scale, despite long-term involvement by and favorable results from PNNL.

Though algae were first researched as a food, the oil crisis of the 1970s led to research into algae's application as a fuel feedstock. By executive order, the U.S. Department of Energy's Office of Fuels Development began the Aquatic Species Program (ASP), which sought to produce biodiesel and renewable transportation fuels from high-lipid-containing algae grown in open ponds and fed with CO_2-rich power plant flue gas. Between 1978 and 1996, the program examined more than 3,000 algae species for productivity, yield, and growing conditions (Sheehan et al. 1998).

The conclusion reached by the ASP in 1998 was that cost-effective production of biofuels from algae was probably infeasible in light of that era's low petroleum prices, and the high capital costs estimated for algae cultivation (Sheehan et al. 1998). The limited genetic engineering technologies available at the time may also have led the program to this conclusion. However, algae remained of interest in other arenas. The biotech industry examined algae as a means of producing novel compounds and as a vector for biological manipulation. While interest still exists, the bioreactors never matured to a cost-effective level and biological research has not yet produced any algae to replace *E. coli* and yeast in chemical and specialty manufacturing of fermentation products. However, the technical lessons learned in these studies inform bioreactor design today.

The current interest surrounding algae biofuels can be considered the third period of increased interest in large-scale cultivation of algae (Lee 2001). In the U.S., with the exception of algae cultivation for food additives, algae have mainly been used in wastewater treatment systems as an inexpensive means to oxygenate water for aerobic digestion. Those systems are either large uncovered ponds or a "race-track" design (see Figure H-23) where the water is advanced by paddlewheels through a system of channels. Such systems typically have a long retention time and only serve small communities. In the algae wastewater treatment facilities built to date, the ponds are naturally populated with algae from the environment and have many species of algae as well as bacterial activity.[111]

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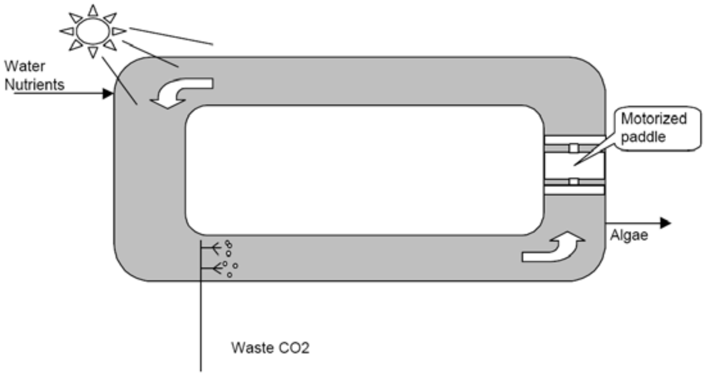
[111] Because of the biological and biochemical diversity of the species present, the most feasible means of energy extraction from existing wastewater systems would be digestion to produce methane. Algal biofuels production and wastewater treatment do not appear to have been coupled in this or any other way in a practical demonstration.

Figure H-23. Race-Track System (Sheehan et al 1998).



Algae cultivation processes in general fall into one of two categories: open or closed systems. The open systems include the race-track system (Figure H-23) and other open-pond processes used in wastewater treatment; the closed photobioreactor system includes vertical growth systems that contain algae in individual bags or closed tank systems that contain algae in spinning drums.

Open Systems

Open systems are easily constructed, but are land- and water-intensive. Their exposure of the algae to ambient air means that temperature is difficult to control and can vary from optimal levels and that carbon dioxide is more difficult to supply in high concentrations. Both of these factors reduce yields significantly from theoretical values. A third yield-reducing factor common to open-pond systems is species contamination. Creating a monoculture in an open pond system would likely prove impossible because other species of algae would become airborne and contaminate the pure culture. Contaminant algae, which aren’t optimized for oil production, compete for light and nutrition, decreasing overall productivity (Benemann 2003). A common way to avoid contamination in an open system is to carefully select the algae species to be cultivated. Species that flourish under certain environmental conditions too harsh for other species—cold temperatures, extreme pH, or low light, for example—are termed “extremophiles.” Selecting an extremophile with a high oil yield allows a pond or racetrack to be maintained at conditions that deter low-yield invasive species but optimize production in the high-yield desired species.

Apart from their applications for wastewater treatment, open cultivation systems have also been used to grow algae for specialty products, such as natural dyes and nutritional supplements. Those cultivation systems are generally between five and 10 acres in size. The only algae strains that have been successfully cultivated commercially are extremophiles. *Dunaliella*, *Spirulina*, and *Chlorella* have been cultivated for their tolerance to elevated salinity, alkalinity and nutrients, respectively. Because of the extreme environments it is possible to cultivate those algae in open pond and “race-track” systems with little risk of contamination by other undesirable algae species (Benemann 2004). Systems of this nature typically produce 10 to 20 tons per acre annually (Vonshak 1997).

Closed Systems

Closed systems, like photobioreactors, are more expensive than open pond systems. A tubular photobioreactor consists of an array of transparent tubes made of either glass or plastic. The tube diameter is generally four inches or less, a size which is determined by optimum light penetration. The tubes can be arranged as a horizontal sheet or as fence-like batteries, but are always directed north-south, so as to maximize sunlight capture. The length of the tubes is controlled by the CO_2 consumption/oxygen generation by the algae. Since the oxygen cannot escape as in the open pond system, while the CO_2 is consumed over the length of the tube, a gradual increase in dissolved oxygen and simultaneous increase in pH will occur. Both inhibit the photosynthesis. The tube length should therefore usually not exceed 80 meters, but biomass concentration, flow rate, light intensity and tube inlet oxygen concentration need to be taken into consideration in tube sizing. A degassing tank is needed to control and optimize oxygen and CO_2 concentrations. Because algae in this tank do not see any light, the volume of the tank needs to be small relative to the volume of the tubes (Chisti 2007).

The costs of an enclosed system are influenced by a number of factors. The degassing tank is relatively small, with a relatively insignificant cost. This means that the cost of an enclosed photobioreactor scales almost linearly with the surface area of its tubes. Land area required is also a significant cost influence. A fence-like array requires the least land acreage of the wide variety of available photobioreactors (Chisti 2007).

In practice, algaculture facilities that use enclosed reactors generally also use open ponds. Aquasearch, for example, operated from 1997 to 2001 a two-hectare combined open pond/photobioreactor facility of >60,000 liter capacity, equally divided between the two systems. Roughly 80% of the production facility area was occupied by the open ponds and 20% by the photobioreactors (Huntley et al. 2006).

Whether in an open or closed system, there are four basic steps in the cultivation of microalgae: inoculation, growth, oil production, and harvesting.

- **Inoculation**: The process of cultivating microalgae begins with inoculation of water with a pure algae culture. In a race-track design the harvesting process lowers the overall concentration of algae in the water but leaves enough to begin the next round of growth. Other systems may rely on enclosed, sterilized tanks in which the culture multiplies. Isolation reduces the risk of contamination from other species, which is a problem for strains that are not extremophiles.

- **Growth**: The algae are allowed to multiply and increase in concentration in the water. During this phase it is important to maintain levels of nutrient, CO_2, temperature, pH, and agitation that maximize the photosynthetic efficiency of the organisms. Some self shading will occur in any bioreactor, but it can be minimized with agitation and reactor design.

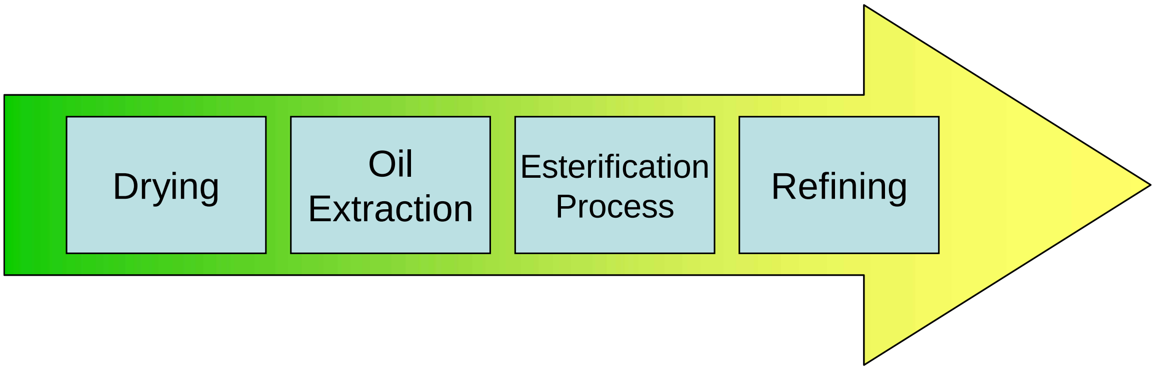
- **In-Cell Oil Production**: Algae can be encouraged to increase their oil yields. Techniques include light deprivation and reducing available nitrogen. Although there is some debate in the literature how this occurs—whether the amount of oil in the cell is increasing or if the mass of the cell is decreasing while the oil content remains constant—the higher oil yield is beneficial to downstream processing.

- **Harvesting**: Extraction of the algae from the water may be achieved by settling, flocculation, centrifugation, filtration or some combination thereof. All methods have advantages and disadvantages and the optimum choice is still under investigation. Some researchers, including Dr. Richard Sayre at the Danforth Plant Science Center, hope to "milk the algae"—harvest oil *in situ* from living algae—attempting to lengthen the productive life of algae cells (Domestic Fuel 2009b).

Conversion to Biofuels

After harvesting, the main steps in conventional algae processing are drying, oil extraction, esterification and refining.[112] These steps are common to most biomass-to-biodiesel facilities. The significant way that this processing differs when applied to algae is during the drying and oil extraction steps. The high water content of the harvested algae mixture requires spray-drying rather than conventional dewatering. The dry algae are then pressed to begin removing the lipids, or oils, from within the cells. Physical pressing, performed with equipment similar to a bean crusher, removes about 75% of the oil present. Subsequent extraction with hexane removes about another 20%, leaving about 5% of the original oil mixed with cellulose and protein solids (Newman 2004). These solids left over from oil-pressing can either be fed to cattle as a protein supplement, or fermented into ethanol.

Figure H-24. Algae Conversion to Biodiesel.



Alternatively, oil can be removed using supercritical fluids. Carbon dioxide (usually high purity "food grade" that contains minimal levels of sulfur and nitrogen compounds) is transformed into a supercritical fluid by application of extremely high pressure. Under these conditions, it is a very good solvent to selectively remove oil from proteins and cellulose. Removal of the CO_2 solvent, unlike hexane solvents, is very easy (the pressure is reduced until the fluid is no longer supercritical), and creates few chemical disposal hazards. The supercritical fluids method can remove nearly 100% of the oil present in algae. (Newman 2004)

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[112] There is not yet a standard method for oil extraction from algae, and widely different methods have benn proposed and used.

Once the oil is extracted and separated from any solvents, it is generally converted into biodiesel through transesterification. Transesterification is common to all oil and fatty acids-to-biodiesel processes, and is explained in more detail in previous sections of this report.

3.7.2 Technology Cost and Performance

Today, biodiesel is predominantly produced from non-algal, plant-derived oils, and in the case of the United States, mainly soybean oil. Biodiesel from microalgae has not been commercialized yet, and there remains a striking lack of pilot scale data to verify yields, chemistry, and cost propositions. There is currently no industry leader, with many variations of systems seen in pilot and demonstration scale facilities.

There are pilot scale facilities in various stages of planning, construction and operation. The emerging industry is largely funded by venture capital and there has been no commercial success to date. Thus, it would be difficult to make any preliminary assessment of the price at which algal biodiesel could be produced in a real system. The costs of making biodiesel from algal oil are relatively clear, as biodiesel facilities are well-characterized in yield, capital, and operating cost. The cost and implications of supplying the algal oil feedstock, however, remain unclear.

Data key to constructing a facility model are unavailable or inconclusive: the capital cost of the cultivation system, the dewatering technology and associated costs, the annual yields from a cultivation system, and the oil produced per ton of biomass. At this stage in development there is inadequate engineering and pilot plant data to model the cost and performance of algae to biodiesel systems at commercial scale. Although several companies have announced the potential of large breakthroughs for scaled-up systems, there appears to be significant R&D and additional investment required to realize those improvements.

The Algae Species and its Oil Content and Yields

Although algae production is conceptually designed to be closely coupled with biofuels conversion, it is helpful to understand the economics of feedstock production and conversion separately.

Different types of algae contain different levels of oil. Oil content in the cells of the algae appears to be a key parameter in the economics of the algae to biodiesel system. A sustainable 50% increase in the oil content of algae can make or break algal fuel economics (Sheehan et al. 1998). Hence bioengineering of algae species to achieve maximal oil content has been suggested and attempted by a number of researchers, among them the U.S.Aquatic Species Program described above (Sheehan et al. 1998). These efforts continue today, with even more powerful bioengineering tools than before.

Of all types of algae known, *Gamophyta* are well suited for biodiesel feedstocks, because they retain particularly high levels of lipids within their cell walls. Up to 80% oil content has been demonstrated, but there is significant variation in oil content between algae species. In Table H-25 below, oil contents are tabulated for some common algae species. Values between 20 and 50% are typical.

Table H-25. Oil Content (Chisti 2007).

Oil Content of Some Microalgae	
Microalga	**Oil Content (% dry wt.)**
Botryococcus braunii	25–75
Chlorella sp.	28–32
Crypthecodinium cohnii	20
Cylindrotheca sp.	16–37
Dunaliella primolecta	23
Gamophyta	up to 80
Isochrysis sp.	25–33
Monallanthus salina	>20
Nannochloris sp.	20–35
Nannochloropsis sp.	31–68
Neochloris oleoabundans	35–54
Nitzschia sp.	45–47
Phaeodactylum tricornutum	20–30
Schizochytrium sp.	50–77
Tetraselmis sueica	50–77

Economic calculations are very much dependent on the sustainable oil content of the algae harvested. Oil contents like 30% (Chisti 2008) or 40% (Huntley et al. 2006) are often quoted in scientific literature. Algae producers hope to double these oil contents and thereby significantly cut the break-even cost per gallon biodiesel. However, as the NREL APS program concluded, it is very difficult to sustain such levels of oil production, as variables such as temperature, sunlight, pH, and contamination (in the case of open ponds) all influence oil content.

Current Feedstock Production Economics

Cultivation and processing of algae in open systems have very different cost structures from enclosed systems. Open systems tend to have high operating cost but low capital cost,[113] while closed systems trade high capital cost for potentially lower operating cost.

Open Pond Systems

Algae have traditionally been cultivated in open-pond systems, with many levels of sophistication. Such facilities range from temperature-controlled, compressor-agitated ponds that supply concentrated carbon dioxide to maximize production, all the way to paddlewheel agitated, ambient-temperature ponds that process atmospheric CO_2 at only a modest rate but with small operating costs. As noted previously, open pond systems of any type risk contamination from ambient algal spores.

Enough open-pond facilities exist that some pilot plant-scale algae cost data has been published. One of the most commonly cultivated species, *Spirulina* (an extremophile cultivated for the nutritional supplement market), grows in open pond systems with yields up to 20 dry tons of biomass per acre per year. Production costs ($2008) have been reported between $11.99 and $22.49 per dry kg equivalent to about $10,795 to $20,390 per ton algae produced (Lee 2001).

The economics of open pond cultivation is strongly influenced by a number of factors, including:

Water loss

Due to their large surface area to volume ratios, one disadvantage of open pond systems is water loss to underlying soil or overlying air. Water loss, due to osmosis through the pond liner or lack of a liner, can increase salinity and pH and reduce algal productivity, and must be carefully controlled. Seepage water loss in unlined ponds is generally between 0.01 and 0.2 in/acre/day for aquaculture farms (Davis 2004). Seepage alone would thus necessitate 271 gallons per acre per day makeup water, a contributor to open ponds' large production costs.[114]

Water can be lost even more significantly from the top of the pond through evaporation. The degree of evaporation varies with climate and level of salinity, and evaporative water loss is thus hard to estimate.

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[113] These comparisons are intended only within the field of algae production, and not the broader field of energy crop production. Farming algae, even in open ponds, is far more capital-intensive than conventional agriculture. This will be discussed later in this section.

[114] Other biomass crops, like corn, alfalfa, or soybeans, have production costs of $146/ton, $327/ton, and $120/ton, respectively, at medium seeding density (Duffy, 2008).

Power

The largest contributor to operating cost is either power to run circulation machinery such as pumps and paddlewheels, and harvesting machinery such as centrifuges (Benemann 2004). Some open pond systems employ CO_2 injection systems, whose compressors also require a large amount of energy.

Capital Costs

Production rates of algae oil in open ponds are expressed as mass per unit surface area per unit time. Similarly, capital costs are expressed in dollars per unit surface area. Reported capital costs for open ponds vary significantly, with values reported as low as ($2008) $14,992/acre (Lee 2001). The single biggest contributor to capital cost appears to be compressors and equipment to distribute additional carbon dioxide to the ponds. Researchers proposing a system of open, 4-hectare, unlined ponds using such compressors estimated a capital investment of ($2008) $24,202/acre (Benemann 2002). Huntley et al. (2006) describe the costs and operation of a two-hectare system of combined open ponds and bioreactors. Calculated capital cost for an open pond system was noted as $44,774 per acre ($11.08 per square meter) – both in $2008. (Huntley et al. 2006)

Closed Systems

When compared to open systems, enclosed bioreactors have a very different cost structure, and have more theoretical cost estimates. Photobioreactors are more expensive than open ponds. Huntley et al (2006), having built several dozen photobioreactors at various scales up to 25,000 liters, suggest that ($2008) $118/$m^2$ ground surface or ten times the costs of an open pond system is a good number. These researchers also suggest that substantial cost reduction for photobioreactors is possible by scaling up the units, particularly in view of the relatively moderate costs of the reactor tubing ($1.18 per square meter—$4,768 per acre—in $2008) and their higher algae production relative to an open pond system. (Huntley et al. 2006).

One theoretical case study used an upper boundary of photosynthetic efficiency and a lower boundary of production costs to calculate that an algae facility could produce biodiesel at a cost of ($2008) $22.23/gal ($947/bbl) (Dimitrov 2007). The analysis based capital and O&M costs on a detailed report on heliostat arrays.[115] It also assumed the land, water, micronutrients and CO_2 were all free, making the analysis a gross underestimate.[116] The study modeled an advanced enclosed cultivation system, estimating a capital cost of ($2008) $811,537/acre ($200/m^2) and an annual yield of 5,000 gallons biodiesel/acre/year (1.24 gal/m^2/year)—about 70 times the yield of soybeans (Dimitrov 2007). The resulting capital cost for an advanced facility would be, conservatively, $161 per gpy capacity in $2008.

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[115] Because they track the sun, heliostat arrays would represent a very advanced, very capital-intensive form of enclosed reactor. Thus, the capital cost calculated on this basis is likely an upper bound rather than a generally applicable estimate for enclosed systems.

[116] Water supply has been identified as a key problem with placing algae farms in non-arable regions. This issue will be addressed later in the report. The authors' assumption of free water, land, and nutrients means that the production cost they generate should be treated as a lower bound, rather than a generally applicable estimate.

While it is useful to compare types of algae cultivation with each other, a more meaningful question is how well algae cultivation compares to conventional agriculture producing energy and oil crops. Cultivation costs for algae are expected to be much higher than for other traditional farm biomass products. One proposed hybrid system (partially open, partially closed) was expected to cost more than ($2008) $17,766 dollars per acre (Benemann, 2003). The example system was 1,100 acres (445 ha) and had several stages of production that used different types of bioreactors. Unlined open ponds, the least expansive form of cultivation, was employed on 988 acres (400 ha) of the proposed system. The author expected unlined ponds to cost approximately $14,805/acre. Assuming that is the minimum cost, it is conceivable that more complex systems that are fully enclosed and have active temperature regulation, nutrient delivery, CO_2 addition and agitation could incur a cost comparable to greenhouse agriculture, which can cost around ($2,008) $388,426/acre (Seckler 2002).

3.7.3 Cost and Performance Summary

System	Capital Costs (per acre)	Operating Costs (per acre-year)	Bio-Oil Yields (gal per acre-year)
Open systems	**$10,000 to $38,000**, depending on location and degree of sophistication	**$6,000**, depending on sophistication and water losses	Commercial Estimate: **6720**
Closed systems	**$100,000 to $250,000**, depending on bioreactor type	**$59,000**	Pilot Target: **2500** Commercial Estimates: **5000 to 6100**
Sources:	*Huntley et al, 2006* *Benemann, 2003* *Seckler, 2002*	*Huntley, 2006* *Solix, 2008*	*Sheehan et al, 1998* *Solix, 2008* *Dimitrov, 2007*

Table H-26 Cost & Performance Summary.

Costs associated with the processing steps—which include harvesting, dewatering, drying, and oil extraction—are not yet found in detailed engineering form for pilot or commercial offerings for any of several proposed processing methods. Based on descriptions for systems under development, key parasitic energy uses will include electricity for pumps and agitators, and energy for dewatering processes that will likely involve some form of mechanical separation. Temperature regulation increases the photosynthetic efficiency of the organisms but could also be a significant energy cost depending on the system and ambient environment. Therefore detailed energy balances will be an important consideration in evaluating the economics and energy efficiency of the proposed technologies. Key indicators for the potential performance of algae-based biofuels are summarized in Table H-26.[117]

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[117] There is significant uncertainty in the estimates for capital and operating costs and yields associated with this emerging technology. Given this fact, the values stated in the Table should only be used as very general, speculative indicators of potential.

There are pilot scale facilities in various stages of planning, construction and operation. The emerging industry is largely funded by venture capital and there has been no commercial success to date. Thus, it would be difficult to make any preliminary assessment of the price at which biodiesel could be produced in a real system. Data key to constructing a facility model are unavailable or inconclusive: the capital cost of the cultivation system, the dewatering technology and associated costs, the annual yields from a cultivation system, and the oil produced per ton of biomass. At this stage in development, there is inadequate engineering and pilot plant data to model the cost and performance of algae to biodiesel systems at commercial scale. The one exception appears to be a detailed process engineering and energy balance reported only in presentation format in September 2008 (Solix, 2008). The absence of peer-reviewed systems analysis and pilot plant data makes it difficult to verify the validity of developers' claims at this stage. Although several companies have announced the potential of large breakthroughs for scaled up systems, there appears to be a long stretch of R&D required to realize those improvements.

3.7.4 Outlook

The cultivation of algae for energy is still a very young industry. There is continuing experimentation in bioreactor design as well as in processing and conversion. Details on this are noted below under the sub-section "R&D."

Many companies and researchers expect the most significant improvements in algae biofuels to come from genetic engineering and species selection to improve the organisms' biochemical properties. Scientists could potentially make organisms that are more productive, easier to grow in culture, or easier to process. However, this raises many environmental and biosecurity concerns. As with any other genetically modified organism, the release of genetically engineered or foreign species of algae into aquatic ecosystems could incur unknown environmental consequences. There are already species of algae known to be invasive and ecologically damaging, which were transported by human actions into foreign ecosystems. Invasive algae species present a significant hazard because there is no good method of removing them and they could potentially add to the environmental problems caused by hypoxic waters. The ecologically safe cultivation of exotic or GMO algae would thus require durable, enclosed bioreactors, at significantly increased capital cost, or the engineering of species that cannot survive in the wild.

Water rights are a growing issue in many parts of the country, and one criticism of algal biofuels is that they could exacerbate water shortages. There are several potential solutions to this problem. One would be to use saltwater microalgae and cultivate the algae in a seawater environment. The other would be to use enclosed bioreactors that would eliminate seepage and evaporative losses. A closed system that recycles water almost completely may be fairly efficient compared to irrigated crops with significant transpiration losses. Algae could also play a role in purifying otherwise unusable wastewater or agricultural runoff. Thus, biodiesel production could be coupled with waste treatment in a unique synergy.

Another potential drawback to algal biofuels is a lack of road testing of existing cars fueled by algal biodiesel. There appears to have been only few such demonstration cars actually on the road. Solazyme—a manufacturer of algal biodiesel, which it terms SolaDiesel—drove an unmodified, algae biodiesel-powered Mercedes-Benz E320 sedan through Park City, Utah at the Sundance Film Festival in January, 2008. Solazyme noted that it planned to produce algal biodiesel price-competitively at a commercial scale within three years. It has worked closely with the National Renewable Energy Laboratories (NREL) and receives funding from Chevron. The scale of its current biodiesel production remains unclear. Although Solazyme claims that its algal biodiesel exceeds the ASTM D6751 specification for use in existing diesel engines, the performance and emissions of the demonstration car were not reported[118] (Squatriglia 2008).

In order to characterize information about other current industry activity in algal biofuels, a distinction must be made between small companies or organizations working with unproven or cutting-edge technologies, and established companies with larger-scale technologies close to or undergoing commercialization. Three sections follow:

- "R&D," which contains information about academic and industrial research into farther-term, advanced technologies[119]
- "Industrial Perspective," which contains information about algal biofuels companies with technologies nearing commercial viability
- "Funding," which contains information about the cost structure of the key companies performing algae-related work

In the first two cases, information is organized by the organization performing the work described.

3.7.4.1 Research and Development

U.S. Air Force. The USAF has funded research into certifying its fleet's operation on a 1:1 blend of conventional jet fuel ("Jet A") and standard military aircraft fuel ("JP-8") made from algae and other biomass feedstocks. The funded researchers claim that "algae could supply enough fuel to meet all of America's transportation needs using a scant 0.2% of the nation's land," but have not addressed the economics of such a scenario. UOP and Syntroleum are taking part in this work. Syntroleum has so far supplied at least 100,000 gallons of "synthetic" fuel—which may or may not be algae-derived—for engine testing (Damron 2009).

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[118] An important complication when discussing published biodiesel performance data is that there is rarely any distinction made between the various types of biodiesels, whether algae-derived or from other plant oil sources. "Biodiesel" can refer to fuels from many sources, including waste oil-based and algae-derived fuels. Though biodiesel can be made from many feedstocks and all biodiesel must meet ASTM specifications, performance and emissions data specific to algal biodiesel is sparse. In the U.S., engine manufacturers have expressed support for blends of soy-derived biodiesel of up to 20%. NREL, in its high altitude engine test laboratory, has performed stationary emission tests on various biodiesels. (Graboski, M.; McCormick, R., 1994). These indicated a marked decrease in hydrocarbon emissions, but a significant increase in NO_x emissions upon switching from petroleum- to plant-derived diesels. Later studies by McCormick (2005) found that in engine-alone tests, NOx emissions increase, particularly in older engines, but that actual vehicle testing shows a modest reduction, rather than increase in NO_x emissions. There does not appear to be a consensus among researchers whether biodiesel in practice elevates or decreases these emissions rates.

[119] Regarding further R&D, http://www.biofuelsdigest.com/blog2/2009/03/10/algae-webinar-highlights-ongoing-research-and-development-activities-as-well-as-science-and-technology-progress-and-challenges/ gives on a regular basis a good overview of the state of the art R&D in the algae-to-fuels industry.

University of Virginia. Professors Lisa Colosi, Andres Clarens, and Mark White received a $30,000 Collaborative Sustainable Energy Seed Grant to determine optimum algae biofuel productivity as a function of carbon dioxide and organic matter available to the growing algae. The researchers hope to increase oil yield as much as 40% by integrating CO_2-rich power plant flue gas or untreated sewage solids into the cultivation process. Their reactors are so far only bench-scale, but they will eventually perform a scenario-based simulation of the overall life-cycle effects of algae cultivation for biodiesel (*ScienceDaily* 2008).

