12-2006

Production of Ethanol from Hardwood

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FROM HARDWOOD

By

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B.S. The University of Maine, 2004

A THESIS

Submitted in Partial Fulfillment of the

Requirements for the Degree of

Master of Science

(in Chemical Engineering)

The Graduate School

The University of Maine

December, 2006

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Diminishing crude oil and natural gas supplies, along with concern about greenhouse gas are major driving forces in the search for efficient renewable energy sources. The conversion of lignocellulosic biomass to energy and useful chemicals is a component of the solution. Ethanol is most commonly produced by enzymatic hydrolysis of complex carbohydrates to simple sugars followed by fermentation using yeast.

$$C_6H_{10}O_5 \quad + \quad H_2O \xrightarrow{Enxymes} C_6H_{12}O_6 \xrightarrow{Yeast} 2\,CH_3CH_2OH \quad + \quad 2\,CO_2$$

In the U.S. corn is the primary starting raw material for commercial ethanol production. However, there is insufficient corn available to meet the future demand for ethanol as a gasoline additive. Consequently a variety of processes are being developed for producing ethanol from biomass; among which is the NREL process for the production of ethanol from white hardwood.

The objective of the thesis reported here was to perform a technical economic analysis of the hardwood to ethanol process. In this analysis a Greenfield plant was

compared to co-locating the ethanol plant adjacent to a Kraft pulp mill. The advantage of the latter case is that facilities can be shared jointly for ethanol production and for the production of pulp. Preliminary process designs were performed for three cases; a base case size of 2205 dry tons/day of hardwood (52 million gallons of ethanol per year) as well as the two cases of half and double this size. The thermal efficiency of the NREL process was estimated to be approximately 36%; that is about 36% of the thermal energy in the wood is retained in the product ethanol and by-product electrical energy.

The discounted cash flow rate of return on investment and the net present value methods of evaluating process alternatives were used to evaluate the economic feasibility of the NREL process. The minimum acceptable discounted cash flow rate of return after taxes was assumed to be 10%. In all of the process alternatives investigated, the dominant cost factors are the capital recovery charges and the cost of wood. The Greenfield NREL process is not economically viable with the cost of producing ethanol varying from $2.58 to $2.08/gallon for the half capacity and double capacity cases respectively.

The co-location cases appear more promising due to reductions in capital costs. The most profitable co-location case resulted in a discounted cash flow rate of return improving from 8.5% for the half capacity case to 20.3% for the double capacity case. Due to economy of scale, the investments become more and more profitable as the size of the plant increases. This concept is limited by the amount of wood that can be delivered to the plant on a sustainable basis as well as the demand for ethanol within a reasonable distance of the plant.

ACKNOWLEDGMENTS

First and foremost I would like to express my deepest gratitude to my thesis advisor, Dr. Joseph M. Genco. This thesis would not have been possible without his hard work, expertise, understanding and guidance. I would like to thank my thesis committee members, Dr. John Hwalek, Dr. Hemant Pendse, and Dr. Adriaan van Heiningen for their insight and assistance. I would also like to thank the University of Maine Pulp and Paper Foundation, the U.S. Department of Energy and the U.S. Department of Agriculture for their financial support.

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BACKGROUND

Introduction

It is widely speculated that within a few decades, the world's crude oil and natural gas supplies will not longer be sufficient to meet global needs for transportation, energy and chemical products ("Biorefinery gets ready", 2006). One potential solution is the development of bio-refineries, or facilities that use thermal and biological processes to convert starch, cellulose and lignin from woody biomass, dedicated annual crops and municipal waste into basic chemicals that can in turn be refined to make fuels, polymers and other consumer products (Ragauskas et al., 2006).

The biorefinery concept impacts directly upon the Forest Products and agricultural Industries, which are predicated upon selling large quantities of commodity products at modest prices. The Forest Products Industry is exploring the biorefinery concept with an eye towards the viability of producing chemical intermediaries in addition to paper and solid wood products. The forest biorefinery concept builds on the principles used by the petrochemical industry. In a petrochemical refinery for example, the raw material is normally crude oil and the end products are gasoline, fuel oils and a variety of petroleum distillates, and chemical feedstocks. In the forest biorefinery concept, the raw material would be wood and woody biomass and the end products would be a variety of chemicals that could be used for energy and as chemical feedstock. The forest biorefinery is not a new concept (Hawley, 1921) and there are numerous examples of chemical pulp mills that produce a variety of organic chemicals in addition to paper products; for example terpenes, resins and fatty acids, fragrances, charcoal, and vanillin. What is new is the scale and variety of products being considered in the modern forest biorefinery concept.

Processing Pathways

There are four pathways under development for the conversion of biomass into useful products and they involve either thermal or biochemical processing (Figure 1-1). The first pathway involves using biomass to produce electrical energy and process steam and is clearly the simplest alternative. Commercial biomass boilers are operated in the Northeast United States to generate electrical energy. In Kraft pulp mills, dissolved wood solids termed "black liquor" is routinely burned in chemical recovery boilers to generate steam and electrical power. Black liquor gasification technology is being developed in an effort to replace the Tomlinson recovery boiler. Black liquor gasification has been under development for several years and holds the potential for increased production of electrical energy and steam that can be exported from Kraft pulp mills.

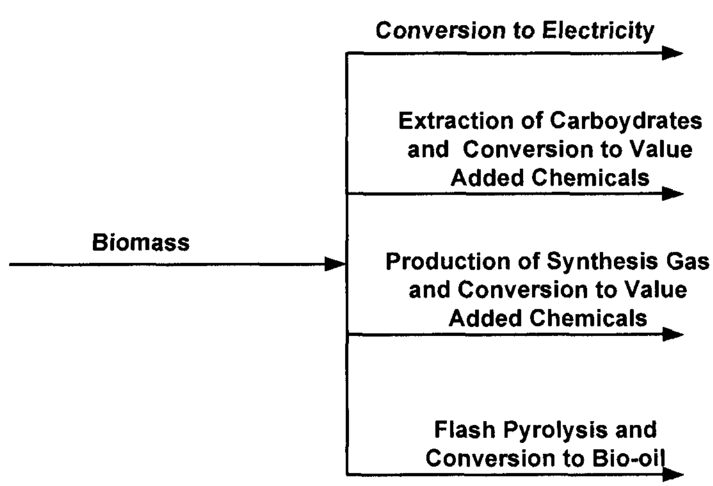


Figure 1-1. Pathways Available for Utilizing Biomass

The second pathway involves extracting carbohydrates from the biomass, converting the complex carbohydrates to simple sugars, and converting them to fuels such as ethanol and potentially to a variety of value added chemicals (Werpy & Petersen, 2004). Component sugars can be derived from woody biomass, starch, and agricultural and municipal waste in this pathway. The criteria for selecting chemicals from biomass sugars are chemical intermediates that have at least two functional groups that can be converted to high value added chemicals. Succinic, fumaric and malic acids are examples of chemical intermediates that have two carboxylic acid groups and can be used as polymer feedstocks.

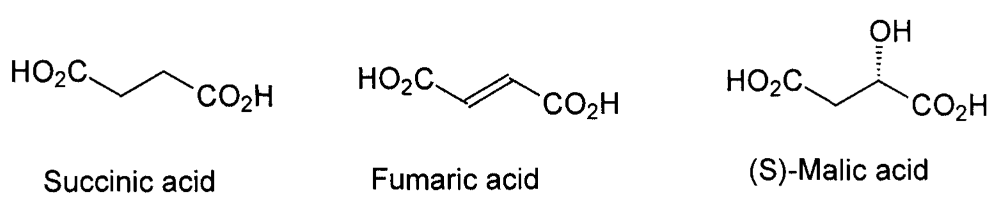


Figure 1-2. Four Carbon Di-Basic Acids Produced from Simple Sugars

The third option involves producing synthesis gas from the biomass, which can then be converted catalytically into hydrogen, methanol, dimethyl either and liquid fuels (Figure 1-3) by what is commonly know as Fischer Tropsch Synthesis. Fischer Tropsch technology was developed in Germany and used during the Second World War to produce liquid fuels. This is currently done commercially in South Africa.

A fourth option related to gasification involves flash pyrolysis or the rapid heating of biomass in the absence of air to produce organic vapors, pyrolysis gases and char (BTG, 2006). The pyrolysis vapors are condensed to oxygenated liquids termed bio-oils that can be used as a fuel. Products in the bio-oil are primarily phenol, levoglucosan, hydroxyl-acetaldehyde and water. Pyrolysis gases include carbon monoxide, carbon dioxide, methane, and hydrogen.

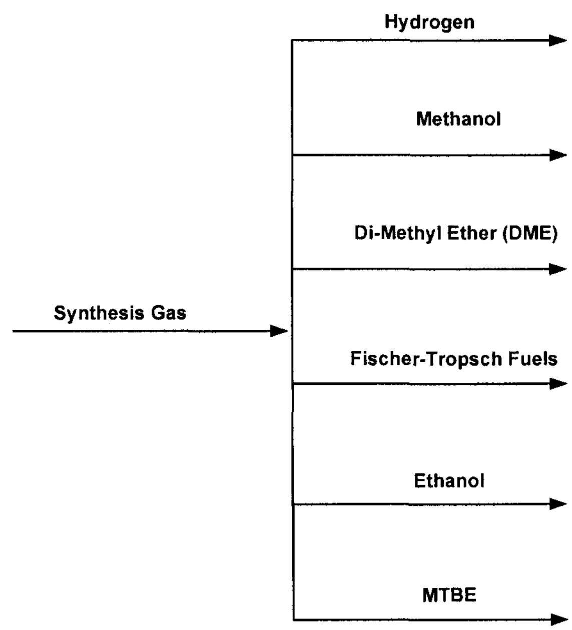


Figure 1-3. Products from Synthesis Gas (Wising and Stuart, 2006)

A variant of flash pyrolysis involves the thermal cracking of lignin, a byproduct in pulping and in future bio-refineries. Residual lignin from pulp production is burned for heat and power. However, lignin thermal-cracking studies using temperatures of 250 to 600 °C have demonstrated the potential of generating low molecular weight feedstocks for further processing into intermediate chemicals (Britt, et al., 2000). Lignin cracking catalysts could lower conversion temperatures and provide tighter control over product distribution. Shabtai et al. (2003) cracked lignin in a two-stage catalytic reactor to produce a reformulated, partially oxygenated gasoline-like product. In the first reactor, lignin is depolymerized catalytically into a mixture of phenols, which are then converted catalytically into a mixture of alkyl-benzenes using hydrogen.

Hemicellulose Extraction Process. A novel biorefinery concept proposed for adaptation to Kraft pulp mills involves the extraction of xylan and mannan hemicelluloses prior to the production of Kraft pulp. The extracted hemicelluloses would then be hydrolyzed to component sugars, which would then be converted to ethanol and acetic acid and higher value intermediate chemicals (van Heiningen, 2006). Resin and fatty acids are of course currently being recovered in large quantities from pine species in southern Kraft mills in tall oil plants. Resin and fatty acids can be converted to biodiesel fuel if economically viable. Lastly, processes for precipitating lignin and conversion of lignin into products such as phenolic resins and carbon fibers are also being considered (Wising & Stuart, 2006).

Biomass as a Raw Material

Biomass can be classified as a wide range of materials including wood, grasses, agricultural crops, mill residues, and other biological material. As oil prices rise and for reasons of national security, it is important than the United States become less dependent on foreign oil. Biomass can be used as a renewable source of fuel and energy and has a positive impact on air quality as well. A 2005 study by the Department of Energy estimates that on a sustainable basis there are over 1.3 billion dry tons per year of biomass available in the United States. Of this value, 368 million dry tons per year (28%) is derived from forest resources and 933 million tons per year arise from agricultural resources (72%). The 1.3 billion tons of material has an energy equivalence that is 75% higher than our current domestic oil production (Kelly, 2006).

If utilized efficiently, the biomass could have a major impact on the U.S. energy and chemical industries.

Forest biomass is usually considered to be the standing inventory in the forest, i.e. the aggregation of tree components found both above and below ground as well as needles and leaves. This involves combinations of live and dead trees, standing and down trees, saplins and shrubs (McWilliams et al., 2005). White wood is derived from the merchantable bole of tree, that is from the stump to the top of the tree exclusive of branches, after it has been debarked. Secondary forest residuals result from bark, sawdust, and wood shavings. Tertiary forest biomass is mulches usually from bark, needles and leaves (McWilliams et al., 2005).

<u>Characterization of Biomass</u>

Biomass can be characterized by its source, elemental composition and energy content (Table 1-1). Table 1-1 illustrates the composition for a few wood species, annual crops and bark. The elemental composition of biomass varies depending upon the source and whether it results from an agricultural residue or from woody biomass. The elemental composition of wood is surprisingly close to 50% carbon (C), 44% oxygen (O), 6% hydrogen (H), 0.1% Nitrogen. The heating value of woody biomass typically is between 8,400 and 8,500 BTU per pound mass on a dry basis depending upon species.

Table 1-1
Elemental Composition of Biomass
(US DOE Energy Efficiency and Renewable Energy)

Biomass Description	Carbon (%C)	Hydrogen (%H)	Nitrogen (%N)	Oxygen (%O) Δ	HHV (BTU/dry lb.)
Monterey Pine *Pinus Radiata*	50.26	5.98	0.03	-42.14	8422
Hybrid Poplar (*Populus deltoids x P.nigra*)	49.75	5.52	0.52	-44.42	8384
Corn Stover *Zea mays*	47.04	5.47	0.68	-41.1	7967
Wheat Straw *Triticum aestivum*	43.88	5.26	0.63	-38.75	7481
Bark (Avg. of pine, oak, and spruce)	51.6	5.6	0.2	-38.5	8713
Δ = By Difference					

The elements are combined to form identifiable biomass substances, the most important of which is cellulose (Table 1-2). The other major constituents are the hemicelluloses or cellulose like polymers, lignin and a group of compounds called "extractives". Annual crops are notably different from woody biomass in that the ash content, especially silica, is extremely high (Table 1-2). On a dry-wood basis, the relative amounts of the major constituents are: cellulose 40 to 45%, hemicelluloses 20 to 30%, lignin 18 to 25% in hardwoods and 25 to 35% in softwoods, and 3 to 8% extractives. Cellulose and hemicelluloses are polymers of simple sugars, termed "polysaccharides". The hemicelluloses and lignin are amorphous while the cellulose is crystalline for the most part (70%). The extractives are low molecular weight materials such as phenols, turpines, resin acids and aliphatic compounds. The amount and types of the extractives removed from biomass will depend upon the solvent used, typically water and organic solvents such as benzene, dimethylchloride and ether.

Table 1-2
Chemical Composition of Biomass
(US DOE Energy Efficiency and Renewable Energy)

Biomass Description	Ash (%)	Extractives (%)	Carbohydrates (%) Δ	Lignin (%)
Monterey Pine *Pinus Radiata*	0.3	2.7	-71.1	25.9
Hybrid Poplar (*Populus deltoids x P.nigra*)	2.03	6.89	-65.9	25.18
Corn Stover *Zea mays*	10.24	7.74	-65.9	17.69
Wheat Straw *Triticum aestivum*	10.22	12.95	-60	16.85
Bark (Softwood)*	Up to 20	2-25	30-48	40-55
Bark (Hardwood)*	Up to 20	5-10	32-45	40-50
Δ = By Difference * = (U.S. Department of Agriculture, 1971)				

The major sugar polymers comprising biomass, whether agricultural or woody, are uronic acids (anhydride), arabinan, xylan, mannan, galactan and glucan (Table 1-3). Acetyl groups are also found in biomass as pendant groups attached to the hemicellulose polymers, principally glucomannan and galactoglucommanan, and glucuronoxylan and glucuronoarabinoxylan. The xylan polymers are principally found in hardwood biomass and annual crops; while the mannan polymers are principally found in softwood biomass (Table 1-3). The acetyl content of biomass varies between 1 and about 6% depending upon the species. Hardwoods and annual crops like cornstalks (*Zea mays*) have greater contents of acetyl groups than softwoods (about 1 to 2%).

Table 1-3
Component Sugar Polymers in Biomass
(US DOE Energy Efficiency and Renewable Energy Program)

Biomass Description	Uronic Acids (%)	Arabinan (%)	Xylan[b] (%)	Mannan[b] (%)	Galactan (%)	Glucan (%)	% Mass Closure (a)
Monterey Pine *Pinus Radiata*	2.5	1.5	5.9	10.7	2.4	41.7	93.6
Hybrid Poplar (*Populus deltoids x P.nigra*)	4.31	0.89	13.07	1.81	0.88	39.23	94.3
Corn Stover *Zea mays*	3.12	2.54	18.32	0.4	0.95	34.61	95.6
Wheat Straw *Triticum aestivum*	2.24	2.35	19.22	0.31	0.75	32.64	97.53

(a) Total Sugar Polymers
(b) Acetyl groups constitute about 1 to 5% of the weight fraction of the biomass and reside as pendant groups on mannan and xylan polymers.

Heating Value

The heating value of biomass (Table 1-1) is considerably lower than the heating value of conventional fossil fuels (Table 1-4). Bituminous coal for example has a heating value of approximately 13,000 BTU/dry lb compared to about 8,400 BTU/lb for woody biomass and about 7,500 BTU per pound for agricultural waste (Table 1-1). By contrast the heating value of black liquor is approximately 6,000 BTU/pound dry solids compared to No. 2 fuel oil which has a heating value of approximately 19,500 BTU/pound. The major difference between conventional fossil fuels and biomass is the higher content of carbon, hydrogen and sulfur in the fossil fuels. In addition fossil fuels such as coal have a significant quantity of ash or inorganic matter. As the data above suggests the energy density of biomass is low when compared to conventional fossil fuels. Since biomass would be converted into other fuel or chemical forms, a larger amount of biomass would have to be used to get an equivalent amount of energy in an alternative form and this would impact the process economics.

Table 1-4
Elemental Composition of Fossil Fuels

Biomass Description	Ash (%)	Carbon (%C)	Hydrogen (%H)	Nitrogen (%N)	Oxygen (%O)	Sulfur (%)	HHV (BTU/dry lb.)
Bituminous Coal	(a)	75	5	1.5	6.7	2.3	13,000
No. 2 Fuel Oil	Nil	87.2	12.5	0.02	0	0.3	19,430
Gasoline	Nil				nil	nil	20,007
Natural Gas	Nil	69.12	23.2	5.76	1.58	0.34	22,077
Black Liquor (a)	(b)	34	3	0	34	5	6,000
(a) Ash content of coal varies between 3% and 12%.							
(b) Black liquor also contains sodium (5%), potassium (1%), and chlorine (0.5%)							

Technologies for Converting Biomass into Energy and Chemicals

A variety of technologies are being developed to convert biomass into useful energy and chemicals. The most basic process is to simply burn the biomass to produce steam, process heat, and electricity. This is already being done commercially and is the standard by which alternative processes are sometimes compared.

Biomass Boilers and Conversion to Steam and Electrical Energy

In 2004 the U.S. energy consumption was 100.3 quadrillion (10^{15}) Btu (EIA, 2005). Renewable energy accounted for 6 percent of the total energy being used. Biomass accounted for 47% of the renewable energy, which amounts to about 2.8 quadrillion Btu. The breakdown of this energy was as follows: 70% from wood, 20%

from municipal waste and %10 from alcohol fuels. Most of the energy from wood, municipal and other wastes resulted when these two fuel sources were burned to produce steam and electrical energy at chemical pulp (Kraft) mills, saw mills and other wood processing facilities. During the period from 2003 to 2004 U.S. industrial energy consumption increased by 2 percent. During the same period, industrial biomass energy consumption increased by approximately 6 percent.

Biomass Boilers. Biomass boilers traditionally burn bark, white wood, and other lignocellulosic materials to produce steam and electrical energy (Figure 1-4) usually in "hog" fuel boilers at pulping facilities and saw mills. There are two situations. First, all of the steam generated in the boiler can be used to generate electrical energy and the steam at low pressure is condensed in a surface condenser. Alternatively the biomass can be burned in a cogeneration boiler. In the cogeneration system, for example in pulp and paper mills, the energy content in the biomass is used to generate electrical energy at high pressure and the low pressure steam leaving the turbine is used for process heating (Huhtinen & Hotta, 1999). The difference between the two processes resides with the pressure of the steam leaving the turbine, typically 27 to 28 inches Hg vacuum in the case of the power boiler and 30 psig in the case of the co-generation system.