The Office of Energy Efficiency and Renewable Energy (EERE). The EERE provides Small Business Innovation Research grants, one category of which is devoted to algal biofuel research. Companies that have received these grants at a Phase I level ($100,000) include:

- *Accent Laboratories, LLC (New York).* Low-cost, high efficiency algae harvesting/dewatering technology
- *Renewable Algal Energy, LLC (Tennessee).* Low-cost algae growing and harvesting method.*Touchstone Research Laboratory, Ltd. (West Virginia)*
- Closed algae cultivation system that cost-competitively averts seasonal temperature fluctuations (Braff 2008)

The National Laboratory System. Four federally-funded research labs in particular are performing algae research. The Pacific Northwest National Laboratory has received over $1 million in funding for basic and applied research in micro-algal biofuels. Sandia National Laboratory is conducting research about water use and resource availability for biofuels and biofuels co-product manufacturing, including algal biodiesel and glycerol. The program’s funding totals $5 million from various governmental sources, and has a cooperative agreement with LiveFuels, Inc. Oak Ridge National Lab also does algal research, partnering with Utah State University to study production problems, like photosynthetic saturation, minimizing surface shading, and scalable photobioreactor design. Los Alamos National Laboratory does similar research and is working closely with companies in the private sector, particularly CEHMM and General Atomics (Braff 2008).

U.S. Universities. Many academic researchers are studying algal cultures, their use as a CO_2 fixer/recycler, and their biofuels production potential. These researchers work for institutions that include:

- U.C. San Diego/Scripps
- Montana State University
- Old Dominion University
- Clemson University
- University of Washington

- University of Hawaii
- California Polytechnic State Univ.
- University of Maryland
- U.C. Berkeley (working closely with BP)
- Arizona State University (working with BP)
- Utah State University
- Ohio State University
- Michigan State University
- Brooklyn College
- U.T./Austin
- Colorado School of Mines
- Princeton University
- Missouri University of Science and Technology
- Lincoln University

3.7.4.2 Industrial Perspectives

In dealing with the industrial perspectives one must consider the bigger "green-minded" companies like BP, Chevron, and Shell Global Solutions, as well as the smaller start-up companies. While some companies are already producing small quantities of algae-derived fuels, most are in the early development stages, with viable—and profitable—production still several years away. The Kiplinger Update Report (2009) provides a wealth of information on the latter field of players in algae technology, as do industry publications and industrial journals.

This section attempts (by no means exhaustively) to give an impression of the efforts put forward by those companies in a market that can be characterized as very competitive, with a tight market for funding and high marketing costs. Key players include:

Aquaflow Bionomics Corporation (ABC; New Zealand). ABC was founded in 2005, and currently operates a facility integrating algae into a wastewater treatment system in order to eventually turn the algae into biodiesel. Founding directors Nick Gerritsen, Barrie Leay and Vicki Buck originally studied algae-rich settling ponds in Blenheim, New Zealand. The company first successfully converted dried algae to biofuels in May 2006, and in 2007 constructed and commissioned a larger-scale "prototype refinery" to accommodate a scaled-up algae harvesting process (ABC 2009).

Boeing, Inc. (Chicago, IL). Boeing is funding several small algae projects, including at Washington State University and the University of Washington. Boeing also gathers samples for internal engine testing and fuel screening (Kiplinger, 2009).

British Petroleum (BP). BP is, in its traditional way, active in the field. In recent years BP has supported outside institutes and universities, rather than perform in-house R&D. Currently, BP sponsors algae R&D work at U.C. Berkley and Arizona State University (Kiplinger, 2009).

Chevron. Chevron has partnered with NREL and private biodiesel firm Solazyme to develop and test algae-based transportation fuels. As part of this effort, Chevron is actively tracking developments in algae and has been testing some of its proprietary hydroprocessing catalysts, which convert "algae-oil" to hydrocarbon diesels rather than fatty acid derivatives (Personal communication, 3/09). Chevron's strength has always been in hydro-treating/hydro-cracking technology, leading the company to play a complementary role to algae technology producers.

Community Fuels (Enctias, CA). Community Fuels hopes to use its proprietary biodiesel production technology to identify new feedstocks and processes to make the biodiesel industry more sustainable and cost-effective. The privately-held company claims to have research and commercialization agreements in place with multiple international research institutes in order to access additional expertise and to expedite projects. Kiplinger (2009) notes that the company "has been awarded a 2007 Phase I Research Grant from the Department of Energy Small Business Innovation Research / Small Business Technology Transfer (SBIR/STTR) Program to evaluate two processes to use agricultural waste as a resource for commercial-scale algal oils development and production into biodiesel. Community Fuels is active in an algal biodiesel project (designed to address key obstacles of commercial-scale biodiesel production from algae) and unique solid catalyst project (anticipated to reduce processing costs associated with alternative feedstocks). Community Fuels has a large biodiesel plant in Stockton, California that uses a proprietary process technology to maximize quality, flexibility and energy efficiency in biodiesel production. Community Fuels, in collaboration with the Hawaii Natural Energy Institute of the University of Hawaii, will investigate the use of agricultural waste products to grow specific algae species."

Donald Danforth Plant Science Center (St. Louis, MO). This not-for-profit research institute hosts a variety of research focusing on plant genetics and optimization for nutrition and pest resistance, but also hosts the research of Dr. Richard Sayre, who is attempting in situ oil extraction from living algae (Domestic Fuel 2009a; Danforth Center 2009).

General Atomics (San Diego, CA). This defense contractor has started collaboration with the Center of Excellence for Hazardous Materials Management (CEHMM). CEHMM is building open ponds for the growing of saltwater micro algae and production of bio diesel on unused, non-arable land in the Carlsbad, NM area (Kiplinger 2009).

Global Green Solutions Inc. (Vancouver, B.C.) and **Valcent Products Inc. (Vancouver, B.C.).** In 2008, this partnership announced its collaboration on "Vertigro," a high-density, vertical bioreactor. The first commercial harvest from the technology is expected at Valcent's El Paso, Texas, facility in March, 2009 (Kiplinger 2009).

Greenfuel Technologies Corp. (Cambridge, MA). This start-up was of note because it was able to continuously raise venture capital and focused on enhancing algae growth with carbon-rich flue gas. In December 2007, Greenfuels entered into a partnership with Aurantia of Spain. The GreenFuel-Aurantia partnership intended to develop a $92 million, 100-hectare algae greenhouse at the Aurantia cement plant in Holcim, Spain. On October 21, 2008, they announced the second stage of their development plan, which would produce 25,000 tons of algae biomass per year. The costs of this development were not stated. In May, 2009, GreenFuel ceased operations (GreenFuel 2007; 2009).

Green Plains Renewable Energy (Omaha, NE). This bioethanol producer operates four plants with a total capacity of 330 million gallons per year. In November, 2008, the company committed $1.4 million to a 25.5% stake in BioProcess Algae LLC, a joint venture between photobioreactor developer BioProcessH_20 and filter manufacturer Clarcor. Green Plains hopes for federal approval to redirect a $2.3-million research grant to fund algae production at its ethanol plant in Shenandoah, Iowa (Kiplinger 2009).

Honeywell (Morristown, NJ). Honeywell's UOP subsidiary is heavily involved in the development of "green" commercial jet fuels from algae and biomass-derived oils. Airbus, JetBlue Airways and General Atomics joined UOP in a joint venture to commercially develop algal biodiesel in 2008 (Kiplinger 2009).

LS9 (San Carlos, CA). LS9 has genetically engineered a trademarked line of DesignerMicrobes™ that can produce biofuels and specialty chemicals from algae and conventional sugars in a single-step fermentation, with no intermediate oil extraction process. The biodiesel product generated by the process has undergone independent ASTM testing and has been certified to meet U.S. and Brazilian emissions standards (LS9 2009).

Neptune Industries (Boca Raton, FL). This marine technology and seafood farming firm accepted a grant from the Florida Farm-to-Fuel program for developing an algae production system optimized for eventual biodiesel production. The firm claims that it has already been contacted by biofuels processors seeking to purchase its algae feedstock (Kiplinger 2009).

OriginOil Inc. (Los Angeles, CA). This start-up hopes to commercialize the hyper-growth algae bioreactor that it has developed, and eventually to license its technology to biodiesel producers. OriginOil's cultivation scheme centers on a 40-foot-long bioreactor module that can hold 10,000 gallons of water. The company grows microalgae in these tanks and uses conventional pressing and extraction to extract oil. According to the company, as much as 60 percent of the dry weight of algae is oil, and as a bonus, a by-product from the process can be used for animal feed (Kiplinger 2009).

Petro Sun (Scottdale, AZ). This privately owned oil company has recently begun working its way into the renewable energy sector. PetroSun BioFuels, a wholly owned subsidiary of PetroSun, recently began operating a 1,100 acre facility growing algae in open saltwater ponds. The facility, located in Rio Hondo, Texas, is designed to produce an estimated 4.4 million gallons of algal oil and 110 million lbs. of biomass co-product per year. Twenty of the operating acres would be reserved for the experimental production of a renewable JP8 jet-fuel for the U.S. military. However, on Aug 5, 2008, PetroSun announced that the company had notified Science Applications International Corporation (SAIC) of its withdrawal as a team member from the Defense Advanced Research Projects Agency (DARPA) BAA08-07 Biofuels-Cellulosic and Algae Feedstocks program. Company executives stated that, "Our withdrawal was prompted by the uncertainty of the award process, combined with an opportunity to initiate a project with a renewable fuel refiner and a commercial jet fuel end-user" (PetroSun 2008).

In a press release in March, 2009, PetroSun communicated that they were providing algal oil feedstock for conversion into B100 biodiesel by a third party biodiesel refiner. This suggests that PetroSun has the ability to produce crude algae oil, but not the capability to hydro-treat, refine or distill the feedstock. The company allocated five drums (275 gallons) of B100 for emissions studies from an initial run of approximately 80,000 gallons of algal oil converted to ASTM biodiesel (PetroSun 2008).

The company's objective is to have three independent emissions studies conducted by Carnegie Mellon, the University of California, Los Angeles (UCLA) and NextEnergy. The studies are intended to benefit both the algae-to-biofuel industry and PetroSun's efforts to establish a carbon credit program for commercial algae operations (Petrosun 2009).

One year earlier, PetroSun announced plans to establish algae farms and algal oil extraction plants in Louisiana, Texas, Arizona, Mexico and Central America. The algal oil product will be marketed as feedstock to existing biodiesel refiners and future company-owned refineries (Petrosun 2008).

Royal Dutch Shell (Netherlands). In October, 2007, Shell announced plans to build a test facility in Hawaii for algae cultivation research, with the ultimate goal of enabling commercial-scale biodiesel production. The six-acre, open-pond demonstration project, funded through a joint venture with HR Biopetroleum, was scheduled for completion in October, 2009 (Herald-Tribune 2007).

Sapphire Energy (San Francisco, CA). Sapphire Energy is a venture capital-funded startup that hopes to commercialize algal production of crude bio-oil with chemical properties and refining techniques identical to those of crude petroleum. Though its goal is to produce renewable hydrocarbon gasoline, its proprietary algae cultivation techniques make it of interest to the biodiesel industry, as well (Sapphire Energy 2009).

Solazyme (San Francisco, CA). In June, 2007, Solazyme and Imperium Renewables, a diesel manufacturing technology developer, entered into partnership. Solazyme plans to grow its proprietary strain of microalgae and extract the oil, which Imperium will convert to biodiesel (Kiplinger 2009).

Solix (Golden, CO). In March, 2006, researchers from Colorado State University and renewable energy engineers from several consulting firms formed Solix to develop and commercialize large-scale manufacturing technologies for algal biodiesel. The 50-person company is headquartered at CSU's Engines and and Energy Conversion Lab and is privately funded (Solix 2008).

The Algal Biomass Association (ABO). This trade association was formed in May, 2008 in order to promote the development of "viable commercial markets for renewable and sustainable commodities and specialty products derived from algae" (Braff 2008). The ABO hopes to serve as the national and international trade association for commercial producers of algae and algal biofuels. The Board of directors contains representatives from corporate, academic, and consulting fields. It includes veteran algae researcher John Benemann (now heading the consulting firm Benemann Associates), as well as leaders from Boeing and NREL (Braff 2008).

3.7.4.3 Funding Sources

The basic funding for capital and research in algae process and development is a complicated matter. There is fierce competition for money. Andrew Braff (2008) categorized companies by their funding type, as shown in Figure H-25. It is important to note the significant amount of public funding that flows into start-up companies and research programs through federal agencies and state governments, either in the form of direct capital investment, or indirectly through subsidies and tax credits.

Figure H-25. Capital Sources for Algal Biodiesel Startups (Braff 2008).

Capital Sources

Venture Funding	Other Private Funding	Public Sector Funding
•Auroa Biofuels	•Algae Biosciences Corp.	•Aquaflow Bionomic
•BioFuelBox	•AlgaTechnologies, Ltd.	•CEHMM
•Cequesta	•A2Be Carbon Capture	•Community Fuels
•Cobalt Technologies	•Bioverda (JV w/ Virgin GF)	•General Atomics
•Euglena	•Blue Sun Biodiesel	•GreenFuel Technologies
•GreenFuel Technologies	•Community Fuels	•Kent SeaTech
•Inventure Chemical	•GreenFuel Technologies	•Patriot Biofuels
•Solazyme	•LiveFuels	
•Synthetic Genomics	•Petro Algae	
	•Primafuel	

Some examples of state government funding organizations are The New Mexico Energy Innovation Fund, the Texas State Emerging Technology Fund, and the Iowa Power Fund, which collectively funded at least $16 million in research and capital funding for algal biofuel applications (Braff 2008).

The algae industry is currently estimated to have a gross impact on the economy and/or revenue in the multi-millions of dollars. It is expected to continue to grow, particularly with the bio-medical industry networking extensively with it. However, should biofuels from algae become cost-competitive with crude oil-derived diesel, an even larger market is available. The gross impact of the industry is still difficult to estimate exactly in tax- and grant-free dollars, but it is expected to be very large.

4 ADDING VALUE TO BIOREFINING: WASTE HIGH-MOISTURE BIOMASS TO BIOFUELS

Manures, wastewater, and biorefinery byproducts are significant energy-rich feedstocks, but their high moisture content has led to their underutilization. Algal biomass is also a wet biomass stream but is not a waste or byproduct and is treated as a separate resource in another section of the report.

A recent limited-distribution study prepared by Antares Group for Battelle Memorial Institute shows the potential market for biosludge as feedstock. On a wet weight basis, the study identified approximately 9.8 million tons/year of biosludge available in the U.S. annually through the pulp and paper making process. For wet milling, Antares discovered 15.2 million and 4.5 million tons per year of potential feedstock from Wet Gluten Feed (WGF) and Wet Gluten Meal (WGM), respectively, for a total supply of approximately 19.8 million tons/year.

As for dry ethanol mills, in the Battelle study Antares estimated the amount of potential feedstock (in the form of wet distillers grains with solubles—WDGS) to be about 60 million tons per year. With continued long-term growth expected in ethanol production, feedstock from dry-grind ethanol mills should rise further in the future. As a conclusion of the Battelle study, Antares estimated a potential total annual feedstock resource from wet and dry mills of approximately 90 million tons in wet weight or about 20 million tons in dry weight.

Given below is a summary of the available feedstock:

- Potential Feedstock = 20 million dry ton / year
- Potential Energy = 320 trillion Btu / year
- Potential Capacity (electric) = 4 to 5 GW electricity
- Potential Capacity (Biofuels) = 1400 MGY ethanol

One viable way to use these otherwise underutilized resources is as a feedstock for biogas production, either by anaerobic digestion or by gasification. "Biogas" is a general term that can be applied to many carbon-rich gas mixtures derived from biomass. In the case of anaerobic digestion, the biogas product is a mixture of methane and carbon dioxide, with a large complement of contaminant gases whose amounts depend on the type of feedstocks digested. Methane derived from anaerobic digestion may be competitive in efficiencies and costs to other biomass energy forms including heat, synthesis gases, and ethanol.

An alternative to digestion is gasification, which can produce a fundamentally different gas, though still considered "biogas." While those types of gasification that can process high-moisture feedstocks are not yet commercial technology, hydrothermal gasification has been demonstrated at lab scale to directly gasify aqueous solutions of biomass (5 to 10% solids) to produce methane or a "syngas" mixture of carbon dioxide, hydrogen and a small amount of methane. Minor traces of hydrocarbons higher than methane can also be formed.

The resulting biogas can have a number of uses, depending on whether the biogas is mainly methane (produced by digestion and less prominently in gasification) or mainly syngas (produced in gasification). Uses for syngas can include direct combustion or conversion to biofuels like ethanol. Uses for methane also include direct combustion, but may include higher-purity applications such as injection into the natural gas pipeline system or compression for use as CNG or LNG vehicle fuels.

Raw biogas produced by digestion is often burned directly, despite high moisture content and a high concentration of sulfur-containing compounds.[120] For most commercial-scale facilities, emissions standards and prime mover[121] specifications will prevent direct combustion without significant gas pretreatment. Methane can also be injected into the natural gas pipeline system, in which case it must meet rigorous purity standards. Those include near-absolute removal of the common contaminant hydrogen sulfide (to 4 ppm by volume, or 0.25 grain per 100 ft^3), CO_2 removal to typically 3% by volume, and water removal to typically < 7 lb per MMSCF. The former and the latter are removed to avoid corrosion in the pipeline system. H_2S is removed in the biogas for most applications, as well. Biogas that is compressed to LNG or CNG and subsequently used in LNG vehicles cannot contain engine-damaging compounds like hydrogen sulfide or mercaptans,[122] and must undergo an energy-expensive compression step (Wardzinski et al. 2004).

Because the U.S. transportation system is largely liquid fuel-based, conversion of methane to a longer-chain liquid hydrocarbons would be a boon to biogas producers. However, gas-to-liquids (GTL) technology works best with CO and hydrogen to produce C_2 and higher hydrocarbons. This means that a gasification syngas product would be the best suited biogas from which to make gasoline-like fuels. Unfortunately, the type of gasification that is required to handle high-moisture biomass (catalytic hydrothermal gasification) is still at a pilot scale and is crippled by sulfur removal constraints and catalyst deactivation.

Condensation reactions with methane are energy-intensive and require specialized, expensive catalysts. A gas-to-liquids infrastructure for methane feedstocks would be challenging since the methane molecule is more stable than any other hydrocarbon species. Scale is also a large factor in the economics of GTL technology and biogas is not produced in the quantities needed for large scale production. However, biogas production can be used to support biorefinery operations by providing a potentially economical source of methane for process heat and power. In this report these technologies are viewed as complementary processes for biorefining, rather than as primary conversion processes.

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[120] Though there is far less sulfur than water or carbon dioxide in the biogas, the sulfur content of biogas, at levels as high as 10,000 ppmv (as hydrogen sulfide) (personal communication with Norm Scott, 8/28/09), can be considered high when compared to pipeline specifications for natural gas, which require less than 4 ppm hydrogen sulfide. According to EMG International (2007), average hydrogen sulfide contents are between 2,000 and 3,000 ppmv.

[121] "Prime mover" refers to the equipment used to generate heat and power from a fuel stream. Such equipment includes gas engines and gas turbines, which are very sensitive to tars (C2 and higher organic compounds) and acidity. Such compounds must be scrubbed from syngas before it can be combusted without damage to the prime mover.

[122] Prior to residential or commercial distribution, mercaptans are added to distribution line gas for ease of detecting gas leaks.

4.1.1 Technology Description

4.1.1.1 Anaerobic Digestion (Biochemical Route)

The process of biogas production from anaerobic digestion begins with feedstock collection, which varies widely depending on the type of feedstock used.[123] Production of methane via anaerobic digestion begins with a very simple pretreatment—mixing in a holding tank prior to entering the digester. This homogenizes the slurry, preventing settling and solids separation, and mildly reduces the particle size of some solids. The resulting sludge is processed in a bioreactor. The anaerobic digestion of biologically volatile solids (VS) produces biogas rich in methane and CO_2 with trace H_2S. The resulting sour gas is scrubbed to remove H_2S and moisture. For biorefinery use, the scrubbed gas can be used to fuel engine generator sets and heating or drying operations. For vehicular use and natural gas pipeline injection, further CO_2 removal is performed to make 95%+ methane. Compression to make renewable compressed natural gas (CNG) or liquefied natural gas (LNG) follows. A general process diagram is shown in Figure H-26.

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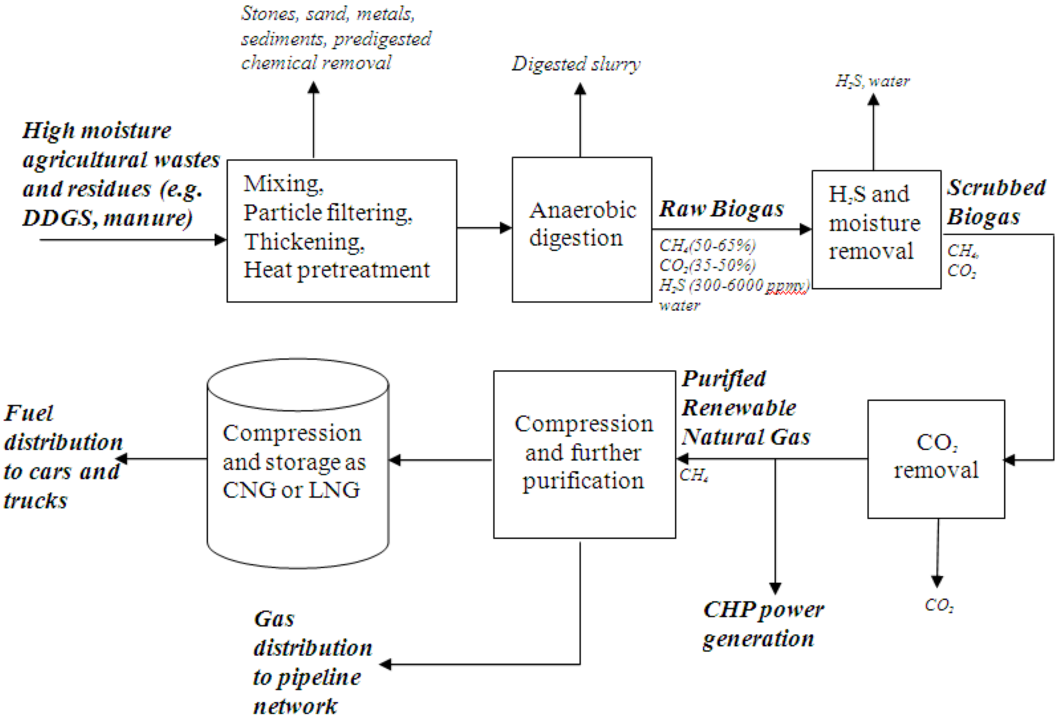
[123] This study will focus selectively on high-moisture wastes, in the form of manures and food processing wastes. However, digesters are not technically limited to waste feedstocks alone, and many proponents argue that their efficiency in converting energy crops to transportable fuel is higher than that of most liquid biofuels technologies. (This may not tell the whole story, as the form that energy takes—whether gas, liquid, heat, or electricity—is also important. Currently, liquids are required in conventional vehicles, and these methods are thus preferred over potentially more efficient technologies.) Digesting wastes averts disposal costs, so a waste's purchase price is often in effect negative. As this study will discuss in more detail, digesters are often economically infeasible even when the feedstock is "free" in the form of a waste. Dedicated crops—which will in the future likely become more expensive as other biofuel technologies begin demanding them as feedstocks—may be even less likely to be economical, and for this reason are not addressed as digester feedstocks in this report.

Figure H-26. Anaerobic Digestion Process Diagram.



For food processing wastes such as cheese whey, corn silage, tomato puree, grape skins and distiller's grain, the organic solids content (expressed by the chemical oxygen demand (COD)) is too high for discharge into regular sewage lines and wastewater treatment facilities. Instead of aerobic treatment to remove the organic solids, anaerobic treatment is used due to the relative purity of the solids that are co-produced with the digester biogas. If the biologically volatile solids (which produce almost all of the COD) are not a large portion of the total solids going into the digester, large volumes of lignocellulosic material and inorganic material (e.g. sand, minerals) will emerge as waste products with the desired biogas.[124] These undigested materials can be composted or further processed for value added products including organic fertilizer, animal bedding, soil amendments, and biomass feedstock for more biofuels.

Apart from organic solids content, an important characteristic of the biofuel feedstock is the total solids when collected. For manures, this is generally a function of livestock type and determines the digestion process and digester design that are most appropriate.

On swine farms, almost all of the manure is flushed into lagoons, which can be covered to collect the biogas released upon decomposition. Flushed dairy cow and flushed broiler chicken manure in warmer climates can also

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[124] Hydrolytic bacteria, which begin the process of digestion, act most quickly on carbohydrates, fats and proteins. They can break down hemicellulose, but this requires a long residence time and a batch process, as other long-chain molecules will be preferentially processed to depletion before hemicellulose will be digested. Lignin is not processed at all by conventional digester cultures, though some xylophageous anaerobes can process it slowly.

be processed in a covered lagoon to make biogas. Broiler chicken and turkey manures, as well as horse and dry feedlot cattle manure, are collected on a packed bed of wood chips, fiber and older manure which periodically is front-loaded into piles. These piles of dry manure are typically not digested because of the vast quantity of water needed to make them pumpable as a slurry. Instead the dry manure is used as fertilizer. In between these ranges of too dilute and too dry material are livestock manure like scraped dairy manure from feedlots, food processing wastewater streams (e.g. whey), and leachates and glycerin from the biodiesel industry. These feedstocks can be mixed and chopped in a slurry of 8-14% total solids, an ideal concentration for heated plug flow (PFR) digesters and continuously stirred (CSTR) digesters.

These two types of digesters both control the temperature of the anaerobic digestion process to tightly regulated mesophilic (37-41°C) temperatures or thermophilic (50-52°C) temperatures to encourage the particular bacteria and particular metabolic processes that produce large amounts of methane. Anaerobic digestion is a multi-stage process in which carbohydrates, fats and proteins are first hydrolyzed to sugars and fatty- and amino acids by hydrolytic bacteria. Acidogenic bacteria convert these to carbonic acid, alcohols, hydrogen gas and carbon dioxide, which acetogenic bacteria convert to acetic acid, hydrogen, and more carbon dioxide. The final—and least stable—step is methanogenesis, in which methane-producing bacteria process acetic acid. Digester operators manage solids loading, feedstock introduction timing, temperature and pH to try to keep these three processes operating at equal rates so that few by-products are evolved and biogas production is continuous.

Thermophilic systems digest more quickly than mesophilic systems, leading to residence times as low as 14 days. They are less stable than mesophilic systems, however. Rapid pH changes can occur and lead to stalled methanogenesis. Therefore many digesters utilize 15-20 day mesophilic digestion instead of faster thermophilic processes. CSTR digesters use impellers, stirrers and mixing pumps to keep the solids in uniform suspension in the liquid, and tend to work more quickly. PFR digesters have a longer residence time, but may offer more complete conversion of volatile solids. It is important to note that the quality of digester management, and such variables as climate can lead to more or less volatile solids conversion, regardless or digester type.

Digesters can be set up in single- or multi-stage configurations. Single-stage processes keep all four types of bacteria in a single reactor; multi-stage processes separate the bacteria into several different reactors that can be individually optimized for the organisms they contain. Multi stage reactors offer better overall pH, temperature, and productivity control, since each segment is "fine-tunable." Multi-stage reactors are in this way much easier to manage and much harder to destabilize.

The U.S. EPA Agstar program (2009) offers guidelines on digestion type selection for biogas production, summarized in Figure H-27 below.

After biogas is collected from the digester, the remaining digestate is collected and often used as an alternative to energy-intensive synthetic fertilizers (Digestate from thermophilic systems is less bacterially active and considered more desirable for food agriculture). The biogas, now mostly carbon dioxide and methane, has significant contamination in the form of hydrogen sulfide and water. These, along with the hydrogen and carbon dioxide, are mostly removed in a series of separations processes intended to make the gas pipeline- or engine-compatible. For CO_2 and H_2S removal either a simple amine wash or physical solvents are used. Many of these separations are proprietary, like Selexol and Rectisol, and use refrigerated solvents that require significant energy input.

Methane at >95% purity, with no water and less than four ppm by volume hydrogen sulfide, can be injected into the existing natural gas pipeline infrastructure. For use as LNG or CNG, biogas requires complete water and H_2S removal prior to compression. LNG and CNG can be used as transportation fuels, but in the U.S., this is currently only applied for large centralized fleets like bus systems due to a lack of CNG/LNG refueling infrastructure. Other countries, like India, have already large urban distribution systems for CNG.

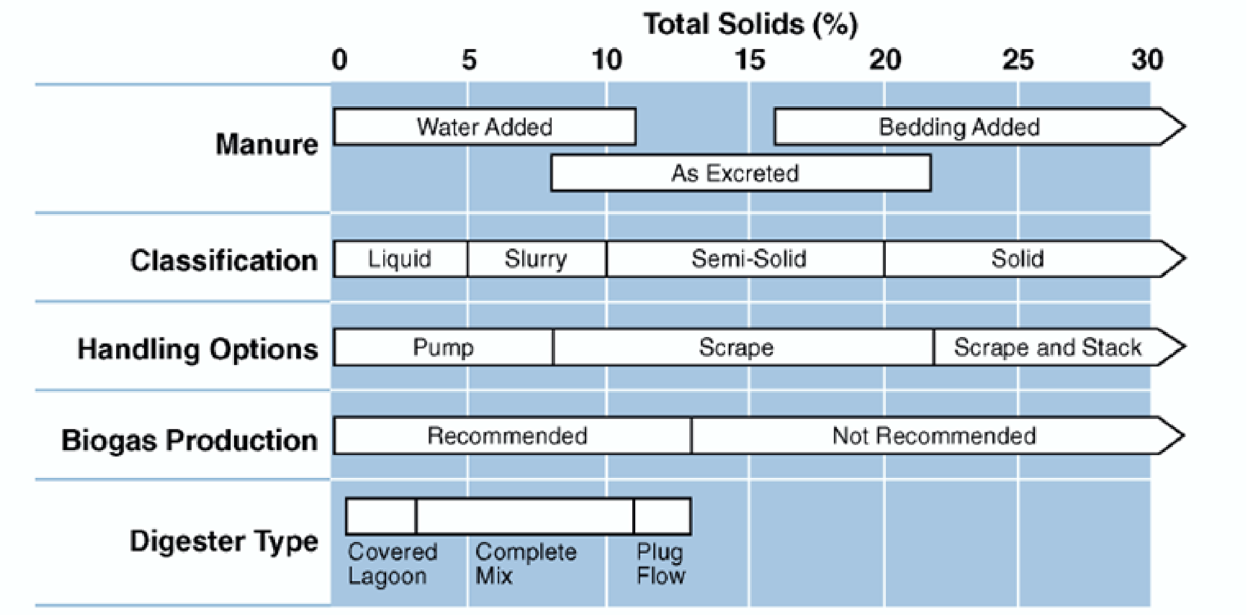

Figure H-27. Appropriate Manure Characteristics and Handling Systems for Specific Types of Biogas Digester Systems (U.S. EPA 2009).

4.1.1.2 Thermochemical Route

As is considered tangentially in the discussion of the biochemical route, digester feedstocks will yield very different amounts of biogas, depending upon their volatile solids content (i.e. biodegradability). A biological digester can only convert biodegradable materials to biogas, meaning that the often large portion of high-moisture feedstocks that is not volatile solids does not get converted to fuel. This inaccessible material—such as manure fibers and agriculture residue fibers—can undergo further treatment after digestion to decompose hemicellulose and cellulose to short-chain hydrocarbons. This increases the amount of volatile solids for conversion to syngas, and can be accomplished with pressure cooking or microwave pretreatment.

One thermochemical technology approach, extensively studied at PNNL and MIT, is gasification in aqueous solution with two key variants proposed: low temperature catalytic hydrothermal gasification (LTCHG) (Elliott

1993) and supercritical water gasification (SCWG) (Peterson 2005). Performed at high pressure, it can process all types of the carbonaceous material, taking it not merely to a reduced chain-length hydrocarbon, but completely to the simplest and most desirable product: a synthetic gas composed out of carbon monoxide, hydrogen and methane. To supply this heat and power, some of the fuel product material can be combusted. LTHG and SCWG do not require moisture removal, as water is itself a reactant and the suspension medium. This technology currently exists only at a laboratory scale.

LTCHG offers an attractive solution for the conversion of wet organic residues to a clean, medium-Btu gas (methane and carbon dioxide). Specifically, the gasification system could operate with streams containing as little as 5-20% by weight dry solids (95% to 80% moisture). Conventional gasification systems are generally inoperable with feedstocks containing more than about 50-60 weight % moisture, and generally uneconomical at moisture contents exceeding 30% by weight. Carbon conversion is near 99% for select feedstocks so that the only remaining byproducts are inert inorganic materials in the waste-stream. The challenge is to broaden the applicability of the process, currently at a demonstration scale, to biorefinery byproducts.

4.1.2 Performance and Cost Analysis

The cost of the digestion system components and feedstock biomass depends on the method used for collecting the high strength liquids and slurries, and upon the size of the farm or food processing plant. Equally importantly, it depends on the volatile solids content or biodegradability of the feedstock. Higher volatile solids contents will indicate better biological conversion of the feedstock to energy, with less inert digestate left unreacted. Volatile solids content, as expressed by chemical oxygen demand, can be used as a direct indicator of the potential biogas output of a given feedstock.

Most units sized for a single commercial animal farm cost between $589,136 and $1,178,273 ($2008)(Mattocks and Wilson 2005). If such a system is used for direct power generation from raw biogas, it will typically cost 1.5-2.3¢ per kWh ($2008) to maintain and operate (Schmidt, 2000). A very recent comprehensive and independent analysis of biogas produced from ten California dairy farms of varying sizes show a wide variation in the economics (Zhang 2008). For example, capital costs ranged from $2500 per kW to $14,000 per kW. Production costs for electricity varied from 180 $/MWh to 450 $/MWh, while production costs for pipeline quality gas varied from $2.10/therm to $5.80/therm. These parameters are all a function of the size of the farming operation, as expressed by the number of animals it contains.

Jewell and Wright (2005) indicate that anaerobic digesters are not economically feasible with less than 400 cows or equivalent animal unit (AU). The cost per cubic foot of methane favors large Confined Animal Feeding Operations (CAFOs) and if possible aggregated biogas production piped to a central processing site. Farmers that are able to accept food processing and other organic wastes from nearby food manufacturing plants may be able to collect a tipping fee, as well as increasing the scale of their energy production in the form of biogas. This may lead to better overall economic feasibility for an anaerobic digester in such a “co-digestion” scenario. Co-digestion may lead to more feedstock variation, depending on the types of additional feedstocks accepted by a farmer, and their regularity of delivery. This may make digester management more difficult, but the economic benefits of co-digestion may outweigh any process control challenges.

The cost of purification of the resulting biogas is less thoroughly studied. It is unclear how economical it is to purify biogas for pipeline injection. Molecular Gate and QuestAir sell molecular sieve/pressure swing absorber (PSA) technology to remove CO_2 from the methane (Molecular Gate 2009). PSA technology is well suited for medium size applications, but has the draw-back of multiple vessels and many moving valve parts. US Filter (now owned by Siemens), Varec and New Energy Solutions sell activated carbon filters that remove much of the hydrogen sulfide and moisture from the gases (Varec Biogas 2009; U.S. Filter 2009; Intrepid Technologies 2009). Activated carbon, or "iron sponge"—iron-impregnated wood chips mixed with the feedstock to the digester—would be a preferred solution for small scale applications. The former is a well proven technology, but requires disposal of the spent adsorbent. For larger applications, regenerative systems as used in conventional hydrocarbon refining technologies (such as Selexol and Rectisol systems) would be preferred. Added costs are involved in the final removal of the H_2S. Possible downstream applications of the removed H_2S are production of liquid or solid sulfur, production of sulfuric acid, or conversion to produce mercaptans.

Biogas yields average 8,000-16,000 cubic feet per ton of feedstock input, compared to a theoretical yield of 38,200 ft^3/ton fatty acids and 22,300 ft^3/ton proteinaceous materials (Ader 1981). Pig manure feedstocks average 9550-14,300 ft^3 biogas per ton feedstock; cattle manure feedstocks 6400 ft^3/ton (DTP 1997). At an average methane content of 65%, biogas has an energy density of 600 BTU/SCF (Pipeline-quality natural gas generally contains >950 BTU/SCF, reemphasizing the magnitude of the required CO_2 removal). This translates to an average energy yield of 13.5 MMBTU/ton feedstock (Biogas Technology 2009). Conversion efficiencies are typically 20-22% for single-farm systems (Schmidt 2000).

4.1.3 Outlook

Due to the success of many large centralized digester systems for swine and industrial bio-wastes in Denmark and Germany, some companies like Microgy and Andigen are building refinery sized manure/food waste digesters to benefit from trucking of energy-rich biodegradable material and economies of scale. This partially removes some of the methane abatement burden from landfills and wastewater treatment plants, which traditionally have the task of collecting and purifying high strength liquids and bio-wastes before discharge into the environment. (Andigen 2009; Microgy 2009; Schmack Biogas 2009).

According to Chandahar et al. (2007), enough manure is produced annually in New York State (one of the nation's biggest dairy producing states) to produce 468 million cubic feet of biogas. Many tons of food processing wastewater are either dumped into rivers or clarified[125] in treatment plants each year. Some facility owners are taking advantage of clean air programs and self-generation energy incentives to convert concentrated biodegradable

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[125] Clarification refers to the reduction of the suspended solids content of a water stream, and thus its opacity and chemical oxygen demand, or organic solids content.

liquids and solids into combined heat and power. However, the use of these wastes need not be limited to combined heat and power and other on-site technologies, and may also extend to the production of vehicular fuels. It all depends on the industry's willingness to invest further in energy conversion and material processing technologies.

There is a steadily rising trend towards larger dairy, swine and poultry livestock production. There is also a growing trend toward centralizing waste treatment. Food processing companies are increasingly diverting waste streams directly to centralized digestion sites or trucking high strength liquid to specialized handling areas in waste water treatment plants. This aggregation is a significant advancement for biogas, because a sufficient volume of biodegradable material can produce enough biogas to make the large capital investment of a digestion facility cost-effective.

At least two clusters of manure producers are planning to build biogas pipelines to increase the cost-effectiveness of refining biogas to CNG. Aggregating smaller biogas streams into larger biogas pipelines would allow for centralized gas processing and could help manure producers in CAFO-intensive states economically convert biogas into pipeline quality natural gas, CNG and combined heat and power. One of these clusters is located in Central Valley, California, the other in Cayuga County, NY (Groom 2009). Such clusters offer an important opportunity for community-based renewable energy in the form of CHP systems, co-producing electricity and district heat from combustion of the biogas in engines or gas turbines. Several such village CHP systems exist in Germany, such as biogas plants in Reiffenhausen, and Krebec/Wollbrandshausen, which each serve about 75 buildings (Fangmeier 2009).

One operational example of the integration of biogas CHP and biorefining is the E3 Biofuels project in Mead, Nebraska. Although the project has been beset by an array of operational problems, including a boiler failure, the plant design does demonstrate the potential benefits of the integration of the technologies. The Mead plant uses both the manure from a large cattle (28,000 head) feedlot operation and the ethanol byproduct thin stillage as the feedstock for the digester. Biogas makes the ethanol facility self sufficient for both power and heat.

5 EMISSIONS AND WATER USE DATA FOR BIOFUEL CONVERSION TECHNOLOGIES

Water use, greenhouse gas and criteria pollutant emissions data were estimated for each modeled conversion technology, in order to be used as inputs for the environmental benefits analysis in Appendix G. These data were calculated based on the conversion technology models or available data in the literature. Those data pertaining to air emissions are presented in Appendix G as part of the lifecycle analysis modeling, while this section addresses the question of water emissions.

Table H-27 summarizes water use for various conversion technologies. The biomass conversion technology data are based on the technology models developed for this effort. The crude oil refining data is included for comparison, based on results from Sheehan et al. (1998).

Table H-27. Water Use of Various Conversion Technologies (Antares 2008).

Technology	Feedstock	Water Use (gal / gal product)
Biomass Conversion Technologies		
Grain to Ethanol - Dry Mill	Grains	4.7
Grain to Ethanol - Wet Mill	Grains	24.4
Fatty Acid to Methyl Ester	Virgin Oil Waste Oil Animal Fat	0.03 0.03 0.03
Lignocellulosics to Ethanol - Enzymatic Hydrolysis/Fermentation (*Mid term facility with dilute acid pretreatment*)	Wood Chips (hardwood) Wood Chips (softwood) Corn Stover / Ag Residues Wheat Straw / Ag Residues Switchgrass / HEC	6.2 5.9 6.6 7.0 6.9
Lignocellulosics to Middle Distillates - Fischer Tropsch	Wood Chips (hardwood) Wood Chips (softwood) Corn Stover / Ag Residues Wheat Straw / Ag Residues Switchgrass / HEC	1.2 1.1 1.3 1.2 1.3
Lignocellulosics to Gasoline - Pyrolysis/Hydrotreating	Clean Wood Chips Softwood Bark	8.2 8.9 9.8
Fatty Acids to Diesel Fuel - Hydrotreatment		
(*stand-alone*)	Vegetable Oil or Animal Fat	0.0
(*co-processing*)	Vegetable Oil or Animal Fat	0.0
Crude Oil Refining		
Gasoline Refining**	Petroleum	0.003
Diesel Refining**	Petroleum	0.001

** Yield reported in gallons per dry ton except for Grains (in Grain to Ethanol), which are in as rec'd tons (15% MC)*
*** Data from Sheehan et al. 1998*

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APPENDIX H-A

Conversion Technology Status Matrix

Reference Name	Feedstock	Conversion Technology	Fuel Type	Benefits	Issues/Challenges	Key Players [1]	Status
PRODUCTION OF BIOFUELS FOR THE GASOLINE FUELS MARKET							
Grain Ethanol - Dry Mill	Grains / Starches	Enzymatic Fermentation	Ethanol	Commercial technology - many facilities already operational and in construction.	Feedstock availability and production is limited and there is competition with food products. Net energy balance lower than projections for cellulosic ethanol. Limited opportunity for improvement.	POET, Dupont, Abengoa, VeraSun, and others	Commercial
Grain Ethanol - Wet Mill	Grains / Starches	Separation and Fermentation	Ethanol	Commercial technology - many facilities already operational and in construction.		ADM, Cargill, Bunge, Pacific Ethanol, and others	Commercial
Sugar to Ethanol Fermentation	Sugars	Fermentation	Ethanol	Commercial technology. More efficient than conversion of starches.	Limited feedstock availability in North America	Imperial Ethanol Inc., Imperial Bioresources LLC	Commercial
Lignocellulosics to Ethanol - Hydrolysis/Fermentation	Lignocellulosic Biomass	Enzymatic Hydrolysis and Fermentation	Ethanol	Relatively low maintenance costs. Potentially very high glucose yield.	Cellulase costs must be reduced. Enzymes are sensitive to poisoning.	POET, Abengoa, Iogen, Catalyst, Mascoma (NREL)	Demonstration/pilot plants currently operating. More planned for operation within five years.
		Acid Hydrolysis and Fermentation	Ethanol	Technically mature - dilute acid process is oldest cellulosic ethanol technology.	Expensive vessels and high maintenance costs (corrosion). Large amount of gypsum for disposal. Requires improved acid recovery (especially strong acid process).	strong acid - Arkenol, BlueFire Ethanol Inc. dilute acid - (Univ. Wisconsin, NREL)	Demonstration plants are planned for operation within five years.
Lignocellulosics to Ethanol - Gasification/Fermentation	Lignocellulosic Biomass	Gasification and Fermentation	Ethanol	Higher yield of ethanol per ton of feedstock than direct fermentation of biomass. See Note #2.	See Note #3.	ALICO, BCT (Univ. of Arkansas)	Pilot / demonstration facilities planned for operation within five years.
Lignocellulosics to Mixed Alcohol - Gasification/Thermocatalysis	Lignocellulosic Biomass	Gasification and Thermochemical Conversion	Mixed Alcohols	Mixed alcohols are easily blended with gasoline. See Note #2.	See Note #3. Catalysts sensitive to deactivation from sintering. Process requires improved yields. Potential groundwater contamination from mixed alcohols usage.	Range Fuels, Pearson (PNNL, NREL)	Pilot / demonstration facilities planned for operation within five years.
Lignocellulosics to Butanol - Enzymatic Hydrolysis/Fermentation	Lignocellulosic Biomass	Thermochemical Conversion or Multistage Fermentation	N-butanol, Iso-butanol	Butanol is high-octane, pipeline-transportable, and energy-dense.	Compatibility with existing motors not proven beyond 16% blends. High production costs.	BP, Gevo, ButylFuel	Demonstration plants operating. Commercial plants to be operating by late 2009.
Hemicellulose to Ethanol: Pulp and Paper Application	Hard- and Soft-woods	Hot Water Extraction Hydrolysis and Fermentation	Ethanol	Increases the value of hemicellulose as a feedstock.	Fermentation studies not yet performed. Probable high water use.	SUNY-ESF, University of Maine	Not yet to pilot phase.
Lignocellulosics to Gasoline - Pyrolysis/Hydrotreating	Lignocellulosic Biomass	Pyrolysis Oil Production and Upgrading via Hydrotreatment / Hydrocracking	Bio-oil, Diesel, Gasoline	Using refinery technologies already in place, ease of adoption into current infrastructure	In early development stages. Refinery integration issues.	BIOCOUP Consortium (UOP, PNNL, NREL)	Development of process chain began in June 2006, concept development by 2011. Industrial demonstration may follow.
High Moisture Biomass: Biorefinery Heat and Power	Biorefinery/Ag Wastes and Biorefinery Byproducts	Digestion or Gasification	Methane, Syngas	Averts disposal costs, allows integration of animal farms and food processing plants with biorefineries.	Significant purification costs for turbine-combustible gas. Gasification technology not well demonstrated.	PNNL, Eastman Chemical	Digesters commercially operating. Gasification still at bench scale.
PRODUCTION OF BIOFUELS FOR THE DIESEL FUELS MARKET							
Fatty Acid Methyl Ester	Seed Oil / Waste Oils / Animal Fats	Esterfication	Methyl Esters	Ease of conversion, biodiesel can be used in diesel engines without adjustments. Significant heating oil market.	Relatively small scale. Limited opportunity for process improvements. Cold weather technical issues.	ADM, Cargill, Bunge, BlueSun Biodiesel, and others	Commercial
LignoCellulosics to Middle Distillates - Fischer Tropsch	Lignocellulosic Biomass	Gasification and Fischer Tropsh Synthesis	Middle Distillates, Gasoline	FT diesel can be substituted directly for conventional diesel with lower emissions. See Note #2.	See Note #3. Catalysts sensitive to poisoning and sintering. Requires improved yields.	Bechtel, ECN, Shell, Choren,	Pilot facilities in operation, demonstration facilities planned for 2009 and beyond.
Black Liquor Conversion to Middle Distillates or DME	Pulp Mill Byproducts	Gasification and Synthesis	FT Liquids, DME	Involves no land conversion events; significant feedstock supply; leverages existing pulp mill equipment.	Complex, capital-intensive process. Pulp industry currently has little available capital. Technology not well demonstrated.	Chemrec, Weyerhauser	Pilot and demonstration facilities operating.
Algae to Biodiesel (Methyl Ester)	Microalgae	Pressing and Esterification	Renewable Diesel	Extremely high potential oil yields per acre; could be integrated with power plants for carbon sequestration.	High capital cost/productivity tradeoff. High production costs. Technology not yet commercialized.	National Laboratories, Solazyme, PetroSun, Solix	Several demonstration facilities in operation. Commercial facilities planned.
Fatty Acids to Diesel Fuel - Hydrotreatment (Green Diesel)	Seed Oil / Waste Oils / Animal Fats	Upgrading via Hydrotreatment	Renewable Diesel	Using refinery technologies already in place, ease of adoption into current infrastructure	In early development stages. Refinery integration issues.	ConocoPhillips, Neste (UOP)	Pilot facilities in operation, others are planned for 2008 and beyond.

1) Examples of some of the key players for each technology type. Universities and National Laboratories are shown in parenthesis.

2) High range of feedstock flexibility. Intermediate product from gasification can be used for broad slate of end-products (i.e. fuels, power, heat, etc.)

3) Gasification requires dried biomass. High level of syngas clean-up required.

APPENDIX H-B

Conversion Technology Pretreatment Matrix

Pretreatment Requirements for Selected Biofuel Conversion Technologies

Technology	**Pretreatment Description**	**Required Moisture Content**	**Size**	**Source**
Grain Ethanol - Dry Mill	Milling to a fine flour	--	fine	*RFA 2005*
Grain Ethanol - Wet Mill	Steeping	--	--	*RFA 2005*
Fatty Acid Methyl Ester	Filtering, dewatering	~ 0%	--	*Zhang et al. 2003a*
Lignocellulosics to Ethanol - Hydrolysis/Fermentation (1)	Cleaning (2), Sizing & Milling, *Dilute Acid Hydrolysis*	--	1-3 mm	*Hamelinck et al. 2005*
	Cleaning (2), Sizing, *Steam Explosion*	--	19 mm	*Hamelinck et al. 2005*
	Cleaning (2), Sizing, *Liquid Hot Water*	--	19 mm	*Hamelinck et al. 2005*
Lignocellulosics to Middle Distillates - Fischer Tropsch (3)	Sizing, Drying	10-15 wt%	6-13 mm	*Tijmensen et al. 2002*
Lignocellulosics to Gasoline - Pyrolysis/Hydrotreating	Sizing, Drying	5-10 wt%	2 mm	*Ringer, Putsche and Scahill 2006*
Fatty Acids to Diesel Fuel - Hydrotreatment (Green Diesel)	Combination of filtering, dewatering, acid washing, ion-exchange, desalting	~ 0%	--	*UOP 2005*

1) This process has a drier to remove moisture from solid residuals before used for heat - usually available at MC of 60%, dried to 15%.

2) Cleaning is not required for all feedstocks. For example, de-barked forest residues do not generally need to be cleaned, while agricultural residues such as corn stover needs to be washed.

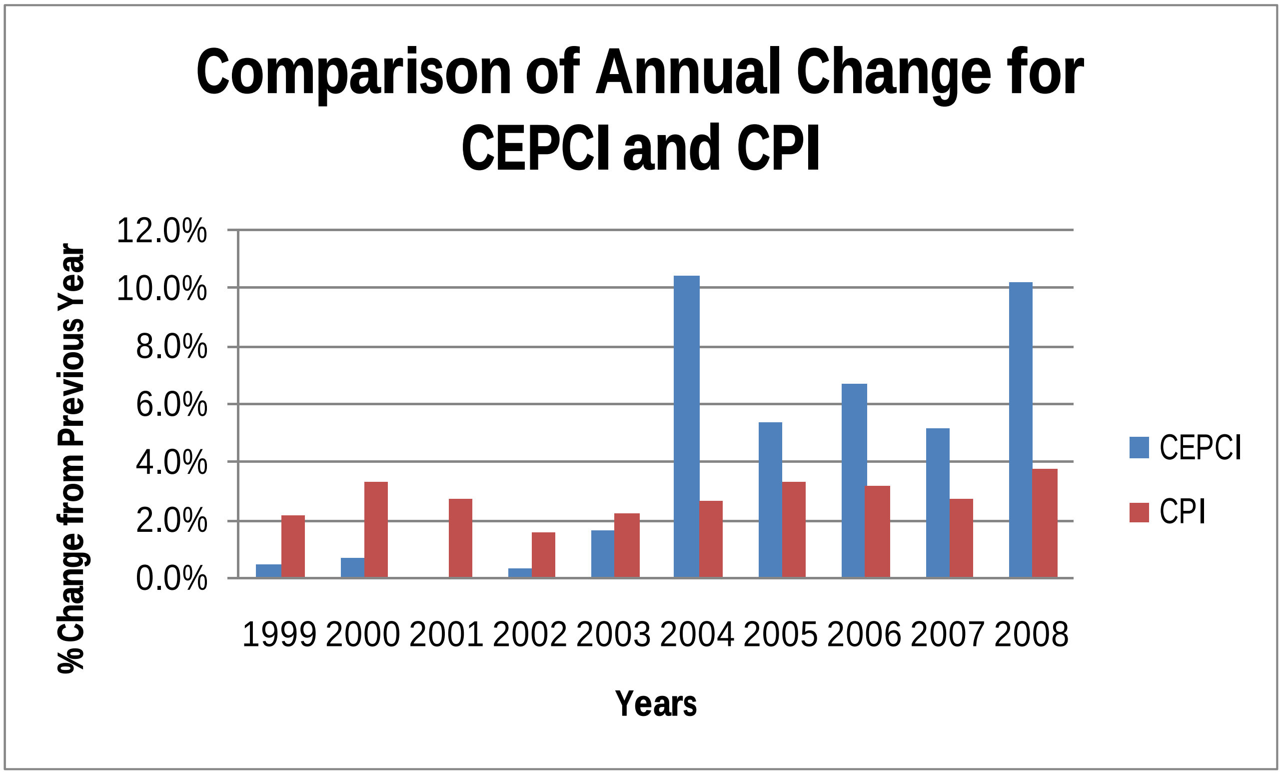
3) The required material size for gasification depends on the gasifier type and the feedstock. Woody biomass is typically sized to 1/4" minus, while agricultural residues are generally 1/2" minus, based on Antares experience.

APPENDIX H-C

Economic Analysis Factors

The economic analyses presented above used a fixed charge rate of 12.3%, based on an assumed economic lifetime of 25 years and a weighted cost of capital of 9.7%. These factors were chosen as a conservative estimate that reflects common biorefinery practices.

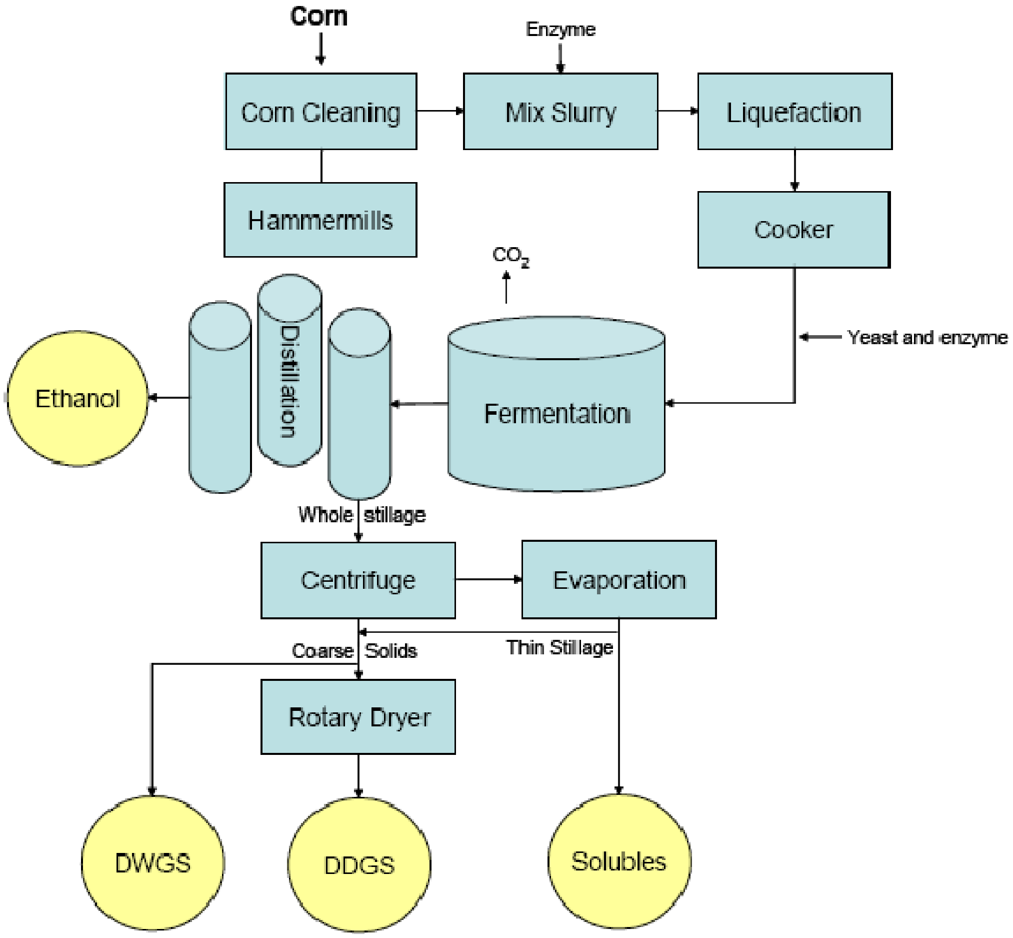
Comparison of Annual Change for CEPCI and CPI



The graph above compares the annual change (from the previous year) for the Chemical Engineering Plant Cost Index (CEPCI) and the Consumer Price Index (CPI). In general, there is not a clear relationship between the CEPCI and the CPI. Furthermore, the CEPCI has seen large increases in recent years, particularly from 2003 to 2004. As a result of these increases, there is some concern within the biofuels industry about capital cost projections that will not adequately account for cost increases of materials used for chemical facilities (such as steel). While this may indeed add significant costs to the capital, the same cost increases will apply to all fuel production industries. As such, the relative capital costs between the biofuel industries and oil refineries will remain the same, despite greater future expansion in the biofuels sector than in the petro-fuels sector.