In a biomass boiler, the biomass is stored in a wood yard where magnets are used to remove tramp metals. It is then reduced in size using a hammer mill and stored. The lignocellulosic material is conveyed from storage to the biomass boiler and injected into the combustion zone of the boiler. In the combustion zone, primary-, secondary- and tertiary-air is introduced and combustion reactions take place that release the energy content in the biomass and produce hot flue gases. The hot flue gas generates superheated steam at high pressure and temperature from boiler feed water that passes through tubes in the boiler. The resultant superheated steam is then sent to a steam turbine. The steam can be taken off before the turbine, de-superheated and sent to the process as high pressure steam. Alternatively it can be taken directly from the turbine at intermediate and low pressures for use as process steam; typically at 150 psig (1.03×10^3 kPa gage) and 30 psig (207 kPa gage). The mechanical energy extracted from the turbine is used to produce electrical power in a generator. Steam exiting from the turbine is condensed in a total condenser using cold water. Condensate from the total condenser is returned to the boiler after it is preheated in the feed water tank. Condensate losses are made up with boiler feed water, which has been treated to remove air and metal ions and then pre-heated.

In conventional biomass boilers where thermal energy is converted into electrical energy, the thermal efficiency (η,%) is considerably lower than the efficiency in large central station fossil power plants (Williams, 2004).

$$\eta(\%) = \left[\frac{\textit{Net Electrical Energy Out}}{\textit{Higher Heating Value of Fuel}} \right] x100\%$$

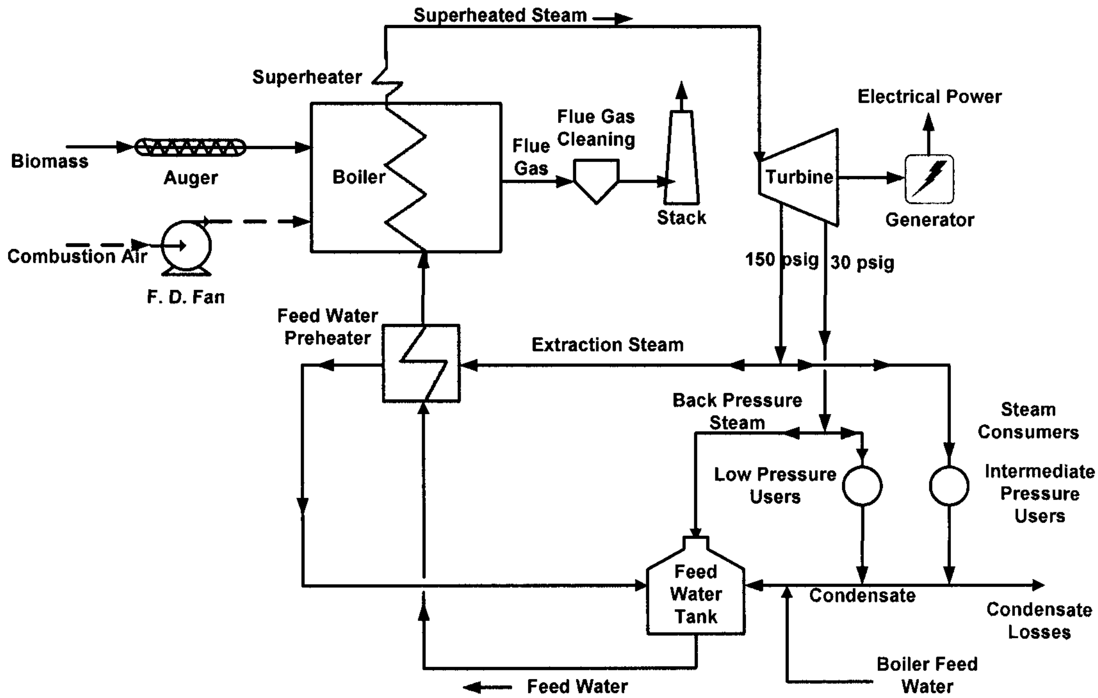


Figure 1-4. Production of Electrical Energy from Biomass Using Rankin Cycle (Huhtinen and Hotta, 2004)

The lower efficiency is due to the smaller facility size and the lower fuel quality, as given by the heating value (Tables 1-1 and 1-4). This latter limitation arises because of the presence of high moisture content in the fuel and because the biomass contains oxygen which normally comes from the air with conventional fossil fuels. The thermal efficiency of conversion for existing biomass based power systems ranges from less than 10% to perhaps as high as 20% depending upon the size and moisture content of the fuel compared to 35 to 40% for large central power faculties. At the lower end of the range for conventional fuels are combustion boiler-steam engine systems, small gasifier-engine systems, and anaerobic digestion-reciprocating engine systems (Williams, 2004). The upper range of efficiency is achieved by larger combustion boiler-steam systems (>40 MW capacity).

Conversion of Biomass to Chemicals

Numerous organizations are working on developing methods of producing value added chemical intermediates from biomass for energy and as a chemical feedstock. Typical biomass conversion technologies include production of ethanol via enzymatic and acid hydrolysis followed by fermentation, gasification of biomass to syngas followed by Fischer Tropsch Synthesis to alcohol and alkanes, fast pyrolysis to produce liquid fuels, aqueous-phase refining of biomass-derived carbohydrates, conversion of biomass to levulinic acid via thermal degradation of cellulose (Fitzpatrick, 2004) and the production of bio-diesel from energy crops. McCloy & O'Connor (1999) and more recently Huber and co-workers (2005 and 2006) review technologies for synthesis of transportation fuels from biomass.

Iogen Process for Production of Ethanol. Iogen Ltd. is a Canadian company that employs an enzymatic hydrolysis process to hydrolyze lignocellulosic materials to simple sugars for the production of ethanol. The process can handle agricultural residues including wheat straw and corn stover as well as hardwood residues. A basic flow diagram of the process is illustrated in Figure 1-5. Iogen employs a steam explosion pretreatment operation that shreds the wood into small matchstick size particles that can be readily digested enzymatically to simple sugars. Iogen has developed proprietary enzymes for the hydrolysis of biomass, which will of course depend upon the composition. The lignin is relatively unharmed during the pretreatment process and is the starting material for lignin based chemicals. Alternatively it can be burned to produce steam and electrical energy. The sugars resulting from the pretreatment and enzymatic hydrolysis are fermented to ethanol using yeast (McCloy & O'Connor, 1999).

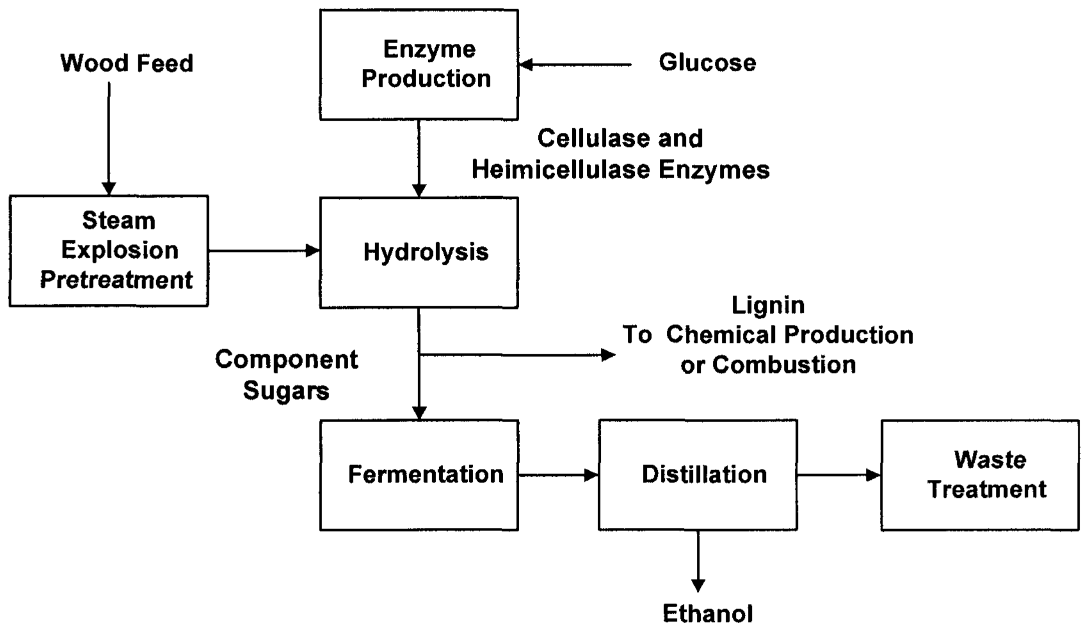


Figure 1-5. Iogen Enzymatic Hydrolysis Process

BC International Process. BC International Corporation (BCI) utilizes acid hydrolysis rather than enzymatic hydrolysis to bring about the dissolution of component sugars. The component sugars are then fermented to produce ethanol. The BC International process is applicable to both agricultural feedstocks and hardwood. The major aspects of the process are outlined in Figure 1-6. The lignocellulosic feed material is hydrolyzed to sugars using a two-stage dilute sulfuric acid hydrolysis at high temperature and elevated pressure. The first stage is hemicellulose hydrolysis and the second stage is cellulose hydrolysis. The aspect of the process that sets BCI apart from others is their proprietary, genetically modified fermentation organism. They claim to use a recombinant organism that is based on multiple organisms; one of which is *E. coli* that has been combined with an ethanol producing gene from *zymomonas*. The organism can ferment hexose and pentose sugars with high efficiency (McCloy & O'Connor, 1999).