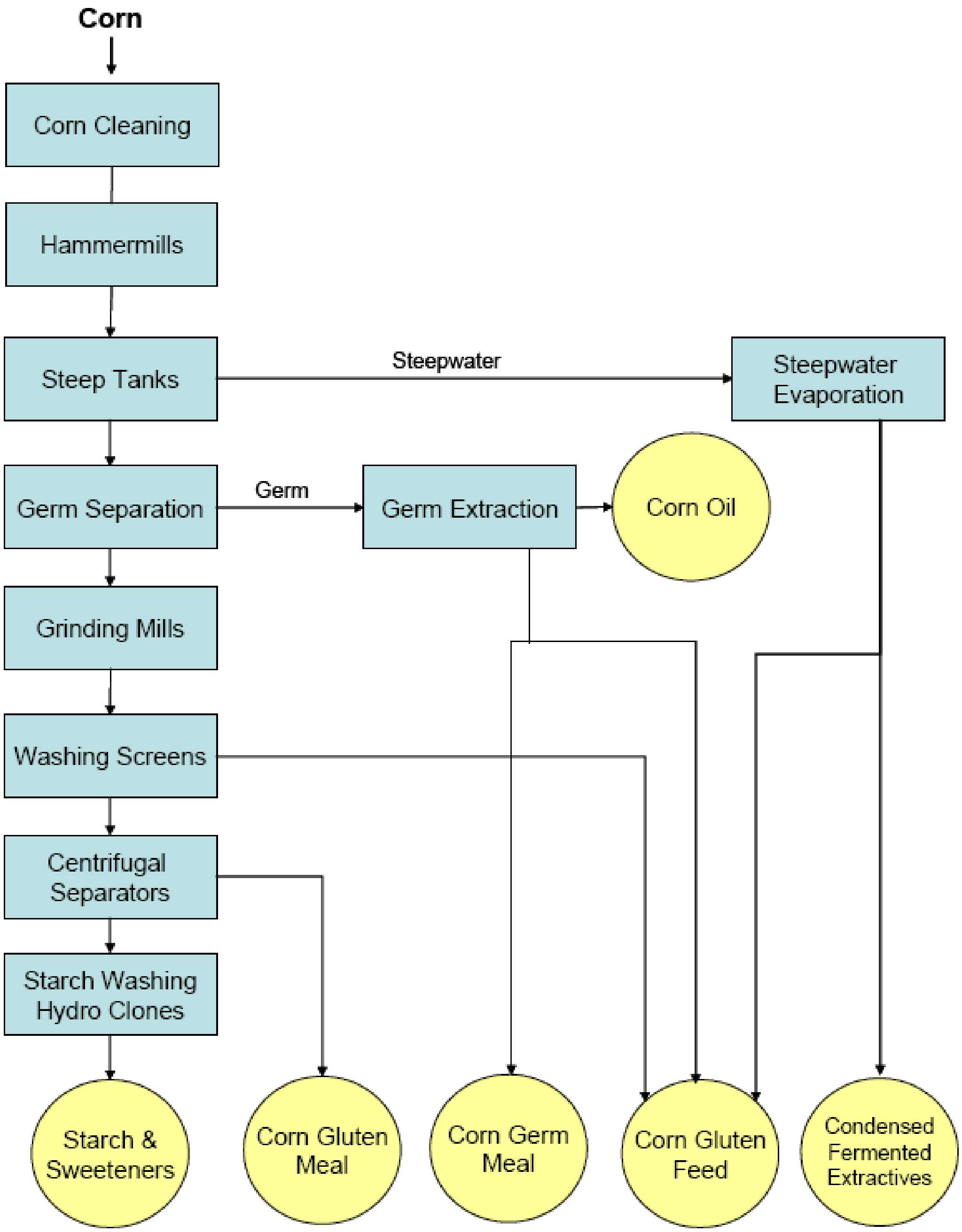
APPENDIX H-D

Additional Information - Grain to Ethanol Processes


Corn Dry-Grind Ethanol Milling Process Overview

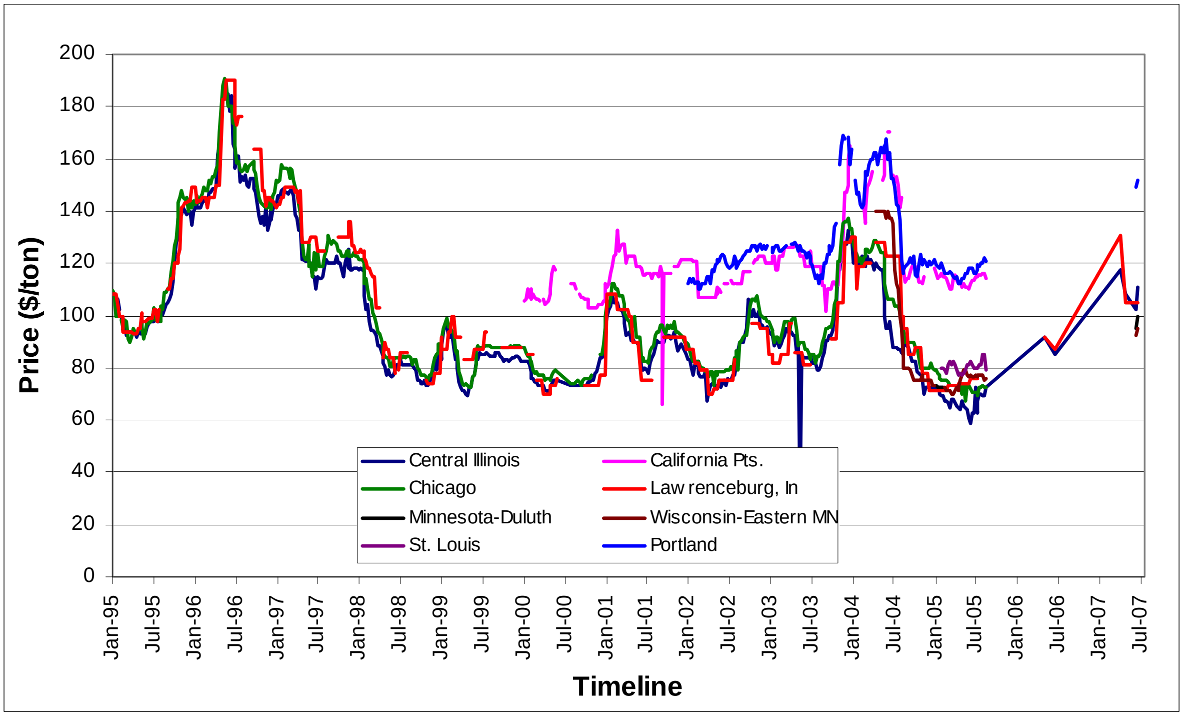
Source: Antares 2005a

Corn Wet Milling Process Overview



Source: Antares 2005a

DDGS 10-Year Historical Wholesale Price History[126]



Source: Antares 2005b and USDA AMS 2007

As shown in the graph above, DDGS prices have fluctuated significantly over the past decade, ranging from $60/ton to more than $170/ton. Current DDGS prices are around $90 to $115 per ton, not including delivery costs.

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[126] Note that the California and Portland prices are consistently higher than the other DDGS prices, as these numbers include the cost of rail delivery.

APPENDIX H-E

Oilseed Crop Feedstock Conversion to Oil

A typical method for conversion of oilseed feedstocks to oils is the n-hexane extraction process (Sheehan et al., 1998). Although alternate processes such as mechanical extraction are also available[127] hexane extraction has been selected for modeling in this analysis due to high oil yield and data availability.

CONVERSION

Table H-28 shows the percentage oil in selected feedstocks. In general, the oil content is 20-40% of the total feedstock. The rest of the oilseed consists of hulls, moisture, meal, and other solids. For example, a typical break-down of soybeans includes about 20% oil, 7% hulls, 14-18% moisture, and 57% meal (approximate weight percentages from Sheehan et al. 1998). A large percentage of this non-oil product, including meal and hulls, is sold as animal feed. This is critical by-product and revenue source for the crushing/extraction process.

Table H-28. Oil Content of Various Oilseed Feedstocks.

Feedstock	Oil Content (Wt%)
Soybean	19%
Canola	40%
Sunflower	40%
Rapeseed	40%
Safflower	25%

Based on data from Tyson et al. 2004

Sheehan et al. (1998) estimate that oil extraction can recover 98.5% of the oil in soybeans, including losses. The end product is pure degummed soybean oil ready for transesterification. For comparison, Tyson et al. (2004) suggest that the hexane extraction method can extract 99.5% of the oil from soybeans. However, this is the maximum extraction potential and does not fully adjust for losses.

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[127] New York State's oilseed extraction processes are mechanical, due in no small part to feedstock specifications from the dairies that purchase the resulting soy and canola meal (Personal communication with Ron Rausch,NYS Division of Agriculture and Markets, 11/4/2009). Due to the national prevalence of chemical extraction, insufficient data was available to model mechanical extraction.

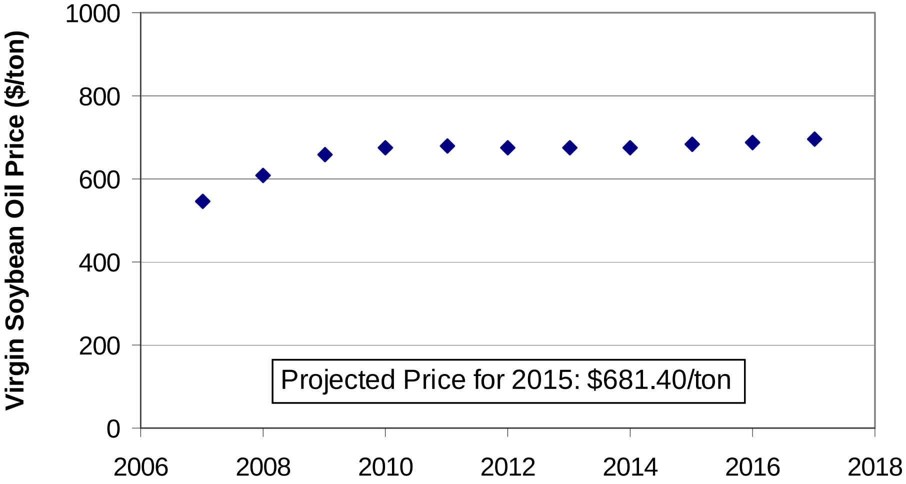
ENERGY REQUIREMENTS

The energy requirements for the extraction process includes: 63.19 kWh/ton of electricity; 958.8 MMBtu/ton of natural gas; and 792.2 MMBtu/ton of steam (Sheehan et al. 1998).[128] Furthermore, extraction requires 4.04 lb hexane and 0.77 gallons of water per ton of feedstock processed (Sheehan et al. 1998).

PROJECTED PRICES

According to the Food and Agricultural Policy Research Institute (FAPRI) 2007 U.S. and World Agricultural Outlook, soybean oil is expected to be the largest source of feedstock for biodiesel in the mid-term. For the 2015 time period, FAPRI estimates that 75% of biodiesel will be produced from soy oil, 16% from canola oil, and the remainder produced from other fats and oils. The FAPRI 2007 price projections for virgin soybean oil and canola oil in the U.S.are shown in Figure H-28 and Figure H-29, respectively. The projected soybean oil price is $681.40/ton in 2015. The projected price for canola oil in 2015 is $753.40/ton.

Figure H-28. Projected Soybean Oil Price.



Data from FAPRI 2007

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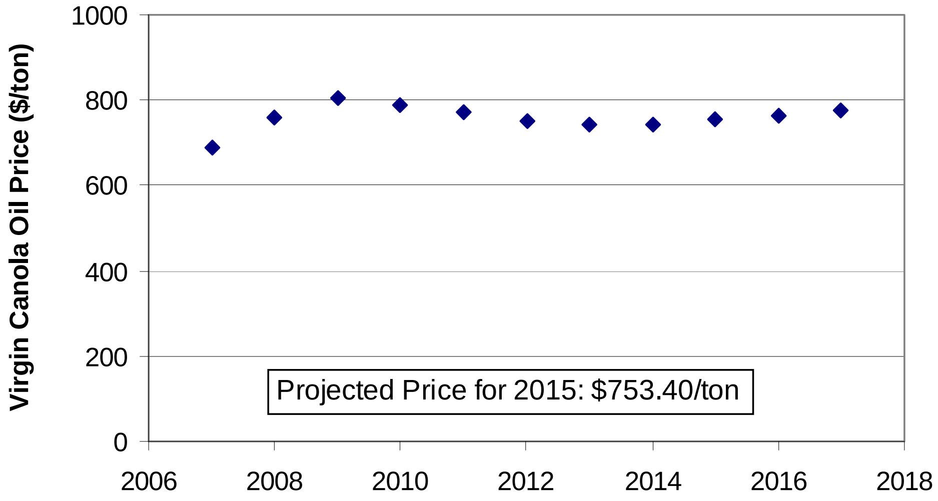
[128] All quantities given per ton of soybeans as received (i.e. with about 15% moisture). The natural gas is used for drying in the bean preparation step.

Figure H-29. Projected Canola Oil Price.



Data from FAPRI 2007

APPENDIX H-F

Sample Model Results

Both large and small-scale plant results are shown for each of the ten models developed according to the standards detailed in the introduction.

Summary Page - Costs & Yield Data, Grain Ethanol Production - *Large Scale Plant*

Inputs

Feedstock Type	**Grain**
Economic Lifetime of Plant (Years)	25
Weighted Cost of Capital	10%

Current Conversion Technology

	Dry Mill	**Wet Mill**
Applicable Feedstocks	Corn, Sorghum	Corn
Applicable Size Range (MGY) (1)	5 - 100	50 - 300
Feedstock Input (ton/yr)	1,000,000	3,359,462
Yield Ethanol (MGY) (2)	100	300
Conversion Efficiency (HHV) (3)	64%	57%
Consumables and By-Products (4)		
DDG Yield (ton/yr)	335,000	--
Corn Gluten Feed Yield (ton/yr)	--	683,890
Corn Gluten Meal Yield (ton/yr)	--	179,971
Corn Oil Yield (ton/yr)	--	95,985
CO2 Stream (ton/yr) (5)	312,500	937,350
Water Consumption (1000 gal/yr) (6)	470,000	7,330,826
Ethanol Production Costs		
Fixed O&M ($/yr)	$ 19,759,960	$ 73,834,528
Variable O&M ($/yr)	$ 40,699,298	$ 135,811,750
Co-product Credit ($/yr)	$ (46,732,500)	$ (231,611,269)
Operating Cost ($/gal/yr)	$ 0.1373	$ (0.0732)
Annual Operating Cost ($/yr)	***$ 13,726,758***	***$ (21,964,991)***
Total Capital Investment ($) (7)	***$ 198,133,767***	***$ 483,229,656***
Total Capital Investment ($/gal/yr)	$ 1.98	$ 1.61
Total Capital Investment ($/ton/yr)	$ 198.13	$ 143.84
Fixed Charge Rate	12.3%	12.3%
Non-Feedstock Production Cost ($/gal)	**$ 0.38**	**$ 0.12**

1) Based on typical sizes for wet and dry mills reported in Gallagher et al. 2005. Dry mill facilities have typically been 5 - 30 MGY, but current expansions are 40 - 100 MGY

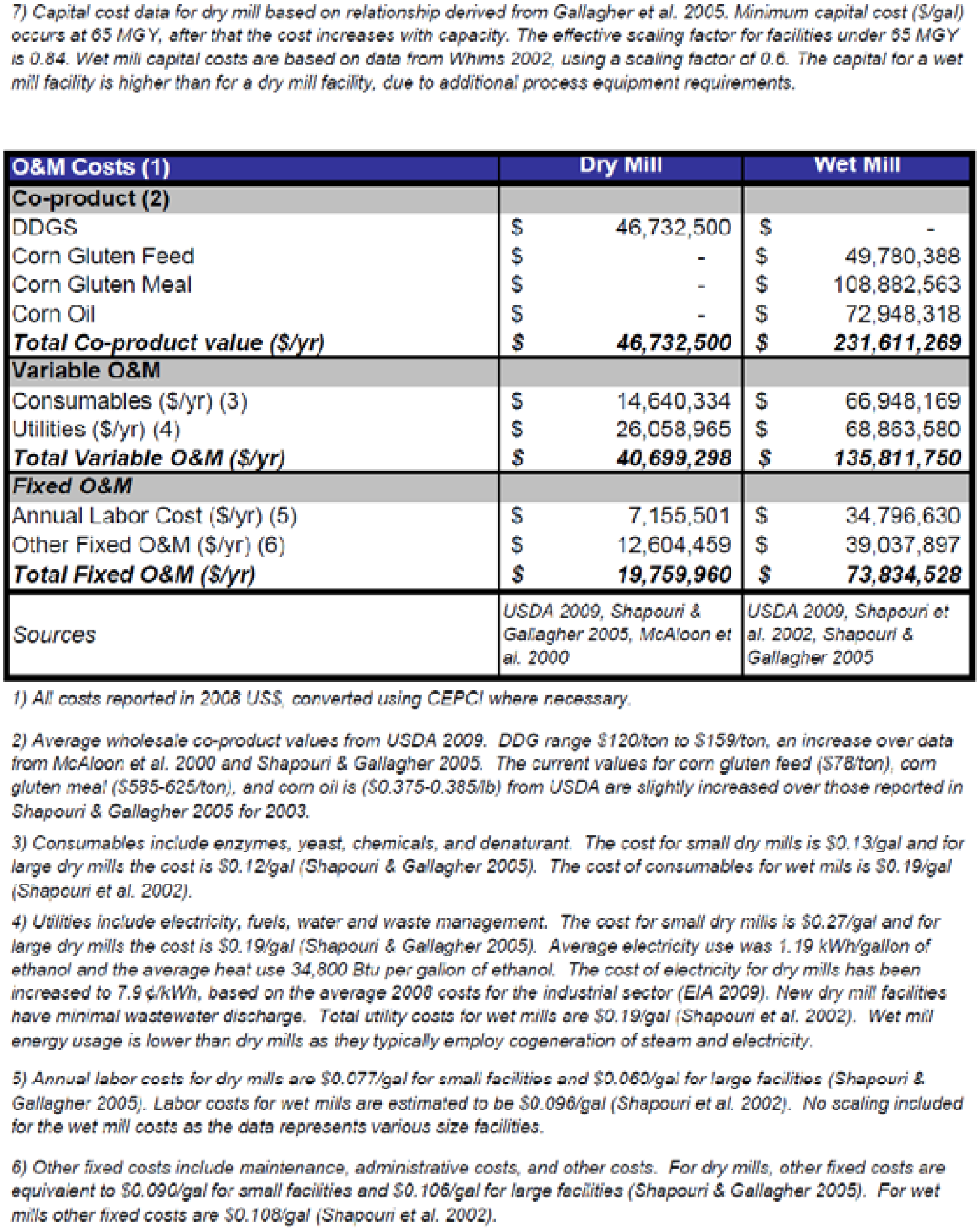
2) A conversion value of 2.8 gallons per bushel of corn (100 gal/ton) is used for dry mills, based on data from Shapouri & Gallagher 2005 for new facilities. The same study also reports averages for existing dry mill plants of 2.7 gal/bu for small facilities (>40 MGY), and 2.6 gal/ton for large facilities (40 - 100 MGY). In comparison, McAloon et al. 2000 uses a conversion of 2.71 gal/ton for dry mills. Wet mill facilities have slightly lower conversion to ethanol, around 2.5 gal/bu (89.3 gal/ton) (Butzen and Hobbs 2002). Note that the USDA standard weight for a bushel of corn is 56 lb.

3) Calculated using a typical ethanol heating value 84,000 Btu/gal, and corn heating value of 369,410 Btu/bushel (6,600 Btu/lb) for product with 15% moisture.

4) Dry mill facilities generate distiller's grains as a by-product. This analysis assumes all distiller's grains are converted to DDG. For comparison, Shapouri & Gallagher 2005 showed that 70% of distiller's grains were converted to DDG in the ethanol plants surveyed. It is estimated that dry mill facilities produce 6.7 lb DDG per gallon of ethanol produced (at 10% moisture content) (McAloon et al. 2000). Wet mills generate several by-products, including corn gluten feed, corn gluten meal, and corn oil. The yields of these products from one bushel of corn are 11.4 lb, 3 lb, and 1.6 lb, respectively (Butzen and Hobbs 2002).

5) CO2 production from fermentation is estimated to be 17.5 lb per bushel of corn for dry mill facilities, based on data from Antares 2005a and McAloon et al. 2000. No data available for wet mill CO2 production. The CO2 production for wet mills is based on the same value, but scaled by the ethanol conversion rate.

6) Average water requirement for dry mills is 4.7 gallons per gallon of ethanol (Shapouri & Gallagher 2005). Based on data from Shapouri & Gallagher 2002, wet mills use 5.2 times as much water as dry mills.



7) Capital cost data for dry mill based on relationship derived from Gallagher et al. 2005. Minimum capital cost ($/gal) occurs at 65 MGY, after that the cost increases with capacity. The effective scaling factor for facilities under 65 MGY is 0.84. Wet mill capital costs are based on data from Whims 2002, using a scaling factor of 0.6. The capital for a wet mill facility is higher than for a dry mill facility, due to additional process equipment requirements.

O&M Costs (1)	**Dry Mill**	**Wet Mill**
Co-product (2)		
DDGS	$ 46,732,500	$ -
Corn Gluten Feed	$ -	$ 49,780,388
Corn Gluten Meal	$ -	$ 108,882,563
Corn Oil	$ -	$ 72,948,318
Total Co-product value ($/yr)	***$ 46,732,500***	***$ 231,611,269***
Variable O&M		
Consumables ($/yr) (3)	$ 14,640,334	$ 66,948,169
Utilities ($/yr) (4)	$ 26,058,965	$ 68,863,580
Total Variable O&M ($/yr)	***$ 40,699,298***	***$ 135,811,750***
Fixed O&M		
Annual Labor Cost ($/yr) (5)	$ 7,155,501	$ 34,796,630
Other Fixed O&M ($/yr) (6)	$ 12,604,459	$ 39,037,897
Total Fixed O&M ($/yr)	***$ 19,759,960***	***$ 73,834,528***
Sources	*USDA 2009, Shapouri & Gallagher 2005, McAloon et al. 2000*	*USDA 2009, Shapouri et al. 2002, Shapouri & Gallagher 2005*

1) All costs reported in 2008 US$, converted using CEPCI where necessary.

2) Average wholesale co-product values from USDA 2009. DDG range $120/ton to $159/ton, an increase over data from McAloon et al. 2000 and Shapouri & Gallagher 2005. The current values for corn gluten feed ($78/ton), corn gluten meal ($585-625/ton), and corn oil is ($0.375-0.385/lb) from USDA are slightly increased over those reported in Shapouri & Gallagher 2005 for 2003.

3) Consumables include enzymes, yeast, chemicals, and denaturant. The cost for small dry mills is $0.13/gal and for large dry mills the cost is $0.12/gal (Shapouri & Gallagher 2005). The cost of consumables for wet mils is $0.19/gal (Shapouri et al. 2002).

4) Utilities include electricity, fuels, water and waste management. The cost for small dry mills is $0.27/gal and for large dry mills the cost is $0.19/gal (Shapouri & Gallagher 2005). Average electricity use was 1.19 kWh/gallon of ethanol and the average heat use 34,800 Btu per gallon of ethanol. The cost of electricity for dry mills has been increased to 7.9 ¢/kWh, based on the average 2008 costs for the industrial sector (EIA 2009). New dry mill facilities have minimal wastewater discharge. Total utility costs for wet mills are $0.19/gal (Shapouri et al. 2002). Wet mill energy usage is lower than dry mills as they typically employ cogeneration of steam and electricity.

5) Annual labor costs for dry mills are $0.077/gal for small facilities and $0.060/gal for large facilities (Shapouri & Gallagher 2005). Labor costs for wet mills are estimated to be $0.096/gal (Shapouri et al. 2002). No scaling included for the wet mill costs as the data represents various size facilities.

6) Other fixed costs include maintenance, administrative costs, and other costs. For dry mills, other fixed costs are equivalent to $0.090/gal for small facilities and $0.106/gal for large facilities (Shapouri & Gallagher 2005). For wet mills other fixed costs are $0.108/gal (Shapouri et al. 2002).

Summary Page - Costs & Yield Data, Grain Ethanol Production - *Small Scale Plant*

Inputs

Feedstock Type	**Grain**
Economic Lifetime of Plant (Years)	25
Weighted Cost of Capital	10%

Current Conversion Technology

	Dry Mill	**Wet Mill**
Applicable Feedstocks	Corn, Sorghum	Corn
Applicable Size Range (MGY) (1)	5 - 100	50 - 300
Feedstock Input (ton/yr)	50,000	559,910
Yield Ethanol (MGY) (2)	5	50
Conversion Efficiency (HHV) (3)	64%	57%
Consumables and By-Products (4)		
DDG Yield (ton/yr)	16,750	--
Corn Gluten Feed Yield (ton/yr)	--	113,982
Corn Gluten Meal Yield (ton/yr)	--	29,995
Corn Oil Yield (ton/yr)	--	15,997
CO2 Stream (ton/yr) (5)	15,625	156,225
Water Consumption (1000 gal/yr) (6)	23,500	1,221,805
Ethanol Production Costs		
Fixed O&M ($/yr)	$ 987,998	$ 12,305,756
Variable O&M ($/yr)	$ 2,545,758	$ 22,635,295
Co-product Credit ($/yr)	$ (2,336,625)	$ (38,601,884)
Operating Cost ($/gal/yr)	$ 0.2394	$ (0.0732)
Annual Operating Cost ($/yr)	***$ 1,197,131***	***$ (3,660,832)***
Total Capital Investment ($) (7)	***$ 14,958,677***	***$ 164,916,028***
Total Capital Investment ($/gal/yr)	$ 2.99	$ 3.30
Total Capital Investment ($/ton/yr)	$ 299.17	$ 294.54
Fixed Charge Rate	12.3%	12.3%
Non-Feedstock Production Cost ($/gal)	**$ 0.61**	**$ 0.33**

1) Based on typical sizes for wet and dry mills reported in Gallagher et al. 2005. Dry mill facilities have typically been 5 - 30 MGY, but current expansions are 40 - 100 MGY

2) A conversion value of 2.8 gallons per bushel of corn (100 gal/ton) is used for dry mills, based on data from Shapouri & Gallagher 2005 for new facilities. The same study also reports averages for existing dry mill plants of 2.7 gal/bu for small facilities (>40 MGY), and 2.6 gal/ton for large facilities (40 - 100 MGY). In comparison, McAloon et al. 2000 uses a conversion of 2.71 gal/ton for dry mills. Wet mill facilities have slightly lower conversion to ethanol, around 2.5 gal/bu (89.3 gal/ton) (Butzen and Hobbs 2002). Note that the USDA standard weight for a bushel of corn is 56 lb.