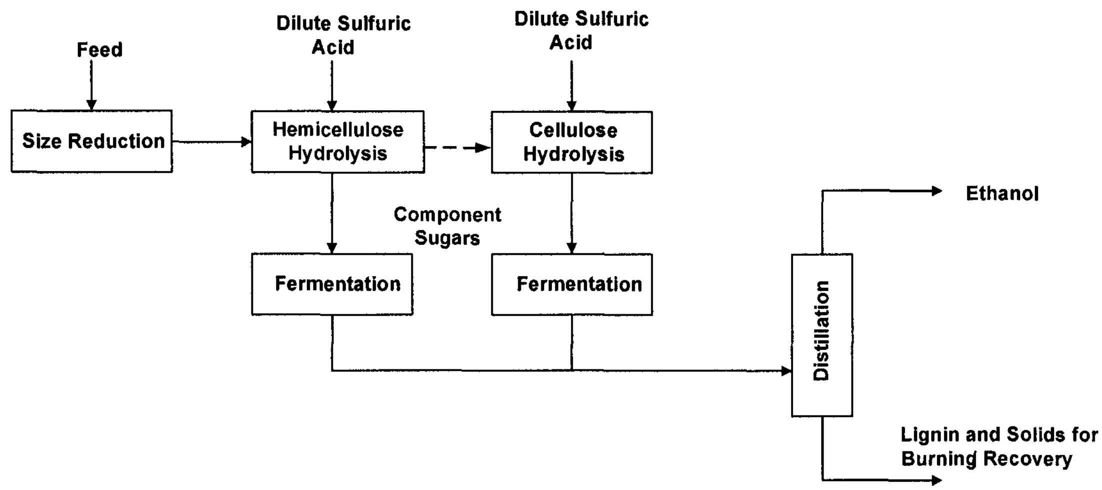


Figure 1-6. BC International Process for the Production of Ethanol

Arkenol Process. Arkenol Inc. is developing a competitive biomass to ethanol process (Figure 1-7). Rice straw is the primary raw material used in the Arkenol process, but woody biomass can also be used. The process utilizes concentrated acid to hydrolyze lignocellulosic biomass and releases condensed lignin. Both a primary and a secondary hydroysis step are used to convert the sugar polymers to component sugars. Concentrated acid is used in the hydrolysis step rather than dilute acid. This leads to faster hydrolysis and is performed at lower temperatures and lower pressures with fewer unwanted byproducts. However, concentrated acid results in higher capital costs, operating costs and waste treatment costs for the process relative to those that use dilute acid. After the second acid hydrolysis step, a fermentation step is used to convert the resulting sugars to ethanol. Lignin and gypsum are also products of the process. The gypsum would be sold and the lignin is burned to produce electricity (McCloy & O'Connor, 1999).

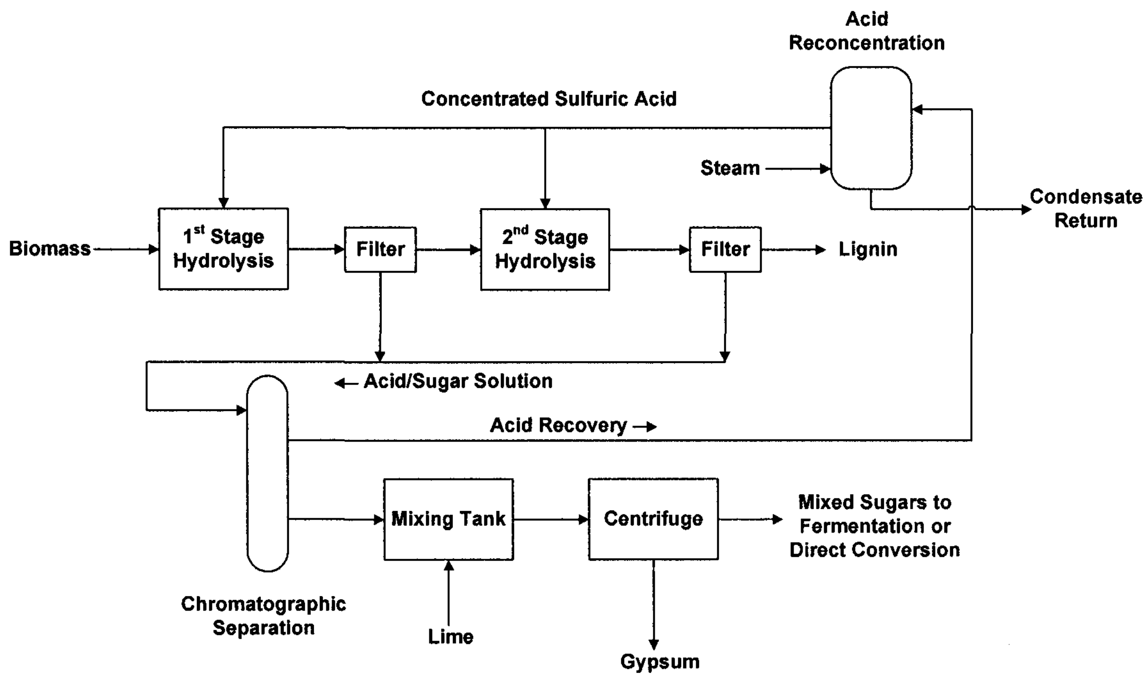


Figure 1-7. Arkenol Process for the Production of Mixed Sugars from Biomass

Acid Catalyzed Organosolv Saccharification Process (ACOS). Figure 1-8 illustrates schematically the Acid Catalyzed Organosolv Saccharification (ACOS) process for producing ethanol from biomass.

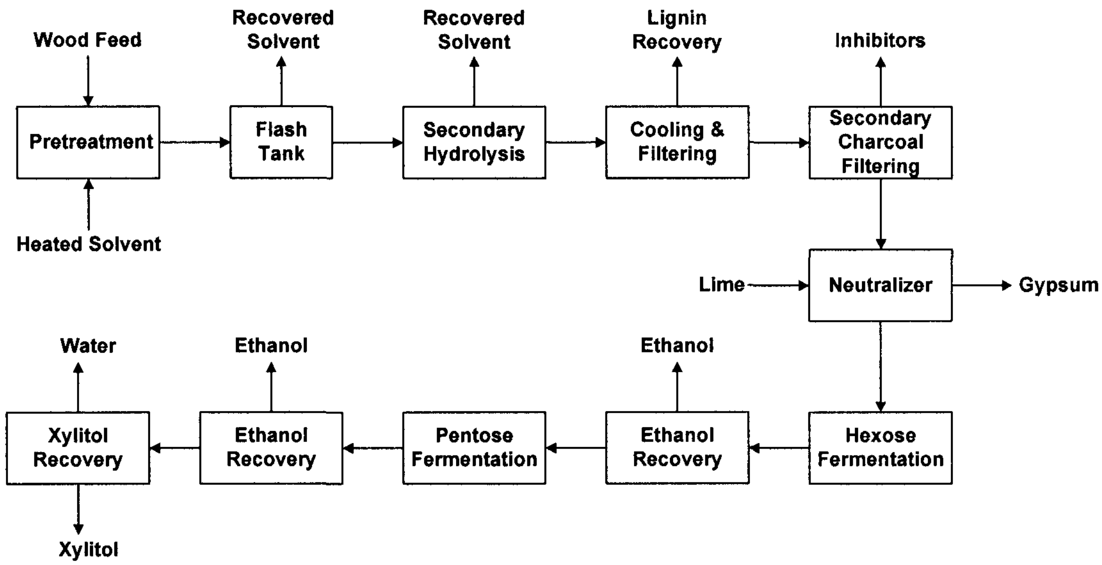


Figure 1-8. ACOS Process for the Production of Ethanol

The ACOS process was developed by Dr. Laszlo Paszner in the early 1980's and is believed to have long-term potential. A unique pretreatment procedure sets the ACOS apart from all the other biomass to ethanol competitors. Lignocellulosic materials are treated with a concentrated acetone solution containing a small quantity of sulfuric acid. The pretreatment takes place at about 200 °C and a pressure of 40 bar. Under these conditions, all of the feedstock components are dissolved and go into solution.

The solution containing the carbohydrates is flashed to recover part of the acetone and the remainder of the acetone is removed during a secondary hydrolysis at approximately 100 °C. The lignin is precipitated and the resulting sugar solution is filtered through charcoal. The result is a very concentrated sugar solution. The hexose sugars are converted to ethanol using fermentation. The ethanol is distilled producing concentrated ethanol and stillage containing pentose sugars. The stillage can then be fermented again producing ethanol and xylitol. Xylitol is a high value sugar that can be used in foods.

The ACOS process has many attractive characteristics. The potential feedstocks include hardwood, softwood, grain, and agricultural residues. The hydrolysis can produce high yields and concentrated sugar solutions. The process also has fewer steps and shorter reaction times than most of the other technologies. The result is a process that could possibly produce more ethanol at a lower cost, not to mention the production of xylitol. Because a non-aqueous solvent is used, the economics will depend strongly on the ability to recover the solvent (McCloy & O'Connor, 1999).

Gasification – Fischer Tropsch Synthesis. Gasification is a process that converts biomass to fuel and value added chemicals. Gasification was developed for coal, oil and natural gas as a method of producing hydrogen and carbon monoxide. This technology has been extended to biomass, municipal waste and sludges. In this process, natural gas, coal, or biomass is heated to high temperatures in a low-oxygen atmosphere and the feed source undergoes partial oxidation (Figure 1-9). Under these conditions the carbonaceous feed source will be gasified to a carbon monoxide and hydrogen mixture known as synthesis gas or syngas.

Natural Gas: $$CH_4 + \frac{1}{2}O_2 \xrightarrow{High\ Temp.} 2\,H_2 + CO$$

Coal: $$CH + \frac{1}{2}O_2 \xrightarrow{High\ Temp.} \frac{1}{2}H_2 + CO$$

Biomass: $$CH_2O \xrightarrow{High\ Temp.} H_2 + CO$$

Converting biomass to syngas provides many advantages over solid fuels. The gaseous mixture can readily mix with oxygen, leading to much greater combustion efficiency than solid biomass. This can be very useful in the case of a biomass power facility. Syngas can also be readily mixed with chemical catalysts, which allows for the conversion to many other fuels and chemical feedstocks.

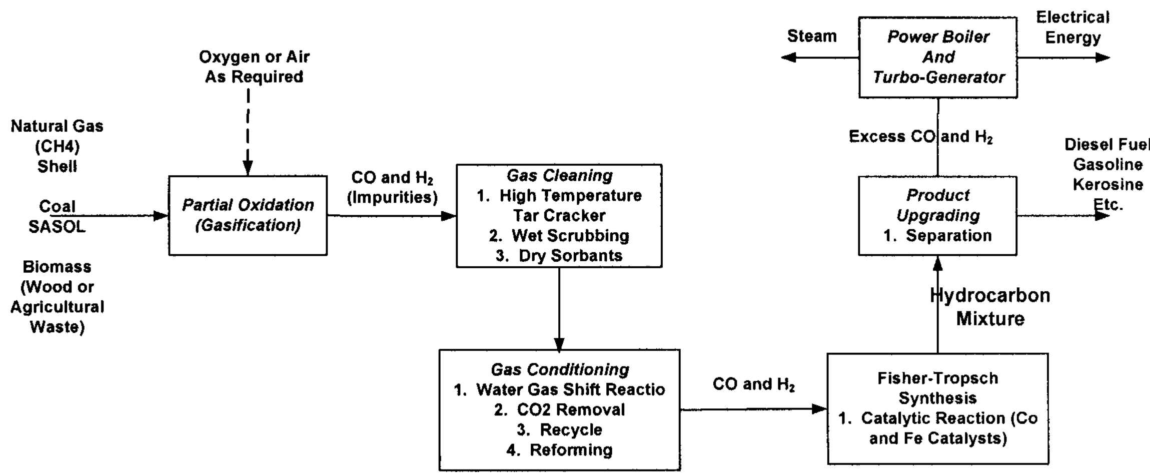


Figure 1-9. Gasification With Fisher Tropsch Synthesis to Hydrocarbon Fuels (Boerrigter, 2002)

The gasified mixture will have a host of impurities as well as tars and must be cleaned. This can be done by using a tar cracker to lower the molecular weight of the tars and convert it to carbon monoxide and hydrogen. Particulate matter is further cleaned by cyclonic separation and gas scrubbing. Sulfur bearing compounds such as hydrogen sulfide (H_2S) and carbonyl sulfide (COS) originating from sulfur bearing fuels are removed by scrubbing and adsorption. Following gas cleanup the carbon monoxide to hydrogen ratio in the synthesis gas may be upgraded using the water gas shift reaction which produces additional hydrogen and carbon dioxide.

Water Gas Shift Reaction: $H_2O + CO = H_2 + CO_2$

If the water gas shift reaction is employed, then the carbon dioxide (CO_2) which is formed must be removed. Purified synthesis gas at the desired CO to Hydrogen ratio is then available for catalytic conversion to fuels.

An example of the catalytic conversion of syngas is the famous Fischer-Tropsch synthesis. Franz Fischer and Hans Tropsch worked to develop this technology in 1923 in Germany where oil was scarce and coal was plentiful. The process converts the synthesis gas into liquid hydrocarbons using iron (Fe) and cobalt (Co) based catalysts. The original Fisher-Tropsch synthesis is described by the following simplified chemical equation:

$$\text{FT Reaction:} \quad (2n+1)\,H_2 \quad + \quad n\,CO \xrightarrow{Catalyst} C_nH_{2n+2} \quad + \quad n\,H_2O$$

Many different products can be made using this process depending on the CO and H_2 ratio, concentrations, temperature and pressure. These products include chemicals used in gasoline and diesel refining, waxes, methanol and other liquid fuels (Clarke, 2006). In the process shown schematically in Figure 1-9, diesel fuel, kerosene and gasoline are being produced (Boerrigter, 2002). Excess carbon monoxide (CO) and hydrogen (H_2) can be used to produce steam and electrical energy for use within the plant or can be exported.

BRI Process. Bioengineering Resources Inc. (BRI) is a process that is predicated on the gasification of biomass followed by fermentation to produce ethanol. In this process (Figure 1-10) biomass is reduced in size, cleaned, and sent to a gasifier (McCloy & O'Connor, 1999). The resulting syngas (CO and H_2) is then fermented directly to ethanol using enzymes developed by BRI. The ethanol is filtered and then separated from stillage by distillation. The crude ethanol at 95% weight percent is dehydrated to produce anhydrous ethanol using molecular sieve technology. Since the biomass is gasified, bark, softwood, sawdust and wood shavings can be used. The reaction time for the fermentation step is reported to be "rapid"; which, if true, is a positive aspect of the process.

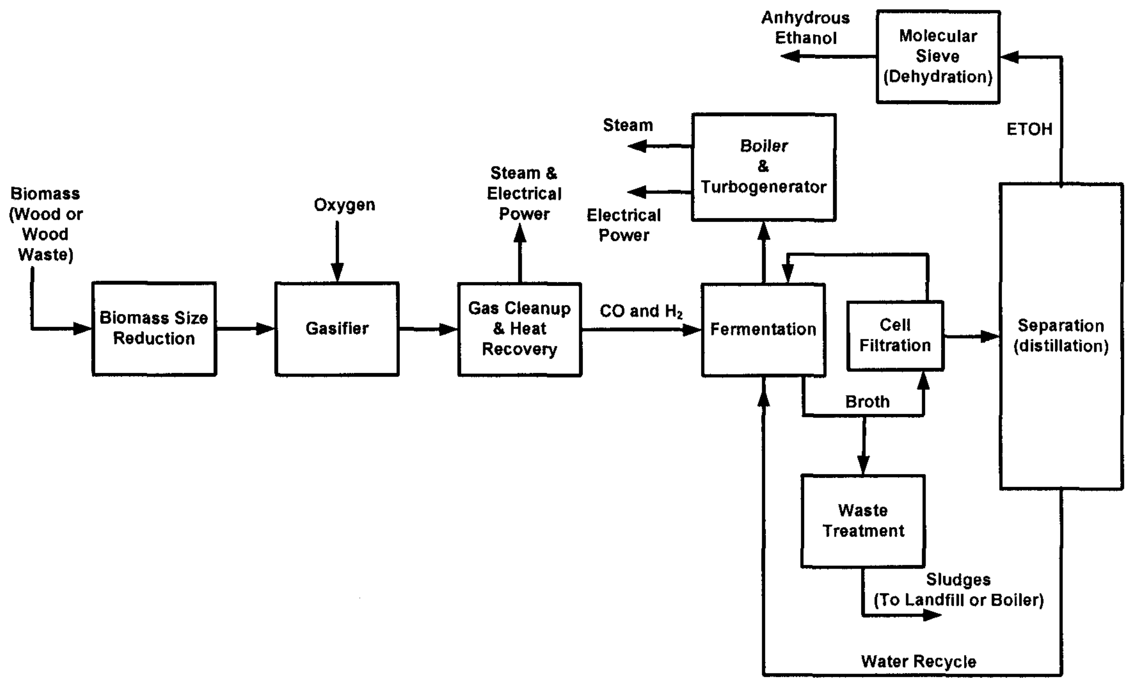


Figure 1-10. Bioengineering Resources Inc. Gasification and Fermentation Process

National Renewable Energy Laboratory (NREL) Ethanol Process. The National Renewable Energy Laboratory (NREL) has led a national effort in the development of processes for conversion of biomass into ethanol. A variety of technical reports are available that summarize research, process economics, and pilot studies on the conversion of hardwood, softwood, and corn stover into ethanol. Figure 1-11 illustrates the NREL process for the conversion of hardwood into ethanol (Wooley, 1999). Corn Stover would follow a similar process (Aden, 2002). In the NREL process white wood is used since technology has not been fully developed for converting biomass that contains bark into ethanol.

The feedstock is washed, screened, and refined to appropriate size. The starting material is then sent to pretreatment where a high temperature dilute sulfuric acid treatment is used to hydrolyze most of the hemicellulose. A cellulase enzyme is produced on site from sugars in the hydrolyzate and is then used to hydrolyze the cellulose and remaining hemicellulose in the feedstock to produce component sugars. Following hydrolysis, the resulting sugars are simultaneously fermented to produce ethanol. The product from fermentation undergoes distillation to produce ethanol at 96%. The ethanol is further upgraded using molecular sieve adsorption to produce approximately 100% ethanol. Byproducts of the NREL process are electricity and gypsum or sodium sulfate, which are sold, CO_2 which is vented to the atmosphere and steam which is used internally in the process as an energy source.

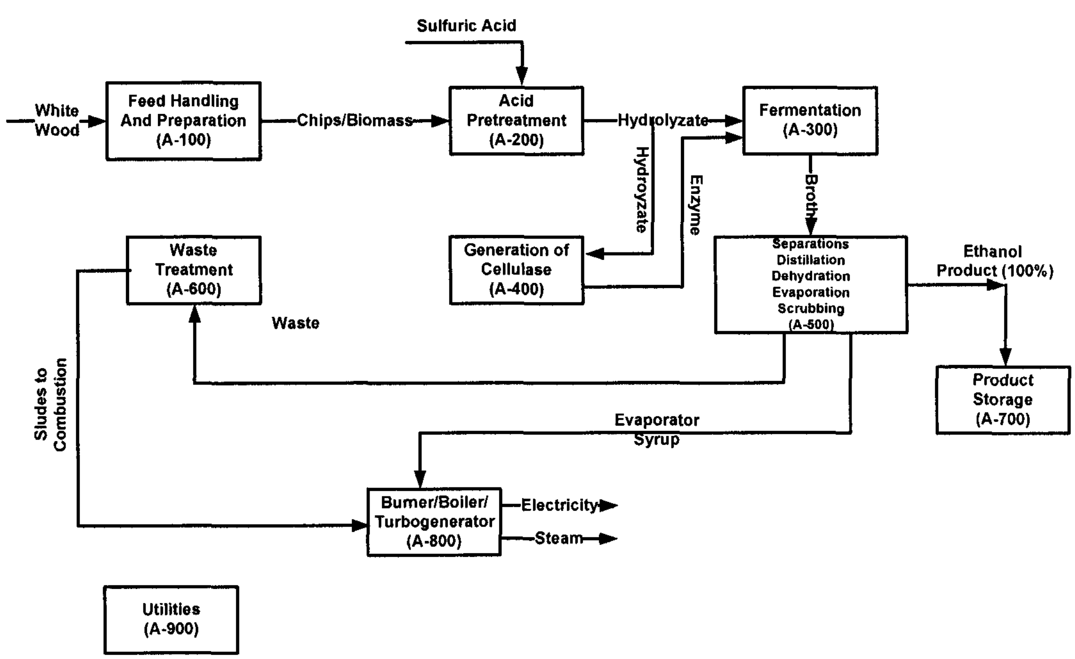


Figure 1-11. NREL Hardwood to Ethanol Process (Wooley, 1999)

The NREL process for softwood is slightly different than the NREL process predicated upon hardwood and corn stover as the feedstock (Merrick & Company, 2004). In the NREL softwood to ethanol process a two-step hydrolysis procedure is used to extract the carbohydrates from the wood chips and then the sugar polymers are converted into component sugars. The two-stage hydrolysis procedure is used because the lignin content is higher in softwoods than in hardwoods and extracting the sugars becomes more difficult.

The softwood chips first enter a prehydrolysis step where they are mixed in an acid impregnator with recycle water and sulfuric acid. The impregnator is run at 20 °C to 50 °C and atmospheric pressure using about 1% acid by weight. A plug screw feeder compresses the chips to approximately 60% water and feeds the chips to the first stage for hydrolysis. The first-stage hydrolyzer is run at pressure and temperature conditions of about 12 atmospheres and 190 °C using direct steam injection. After three minutes the wood is cooled in a flash tank and oligomers are converted to monomers. The wood is then dewatered to 60% by using a screw press. The extracted wood is then sent to a second acid impregnator that impregnates the chips with 1.6% sulfuric acid by weight. A plug screw feeder sends the chips to the second stage hydrolysis reactor which is operated at that is run at 50 °C and 40% solids. The pressure is now about 22.5 atmospheres and steam is injected to reach a temperature of 220 °C. The resulting material goes to a second flash stage. The product is then neutralized using lime, which precipitates gypsum and calcium oxalate. Solids are removed using a rotary drum filter and the result is a liquid feedstock containing five (C5) and six carbon (C6) sugars. Special yeast is used to ferment the five and six carbon sugars to ethanol in a two stage fermentation process. The difference between the NREL hardwood and softwood processes for producing ethanol is that for the hardwood process, some of the hydrolyzate for use in the fermentation step is used to produce cellulase enzymes.