3) Calculated using a typical ethanol heating value 84,000 Btu/gal, and corn heating value of 369,410 Btu/bushel (6,600 Btu/lb) for product with 15% moisture.

4) Dry mill facilities generate distiller's grains as a by-product. This analysis assumes all distiller's grains are converted to DDG. For comparison, Shapouri & Gallagher 2005 showed that 70% of distiller's grains were converted to DDG in the ethanol plants surveyed. It is estimated that dry mill facilities produce 6.7 lb DDG per gallon of ethanol produced (at 10% moisture content) (McAloon et al. 2000). Wet mills generate several by-products, including corn gluten feed, corn gluten meal, and corn oil. The yields of these products from one bushel of corn are 11.4 lb, 3 lb, and 1.6 lb, respectively (Butzen and Hobbs 2002).

5) CO2 production from fermentation is estimated to be 17.5 lb per bushel of corn for dry mill facilities, based on data from Antares 2005a and McAloon et al. 2000. No data available for wet mill CO2 production. The CO2 production for wet mills is based on the same value, but scaled by the ethanol conversion rate.

6) Average water requirement for dry mills is 4.7 gallons per gallon of ethanol (Shapouri & Gallagher 2005). Based on data from Shapouri & Gallagher 2002, wet mills use 5.2 times as much water as dry mills.

7) Capital cost data for dry mill based on relationship derived from Gallagher et al. 2005. Minimum capital cost ($/gal) occurs at 65 MGY, after that the cost increases with capacity. The effective scaling factor for facilities under 65 MGY is 0.84. Wet mill capital costs are based on data from Whims 2002, using a scaling factor of 0.6. The capital for a wet mill facility is higher than for a dry mill facility, due to additional process equipment requirements.

O&M Costs (1)	**Dry Mill**	**Wet Mill**
Co-product (2)		
DDGS	$ 2,336,625	$ -
Corn Gluten Feed	$ -	$ 8,296,733
Corn Gluten Meal	$ -	$ 18,147,096
Corn Oil	$ -	$ 12,158,055
Total Co-product value ($/yr)	***$ 2,336,625***	***$ 38,601,884***
Variable O&M		
Consumables ($/yr) (3)	$ 750,729	$ 11,158,030
Utilities ($/yr) (4)	$ 1,795,029	$ 11,477,265
Total Variable O&M ($/yr)	***$ 2,545,758***	***$ 22,635,295***
Fixed O&M		
Annual Labor Cost ($/yr) (5)	$ 455,826	$ 5,799,439
Other Fixed O&M ($/yr) (6)	$ 532,172	$ 6,506,317
Total Fixed O&M ($/yr)	***$ 987,998***	***$ 12,305,756***
Sources	*USDA 2009, Shapouri & Gallagher 2005, McAloon et al. 2000*	*USDA 2009, Shapouri et al. 2002, Shapouri & Gallagher 2005*

1) All costs reported in 2008 US$, converted using CEPCI where necessary.

2) Average wholesale co-product values from USDA 2009. DDG range $120/ton to $159/ton, an increase over data from McAloon et al. 2000 and Shapouri & Gallagher 2005. The current values for corn gluten feed ($78/ton), corn gluten meal ($585-625/ton), and corn oil is ($0.375-0.385/lb) from USDA are slightly increased over those reported in Shapouri & Gallagher 2005 for 2003.

3) Consumables include enzymes, yeast, chemicals, and denaturant. The cost for small dry mills is $0.13/gal and for large dry mills the cost is $0.12/gal (Shapouri & Gallagher 2005). The cost of consumables for wet mils is $0.19/gal (Shapouri et al. 2002).

4) Utilities include electricity, fuels, water and waste management. The cost for small dry mills is $0.27/gal and for large dry mills the cost is $0.19/gal (Shapouri & Gallagher 2005). Average electricity use was 1.19 kWh/gallon of ethanol and the average heat use 34,800 Btu per gallon of ethanol. The cost of electricity for dry mills has been increased to 7.9 ¢/kWh, based on the average 2008 costs for the industrial sector (EIA 2009). New dry mill facilities have minimal wastewater discharge. Total utility costs for wet mills are $0.19/gal (Shapouri et al. 2002). Wet mill energy usage is lower than dry mills as they typically employ cogeneration of steam and electricity.

5) Annual labor costs for dry mills are $0.077/gal for small facilities and $0.060/gal for large facilities (Shapouri & Gallagher 2005). Labor costs for wet mills are estimated to be $0.096/gal (Shapouri et al. 2002). No scaling included for the wet mill costs as the data represents various size facilities.

6) Other fixed costs include maintenance, administrative costs, and other costs. For dry mills, other fixed costs are equivalent to $0.090/gal for small facilities and $0.106/gal for large facilities (Shapouri & Gallagher 2005). For wet mills other fixed costs are $0.108/gal (Shapouri et al. 2002).

Summary Page - Costs & Yield Data, Future Black Liquor Gasification Biorefining - *Small and Large Scale Plants*

Please note that this model yields both incremental and total cost and material values. Incremental values can be applied to the case of retrofitting a gasifier to an existing pulp mill, or total values to the case of constructing a complete facility. All costs noted are scaled to 2006 via CEPCI unless otherwise noted. Also, due to the lack of any demonstration plants in operation, data is based on a detailed series of AspenPlus simulations by Larson et al, 2006. It examines two types of end products--dimethyl ether (DME) and Fischer Tropsch Liquids (FTL)--each achieved in three different plant configurations. The biofuel product yields are stated as undenatured, without additives. The FTL case assumes no on-site refining apart from removal of light gases. The cost of refining and transportation to an oil refinery are not included in this model.

Inputs

Feedstock Type: Percent of Hardwood in Feed	100
Plant Type	FTL (medium gas turbine, all syngas used as fuel feedstock)
Economic Lifetime of Plant (Years) (1)	25
Weighted Cost of Capital (2)	10%

1), 2) These assumptions were made by Larson et al, 2006 in calculating operating cost. Updating these cells does not automatically.update the cost information below.

Sample Model Results				
Fuel Type	FTL	FTL	DME	DME
Feedstock	Hardwoods	Hardwoods	Hardwoods	Hardwoods
Feedstock Input (ton/yr)	750,000	250,000	750,000	250,000
Yield Biofuel(MGY)	30.3	10.1	37.5	12.5
Conversion Efficiency (HHV)	38.43%	38.43%	25.47%	25.47%
Displaced Fuel Type	Crude Oil	Crude Oil	Diesel	Diesel
Consumables and By-Products				
Pulp Yield (dry ton/yr)	193,222	64,407	299,821	99,940
Electricity Exported to Grid (MWh/yr)	(65,609)	(21,870)	(447,515)	(149,172)
Incremental Bioethanol Production Costs				
Annual Operating Cost ($/yr)	***$ 8,228,257***	***$ 2,742,752***	***$ 6,919,551***	***$ 2,306,517***
Incremental Annual Operating Cost ($/yr)	***$ 4,284,379***	***$ 1,428,126***	***$ 2,975,673***	***$ 991,891***
Total Capital Investment ($)	***$ 313,080,448***	***$ 161,950,836***	***$ 220,849,359***	***$ 114,241,367***
Incremental Capital Investment ($)	***$ 189,840,156***	***$ 98,200,869***	***$ 97,609,067***	***$ 50,491,400***
Non-Feedstock Production Cost, Entire Plant ($/gal)	**$ 1.54**	**$ 2.24**	**$ 0.91**	**$ 1.31**
Non-Feedstock Production Retrofitted Gasifier ($/gal)	**$ 0.91**	**$ 1.33**	**$ 0.40**	**$ 0.58**

1) This feedstock input includes both undebarked pulp feedstock and wood residuals purchased solely for boiler fuel.

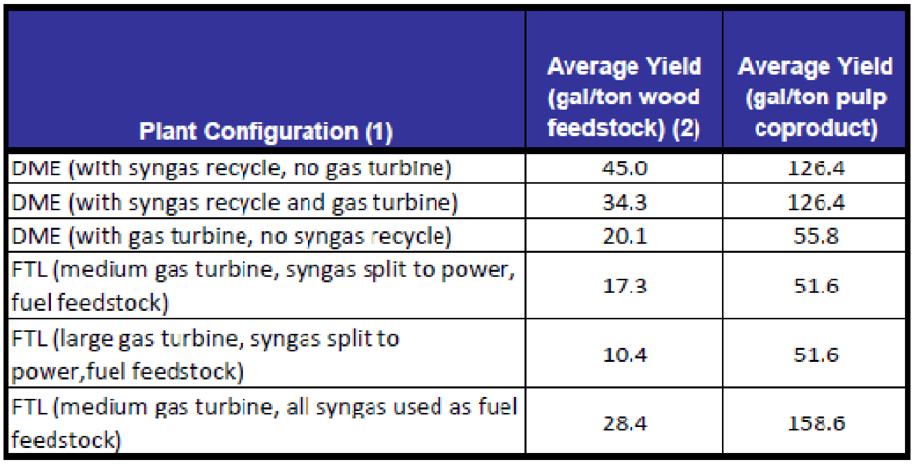
2) Calculated based on the total amount of feedstock input, including forest residuals purchased for fuel. Total annual fuel energy production for each plant configuration was scaled from Larson et al, 2006. A hardwood HHV of 7817 BTU/lb and a softwood HHV of 8127 BTU/ lb were calculated from Yin, 2006, for a 50% moisture content.

3) Assumes 48.75 wt% (softwood) and 49.75 wt% (hardwood) digester yields, as per Larson et al, 2006.

4) Applies to entire Kraft mill, not fuel production exclusively. Assumed to be 4% of total installed cost as per Larson et al, 2006, but inflated from Larson's original 2005$ estimate with 2.67% annual inflation as per the authors' original study. The Incremental operating cost compares total facility to a modern Tomlinson boiler-driven Kraft mill of the same pulp capacity, with no biofuel production.

5) A scaling factor of 0.6 was used to extrapolate from capital cost estimates by Larson et al, 2006.

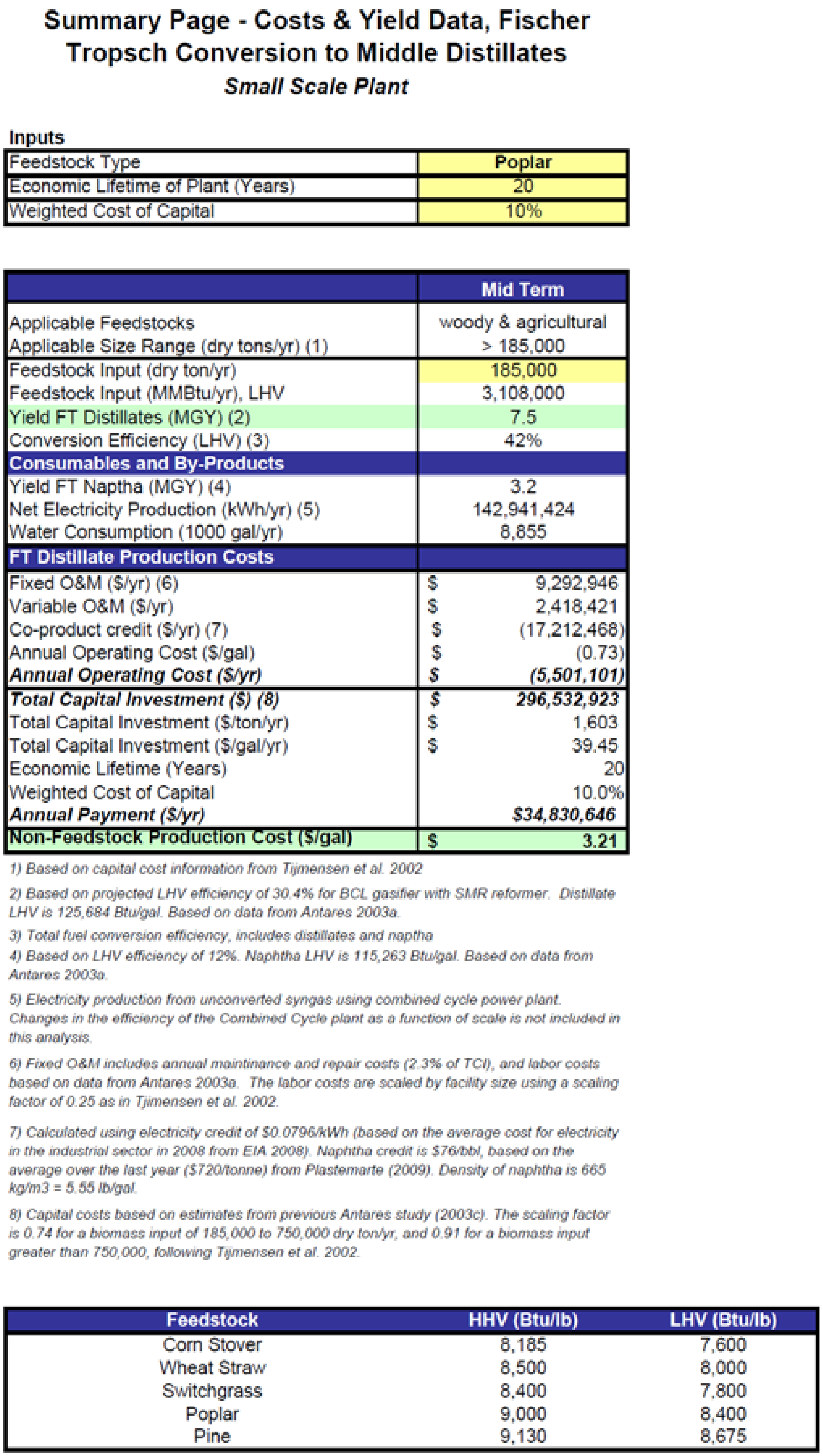
6) The entire production cost is charged to the biofuel produced, rather than to both the pulp and the biofuel; the incremental production cost follows.



Plant Configuration (1)	Average Yield (gal/ton wood feedstock) (2)	Average Yield (gal/ton pulp coproduct)
DME (with syngas recycle, no gas turbine)	45.0	126.4
DME (with syngas recycle and gas turbine)	34.3	126.4
DME (with gas turbine, no syngas recycle)	20.1	55.8
FTL (medium gas turbine, syngas split to power, fuel feedstock)	17.3	51.6
FTL (large gas turbine, syngas split to power,fuel feedstock)	10.4	51.6
FTL (medium gas turbine, all syngas used as fuel feedstock)	28.4	158.6

1) Assumes 65:35 distribution of hardwood to softwood feedstock by mass. Calculated from Larson et al, 2006.

2) Calculated based on the total amount of feedstock input, including forest residuals purchased for fuel.



Summary Page - Costs & Yield Data, Fischer Tropsch Conversion to Middle Distillates

Small Scale Plant

Inputs

Feedstock Type	**Poplar**
Economic Lifetime of Plant (Years)	20
Weighted Cost of Capital	10%

	Mid Term
Applicable Feedstocks	woody & agricultural
Applicable Size Range (dry tons/yr) (1)	> 185,000
Feedstock Input (dry ton/yr)	185,000
Feedstock Input (MMBtu/yr), LHV	3,108,000
Yield FT Distillates (MGY) (2)	7.5
Conversion Efficiency (LHV) (3)	42%
Consumables and By-Products	
Yield FT Naptha (MGY) (4)	3.2
Net Electricity Production (kWh/yr) (5)	142,941,424
Water Consumption (1000 gal/yr)	8,855
FT Distillate Production Costs	
Fixed O&M ($/yr) (6)	$ 9,292,946
Variable O&M ($/yr)	$ 2,418,421
Co-product credit ($/yr) (7)	$ (17,212,468)
Annual Operating Cost ($/gal)	$ (0.73)
Annual Operating Cost ($/yr)	***$ (5,501,101)***
Total Capital Investment ($) (8)	***$ 296,532,923***
Total Capital Investment ($/ton/yr)	$ 1,603
Total Capital Investment ($/gal/yr)	$ 39.45
Economic Lifetime (Years)	20
Weighted Cost of Capital	10.0%
Annual Payment ($/yr)	***$34,830,646***
Non-Feedstock Production Cost ($/gal)	$ 3.21

1) Based on capital cost information from Tijmensen et al. 2002

2) Based on projected LHV efficiency of 30.4% for BCL gasifier with SMR reformer. Distillate LHV is 125,684 Btu/gal. Based on data from Antares 2003a.

3) Total fuel conversion efficiency, includes distillates and naptha

4) Based on LHV efficiency of 12%. Naphtha LHV is 115,263 Btu/gal. Based on data from Antares 2003a.

5) Electricity production from unconverted syngas using combined cycle power plant. Changes in the efficiency of the Combined Cycle plant as a function of scale is not included in this analysis.

6) Fixed O&M includes annual maintinance and repair costs (2.3% of TCI), and labor costs based on data from Antares 2003a. The labor costs are scaled by facility size using a scaling factor of 0.25 as in Tjimensen et al. 2002.

7) Calculated using electricity credit of $0.0796/kWh (based on the average cost for electricity in the industrial sector in 2008 from EIA 2008). Naphtha credit is $76/bbl, based on the average over the last year ($720/tonne) from Plastemarte (2009). Density of naphtha is 665 kg/m3 = 5.55 lb/gal.

8) Capital costs based on estimates from previous Antares study (2003c). The scaling factor is 0.74 for a biomass input of 185,000 to 750,000 dry ton/yr, and 0.91 for a biomass input greater than 750,000, following Tijmensen et al. 2002.

Feedstock	HHV (Btu/lb)	LHV (Btu/lb)
Corn Stover	8,185	7,600
Wheat Straw	8,500	8,000
Switchgrass	8,400	7,800
Poplar	9,000	8,400
Pine	9,130	8,675

Summary Page - Costs & Yield Data, Fischer Tropsch Conversion to Middle Distillates

Large Scale Plant

Inputs

Feedstock Type	**Poplar**
Economic Lifetime of Plant (Years)	20
Weighted Cost of Capital	10%

	Mid Term
Applicable Feedstocks	woody & agricultural
Applicable Size Range (dry tons/yr) (1)	> 185,000
Feedstock Input (dry ton/yr)	1,500,000
Feedstock Input (MMBtu/yr), LHV	25,200,000
Yield FT Distillates (MGY) (2)	61.0
Conversion Efficiency (LHV) (3)	42%
Consumables and By-Products	
Yield FT Naptha (MGY) (4)	26.2
Net Electricity Production (kWh/yr) (5)	1,158,984,518
Water Consumption (1000 gal/yr)	71,798
FT Distillate Production Costs	
Fixed O&M ($/yr) (6)	$ 48,792,011
Variable O&M ($/yr)	$ 19,608,821
Co-product credit ($/yr) (7)	$ (139,560,548)
Annual Operating Cost ($/gal)	$ (1.17)
Annual Operating Cost ($/yr)	***$ (71,159,715)***
Total Capital Investment ($) (8)	***$ 1,604,819,355***
Total Capital Investment ($/ton/yr)	$ 1,070
Total Capital Investment ($/gal/yr)	$ 26.33
Economic Lifetime (Years)	20
Weighted Cost of Capital	10.0%
Annual Payment ($/yr)	***$188,501,479***
Non-Feedstock Production Cost ($/gal)	**$ 1.47**

1) Based on capital cost information from Tijmensen et al. 2002

2) Based on projected LHV efficiency of 30.4% for BCL gasifier with SMR reformer. Distillate LHV is 125,684 Btu/gal. Based on data from Antares 2003a.

3) Total fuel conversion efficiency, includes distillates and naptha

4) Based on LHV efficiency of 12%. Naphtha LHV is 115,263 Btu/gal. Based on data from Antares 2003a.

5) Electricity production from unconverted syngas using combined cycle power plant. Changes in the efficiency of the Combined Cycle plant as a function of scale is not included in this analysis.

6) Fixed O&M includes annual maintinance and repair costs (2.3% of TCI), and labor costs based on data from Antares 2003a. The labor costs are scaled by facility size using a scaling factor of 0.25 as in Tjimensen et al. 2002.

7) Calculated using electricity credit of $0.0796/kWh (based on the average cost for electricity in the industrial sector in 2008 from EIA 2008). Naphtha credit is $76/bbl, based on the average over the last year ($720/tonne) from Plastemarte (2009). Density of naphtha is 665 kg/m3 = 5.55 lb/gal.

8) Capital costs based on estimates from previous Antares study (2003c). The scaling factor is 0.74 for a biomass input of 185,000 to 750,000 dry ton/yr, and 0.91 for a biomass input greater than 750,000, following Tijmensen et al. 2002.

Feedstock	HHV (Btu/lb)	LHV (Btu/lb)
Corn Stover	8,185	7,600
Wheat Straw	8,500	8,000
Switchgrass	8,400	7,800
Poplar	9,000	8,400
Pine	9,130	8,675

Summary Page - Costs & Yield Data, Lignocellulosic Biobutanol Production

Small Scale Plant

The model--for a greenfield, stand-alone lignocellulosic n-butanol plant--is based on three detailed economic assessments of butanol and ethanol production: Aden et al's (2000) lignocellulosic ethanol model, Wu et al's (2007) corn ethanol assessment, and McAloon et al's (1999) lignocellulosic ethanol simulation. All costs were inflated to 2008$ using the CEPCI index. Yields were calculated based on the relative amounts of hemicellulose and cellulose in each feedstock, as fermented with C. beijerinckii P260 *as per Qureshi et al (2008).*

Inputs

Feedstock Type	Switchgrass
Economic Lifetime of Plant (Years)	25
Weighted Cost of Capital	10%

		Estimate for Future Plants
Applicable Feedstocks		Cellulosic Biomass: Wheat Straw, Bagasse, Mixed Paper, Hybrid Poplar, Switchgrass
Applicable Size Range (MGY)		25-75 MGY
Feedstock Input (ton/yr) (1)		515,000
Yield Biobutanol (ton/yr)		84,432
Yield Biobutanol (MGY) (2)		25.0
Conversion Efficiency (HHV) (3)		30.8%
Consumables and By-Products		
Acetone Yield (99.5%; tons/yr)		37,644
Ethanol Yield (99%; tons/yr)		1,996
Carbon Dioxide Emissions (tons/yr) (4)		230,670
Biobutanol Production Costs		
Fixed O&M ($/yr) (5)	$	51,306,288
Variable O&M ($/yr; does not include feedstock cost)	$	9,231,782
Variable O&M ($/gal)	$	0.369
Byproduct Credit ($/yr) (6)	$	35,091,939
Annual Operating Cost ($/yr)	***$***	***25,446,131***
Total Capital Investment ($) (7)	***$***	***205,673,209***
Unit Capital Cost ($/gal/yr)	***$***	***8.23***
Unit Capital Cost ($/ton feed/yr)	***$***	***399***
Fixed Charge Rate		12%
Non-Feedstock Production Cost ($/gal)	$	2.03

1) Dry tons; pricing and materials costs reflect purchasing at assumed 30% moisture.

2) This yield is calculated based on laboratory studies on fermentation of wheat straw published by Qureshi et al, 2006. The fermentation is assumed to be fed-batch with gas stripping to remove ABE, and is performed with C. beijerinckii P260 at an assumed yield of 0.44 g ABE/g wheat straw, and productivity of 0.36 g/L/hr. Relative ABE productivities and product distributions from hemicellulose and cellulose were calculated from this laboratory data and used to calculate the expected yields from the other listed feedstocks.

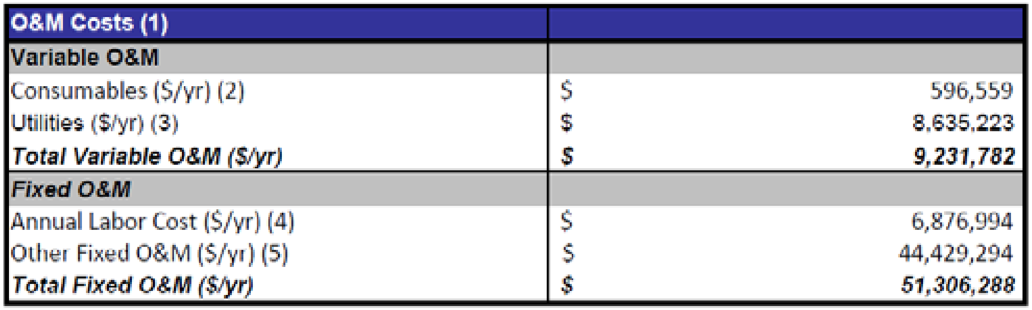
3) Based on the HHV data tabulated to the right, on dry mass basis. Only the desired liquid fuel products--butanol and ethanol--are counted toward the output energy value. Acetone is considered a byproduct for the purposes of this study. N-butanol HHV: 37.5 MJ/kg. Ethanol HHV: 29.2 MJ/kg. (Qureshi et al, 2000)

4) CO2 emissions are estimated based on stoichiometric production of CO2 during fermentation. Boiler emissions are not included in this value.

5) Operating costs were scaled from a detailed economic of a corn butanol plant by Wu et al (2007), either by feedstock or product volume as was most applicable.

6) Acetone was given a value of $900/ton as a solvent, ethanol a value of $2.00/gallon as a fuel, based on market trends.

7) Capital cost data was drawn from three detailed economic assessments of butanol and ethanol production. Front-end major equipment costs were individually scaled from a study by Aden et al (2000) of a lignocellulosic ethanol plant, on the assumption that feed handling and pretreatment should be comparable for all lignocellulosic feedstocks, regardless of downstream fermentation type. Saccharification and later process step equipment were modeled after a corn butanol cost study done by Wu et al (2000). Fermentation volume requirements were assumed to be the same for both types of feedstock, since residence time and butanol tolerance were similar for the two cases. When no more detailed capital scaling factor was available, 0.6 was used as per McAloon et al (1999). All costs were inflated to 2008$ using the CEPCI index listed in the sheet "Detailed Calculations." A Lang factor of 3.15 was used, based on the implied Lang factor used by Aden et al (2002), who used installation factors and contingincy factors to generate a total installed cost. This was in good agreement with the effective Lang factor generated from the major equipment roster by the method of Peters and Timmerhaus (1980), which was 3.08. The larger result was used in order to give a more conservative estimate.



O&M Costs (1)		
Variable O&M		
Consumables ($/yr) (2)	$	596,559
Utilities ($/yr) (3)	$	8,635,223
Total Variable O&M ($/yr)	***$***	***9,231,782***
Fixed O&M		
Annual Labor Cost ($/yr) (4)	$	6,876,994
Other Fixed O&M ($/yr) (5)	$	44,429,294
Total Fixed O&M ($/yr)	***$***	***51,306,288***

1) All costs reported in 2008 US$. Where necessary, costs have been converted using CEPCI.

2) Consumables include sulfuric acid, buffers, nutrients, lime, enzymes, and water treatment chemicals. Utilities include natural gas, makeup water, wastewater treatment, and electricity. Feedstock prices are not included in this value, but typical prices were estimated from the Western N.C. Timber Report, the online pulp and paper exchange, and agricultural studies. These are noted in "Detailed Calculations" sheet.

3) Other fixed O&M costs include taxes, insurance, and supplies, and are estimated as in Wu et al (2007).

Summary Page - Costs & Yield Data, Lignocellulosic Biobutanol Production

Large Scale Plant

The model--for a greenfield, stand-alone lignocellulosic n-butanol plant--is based on three detailed economic assessments of butanol and ethanol production: Aden et al's (2000) lignocellulosic ethanol model, Wu et al's (2007) corn ethanol assessment, and McAloon et al's (1999) lignocellulosic ethanol simulation. All costs were inflated to 2008$ using the CEPCI index. Yields were calculated based on the relative amounts of hemicellulose and cellulose in each feedstock, as fermented with C. beijerinckii P260 as per Qureshi et al (2008).