The liquid product following fermentation proceeds to distillation and dehydration and is similar to the hardwood and corn Stover processes and results in ethanol at approximately 100%.

Integrated Forest Products Biorefinery Concept. The concept of an Integrated Forest Product Biorefinery (IFBR) is being advanced by a number of investigators (Wising and Stuart, 2005; Pervait and Sain, 2005; and Maybee and Saddler, 2005). Notable among the advocates for this concept is Adriaan van Heiningen of the University of Maine (2005). The van Heiningen IFBR concept involves using biomass to produce pulp and a number of by-products in an integrated manner. The cellulose contained in the woody biomass would be used to produce bleached pulp (Table 1-5) since producing pulp is more advantageous than producing ethanol (Table 1-6). The results of an input/output analysis comparing the production of pulp to the production of ethanol and diesel fuel are illustrated in Tables 1-5 and 1-6 respectively (van Heiningen, 2006). In an input/output analysis the value added per metric ton of oven dried wood (ODMT) is taken to be the difference between the total value of the products and the total cost of the feed; $75 and $55 per ODMT for hardwood white wood and biomass respectively. An input/output analysis ignores capital and operating charges for the plant.

Table 1-5
Present Value of Hardwood Kraft Mill Products. (van Heiningen, 2006)

Product	Price (U.S.$/ODMT	Yield(%)	Value (U.S.$/ODMT Wood)
Bleached Kraft Pulp	500	45	225
Fuel Value of Black Liq.	55	55	30
Total		100	255
Value Added			255-75[a]=180
(a) Cost of hardwood chips assumed to be $75/ODMT			

Table 1-6
Value of IFBR Producing Ethanol and Transportation Fluid Products
Rather Than Pulp ($55/ODMT). (van Heiningen, 2006)

Product	Price (US$)	Yield(%)	Conversion Yield (%)	Value (U.S.$/ODMT Wood)
Ethanol from Cellulose	$420/MT or $1.25/Gallon	40	47	76
Ethanol from Hemicelluloses	$420/MT or $1.25/Gallon	25	43	45
Diesel Fuel	$630/MT $2.00/Gallon	35	40	88
Total		100		209
Value Added				209-55[a]=154
(a) Cost of hardwood biomass assumed to be				

In the U. Maine process, sodium hydroxide and anthraquinone, a pulping catalyst, would be used as the pulping liquor and sulfur bearing compounds such as sodium sulfide (Na_2S) would be excluded from the process. Ethanol would be produced only from the hemicellulose portion of wood (Table 1-7) which has a low heating value and is of little value in black liquor as a fuel. Hemicellulose polymers such as glucuronoxylan would be extracted from chips using dilute alkali and converted to component sugars. The extracted wood would then be processed into pulp and the residual lignin processed into diesel fuel (Table 1-7) or electrical energy. The chips could originate from the pulp mill or from wood used in an oriented strand board plant.

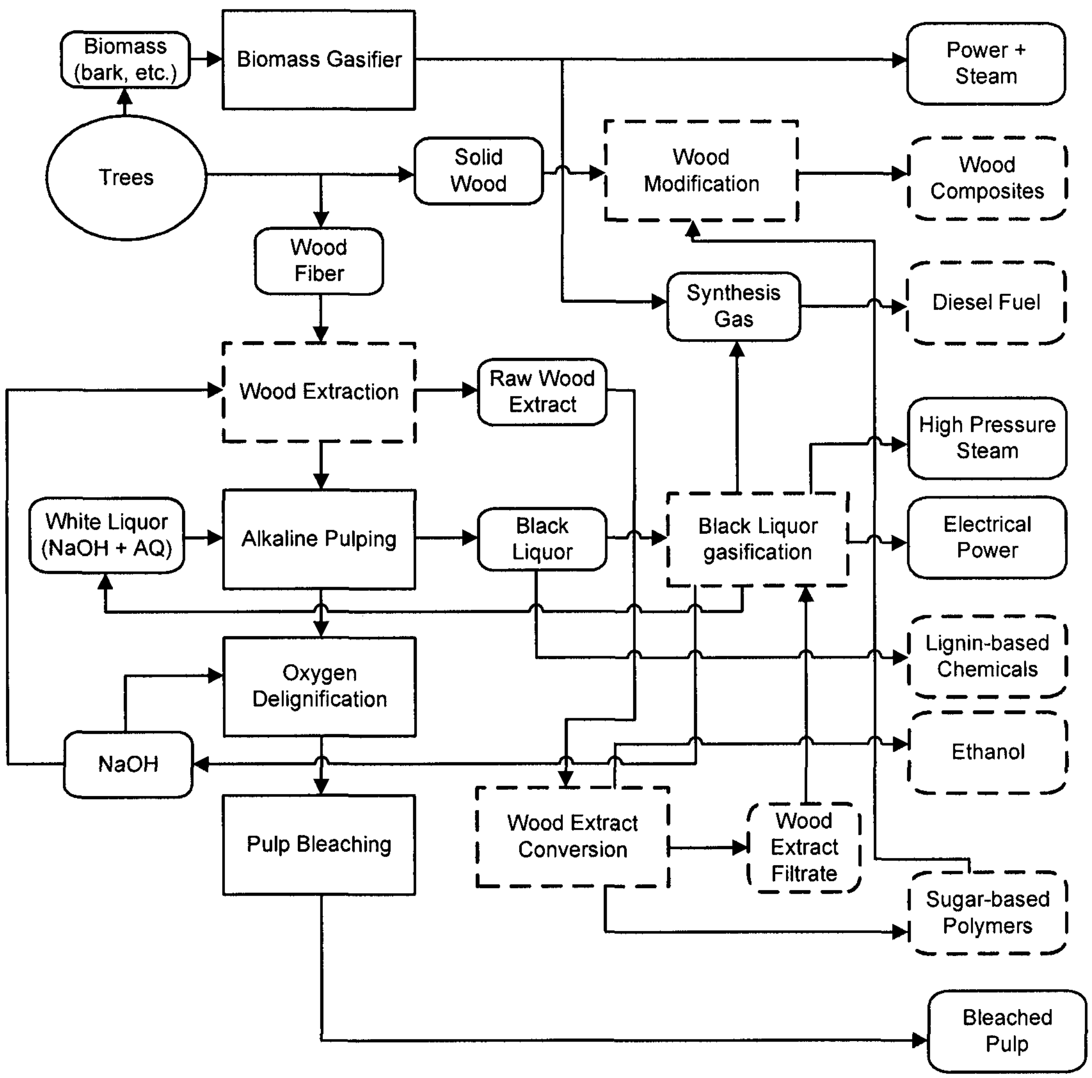


Figure 1-12. University of Maine IFBR Concept Based on Alkaline Pulping

Table 1-7
Value of IFBR Producing Ethanol and Transportation Fluid Products Rather Than Pulp ($75/ODMT). (van Heiningen, 2006)

Product	**Price (US$)**	**Yield(%)**	**Conversion Yield (%)**	**Value (U.S.$/ODMT Wood)**
Alkaline Pulp	$500/ODMT	47	100	235
Ethanol from Hemicelluloses	$420/MT or $1.25/Gallon	10	43	18
Diesel Fuel	$630/MT $2.00/Gallon	43	40	108
Total		100		361
Value Added				361-75[a]=286
(a) Cost of hardwood biomass assumed to be $75/ODMT				

Another variant of the IFBR is the replacement of a conventional Kraft recovery boiler with a black liquor gasification system. This would increase the thermal efficiency of energy conversion and improve the production of steam and electrical power. Sugar based polymers such as itaconic and other dibasic acids (see Figure 1-2) could also be produced from the dissolved hemicelluloses and lignin could be dissolved and partially converted to polyurethane foams (Table 1-8). This later scenario was shown to be the most profitable in the economic analysis (van Heiningen, 2006).

Table 1-8
Value of IFPR Producing Higher Valued Products and Diesel Fuel (van Heiningen, 2006)

Product	Price (US$)	Yield(%)	Conversion Yield (%)	Value (U.S.$/ODMT Wood)
Alkaline Pulp	$500/ODMT	45	100	225
Poly Itaconic Acid	$3000/MT	10	50	150
Polyurethane	$2000.MT	10	50	150
Diesel Fuel	$630/MT $2.00/Gallon	35	40	88
Total		100		613
Value Added				613-75[(a)]=538
(a) Cost of hardwood biomass assumed to be $75/ODMT				

The Biofine Process. The Biofine process (Figure 1-13) utilizes a high temperature, dilute acid catalyzed hydrolysis to convert lignocellulosic biomass into levulinic acid, formic acid, furfural, and a carbon rich char powder (Fitzpatrick, 2004). The biomass feedstock is sent through a hammer mill to reduce the size and mixed with recycled acid. The material is then hydrolyzed using dilute sulfuric acid at about 200 °C. In the hydrolysis step cellulose is converted to levulinic acid, formic acid and char. The hemicellulsoses in the feedstock are converted to furfural while the lignin in the raw material would be converted to char.

Cellulose + H_2SO_4 (2%) → Levulinic Acid + Formic Acid + Char (Carbon)

Hemicelluloses + H_2SO_4 (2%) at 220 C → Furfural

Lignin + H_2SO_4 (2%) at 220 C → Char and Condensed Lignin

Ligneous char is separated from the mixture and burned to produce steam or electrical power. Formic acid and furfural are also recovered at this point. The remaining mixture is then concentrated and the acid is separated out to be recycled. The levulinic acid stream is then purified. The levulinic acid can then be converted to a variety of useful chemicals including methyltetrahydrofuran (MTHF) and ethyl levulinate. Also furfural can be converted to levulinic acid to improve the process yield.

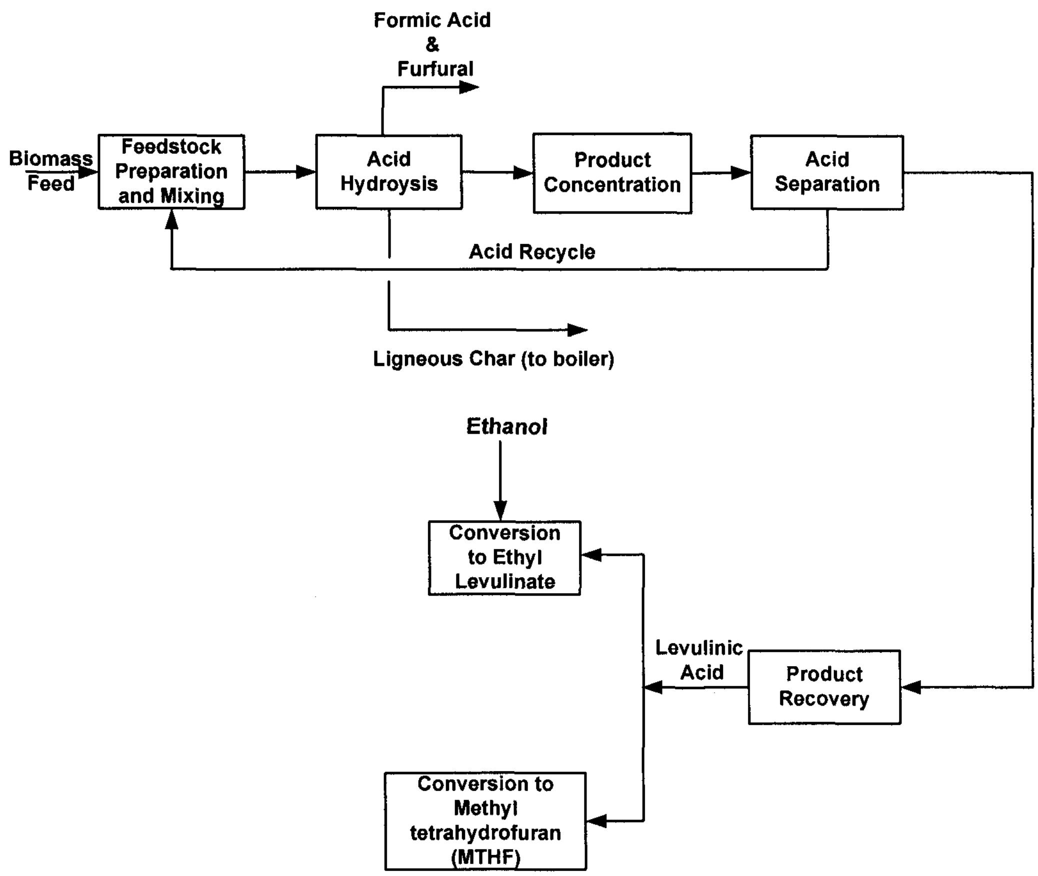


Figure 1-13. The Biofine Process for the Production of Levulinic Acid

The Biofine process can produce a wide range of useful chemicals. (Fitzpatrick, 2004). Formic acid and furfural are both commodity chemicals. Formic acid is used in the production of rubber, pharmaceuticals, textiles, and catalysts including nickel and aluminum. Furfural is used in the production of furan resins, lubricating oils and textiles for clothing. Levulinic acid can be used as a starting material for fuel substitutes, monomers, pesticides and many commodity chemicals. MTHF is very useful as a gasoline additive due to its anti-knock properties, energy density and low volatility. It is also used as a co-solvent for ethanol for fuels that blend ethanol and gasoline. Ethyl levulinate, which can be produced from levulinic acid using ethanol, has shown potential as a possible diesel additive to lower emissions and improve lubricity. A 3,000 ton per year demonstration plant that produces levulinic acid using thermal processing is due to start up in Italy ("Biorefinery gets ready", 2006).

Fast Pyrolysis. Fast pyrolysis is another method of producing fuel from biomass. In this process, biomass is heated quickly to 450-600 °C in the absence of air. The high temperatures and heavy vibrations cause the material to decompose and break down in random positions (Figure 1-14). The result is the production of organic vapors, pyrolysis gases and char. The primary reaction products are phenol, methane, water vapor, carbon monoxide, carbon dioxide, levo-glucosan, hydroxyl acetaldehyde, hydrogen and aromatic carbon or char. The pyrolytic vapors are condensed to pyrolysis oil referred to as bio-oil. The phenol, levoglucosan, and hydroxyl acetaldehyde constitute the pyrolysis oil. The methane, carbon monoxide, carbon dioxide, and hydrogen constitute the pyrolysis gas. About 70-75% of the biomass is converted to pyrolysis oil, which is a clean liquid with many possible functions (BTG World, 2006). The oil can be used as a feed for a petroleum refinery, as an intermediate for many applications, or directly as a fuel (Kelly, 2006).

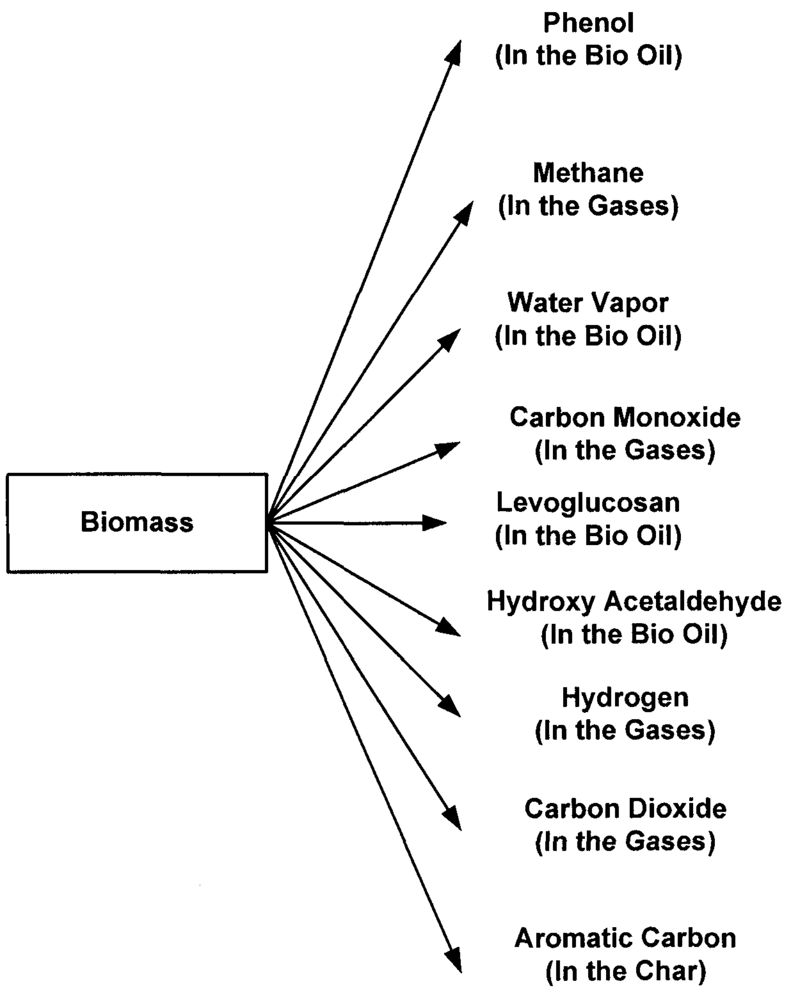


Figure 1-14. Pyrolysis Products

In the BTG fast pyrolysis process (Figure 1-15), biomass is reduced in size and fed from a lock hopper into a fluidized pyrolysis reactor together with hot sand being recycled from a char bed combustor. The function of the sand is to provide a heat transfer media in the fluidized bed. In the pyrolysis reactor, the biomass is rapidly heated to a temperature between 450 and 600 °C. Due to the heavy vibrational energy in the biomass, caused by the rapid heating, the atoms vibrate apart at random positions to give the spectrum of products shown in Figure 1-14. Three primary products are produced when the biomass decomposes at elevated temperature; namely, a pyrolysis gas or low molecular weight molecules (H_2, CH_4, CO_2 and CO_2) that are above their critical temperatures, vapors of phenol, water, levoglucosan, and hydroxyl acetaldehyde that are below their critical temperatures that can be condensed to form a bio-oil, and char or aromatic carbon that is deposited on the sand. The pyrolysis vapors go to a direct contact condenser where the bio-oil is condensed and taken off as the product for further purification. Heat is rejected to the environment by using a cooling tower in conjunction with a water cooled tube and shell heat exchange. Since char builds up on the sand in the pyrolysis reactor, a portion of the bed is continuously removed and sent to a char combustion reactor, which is a second fluid bed. Inside the char combustion reactor, primary and secondary air is used to burn the char from the sand and heats the sand particles. The hot sand is then returned to the fluidized bed pyrolysis reactor where it supplies the heat of decomposition for the biomass. Ash is eluted from the reactor with the combustion gases and passes thru a series of cyclones where the ash is separated from the flue gas. The hot flue gases are either sent to the stack or to an economizer for removing additional heat energy from the flue gases. Vapors off the direct contact condenser can be further treated or sent to a waste heat boiler.