Inputs

Feedstock Type	Switchgrass
Economic Lifetime of Plant (Years)	25
Weighted Cost of Capital	10%

	Estimate for Future Plants	
Applicable Feedstocks	Cellulosic Biomass: Wheat Straw, Bagasse, Mixed Paper, Hybrid Poplar, Switchgrass	
Applicable Size Range (MGY)		25-75 MGY
Feedstock Input (ton/yr) (1)		1,545,000
Yield Biobutanol (ton/yr)		253,296
Yield Biobutanol (MGY) (2)		75.0
Conversion Efficiency (HHV) (3)		30.8%
Consumables and By-Products		
Acetone Yield (99.5%; tons/yr)		112,931
Ethanol Yield (99%; tons/yr)		5,988
Carbon Dioxide Emissions (tons/yr) (4)		692,010
Biobutanol Production Costs		
Fixed O&M ($/yr) (5)	$	108,870,521
Variable O&M ($/yr; does not include feedstock cost)	$	27,695,347
Variable O&M ($/gal)	$	0.369
Byproduct Credit ($/yr) (6)	$	105,275,817
Annual Operating Cost ($/yr)	***$***	***31,290,051***
Total Capital Investment ($) (7)	***$***	***401,093,356***
Unit Capital Cost ($/gal/yr)	***$***	***5.35***
Unit Capital Cost ($/ton feed/yr)	***$***	***260***
Fixed Charge Rate		12%
Non-Feedstock Production Cost ($/gal)	$	1.07

1) Dry tons; pricing and materials costs reflect purchasing at assumed 30% moisture.

2) This yield is calculated based on laboratory studies on fermentation of wheat straw published by Qureshi et al, 2006. The fermentation is assumed to be fed-batch with gas stripping to remove ABE, and is performed with C. beijerinckii P260 at an assumed yield of 0.44 g ABE/g wheat straw, and productivity of 0.36 g/L/hr. Relative ABE productivities and product distributions from hemicellulose and cellulose were calculated from this laboratory data and used to calculate the expected yields from the other listed feedstocks.

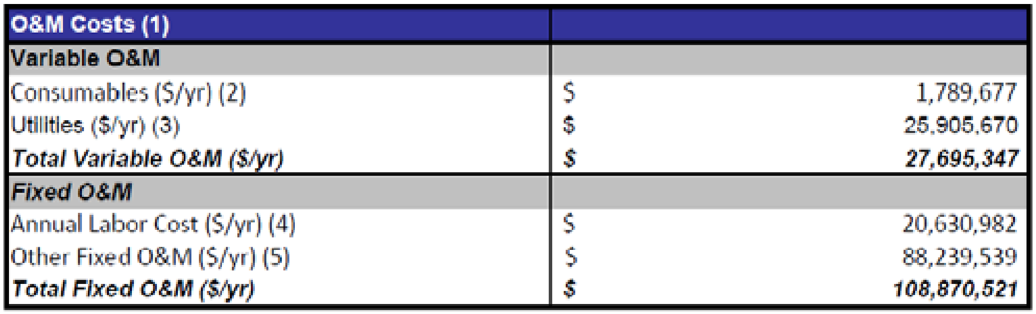
3) Based on the HHV data tabulated to the right, on dry mass basis. Only the desired liquid fuel products--butanol and ethanol--are counted toward the output energy value. Acetone is considered a byproduct for the purposes of this study. N-butanol HHV: 37.5 MJ/kg. Ethanol HHV: 29.2 MJ/kg. (Qureshi et al, 2000)

4) CO2 emissions are estimated based on stoichiometric production of CO2 during fermentation. Boiler emissions are not included in this value.

5) Operating costs were scaled from a detailed economic of a corn butanol plant by Wu et al (2007), either by feedstock or product volume as was most applicable.

6) Acetone was given a value of $900/ton as a solvent, ethanol a value of $2.00/gallon as a fuel, based on market trends.

7) Capital cost data was drawn from three detailed economic assessments of butanol and ethanol production. Front-end major equipment costs were individually scaled from a study by Aden et al (2000) of a lignocellulosic ethanol plant, on the assumption that feed handling and pretreatment should be comparable for all lignocellulosic feedstocks, regardless of downstream fermentation type. Saccharification and later process step equipment were modeled after a corn butanol cost study done by Wu et al (2000). Fermentation volume requirements were assumed to be the same for both types of feedstock, since residence time and butanol tolerance were similar for the two cases. When no more detailed capital scaling factor was available, 0.6 was used as per McAloon et al (1999). All costs were inflated to 2008$ using the CEPCI index listed in the sheet "Detailed Calculations." A Lang factor of 3.15 was used, based on the implied Lang factor used by Aden et al (2002), who used installation factors and contingincy factors to generate a total installed cost. This was in good agreement with the effective Lang factor generated from the major equipment roster by the method of Peters and Timmerhaus (1980), which was 3.08. The larger result was used in order to give a more conservative estimate.



O&M Costs (1)		
Variable O&M		
Consumables ($/yr) (2)	$	1,789,677
Utilities ($/yr) (3)	$	25,905,670
Total Variable O&M ($/yr)	***$***	***27,695,347***
Fixed O&M		
Annual Labor Cost ($/yr) (4)	$	20,630,982
Other Fixed O&M ($/yr) (5)	$	88,239,539
Total Fixed O&M ($/yr)	***$***	***108,870,521***

1) All costs reported in 2008 US$. Where necessary, costs have been converted using CEPCI.

2) Consumables include sulfuric acid, buffers, nutrients, lime, enzymes, and water treatment chemicals. Utilities include natural gas, makeup water, wastewater treatment, and electricity. Feedstock prices are not included in this value, but typical prices were estimated from the Western N.C. Timber Report, the online pulp and paper exchange, and agricultural studies. These are noted in "Detailed Calculations" sheet.

3) Other fixed O&M costs include taxes, insurance, and supplies, and are estimated as in Wu et al (2007).

Feedstocks and Yields

Feedstock (1)	Average Yield (gal BuOH/ton feedstock)	Average Yield Coproducts (tons/ton feedstock)
Switchgrass	48.5	0.0770
Hybrid Poplar	51.9	0.0859
Pine	55.0	0.0901
Mixed Paper	51.3	0.0863
Wheat Straw	56.5	0.0764

1) Feedstock characteristics from EERE biomass database (2009). Yields are estimates based on hydrolysis and theoretical fermentation efficiencies reported by Qureshi et al (2007).

Feedstock Composition

Feedstock	Hemicellulose (wt%)	Cellulose (wt%)	Lignin (wt%)	HHV (BTU/lb) (1)
Agricultural Resources				
Corn Stover (2)	23.23%	36.20%	18.50%	8,185
Wheat Straw (2)	23.20%	33.47%	17.28%	8,500
Bagasse (2)	25.79%	40.37%	23.90%	8,500
Grass Resources				
Switchgrass (3)	25.19%	31.98%	18.13%	8,400
Woody Resources				
Hybrid poplar (3)	18.55%	44.70%	26.44%	9,000
Black locust (3)	17.66%	41.61%	26.70%	9,000
Eucalyptus (3)	13.07%	49.50%	27.71%	9,000
Pine (3)	21.90%	44.55%	27.67%	9,130
Other Resources				
Yard Trimmings / Green Waste (4)	31.80%	20.00%	20.60%	8,500
Mixed Paper 4)	15.20%	48.10%	22.60%	8,200

1) Higher Heating Value (HHV) given for dry feedstocks. Data from (Energy Research Centre of the Netherlands, n.d.).

2) Biochemical composition data from U.S. DOE Biomass Feedstock Composition and Property Database, adjusted to 100% mass closure

3) Biochemical composition data from Hamelinck et al. 2005

4) Biochemical composition data from Energy Research Centre of the Netherlands (n.d.). Yard trimmings consists of 60% woody residues and 40% herbaceous residues (mix of leaves and grass), based on data from McKeever and Skog (2003). Yard trimmings theoretical ethanol yield based on average of similar materials.

Summary Page - Costs & Yield Data, Lignocellulosic Ethanol Production

Small Scale Plant

Inputs

Feedstock Type	**Switchgrass**
Economic Lifetime of Plant (Years)	25
Weighted Cost of Capital	10%

	Short Term (2010)	**Mid Term (2015-2020)**	**Long Term (2025+)**
Applicable Feedstocks	clean herbaceous feedstocks (agricultural residues and grasses) and de-barked wood chips		
Applicable Size Range (MGY)	25 - 60	60 - 150	60 - 300
Feedstock Input (dry ton/yr)	357,083	778,452	731,208
Feedstock Input (MMBtu/yr)	5,999,000	13,077,987	12,284,300
Yield - Dilute Acid (gal/dry ton)	70.0	77.4	82.6
Yield - Advanced Tech (gal/dry ton)	--	74.4	78.7
Conversion efficiency (HHV)	35%	39%	41%
Ethanol Production (MGY)	**25.0**	**60**	**60**
Pretreatment Technology	Dilute Acid	Dilute Acid	Dilute Acid
Consumables and By-Products			
CO2 Stream (ton/yr) (1)	79,100	191,000	191,400
Water Consumption (1000 gal/yr) (2)	190,700	415,800	390,600
Net Electricity Production (kWh/gal) (3)	2.62	2.01	0.00
Ethanol Production Costs			
Fixed O&M ($/yr) (3)	$ 12,353,672	$ 8,495,487	$ 6,329,695
Fixed O&M ($/gal/yr)	$ 0.49	$ 0.14	$ 0.10
Variable O&M ($/gal/yr) (3)	$ 0.29	$ 0.21	$ 0.16
Electricity Value ($/gal/yr) (4)	$ (0.21)	$ (0.16)	$ -
Net Electricity Value ($/yr) - credit	$ 5,209,570	$ 9,656,193	$ -
Operating cost ($/gal/yr)	$ 0.58	$ 0.19	$ 0.27
Annual Operating Cost ($/yr)	***$ 14,409,111***	***$ 11,390,510***	***$ 16,199,727***
Total Capital Cost ($) (5)	***$ 230,234,393***	***$ 241,714,842***	***$ 181,339,321***
Capital Unit Cost ($/gal/yr)	$ 9.22	$ 4.01	$ 3.00
Capital Unit Cost ($/ton feed/yr)	$ 645	$ 311	$ 248
Fixed Charge Rate	12.3%	12.3%	12.3%
Annual Capital Payment ($/gal/yr)	$ 1.13	$ 0.49	$ 0.37
Non-Feedstock Production Cost ($/gal)	**$ 1.71**	**$ 0.68**	**$ 0.64**

1) Estimated by assuming that the amount of CO2 produced is approximately 96% of the weight of ethanol produced, since the theoretical max yield is 51% ethanol and 49% CO2 (by weight) from sugars.

2) Calculated based on the annual feedstock input, using data from Wooley et al. (1999) for near-term dilute acid process. Potential changes in water consumption for different time periods or for different pretreatment processes are not included in this analysis.

3) Short term value is average of values reported in the literature - see table below for details.

4) Based on average electricity price for industrial consumers in 2008, from EIA (2009).

5) Total capital cost is based on the average TCI values from the literature (shown in the table below), scaled by facility size within a particular size range, using a scaling factor of 0.8.

Plant Technology/Capacity (MGY)	**Total Capital Investment ($/gallon) (1)**		
	Short Term	**Mid Term**	**Long Term**
25 - 60 Dilute Acid	$ 8.36	--	--
60 - 100 Dilute Acid	--	$ 3.92	$ 2.78
> 100 (2) Steam Explosion, LHW	--	$ 5.46	$ 4.09

1) Based on average of results from Hamelinck et al. 2005, Wooley et al. 1999, Aden et al. 2002, and McAloon et al. 2000. BIGCC case and best of industry are excluded from capital cost estimates, as they are not representative of an average facility.

2) Trend suggests that large ethanol facilities (>100 MGY) have higher capital costs than medium facilities, similar to dry mill corn ethanol plants (Gallagher et al. 2005). However, the larger facilities are based on advanced pretreatment technologies (steam explosion and LHW for mid and long term), while the small and medium plants (<100 MGY) are based on dilute acid pretreatment. Ethanol production costs will vary by technology.

O&M Costs (1)	**Short Term**		
Ethanol Production (MGY)	24	45	52.2
Pretreatment		dilute acid	
Fixed O&M (3) (million$/yr)	$ 13.45	$ 12.58	$ 11.03
Fixed O&M ($/gal/yr)	$ 0.57	$ 0.28	$ 0.21
Variable O&M (4) ($/gal/yr)	$ 0.31	$ 0.28	$ 0.28
Total ($/gal/yr)	$ 0.88	$ 0.56	$ 0.49
Co-Products (5)			
Net Electricity Production (kWh/gal)	2.92	3.16	1.76
Source	*McAloon et al. 2000*	*Hamelinck et al. 2005*	*Wooley et al. 1999*

O&M Costs (1)	**Mid Term**	
Ethanol Production (MGY)	67.2	130
Pretreatment	dilute acid	steam explosion
Fixed O&M (3) (million$/yr)	$ 9.47	$ 27.27
Fixed O&M ($/gal/yr)	$ 0.14	$ 0.21
Variable O&M (4) ($/gal/yr)	$ 0.21	$ 0.25
Total ($/gal/yr)	$ 0.35	$ 0.46
Co-Products (5)		
Net Electricity Production (kWh/gal)	2.01	6.33
Source	*Wooley et al. 1999*	*Hamelinck et al. 2005*

O&M Costs (1)	**Long Term (2)**	
Ethanol Production (MGY)	87.5	311
Pretreatment	dilute acid	LHW
Fixed O&M (3) (million$/yr)	$ 9.17	$ 36.23
Fixed O&M ($/gal/yr)	$ 0.10	$ 0.12
Variable O&M (4) ($/gal/yr)	$ 0.16	$ 0.13
Total ($/gal/yr)	$ 0.27	$ 0.25
Co-Products (5)		
Net Electricity Production (kWh/gal)	0.00	2.11
Source	*Wooley et al. 1999*	*Hamelinck et al. 2005*

1) All costs reported in 2008$, converted using CEPCI where needed. All sources use poplar feedstock except McAloon et. al, which uses corn stover.

2) Long term Variable O&M estimated based on decreased cellulase costs from mid term, equivalent to $0.12/gal/yr

3) Fixed O&M includes labor, maintenance supplies, insurance and property taxes

4) Variable O&M includes raw materials (sulfuric acid & lime for dilute acid, cellulase or ammonia & CSL for cellulase production, other chemicals) and solids disposal (i.e. ash & gypsum).

5) Net electricity production based on heat and power production with partially dried solid residuals (lignin, cell mass) and unfermented sugars (dried to a syrup), fired in a boiler. Note that the net electricity production for the steam explosion pretreatment is much higher than other pretreatment, as this method has lower steam requirements. Furthermore, cellulase is not made on site in the steam explosion case, which means a larger quantity of non-fermented sugars and residuals are available for power generation.

Summary Page - Costs & Yield Data, Lignocellulosic Ethanol Production

Large Scale Plant

Inputs

Feedstock Type	**Switchgrass**
Economic Lifetime of Plant (Years)	25
Weighted Cost of Capital	10%

	Short Term (2010)	**Mid Term (2015-2020)**	**Long Term (2025+)**
Applicable Feedstocks	clean herbaceous feedstocks (agricultural residues and grasses) and de-barked wood chips		
Applicable Size Range (MGY)	25 - 60	60 - 150	60 - 300
Feedstock Input (dry ton/yr)	857,000	2,011,000	3,815,000
Feedstock Input (MMBtu/yr)	14,397,600	33,784,800	64,092,000
Yield - Dilute Acid (gal/dry ton)	70.0	77.4	82.6
Yield - Advanced Tech (gal/dry ton)	--	74.4	78.7
Conversion efficiency (HHV)	35%	37%	39%
Ethanol Production (MGY)	**60.0**	**150**	**300**
Pretreatment Technology	Dilute Acid	Steam Explosion	LHW
Consumables and By-Products			
CO2 Stream (ton/yr) (1)	190,000	473,700	951,600
Water Consumption (1000 gal/yr) (2)	457,800	1,074,200	2,037,900
Net Electricity Production (kWh/gal) (3)	2.62	6.33	2.11
Ethanol Production Costs			
Fixed O&M ($/yr) (3)	$ 12,353,672	$ 31,362,083	$ 34,992,534
Fixed O&M ($/gal/yr)	$ 0.21	$ 0.21	$ 0.12
Variable O&M ($/gal/yr) (3)	$ 0.29	$ 0.25	$ 0.13
Electricity Value ($/gal/yr) (4)	$ (0.21)	$ (0.50)	$ (0.17)
Net Electricity Value ($/yr) - credit	$ 12,502,969	$ 75,405,188	$ 50,496,133
Operating cost ($/gal/yr)	$ 0.29	$ (0.04)	$ 0.08
Annual Operating Cost ($/yr)	***$ 17,286,726***	***$ (6,299,317)***	***$ 24,276,093***
Total Capital Cost ($) (5)	***$ 463,809,024***	***$ 793,793,228***	***$ 1,238,552,957***
Capital Unit Cost ($/gal/yr)	$ 7.74	$ 5.31	$ 4.12
Capital Unit Cost ($/ton feed/yr)	$ 541	$ 395	$ 325
Fixed Charge Rate	12.3%	12.3%	12.3%
Annual Capital Payment ($/gal/yr)	$ 0.95	$ 0.65	$ 0.51
Non-Feedstock Production Cost ($/gal)	**$ 1.24**	**$ 0.61**	**$ 0.59**

1) Estimated by assuming that the amount of CO2 produced is approximately 96% of the weight of ethanol produced, since the theoretical max yield is 51% ethanol and 49% CO2 (by weight) from sugars.

2) Calculated based on the annual feedstock input, using data from Wooley et al. (1999) for near-term dilute acid process. Potential changes in water consumption for different time periods or for different pretreatment processes are not included in this analysis.

3) Short term value is average of values reported in the literature - see table below for details.

4) Based on average electricity price for industrial consumers in 2008, from EIA (2009).

5) Total capital cost is based on the average TCI values from the literature (shown in the table below), scaled by facility size within a particular size range, using a scaling factor of 0.8.

Plant Technology/Capacity (MGY)	**Total Capital Investment ($/gallon) (1)**		
	Short Term	**Mid Term**	**Long Term**
25 - 60 Dilute Acid	$ 8.36	--	--
60 - 100 Dilute Acid	--	$ 3.92	$ 2.78
> 100 (2) Steam Explosion, LHW	--	$ 5.46	$ 4.09

1) Based on average of results from Hamelinck et al. 2005, Wooley et al. 1999, Aden et al. 2002, and McAloon et al. 2000. BIGCC case and best of industry are excluded from capital cost estimates, as they are not representative of an average facility.

2) Trend suggests that large ethanol facilities (>100 MGY) have higher capital costs than medium facilities, similar to dry mill corn ethanol plants (Gallagher et al. 2005). However, the larger facilities are based on advanced pretreatment technologies (steam explosion and LHW for mid and long term), while the small and medium plants (<100 MGY) are based on dilute acid pretreatment. Ethanol production costs will vary by technology.

O&M Costs (1)	**Short Term**		
Ethanol Production (MGY)	24	45	52.2
Pretreatment		dilute acid	
Fixed O&M (3) (million$/yr)	$ 13.45	$ 12.58	$ 11.03
Fixed O&M ($/gal/yr)	$ 0.57	$ 0.28	$ 0.21
Variable O&M (4) ($/gal/yr)	$ 0.31	$ 0.28	$ 0.28
Total ($/gal/yr)	$ 0.88	$ 0.56	$ 0.49
Co-Products (5)			
Net Electricity Production (kWh/gal)	2.92	3.16	1.76
Source	*McAloon et al. 2000*	*Hamelinck et al. 2005*	*Wooley et al. 1999*

O&M Costs (1)	**Mid Term**	
Ethanol Production (MGY)	67.2	130
Pretreatment	dilute acid	steam explosion
Fixed O&M (3) (million$/yr)	$ 9.47	$ 27.27
Fixed O&M ($/gal/yr)	$ 0.14	$ 0.21
Variable O&M (4) ($/gal/yr)	$ 0.21	$ 0.25
Total ($/gal/yr)	$ 0.35	$ 0.46
Co-Products (5)		
Net Electricity Production (kWh/gal)	2.01	6.33
Source	*Wooley et al. 1999*	*Hamelinck et al. 2005*

O&M Costs (1)	**Long Term (2)**	
Ethanol Production (MGY)	87.5	311
Pretreatment	dilute acid	LHW
Fixed O&M (3) (million$/yr)	$ 9.17	$ 36.23
Fixed O&M ($/gal/yr)	$ 0.10	$ 0.12
Variable O&M (4) ($/gal/yr)	$ 0.16	$ 0.13
Total ($/gal/yr)	$ 0.27	$ 0.25
Co-Products (5)		
Net Electricity Production (kWh/gal)	0.00	2.11
Source	*Wooley et al. 1999*	*Hamelinck et al. 2005*

1) All costs reported in 2008S, converted using CEPCI where needed. All sources use poplar feedstock except McAloon et. al, which uses corn stover.

2) Long term Variable O&M estimated based on decreased cellulase costs from mid term, equivalent to S0.12/gal/yr

3) Fixed O&M includes labor, maintenance supplies, insurance and property taxes

4) Variable O&M includes raw materials (sulfuric acid & lime for dilute acid, cellulase or ammonia & CSL for cellulase production, other chemicals) and solids disposal (i.e. ash & gypsum).

5) Net electricity production based on heat and power production with partially dried solid residuals (lignin, cell mass) and unfermented sugars (dried to a syrup), fired in a boiler. Note that the net electricity production for the steam explosion pretreatment is much higher than other pretreatment, as this method has lower steam requirements. Furthermore, cellulase is not made on site in the steam explosion case, which means a larger quantity of non-fermented sugars and residuals are available for power generation.

	Feedstock Composition			Yield	Actual Yield (gal/ton) (2)								
Feedstock	Hemicellulose (wt%)	Cellulose (wt%)	Lignin (wt%)	Theoretical Yield (1) (gal/dry ton)	Short Term (Dilute Acid)	Comparative Values (3)	Mid Term (Dilute Acid)	Mid Term (Steam Explosion)	Comparative Values (3)	Long Term (Dilute Acid)	Long Term (LHW)	Comparative Values (3)	HHV (Btu/lb) (4)
Agricultural Resources													
Corn Grain				124.4									
Corn Stover (5)	23.23%	36.20%	18.50%	107.6	72.6	68.5	80.6	77.3	90.6	86.0	82.0		8,185
Wheat Straw (5)	23.20%	33.47%	17.28%	98.8	69.3		76.8	73.6		82.0	77.9		8,500
Bagasse (5)	25.79%	40.37%	23.90%	116.2	80.8		89.7	85.9		95.8	91.0		8,500
Grass Resources													
Switchgrass (6)	25.19%	31.98%	18.13%	99.7	70.0		77.4	74.4		82.6	78.7		8,400
Woody Resources													
Hybrid Poplar (6)	18.55%	44.70%	26.44%	109.9	76.9	73.0	85.9	81.2	86.9	91.9	98.6	107.4	9,000
Black Locust (6)	17.66%	41.61%	26.70%	103.0	72.1		80.4	75.9		86.1	92.4		9,000
Eucalyptus (6)	13.07%	49.50%	27.71%	108.6	75.8		85.0	83.8		91.1	97.8		9,000
Pine (6)	21.90%	44.55%	27.67%	115.1	80.9		90.2	83.7		96.4	103.5		9,130
Other Resources													
Yard Trimmings / Green Waste (7)	31.8%	20.0%	20.6%	108.0	63.9		70.0	68.6		74.4	73.4		8,500
Mixed Paper (7)	15.2%	48.1%	22.6%	116.2	76.8		86.0	83.5		92.1	98.8		8,200

1) Yield values from EERE's Theoretical Ethanol Yield Calculator (http://www1.eere.energy.gov/biomass/ethanol_yield_calculator.html), based on sugar content

2) Actual yields calculated based on average conversion values for all feedstocks in short term, and wood for all terms. Agricultural and grass feedstock yields for advanced mid and long term pretreatment technologies based on average increase of wood feedstock yields from short to mid term (4.4% of theoretical), as the conversion technology information does not apply to these materials, but other comparable advanced technologies may apply.

3) Comparative values based on average values from published research, only available for select feedstocks. Includes various pretreatment technologies.

4) Higher Heating Value (HHV) given for dry feedstocks. Data from (Energy Research Centre of the Netherlands, n.d.)

5) Biochemical composition data from U.S. DOE Biomass Feedstock Composition and Property Database, adjusted to 100% mass closure

6) Biochemical composition data from Hamelinck et al. 2005

7) Biochemical composition data from Energy Research Centre of the Netherlands (n.d.). Yard trimmings consists of 60% woody residues and 40% herbaceous residues (mix of leaves and grass), based on data from McKeever and Skog (2003). Yard trimmings theoretical ethanol yield based on average of similar materials.

	Conversion (% of theoretical) (1)				
	Short Term	*Mid Term*		*Long Term*	
Pretreatment Technology	*Dilute Acid*	*Dilute Acid*	*Steam Explosion*	*Dilute Acid*	*LHW*
Saccharification					
Hemicellulose	83%	85%	55%	85%	93%
Cellulose	75%	85%	93%	90%	98%
Fermentation					
Xylose & other sugars (2)	86%	90%	85%	95%	94%
Glucose (2)	93%	94%	93%	95%	94%
Source	*Hamelinck et al 2005, Wooley et al. 1999*	*Wooley et al 1999*	*Hamelinck et al. 2005*	*Wooley et al 1999*	*Hamelinck et al 2005*

1) Average conversion estimates based on poplar feedstock and the reported pretreatment technologies. The advanced technologies for mid and long term pretreatments may not be suitable for all feedstocks.

2) Cellulose is converted to glucose during saccharification, while hemicellulose is converted to xylose and other sugars

Summary Page - Costs & Yield Data, Thermochemical Conversion to Ethanol

Small Scale Plant

Inputs

Feedstock Type	**Poplar**
Economic Lifetime of Plant (Years)	25
Weighted Cost of Capital	12.3%

	Mid Term
Applicable Feedstocks	woody & agricultural; moisture content ≤ 50%
Applicable Size Range (dry tons/yr) (1)	> 185,000
Feedstock Input (dry ton/yr)	185,000
Feedstock Input (MMBtu/yr), HHV	3,330,000
Yield Ethanol (MGY) (2)	15.4
Conversion Efficiency (HHV) (3)	46.5%
Consumables and By-Products	
Yield Other Alcohols (MGY) (4)	2.6
Water Consumption (1000 gal/yr) (5)	29,917
Ethanol Production Costs	
Fixed O&M ($/yr) (6)	$ 11,080,546
Variable O&M ($/yr) (7)	$ 363,725
Co-product credit ($/yr) (8)	$ 4,863,922
Operating Cost ($/gal)	$ 0.427
Annual Operating Cost ($/yr)	***$ 6,580,349***
Total Capital Investment ($) (9)	***$ 76,958,910***
Capital Investment ($/gal/yr)	$ 4.99
Capital Investment ($/ton/yr)	$ 416
Fixed Charge Rate	12.3%
Non-Feedstock Production Cost ($/gal)	**$ 1.04**

1) Based on size range for Fischer Tropsch analysis, which has similar technology characteristics.

2) Yield based on overall conversion efficiency of 38.9% to ethanol (HHV) from Phillips et al. 2007, and estimated ethanol HHV of 84,000 Btu/gal.

3) Total fuel conversion efficiency, includes ethanol and higher alcohols.

4) Based on HHV efficiency of 7.6%, following Phillips et al. 2007. Nearly all of this stream is made of propanol, which has an HHV of about 97,300 Btu/gal.

5) Based on water use of 1.94 gal H2O per gal ethanol, which includes cooling tower make-up, process water, and other system losses.

6) Fixed O&M includes annual maintinance costs, insurance and taxes (4% of TCI), and labor costs based on data from Phillips et al. 2007. The labor costs are scaled by facility size using a scaling factor of 0.25, as in the Fischer Tropsch analysis.

7) Variable O&M includes consumables (catalysts, gasifier bed material, and other raw materials) and waste disposal, and is calculated on a feedstock input basis. No electricity is required as a portion of the syngas is diverted for heat and power production.

8) Calculated following the methodology in Phillips et al. 2007, but with updated gasoline prices from EIA (average wholesale gasoline price for January 2008 to February 2009 is $2.39/gal). Value of mixed alcohols estimated to be $1.87/gal, based on 78% discount for lower energy density than gasoline, and a 10 cent/gal reduction for potential additional processing requirements before use.

9) Capital costs based on estimates from Phillips et al. 2007, and a scaling factor of 0.8, which is typical for biofuel conversion technologies.

Summary Page - Costs & Yield Data, Thermochemical Conversion to Ethanol

Large Scale Plant

Inputs

Feedstock Type	Poplar
Economic Lifetime of Plant (Years)	25
Weighted Cost of Capital	12.3%

	Mid Term
Applicable Feedstocks Applicable Size Range (dry tons/yr) (1)	woody & agricultural; moisture content ≤ 50% > 185,000
Feedstock Input (dry ton/yr)	1,500,000
Feedstock Input (MMBtu/yr), HHV	27,000,000
Yield Ethanol (MGY) (2)	125.0
Conversion Efficiency (HHV) (3)	46.5%
Consumables and By-Products	
Yield Other Alcohols (MGY) (4)	21.1
Water Consumption (1000 gal/yr) (5)	242,569
Ethanol Production Costs	
Fixed O&M ($/yr) (6)	$ 21,165,267
Variable O&M ($/yr) (7)	$ 2,949,118
Co-product credit ($/yr) (8)	$ 39,437,205
Operating Cost ($/gal)	$ (0.123)
Annual Operating Cost ($/yr)	***$ (15,322,819)***
Total Capital Investment ($) (9)	***$ 410,576,919***
Capital Investment ($/gal/yr)	$ 3.28
Capital Investment ($/ton/yr)	$ 274
Fixed Charge Rate	12.3%
Non-Feedstock Production Cost ($/gal)	**$ 0.28**

1) Based on size range for Fischer Tropsch analysis, which has similar technology characteristics.

2) Yield based on overall conversion efficiency of 38.9% to ethanol (HHV) from Phillips et al. 2007, and estimated ethanol HHV of 84,000 Btu/gal.

3) Total fuel conversion efficiency, includes ethanol and higher alcohols.

4) Based on HHV efficiency of 7.6%, following Phillips et al. 2007. Nearly all of this stream is made of propanol, which has an HHV of about 97,300 Btu/gal.

5) Based on water use of 1.94 gal H2O per gal ethanol, which includes cooling tower make-up, process water, and other system losses.

6) Fixed O&M includes annual maintinance costs, insurance and taxes (4% of TCI), and labor costs based on data from Phillips et al. 2007. The labor costs are scaled by facility size using a scaling factor of 0.25, as in the Fischer Tropsch analysis.

7) Variable O&M includes consumables (catalysts, gasifier bed material, and other raw materials) and waste disposal, and is calculated on a feedstock input basis. No electricity is required as a portion of the syngas is diverted for heat and power production.

8) Calculated following the methodology in Phillips et al. 2007, but with updated gasoline prices from EIA (average wholesale gasoline price for January 2008 to February 2009 is $2.39/gal). Value of mixed alcohols estimated to be $1.87/gal, based on 78% discount for lower energy density than gasoline, and a 10 cent/gal reduction for potential additional processing requirements before use.

9) Capital costs based on estimates from Phillips et al. 2007, and a scaling factor of 0.8, which is typical for biofuel conversion technologies.

Summary Page - Costs & Yield Data, Future Hemicellulose Extraction Biorefining

Small Scale Plant

This facility was modeled as an addition to an existing paper mill, not as a stand-alone greenfield project. All costs noted are incremental to the costs of the existing mill, rather than absolute. Also, due to the lack of any demonstration plants in operation, data is based on two diffierent AspenPlus simulations by researchers: Mao et al (2008) and McAloon et al (1999). A laboratory study by Amidon et al (2006) was used as a reference for mass balances and overall process requipments. McAloon's study dealt with a cellulosic ethanol rather than a hemicellulosic ethanol plant, but similarities in major equipment have allowed some extrapolation of costs. The bioethanol product yields are stated as denatured ethanol with the addition of gasoline additives. The cost of gasoline for denaturation is included in O&M costs.

Inputs

Feedstock Type	Scots Pine
Economic Lifetime of Plant (Years)	25
Weighted Cost of Capital	10%

	Estimate for Future Plants
Applicable Feedstocks	Scots Pine, Silver Birch, Spruce, Eucalyptus
Applicable Size Range (MGY)	4-16
Feedstock Input (ton/yr)	700,000
Yield Bioethanol (ton/yr)	12,924
Yield Bioethanol (MGY)	3.9
Conversion Efficiency (HHV) (1)	9.9%
Consumables and By-Products	
Pulp Yield (dry ton/yr) (2)	296,019
Water Consumption (1000 gal/yr) (3)	33,019
Incremental Bioethanol Production Costs	
Fixed O&M ($/yr)	$ 3,948,942
Variable O&M ($/yr)	$ 4,753,027
Reduction in paper yield ($/yr) (4)	$ 4,797,120
Operating cost ($/gal/yr)	$ 3.44
Annual Operating Cost ($/yr)	***$ 13,499,089***
Total Capital Investment ($) (5)	***$ 24,675,062***
Unit Capital Cost ($/gal/yr)	$ 6.28
Unit Capital Cost ($/ton feed/yr)	$ 35.25
Fixed Charge Rate	12%
Non-Feedstock Production Cost ($/gal)	**$ 4.21**

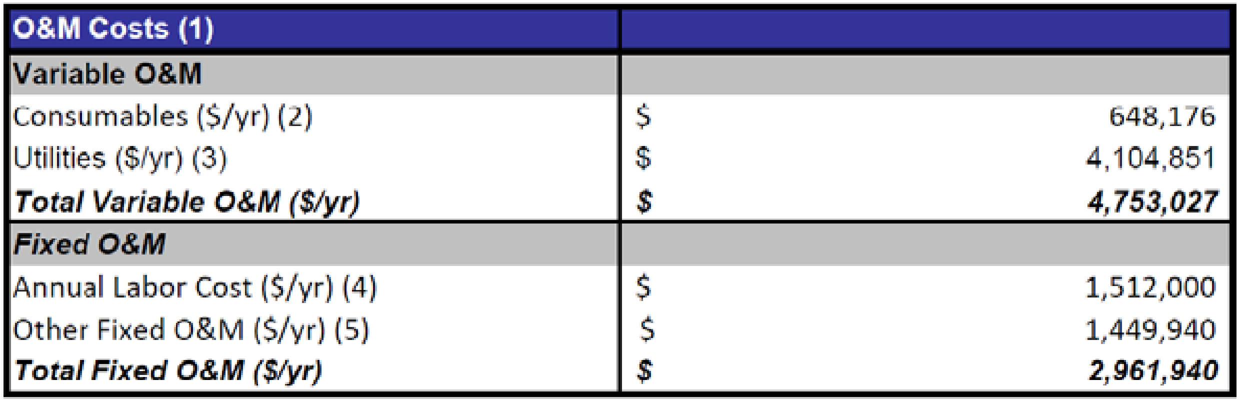
1) Calculated using a heating value of hemicellulosic feedstocks of 8000 Btu/lb, and a typical bioethanol heating value of 84,000 Btu/gal. (White, 1999). Energy input is taken as total hemicellulose initially present in the wood feedstock, not as hemicellulose extracted.

2) Based on typical hemicellulose content of 12.5% in conventional kraft pulp.

3) Scaled based on data from McAloon et al (1999). Annual water use is 70 lb water per gallon bioethanol produced.

4) Current U.S. market prices for bleached pulp (NBSK, or long-fiber northern bleached softwood kraft pulp) range between $700 and $800/ton based on FOEX PIX delivered prices. $750 per ton, discounted 20% to account for the lack of transportation costs in the value of the pulp to the mill, was taken as the value of the pulp, and reduction in pulp yield was calculated based on a digestion hemicellulose removal efficiency of 80%, based on average kraft pulp hemicellulose content of 12.5%.

5) Capital cost data for extraction and ethanol production facility added to an existing pulp and paper mill from Frederick et al (2006). A scaling factor of 0.699 was used as per Mao et al (2008). This factor agrees with the capital costs presented in Frederick et al (2006).



O&M Costs (1)		
Variable O&M		
Consumables ($/yr) (2)	$	648,176
Utilities ($/yr) (3)	$	4,104,851
Total Variable O&M ($/yr)	***$***	***4,753,027***
Fixed O&M		
Annual Labor Cost ($/yr) (4)	$	1,512,000
Other Fixed O&M ($/yr) (5)	$	1,449,940
Total Fixed O&M ($/yr)	***$***	***2,961,940***

1) All costs reported in 2008 US$. Where necessary, costs have been converted using CEPCI. Data source used for utility costs is Mao et al (2008); data source for labor,fixed O&M costs, and chemical requirements is McAloon et al (1999).

2) Consumables include sodium hydroxide, sulfuric acid, buffers, nutrients, gasoline denaturant, and water treatment chemicals. Utilities include natural gas, makeup water, and electricity.

3) Utilities were not fully heat integrated, so steam demand may be overestimated.

4) Based on data for a lignocellulosic ethanol facility from McAloon et al (1999), but increased operator salary to $20/hr from $18/hr, based on Antares experience.

5) Other fixed O&M costs include taxes, insurance, and supplies, and are estimated as in McAloon et al (1999).

Summary Page - Costs & Yield Data, Future Hemicellulose Extraction Biorefining

Large Scale Plant

This facility was modeled as an addition to an existing paper mill, not as a stand-alone greenfield project. All costs noted are incremental to the costs of the existing mill, rather than absolute. Also, due to the lack of any demonstration plants in operation, data is based on two diffierent AspenPlus simulations by researchers: Mao et al (2008) and McAloon et al (1999). A laboratory study by Amidon et al (2006) was used as a reference for mass balances and overall process requipments. McAloon's study dealt with a cellulosic ethanol rather than a hemicellulosic ethanol plant, but similarities in major equipment have allowed some extrapolation of costs. The bioethanol product yields are stated as denatured ethanol with the addition of gasoline additives. The cost of gasoline for denaturation is included in O&M costs.

Inputs

Feedstock Type	Scots Pine
Economic Lifetime of Plant (Years)	25
Weighted Cost of Capital	10%

		Estimate for Future Plants
Applicable Feedstocks		Scots Pine, Silver Birch, Spruce, Eucalyptus
Applicable Size Range (MGY)		4-16
Feedstock Input (ton/yr)		2,500,000
Yield Bioethanol (ton/yr)		46,158
Yield Bioethanol (MGY)		14.0
Conversion Efficiency (HHV) (1)		9.9%
Consumables and By-Products		
Pulp Yield (dry ton/yr) (2)		1,057,212
Water Consumption (1000 gal/yr) (3)		117,924
Incremental Bioethanol Production Costs		
Fixed O&M ($/yr)	$	6,515,448
Variable O&M ($/yr)	$	16,975,096
Reduction in paper yield ($/yr) (4)	$	17,132,571
Operating cost ($/gal/yr)	$	2.90
Annual Operating Cost ($/yr)	***$***	***40,623,115***
Total Capital Investment ($) (5)	***$***	***60,075,140***
Unit Capital Cost ($/gal/yr)	$	4.28
Unit Capital Cost ($/ton feed/yr)	$	24.03
Fixed Charge Rate		12%
Non-Feedstock Production Cost ($/gal)	**$**	**3.42**

1) Calculated using a heating value of hemicellulosic feedstocks of 8000 Btu/lb, and a typical bioethanol heating value of 84,000 Btu/gal. (White, 1999). Energy input is taken as total hemicellulose initially present in the wood feedstock, not as hemicellulose extracted.

2) Based on typical hemicellulose content of 12.5% in conventional kraft pulp.

3) Scaled based on data from McAloon et al (1999). Annual water use is 70 lb water per gallon bioethanol produced.

4) Current U.S. market prices for bleached pulp (NBSK, or long-fiber northern bleached softwood kraft pulp) range between $700 and $800/ton based on FOEX PIX delivered prices. $750 per ton, discounted 20% to account for the lack of transportation costs in the value of the pulp to the mill, was taken as the value of the pulp, and reduction in pulp yield was calculated based on a digestion hemicellulose removal efficiency of 80%, based on average kraft pulp hemicellulose content of 12.5%.

5) Capital cost data for extraction and ethanol production facility added to an existing pulp and paper mill from Frederick et al (2006). A scaling factor of 0.699 was used as per Mao et al (2008). This factor agrees with the capital costs presented in Frederick et al (2006).

O&M Costs (1)		
Variable O&M		
Consumables ($/yr) (2)	$	2,314,914
Utilities ($/yr) (3)	$	14,660,182
Total Variable O&M ($/yr)	***$***	***16,975,096***
Fixed O&M		
Annual Labor Cost ($/yr) (4)	$	1,512,000
Other Fixed O&M ($/yr) (5)	$	2,600,442
Total Fixed O&M ($/yr)	***$***	***4,112,442***

1) All costs reported in 2008 US$. Where necessary, costs have been converted using CEPCI. Data source used for utility costs is Mao et al (2008); data source for labor,fixed O&M costs, and chemical requirements is McAloon et al (1999).

2) Consumables include sodium hydroxide, sulfuric acid, buffers, nutrients, gasoline denaturant, and water treatment chemicals. Utilities include natural gas, makeup water, and electricity.

3) Utilities were not fully heat integrated, so steam demand may be overestimated.

4) Based on data for a lignocellulosic ethanol facility from McAloon et al (1999), but increased operator salary to $20/hr from $18/hr, based on Antares experience.

5) Other fixed O&M costs include taxes, insurance, and supplies, and are estimated as in McAloon et al (1999).

Feedstocks and Yields

Feedstock (1)	**Average Yield (gal/ton wood feedstock) (2)**	**Average Yield (gal/ton pulp coproduct) (3)**
Spruce	5.760	13.8
Silver Birch	6.099	14.1
Eucalyptus	3.614	7.6
Scots Pine	5.365	12.7

1) Feedstock characteristics given by Amidon (2008). Yields are estimates based on 90% hydrolysis efficiency and 90% of theoretical fermentation efficiency (Amidon, 2008).

2) At 15% removal of hemicellulcse, an operating point that appears to minimize cellulose loss during extraction as per Frederick et al, 2006.

3) Assumes current digestion hemicellulose removal efficiencies (80%), based on a conventional paper composition of 12.5% hemicellulose.

Summary Page - Costs & Yield Data, Pyrolysis Oil Production & Upgrading - *Small Scale Plant*

Inputs

Feedstock Type	**Bark**
Economic Lifetime of Plant (Years)	25
Weighted Cost of Capital	10%

	Mid Term (2015-2020)
Applicable Feedstocks	woody & agricultural
Applicable Size Range (dry tons/yr) (1)	30,000 - 800,000
Feedstock Input (dry ton/yr)	30,000
Yield Bio-oil (ton/yr)	19,050
Yield Bio-oil (MGY)	3.8
Conversion Efficiency (HHV)	65.9%
Pyrolytic Lignin (ton/yr)	4,705
Pyrolytic Lignin (MGY)	1.0
Capital Cost Pyrolysis Production (Mil $) (2)	$ 7.78
Hydrotreatment Process (3)	
H2 consumption (ton/yr)	212
Light hydrocarbons produced (ton/yr)	706
Gasoline produced (MGY)	0.4
Diesel produced (MGY)	0.1
Capital Cost Hydrotreatment (Mil $) (4)	$ 4.38
Yield of By-Products (5)	
Solid Char (ton/yr)	6,600
Non-condensable gas (ton/yr)	3,750
Water-soluble bio-oil (ton/yr)	14,345
CO2 Stream (ton/yr)	6,390
Water consumption (1000 gal/yr) (6)	4,178
Gasoline Production Costs	
Fixed O&M ($/yr)	$ 1,154,397
Variable O&M ($/yr)	$ 1,158,555
Co-product credit ($/yr)	$ (686,449)
Annual Operating Cost ($/gal)	$ 0.423
Annual Operating Cost ($/yr)	***$ 1,626,503***
Total Capital Investment ($)	***$ 12,157,877***
Total Capital Investment ($/gpy)	$ 3.16
Total Capital Investment ($/tpy)	$ 405
Fixed Charge Rate	12.3%
Non-Feedstock Production Cost ($/gal)	**$ 7.30**

1) Based on reported sizes for pyrolysis oil production facilities in UOP 2005, Ringer et al. 2006, and Polagye et al. 2007.

2) Pyrolysis oil production capital cost based on data from: Cole Hill Associates 2004, Ringer et al. 2006, Polagye et al. 2007, UOP 2005. Total capital cost is scaled by facility size using a scaling factor of 0.77, as estimated from the data.

3) Calculated as percentages (by weight) of pyrolytic lignin, based on yields from UOP 2005. H2 consumption is 4.5%, light ends production 15%, gasoline production 30%, diesel production 8%.

4) Hydrotreatment cost is based on data from UOP 2005, and scaled using a standard scaling factor of 0.6.

5) The solid char and non-condensible gas are not assigned a value, as the char is combusted to provide heat for the pyrolysis reaction and the gas is recycled as a fluidizing gas. The heating value of the char (HHV) is ~12,900 Btu/lb (Mohan et al. 2006). Hydrotreatment generates a concentrated CO2 stream, which could have a market value (not considered here). Data for the CO2 yield is derived from UOP 2005.

6) Based on data from Ringer et al. 2006, a 200,750 dry ton per year facility uses about 160 thousand gal/hr water, mostly for cooling. Assuming only 2% of this water is lost in blowdown, the annual consumption is about 28,000 thousand gallons per year. This value is scaled based on feedstock input. No additional water consumption is added for hydrotreatment.

O&M Costs (1)	**Value**	***Source***
Variable O&M (2)		
Waste Treatment ($/yr)	$ 243,493	*Ringer et al. 2006*
Electricity Consumption (kWh/yr)	6,691,054	*Ringer et al. 2006*
Total Utility Cost ($/yr)	$ 593,679	*UOP 2005,* *EIA AEO 2007*
Total Variable O&M ($/yr)	***$ 1,158,555***	
Fixed O&M (3)		
Labor ($/yr)	$ 485,714	*Polagye et al. 2007,* *Ringer et al. 2006*
Other Fixed Costs ($/yr)	$ 668,683	*Polagye et al. 2007*
Total Fixed O&M ($/yr)	***$ 1,154,397***	
Co-Products (4)		
Water-Soluble Bio-oil ($/yr)	$ -	
Diesel ($/yr)	$ 316,352	*EIA AER 2005*
Light Hydrocarbons (LPG) ($/yr)	$ 370,097	*EIA AER 2005*
Total Co-Product Value ($/yr)	***$ 686,449***	

1) All costs reported in 2008$, converted using CEPCI where needed.

2) Electricity consumption is for pyrolysis oil production portion process only. Total utility cost includes electricity consumption for entire process,using an electricity cost of 7.9 cents/kWh for pyrolysis production (based on the 2008 average cost for electricity in the industrial sector from EIA 2009) and utility cost for hydrotreatment from UOP 2005. Hydrogen cost is based on rate of hydrogen consumption and cost ($1,200/ton) from UOP 2005.

3) The labor cost is scaled using a factor of 0.31, which was derived using data from listed sources for pyrolysis oil production facilities. The base labor cost is from a plant with 550 dry ton per day biomass input from Ringer et al. 2006. Other fixed costs include maintenance, insurance, and overhead, with a total annual cost equivalent to 5.5% of the total capital investment for the entire facility (including pyrolysis oil production and hydrotreatment).

4) Assume non-condensable gases and char recycled into pyrolysis reaction, per Ringer et al. 2006. Assume water-soluble bio-oil has a net zero value, as it has a low heating value and needs to be co-fired with natural gas for use. Value of diesel and light hydrocarbons based on average 2008 wholesale price from EIA AER 2009 for No. 2 diesel fuel ($2.99/gal) and Propane ($1.11/gal). Antares estimates of the added cost for compressing light hydrocarbons is negligible and is not included.

Summary Page - Costs & Yield Data, Pyrolysis Oil Production & Upgrading - *Large Scale Plant*

Inputs

Feedstock Type	**Bark**
Economic Lifetime of Plant (Years)	25
Weighted Cost of Capital	10%

	Mid Term (2015-2020)
Applicable Feedstocks	woody & agricultural
Applicable Size Range (dry tons/yr) (1)	30,000 - 800,000
Feedstock Input (dry ton/yr)	800,000
Yield Bio-oil (ton/yr)	508,000
Yield Bio-oil (MGY)	102.6
Conversion Efficiency (HHV)	65.9%
Pyrolytic Lignin (ton/yr)	125,476
Pyrolytic Lignin (MGY)	25.3
Capital Cost Pyrolysis Production (Mil $) (2)	$ 97.96
Hydrotreatment Process (3)	
H2 consumption (ton/yr)	5,646
Light hydrocarbons produced (ton/yr)	18,821
Gasoline produced (MGY)	11.4
Diesel produced (MGY)	2.8
Capital Cost Hydrotreatment (Mil $) (4)	$ 31.40
Yield of By-Products (5)	
Solid Char (ton/yr)	176,000
Non-condensable gas (ton/yr)	100,000
Water-soluble bio-oil (ton/yr)	382,524
CO2 Stream (ton/yr)	170,413
Water consumption (1000 gal/yr) (6)	111,422
Gasoline Production Costs	
Fixed O&M ($/yr)	$ 11,734,390
Variable O&M ($/yr)	$ 30,894,793
Co-product credit ($/yr)	$ (18,305,309)
Annual Operating Cost ($/gal)	$ 0.237
Annual Operating Cost ($/yr)	***$ 24,323,875***
Total Capital Investment ($)	***$ 129,361,673***
Total Capital Investment ($/gpy)	$ 1.26
Total Capital Investment ($/tpy)	$ 162
Fixed Charge Rate	12.3%
Non-Feedstock Production Cost ($/gal)	**$ 3.53**

1) Based on reported sizes for pyrolysis oil production facilities in UOP 2005, Ringer et al. 2006, and Polagye et al. 2007.

2) Pyrolysis oil production capital cost based on data from: Cole Hill Associates 2004, Ringer et al. 2006, Polagye et al. 2007, UOP 2005. Total capital cost is scaled by facility size using a scaling factor of 0.77, as estimated from the data.

3) Calculated as percentages (by weight) of pyrolytic lignin, based on yields from UOP 2005. H2 consumption is 4.5%, light ends production 15%, gasoline production 30%, diesel production 8%.

4) Hydrotreatment cost is based on data from UOP 2005, and scaled using a standard scaling factor of 0.6

5) The solid char and non-condensible gas are not assigned a value, as the char is combusted to provide heat for the pyrolysis reaction and the gas is recycled as a fluidizing gas. The heating value of the char (HHV) is ~12,900 Btu/lb (Mohan et al. 2006). Hydrotreatment generates a concentrated CO2 stream, which could have a market value (not considered here). Data for the CO2 yield is derived from UOP 2005.

6) Based on data from Ringer et al. 2006, a 200,750 dry ton per year facility uses about 160 thousand gal/hr water, mostly for cooling. Assuming only 2% of this water is lost in blowdown, the annual consumption is about 28,000 thousand gallons per year. This value is scaled based on feedstock input. No additional water consumption is added for hydrotreatment.

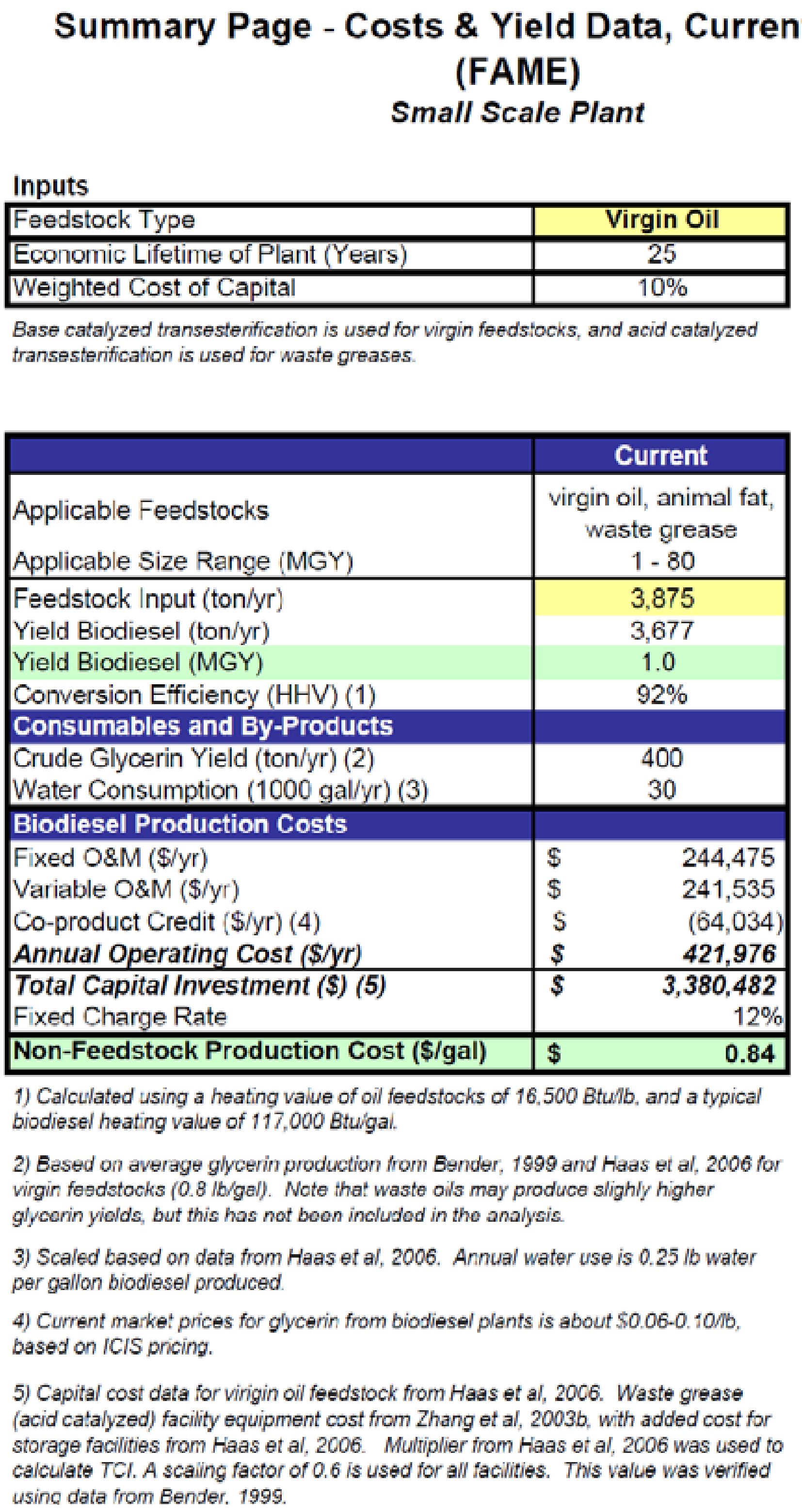
O&M Costs (1)	**Value**	***Source***
Variable O&M (2)		
Waste Treatment ($/yr)	$ 6,493,154	*Ringer et al. 2006*
Electricity Consumption (kWh/yr)	178,428,106	*Ringer et al. 2006*
Total Utility Cost ($/yr)	$ 15,831,430	*UOP 2005, EIA AEO 2007*
Total Variable O&M ($/yr)	***$ 30,894,793***	
Fixed O&M (3)		
Labor ($/yr)	$ 4,619,498	*Polagye et al. 2007, Ringer et al. 2006*
Other Fixed Costs ($/yr)	$ 7,114,892	*Polagye et al. 2007*
Total Fixed O&M ($/yr)	***$ 11,734,390***	
Co-Products (4)		
Water-Soluble Bio-oil ($/yr)	$ -	
Diesel ($/yr)	$ 8,436,044	*EIA AER 2005*
Light Hydrocarbons (LPG) ($/yr)	$ 9,869,265	*EIA AER 2005*
Total Co-Product Value ($/yr)	***$ 18,305,309***	

1) All costs reported in 2008$, converted using CEPCI where needed.

2) Electricity consumption is for pyrolysis oil production portion process only. Total utility cost includes electricity consumption for entire process,using an electricity cost of 7.9 cents/kWh for pyrolysis production (based on the 2008 average cost for electricity in the industrial sector from EIA 2009) and utility cost for hydrotreatment from UOP 2005. Hydrogen cost is based on rate of hydrogen consumption and cost ($1,200/ton) from UOP 2005.

3) The labor cost is scaled using a factor of 0.31, which was derived using data from listed sources for pyrolysis oil production facilities. The base labor cost is from a plant with 550 dry ton per day biomass input from Ringer et al. 2006. Other fixed costs include maintenance, insurance, and overhead, with a total annual cost equivalent to 5.5% of the total capital investment for the entire facility (including pyrolysis oil production and hydrotreatment).

4) Assume non-condensable gases and char recycled into pyrolysis reaction, per Ringer et al. 2006. Assume water-soluble bio-oil has a net zero value, as it has a low heating value and needs to be co-fired with natural gas for use. Value of diesel and light hydrocarbons based on average 2008 wholesale price from EIA AER 2009 for No. 2 diesel fuel ($2.99/gal) and Propane ($1.11/gal). Antares estimates of the added cost for compressing light hydrocarbons is negligible and is not included.



Summary Page - Costs & Yield Data, Current Biodiesel (FAME)

Small Scale Plant

Inputs

Feedstock Type	**Virgin Oil**
Economic Lifetime of Plant (Years)	25
Weighted Cost of Capital	10%

Base catalyzed transesterification is used for virgin feedstocks, and acid catalyzed transesterification is used for waste greases.

	Current	
Applicable Feedstocks	virgin oil, animal fat, waste grease	
Applicable Size Range (MGY)	1 - 80	
Feedstock Input (ton/yr)	3,875	
Yield Biodiesel (ton/yr)	3,677	
Yield Biodiesel (MGY)	1.0	
Conversion Efficiency (HHV) (1)	92%	
Consumables and By-Products		
Crude Glycerin Yield (ton/yr) (2)	400	
Water Consumption (1000 gal/yr) (3)	30	
Biodiesel Production Costs		
Fixed O&M ($/yr)	$	244,475
Variable O&M ($/yr)	$	241,535
Co-product Credit ($/yr) (4)	$	(64,034)
Annual Operating Cost ($/yr)	***$***	***421,976***
Total Capital Investment ($) (5)	***$***	***3,380,482***
Fixed Charge Rate		12%
Non-Feedstock Production Cost ($/gal)	**$**	**0.84**

1) Calculated using a heating value of oil feedstocks of 16,500 Btu/lb, and a typical biodiesel heating value of 117,000 Btu/gal.

2) Based on average glycerin production from Bender, 1999 and Haas et al, 2006 for virgin feedstocks (0.8 lb/gal). Note that waste oils may produce slighly higher glycerin yields, but this has not been included in the analysis.

3) Scaled based on data from Haas et al, 2006. Annual water use is 0.25 lb water per gallon biodiesel produced.

4) Current market prices for glycerin from biodiesel plants is about $0.06-0.10/lb, based on ICIS pricing.

5) Capital cost data for virigin oil feedstock from Haas et al, 2006. Waste grease (acid catalyzed) facility equipment cost from Zhang et al, 2003b, with added cost for storage facilities from Haas et al, 2006. Multiplier from Haas et al, 2006 was used to calculate TCI. A scaling factor of 0.6 is used for all facilities. This value was verified using data from Bender, 1999.

O&M Costs (1)	
Variable O&M (2)	
Consumables ($/yr) (3)	$ 191,724
Utilities ($/yr)	$ 49,811
Total Variable O&M ($/yr)	***$ 241,535***
Fixed O&M	
Annual Labor Cost ($/yr) (4)	$ 173,485
Other Fixed O&M ($/yr) (5)	$ 70,990
Total Fixed O&M ($/yr)	***$ 244,475***

1) All costs reported in 2008 US$. Where necessary, costs have been converted using CEPCI. Data source used for O&M costs for virgin feedstocks is Haas et al, 2006. Waste grease facility data also from Zhang et al, 2003b.

2) Consumables include methanol, sodium methoxide, hydrochloric acid, sodium hydroxide, and water. Utilities include natural gas, WWT and waste disposal, and electricity.

3) Data from Haas et al, 2006 used for virgin feedstocks. Waste grease cost based on Haas et al, 2006 data, but adjusted to incorporate increased quantity of methanol and different catalyst using data from Zhang et al, 2003.

4) Based on data for a virgin oil facility from Haas et al, 2006, but increased operator salary to $20/hr from $12.5/hr, based on Antares experience. Values in the literature suggest that acid catalyzed facilities have higher labor requirements, so Antares estimated there would be one additional operator per shift (for a 10 MGY facility). For comparison, the base catalyzed facility had two operators per shift. Labor cost includes scaling by facility size (based on available data from pyrolysis oil production facilities).

5) Other fixed O&M costs (including taxes, insurance, and supplies) are 2.1% of capital cost.

Feedstock (1)	Yield range (gal/ton)	Average Yield (gal/ton)	*Sources*
Virgin Oil (2)	246.4-270.0	258.2	*Haas et. al 2006, Sheehan et al. 1998, Zhang et al. 2003a, Canakci and van Gerpen 2001*
Animal Fats	266.3	266.3	*Bender 1999*
Yellow Grease (3)	234.9 - 263.2	249.1	*Zhang et al. 2003a, Canakci and van Gerpen 2001*

1) All data based on feedstocks with negligible water content.

2) Based on data for degummed soybean oil. High end of the yield range from Haas et al, 2006, which assumes a negligible free fatty acid content in the feedstock.

3) Conversion yield for waste greases based on acid catalyzed reaction. For comparison, the transesterification reaction for virgin feedstocks use base catalysts.

Summary Page - Costs & Yield Data, Current Biodiesel (FAME)

Large Scale Plant

Inputs

Feedstock Type	**Virgin Oil**
Economic Lifetime of Plant (Years)	25
Weighted Cost of Capital	10%

Base catalyzed transesterification is used for virgin feedstocks, and acid catalyzed transesterification is used for waste greases.

	Current
Applicable Feedstocks	virgin oil, animal fat, waste grease
Applicable Size Range (MGY)	1 - 80
Feedstock Input (ton/yr)	310,000
Yield Biodiesel (ton/yr)	294,154
Yield Biodiesel (MGY)	80.0
Conversion Efficiency (HHV) (1)	92%
Consumables and By-Products	
Crude Glycerin Yield (ton/yr) (2)	32,017
Water Consumption (1000 gal/yr) (3)	2,396
Biodiesel Production Costs	
Fixed O&M ($/yr)	$ 4,489,874
Variable O&M ($/yr)	$ 19,322,820
Co-product Credit ($/yr) (4)	$ (5,122,688)
Annual Operating Cost ($/yr)	***$ 18,690,006***
Total Capital Investment ($) (5)	***$ 46,863,271***
Fixed Charge Rate	12%
Non-Feedstock Production Cost ($/gal)	**$ 0.31**

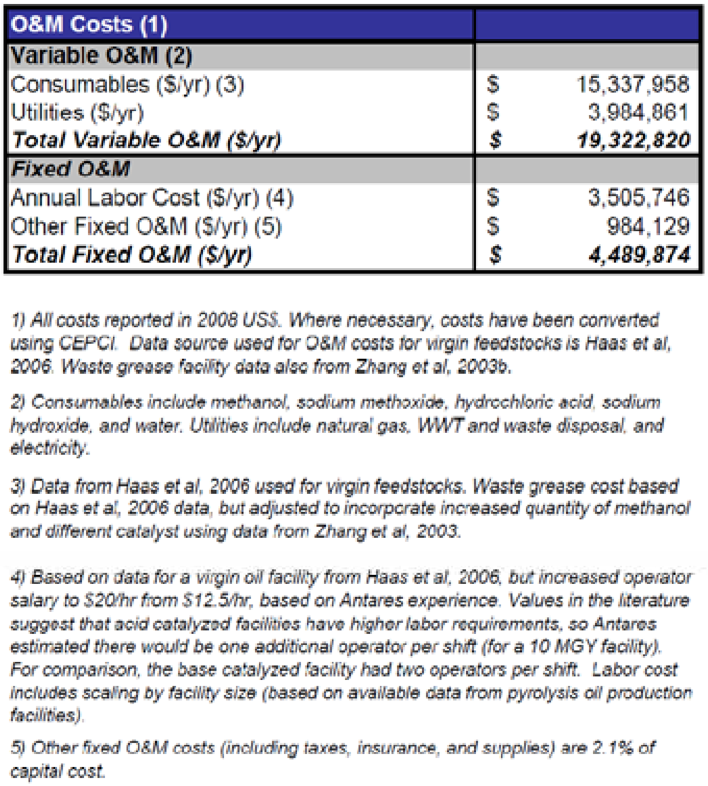
1) Calculated using a heating value of oil feedstocks of 16,500 Btu/lb, and a typical biodiesel heating value of 117,000 Btu/gal.

2) Based on average glycerin production from Bender, 1999 and Haas et al, 2006 for virgin feedstocks (0.8 lb/gal). Note that waste oils may produce slighly higher glycerin yields, but this has not been included in the analysis.

3) Scaled based on data from Haas et al, 2006. Annual water use is 0.25 lb water per gallon biodiesel produced.

4) Current market prices for glycerin from biodiesel plants is about $0.06-0.10/lb, based on ICIS pricing.

5) Capital cost data for virigin oil feedstock from Haas et al, 2006. Waste grease (acid catalyzed) facility equipment cost from Zhang et al, 2003b, with added cost for storage facilities from Haas et al, 2006. Multiplier from Haas et al, 2006 was used to calculate TCI. A scaling factor of 0.6 is used for all facilities. This value was verified using data from Bender, 1999.



O&M Costs (1)	
Variable O&M (2)	
Consumables ($/yr) (3)	$ 15,337,958
Utilities ($/yr)	$ 3,984,861
Total Variable O&M ($/yr)	***$ 19,322,820***
Fixed O&M	
Annual Labor Cost ($/yr) (4)	$ 3,505,746
Other Fixed O&M ($/yr) (5)	$ 984,129
Total Fixed O&M ($/yr)	***$ 4,489,874***

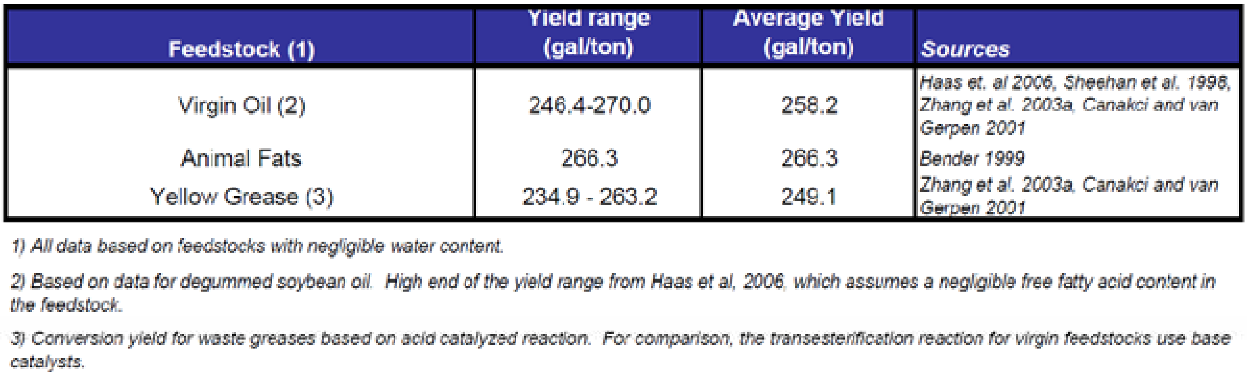
1) All costs reported in 2008 US$. Where necessary, costs have been converted using CEPCI. Data source used for O&M costs for virgin feedstocks is Haas et al, 2006. Waste grease facility data also from Zhang et al, 2003b.

2) Consumables include methanol, sodium methoxide, hydrochloric acid, sodium hydroxide, and water. Utilities include natural gas, WWT and waste disposal, and electricity.

3) Data from Haas et al, 2006 used for virgin feedstocks. Waste grease cost based on Haas et al, 2006 data, but adjusted to incorporate increased quantity of methanol and different catalyst using data from Zhang et al, 2003.

4) Based on data for a virgin oil facility from Haas et al, 2006, but increased operator salary to $20/hr from $12.5/hr, based on Antares experience. Values in the literature suggest that acid catalyzed facilities have higher labor requirements, so Antares estimated there would be one additional operator per shift (for a 10 MGY facility). For comparison, the base catalyzed facility had two operators per shift. Labor cost includes scaling by facility size (based on available data from pyrolysis oil production facilities).

5) Other fixed O&M costs (including taxes, insurance, and supplies) are 2.1% of capital cost.



Feedstock (1)	**Yield range (gal/ton)**	**Average Yield (gal/ton)**	***Sources***
Virgin Oil (2)	246.4-270.0	258.2	*Haas et. al 2006, Sheehan et al. 1998, Zhang et al. 2003a, Canakci and van Gerpen 2001*
Animal Fats	266.3	266.3	*Bender 1999*
Yellow Grease (3)	234.9 - 263.2	249.1	*Zhang et al. 2003a, Canakci and van Gerpen 2001*

1) All data based on feedstocks with negligible water content.

2) Based on data for degummed soybean oil. High end of the yield range from Haas et al, 2006, which assumes a negligible free fatty acid content in the feedstock.

3) Conversion yield for waste greases based on acid catalyzed reaction. For comparison, the transesterification reaction for virgin feedstocks use base catalysts.

Summary Page - Costs & Yield Data, Renewable Diesel Co-processing - *Small Scale Plant*

Inputs

Feedstock Type (1)	**Virgin Oil**
Economic Lifetime of Plant (Years)	25
Weighted Cost of Capital	10%

1) Feedstock type does not affect the yield of the processes in this model. Typical heating value of these feedstocks are very similar, generally about 16,500 to 17,000 Btu/lb HHV (US EPA 2001). Although processing of yellow grease feedstocks may lead to higher capital costs due to additional pre-processing (such as desalting) and different metallurgy requirements, this has not been included as the added cost is expected to be minimal (a few percent according to discussions with industry contacts).

	Stand Alone	**Co-processing**
Applicable Feedstocks	virgin oil, animal fat, waste grease	virgin oil, animal fat, waste grease
Applicable Size Range (MGY) (1)	15 - 200	15 - 200
Feedstock Input (ton/yr)	58,735	59,895
Yield Renewable Diesel (ton/yr) (2)	48,750	48,755
Yield Renewable Diesel (MGY)	15.0	15.0
Conversion Efficiency (LHV) (3)	98%	96%
Consumables and By-Products		
H2 consumption (ton/yr) (4)	881	2,276
Light hydrocarbons / propane (ton/yr)	2,056	2,635
CO2 / Water Generated (ton/yr) (5)	8,810	10,781
Renewable Diesel Production Costs		
Fixed O&M ($/yr)	$ 4,008,025	$ 641,382
Variable O&M ($/yr)	$ 1,340,226	$ 3,457,611
Co-product credit ($/yr)	$ (1,077,948)	$ (1,381,898)
Annual Operating Cost ($/gal)	$ 0.285	$ 0.181
Annual Operating Cost ($/yr)	***$ 4,270,303***	***$ 2,717,096***
Total Capital Investment - Near Term ($) (6)	***$ 72,873,189***	***$ 11,661,499***
Total Capital Investment - Mid Term ($) (7)	***$ 58,376,516***	***$ 9,341,675***
Total Capital Investment - Near Term ($gal/yr)	$ 4.86	$ 0.777
Total Capital Investment - Mid Term ($/gal/yr)	$ 3.89	$ 0.623
Total Capital Investment - Near Term ($/ton/yr)	$ 1,241	$ 194.7
Total Capital Investment - Mid Term ($/ton/yr)	$ 994	$ 156.0
Fixed Charge Rate	12.3%	12.3%
Non-Feedstock Production Cost Near Term ($/gal)	**$ 0.88**	**$ 0.28**
Non-Feedstock Production Cost Mid Term ($/gal)	**$ 0.76**	**$ 0.26**

1) Estimated based on current and planned facility sizes for ConocoPhillips, Neste, and Petrobras. The smallest faciliy sizes produce around 1,000 barrels per day of renewable diesel, and the largest will be able to generate about 12,000 barrels per day.

2) Stand-alone yield is based on data from UOP (2005), co-processing yield is based on data for Petrobras H-BIO system (source: Petrobras n.d.). Yields are converted to volumetric quantities using a typical green diesel density of 6.5 lb/gal.

3) Calculated using a lower heating value of 16,000 Btu/lb for the oil/grease feedstock, and 123,200 Btu/lb for renewable diesel, based on data from UOP 2005.

4) Hydrogen consumption is based on feedstock input (by weight), and is about 1.5% for stand-alone and 3.8% for co-processing (UOP 2005).

5) Stand alone processing conditions lead to production of CO2 primarily, while co-processing conditions typically produce water. The amount of CO2 and water produced is calculated by difference of the inputs and other outputs. Result is similar to UOP 2005 estimate that water and CO2 output is equivalent to 12-16 wt% of feedstock input.

6) Capital costs based on reported costs of facilities under development from Neste (stand alone) and Petrobras (co-processing), using a scaling factor of 0.6 (equivalent to the scaling factor for pyrolysis hydrotreatment following UOP 2005).

7) Cost decreased based on next generation facility development with a learning rate of 20%, following McDonald and Schrattenholzer (2002). This rate is similar to that found for ethanol production facilities and retail petrol processing.

O&M Costs (1)	**Stand Alone**	**Co-processing**
Variable O&M (2)		
Total Utility Cost ($/yr)	$ 2,995	$ 3,054
Hydrogen ($/yr)	$ 1,337,231	$ 3,454,557
Total Variable O&M ($/yr)	***$ 1,340,226***	***$ 3,457,611***
Fixed O&M		
Total Fixed O&M ($/yr) (3)	$ 4,008,025	$ 641,382
Co-Product Value		
Propane ($/yr) (4)	$ 1,077,948	$ 1,381,898

1) All costs reported in 2008$, converted using CEPCI where needed.

2) Total utility cost is based on data for pyrolysis oil hydrotreatment from UOP 2005. Annual hydrogen cost is based on hydrogen consumption and cost data from UOP 2005. Any additional costs for catalysts and waste disposal are not included here due to lack of available data.

3) Fixed O&M costs include maintenance, insurance, and overhead. These costs are estimated to be 5.5% of near term capital cost, based on pyrolysis oil hydrotreatment O&M. Assume no added labor cost as process is part of larger refinery operation.

4) Based on reported wholesale price of propane ($1.11/gal) in 2006, from EIA AER 2009. Estimated added cost for compressing propane is negligible and has not been included.

Summary Page - Costs & Yield Data, Renewable Diesel Co-processing - *Large Scale Plant*

Inputs

Feedstock Type (1)	**Virgin Oil**
Economic Lifetime of Plant (Years)	25
Weighted Cost of Capital	10%

1) Feedstock type does not affect the yield of the processes in this model. Typical heating value of these feedstocks are very similar, generally about 16,500 to 17,000 Btu/lb HHV (US EPA 2001). Although processing of yellow grease feedstocks may lead to higher capital costs due to additional pre-processing (such as desalting) and different metallurgy requirements, this has not been included as the added cost is expected to be minimal (a few percent according to discussions with industry contacts).

	Stand Alone	**Co-processing**
Applicable Feedstocks	virgin oil, animal fat, waste grease	virgin oil, animal fat, waste grease
Applicable Size Range (MGY) (1)	15 - 200	15 - 200
Feedstock Input (ton/yr)	783,000	798,500
Yield Renewable Diesel (ton/yr) (2)	649,890	649,979
Yield Renewable Diesel (MGY)	200.0	200.0
Conversion Efficiency (LHV) (3)	98%	96%
Consumables and By-Products		
H2 consumption (ton/yr) (4)	11,745	30,343
Light hydrocarbons / propane (ton/yr)	27,405	35,134
CO2 / Water Generated (ton/yr) (5)	117,450	143,730
Renewable Diesel Production Costs		
Fixed O&M ($/yr)	$ 18,960,480	$ 3,034,224
Variable O&M ($/yr)	$ 17,866,639	$ 46,095,713
Co-product credit ($/yr)	$ (14,370,195)	$ (18,423,005)
Annual Operating Cost ($/gal)	$ 0.112	$ 0.154
Annual Operating Cost ($/yr)	***$ 22,456,924***	***$ 30,706,933***
Total Capital Investment - Near Term ($) (6)	***$ 344,736,007***	***$ 55,167,713***
Total Capital Investment - Mid Term ($) (7)	***$ 276,157,631***	***$ 44,193,193***
Total Capital Investment - Near Term ($gal/yr)	$ 1.72	$ 0.276
Total Capital Investment - Mid Term ($/gal/yr)	$ 1.38	$ 0.221
Total Capital Investment - Near Term ($/ton/yr)	$ 440	$ 69.1
Total Capital Investment - Mid Term ($/ton/yr)	$ 353	$ 55.3
Fixed Charge Rate	12.3%	12.3%
Non-Feedstock Production Cost Near Term ($/gal)	**$ 0.32**	**$ 0.19**
Non-Feedstock Production Cost Mid Term ($/gal)	**$ 0.28**	**$ 0.18**

1) Estimated based on current and planned facility sizes for ConocoPhillips, Neste, and Petrobras. The smallest faciliy sizes produce around 1,000 barrels per day of renewable diesel, and the largest will be able to generate about 12,000 barrels per day.

2) Stand-alone yield is based on data from UOP (2005), co-processing yield is based on data for Petrobras H-BIO system (source: Petrobras n.d.). Yields are converted to volumetric quantities using a typical green diesel density of 6.5 lb/gal.

3) Calculated using a lower heating value of 16,000 Btu/lb for the oil/grease feedstock, and 123,200 Btu/lb for renewable diesel, based on data from UOP 2005.

4) Hydrogen consumption is based on feedstock input (by weight), and is about 1.5% for stand-alone and 3.8% for co-processing (UOP 2005).

5) Stand alone processing conditions lead to production of CO2 primarily, while co-processing conditions typically produce water. The amount of CO2 and water produced is calculated by difference of the inputs and other outputs. Result is similar to UOP 2005 estimate that water and CO2 output is equivalent to 12-16 wt% of feedstock input.

6) Capital costs based on reported costs of facilities under development from Neste (stand alone) and Petrobras (co-processing), using a scaling factor of 0.6 (equivalent to the scaling factor for pyrolysis hydrotreatment following UOP 2005).

7) Cost decreased based on next generation facility development with a learning rate of 20%, following McDonald and Schrattenholzer (2002). This rate is similar to that found for ethanol production facilities and retail petrol processing.

O&M Costs (1)	**Stand Alone**	**Co-processing**
Variable O&M (2)		
Total Utility Cost ($/yr)	$ 39,926	$ 40,717
Hydrogen ($/yr)	$ 17,826,713	$ 46,054,997
Total Variable O&M ($/yr)	***$ 17,866,639***	***$ 46,095,713***
Fixed O&M		
Total Fixed O&M ($/yr) (3)	$ 18,960,480	$ 3,034,224
Co-Product Value		
Propane ($/yr) (4)	$ 14,370,195	$ 18,423,005

1) All costs reported in 2008$, converted using CEPCI where needed.

2) Total utility cost is based on data for pyrolysis oil hydrotreatment from UOP 2005. Annual hydrogen cost is based on hydrogen consumption and cost data from UOP 2005. Any additional costs for catalysts and waste disposal are not included here due to lack of available data.

3) Fixed O&M costs include maintenance, insurance, and overhead. These costs are estimated to be 5.5% of near term capital cost, based on pyrolysis oil hydrotreatment O&M. Assume no added labor cost as process is part of larger refinery operation.

4) Based on reported wholesale price of propane ($1.11/gal) in 2008, from EIA AER 2009. Estimated added cost for compressing propane is negligible and has not been included.

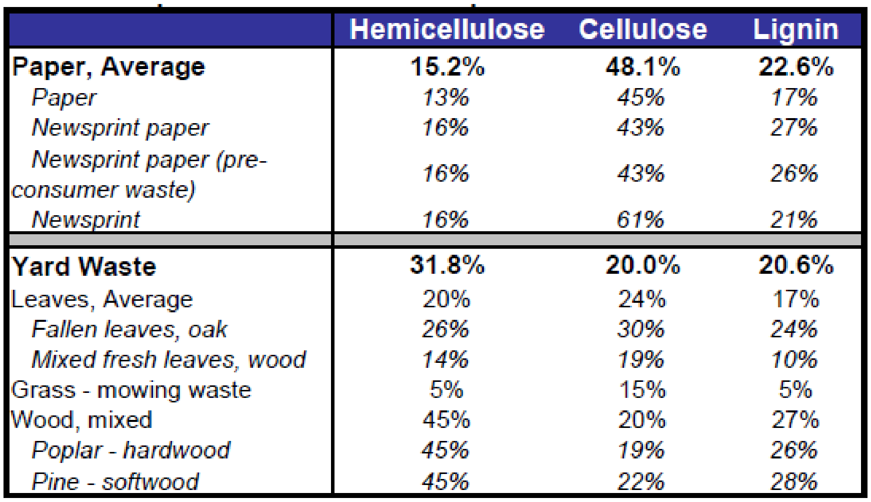
APPENDIX H-G
BIOMASS RECOVERED FROM MSW

Biomass materials in the MSW stream include paper and cardboard, green waste, food waste, construction wood waste. Engineers at Taylor Recycling estimate that today's systems for culling biomass from the mixed stream will add about $25 to $30 per ton to the cost of biomass recovered from a mixed waste stream.[129] This cost covers the labor for hand picking clean materials or alternatively for rejecting visibly tainted materials, as well as O&M and capital recovery. Equipment for biomass recovery can include trommel screens, vibratory screens, air clarifiers, finger screen separators and shredders.

While much of the recovered biomass material can potentially be used for gasification-based processes like Fischer Tropsch (LCMD – G/FT), the same is not true for biochemical processes such as LCEt Fermentation/Hydrolysis. The LCEt Fermentation/Hydrolysis process is particularly sensitive to feedstock composition quality and consistency as well as poisoning from contaminants, which limits the type of recovered materials that could be used. Based on discussions with industry experts, it has been determined that at least a portion of the recovered paper products and yard trimmings could potentially be used for the LCE Fermentation/Hydrolysis process, in addition to clean wood waste residues.

The compositions of the MSW materials applicable for the LCEt Fermentation/Hydrolysis process are shown in Table H-29. The paper composition is an average for newsprint and other paper. The yard waste is assumed to be a mix of leaves (20% by weight), grass (20%) and woody residues (60%), based on data from McKeever and Skog (2003). The yard waste composition is calculated as a weighted average of these materials.

Table H-29. Composition of Some Components of MSW.



	Hemicellulose	Cellulose	Lignin
Paper, Average	**15.2%**	**48.1%**	**22.6%**
Paper	*13%*	*45%*	*17%*
Newsprint paper	*16%*	*43%*	*27%*
Newsprint paper (pre-consumer waste)	*16%*	*43%*	*26%*
Newsprint	*16%*	*61%*	*21%*
Yard Waste	**31.8%**	**20.0%**	**20.6%**
Leaves, Average	20%	24%	17%
Fallen leaves, oak	*26%*	*30%*	*24%*
Mixed fresh leaves, wood	*14%*	*19%*	*10%*
Grass - mowing waste	5%	15%	5%
Wood, mixed	45%	20%	27%
Poplar - hardwood	*45%*	*19%*	*26%*
Pine - softwood	*45%*	*22%*	*28%*

Data from (Energy Research Centre of the Netherlands, n.d.)

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[129] Teleconference with Mark Paisley, Director of R&D for Taylor Recycling, on December 10, 2007