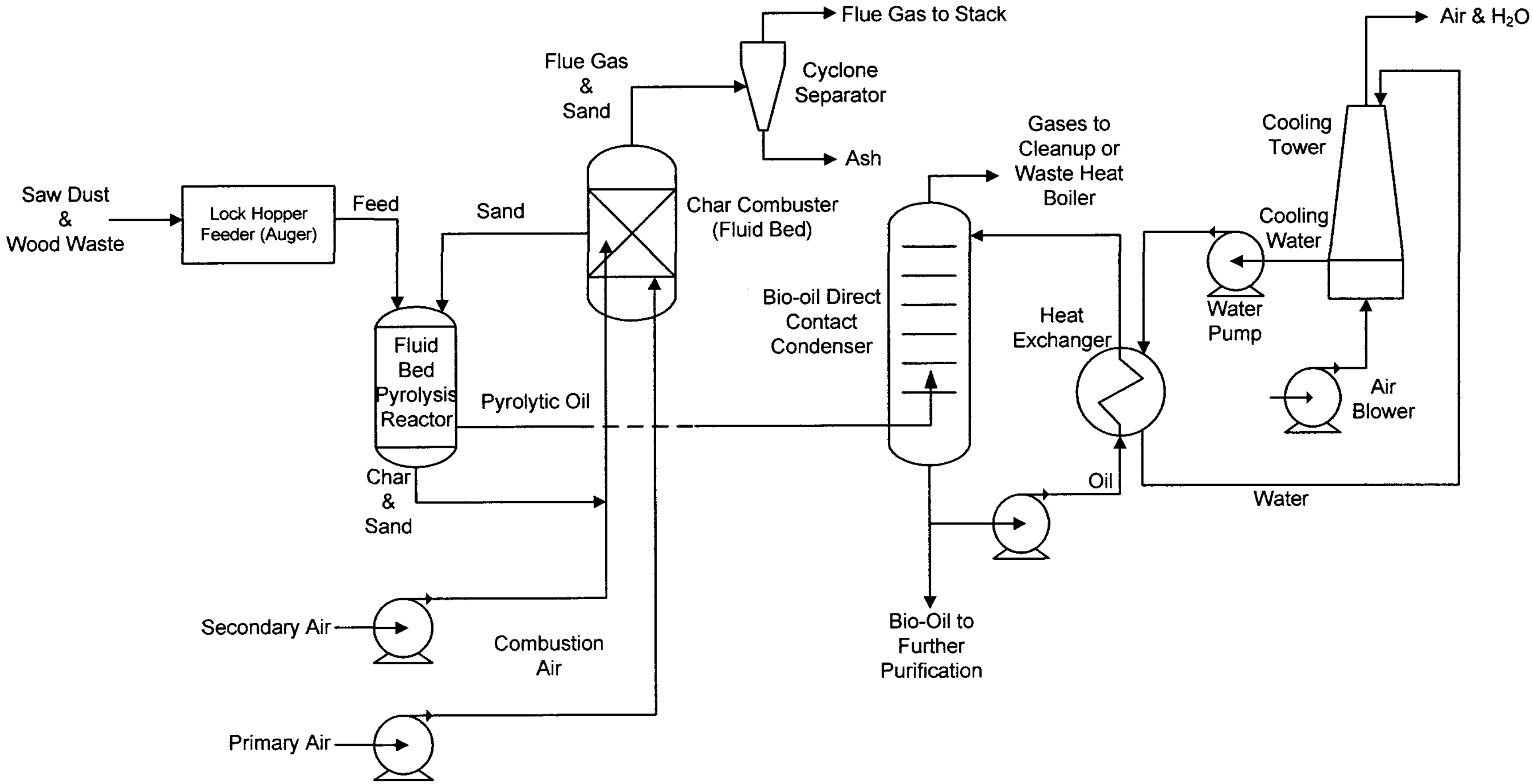


Figure 1-15. BTG Fast Pyrolysis Process

Thermal Depolymerization Process. Changing World Technologies Inc. has developed a thermal depolymerization process (TDP) that can convert a wide variety of waste materials into useful chemicals (Figure 1-16). It is a variant of the pyrolysis technology discussed in Figure 1-15. This process is reported to be able to handle turkey waste, tires, plastic bottles, garbage, paper mill effluent, medical waste, oil refinery residues, and many other forms of waste. Products are high-quality oil, clean-burning gas, and purified minerals that can be used in a variety of ways (Lemley, 2003).

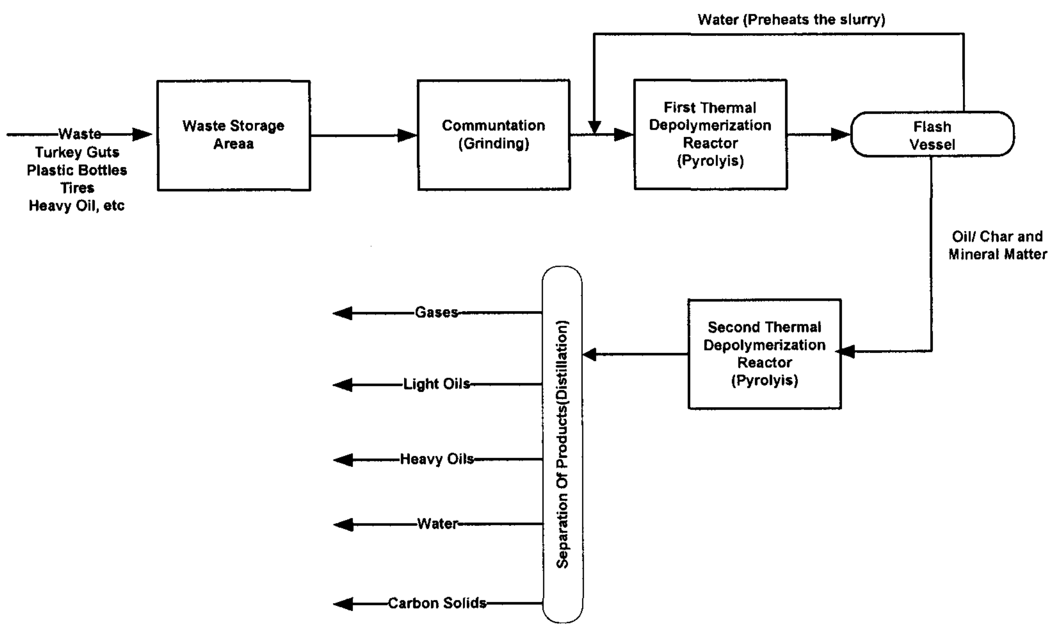


Figure 1-16. Changing World Technologies' Thermal Depolymerization Process

The first step in the process (Figure 1-16) is to grind the waste material and mix it with water to form slurry. The slurry is then heated to 260 °C at a pressure of 40 bar in the first stage TDP reactor causing the long molecular chains to partially break apart. A flash vessel is used to remove water by rapidly dropping the pressure.

The resulting hot water is used to preheat the stream to the first depolymerization reactor. The second stage reactor is run at 482 °C and breaks the molecular chains further. Vertical distillation columns are then used to separate the mixture into gases, light oils, heavy oils, water, and carbon solids respectively from top to bottom as shown in Figure 1-16. The gases are burned and the steam generated is used in the process. The oils, minerals and carbons are all sold.

Changing World Technologies developed their technology in a pilot plant and now have an industrial sized plant in Missouri. The plant is located adjacent to a Butterball Turkey plant operated by ConAgra Foods. The plant will produce 10 tons of gas, 11 tons of minerals, and 600 barrels of oil daily (Lemley, 2003). The gas is used to help heat the plant and the minerals can be used as fuels, fertilizers, or specialty chemicals. The oil is said to be almost identical to No. 2 fuel oil.

Chapter 2

ETHANOL PRODUCTION ROUTES

Ethanol Introduction

The United States ethanol industry is currently considered to be the fastest growing energy industry in the world. There are 101 corn to ethanol plants in operation with a total annual production capacity of 4.8 billion gallons. There are an additional 44 plants currently under construction and 7 plants being expanded. When these projects are completed the U.S. ethanol production will increase by 2.9 billion gallons to 7.7 billion gallons per year (RFA, 2006). A major reason for this growth in capacity is that ethanol is currently being used to replace the oxygenate methyl tertiary butyl ether (MTBE) in gasoline ("Chemical commodities", 2006). As a fuel additive, ethanol can be used to lower the emissions of unburned hydrocarbons and increase the octane number of the fuel. It is also possible to use the ethanol as more than just an additive. Some automobile companies already sell vehicles that can run on 85% ethanol and 15% gasoline. This mixture is known as E85 and is considered an alternative fuel (Fialka, 2005). High gasoline prices and ever-increasing environmental awareness will ensure that production figures will continue to rise. According to the *Wall Street Journal* (June 15, 2005), there have been discussions of a federal mandate that would effectively double the use of ethanol as a fuel additive within the next seven years (Fialka, 2005). Aside from its obvious value as a fuel, ethanol has many other uses. It is used as a solvent in the production of perfumes, pharmaceuticals, detergents, inks, and coatings. It is also found in many beverages.

Production Routes

The first commercial route for the production of ethanol, in the United States was initiated by Shell Oil Ltd. in 1947 (Karas & Piel, 1994). The Shell process involved the synthesis of ethanol from ethylene by a technology known as direct hydration. Direct hydration plants are no longer operating in the United States because the technology is not profitable due to the high cost of ethylene. Ethanol can also be obtained by the fermentation of simple sugars. Any material that contains sugar or compounds that can be readily converted to sugar can be used. Starting raw materials for fermentation processes include sugars, starches, and cellulose. Sugars can come from sugar cane, sugar beets, molasses, or fruit. Starches are derived from corn, grains, potatoes, or root crops. Cellulose comes from wood, agricultural residues such as corn stover and waste liquors. The commercial production of ethanol in the U.S. is currently being carried out almost exclusively by fermentation processes using corn (RFA, 2006). Fermentation from corn can be done by two methods known as the dry milling and the wet milling processes. Several organizations are developing technology to produce ethanol from biomass, but none of the processes under development have been commercialized.

Direct Hydration of Ethylene

The key to direct hydration of ethylene is to contact ethylene gas with a catalyst to produce ethanol. A basic flow diagram of the process can be seen in Figure 2-1. A combination of ethylene gas and process water is heated to about 265 °C and fed to a fixed-bed catalytic reactor where it is contacted with a phosphoric acid and hydrochloric acid based catalyst. The reactor is operated at about 70 atm. The product from the reactor must then be cooled and separated into liquid and vapor components. The liquid goes straight to ethanol refining steps, while the vapor is scrubbed with water. The scrubbing process removes ethanol and the leftover vapor from the scrubber is recycled and enriched with fresh ethylene. If the liquid streams contain aldehyde impurities due to side reactions, then they must be hydrogenated catalytically before distillation. This is done to make sure the ethanol is not contaminated. The two liquid products then undergo distillation to remove other impurities and to concentrate the solution. The result is a 95% ethanol 5% water azeotrope. This product is sent to a dehydration system where molecular sieves are used to dehydrate the azeotrope into anhydrous ethanol (Karas & Piel, 1994).

Basic Chemistry. The main chemical reaction in the direct hydration of ethylene is the addition of water to ethylene to form ethanol as seen in equation 2-1 (Karas & Piel, 1994).

$$C_2H_4 + H_20 \Leftrightarrow CH_3CH_2OH + Energy \tag{2-1}$$

However, there are several side reactions that must be avoided or the ethanol will become contaminated. Diethyl ether can be formed directly from ethanol

$$2CH_3CH_2OH \Leftrightarrow (CH_3CH_2)_2O + H_2O \tag{2-2}$$

and acetaldehyde can be formed from trace amounts of acetylene in the ethylene stream.

$$C_2H_2 \xrightarrow{H_2O} CH_3CHO \tag{2-3}$$

Formation of more complex aldehydes such as crotonaldeyde is also possible.

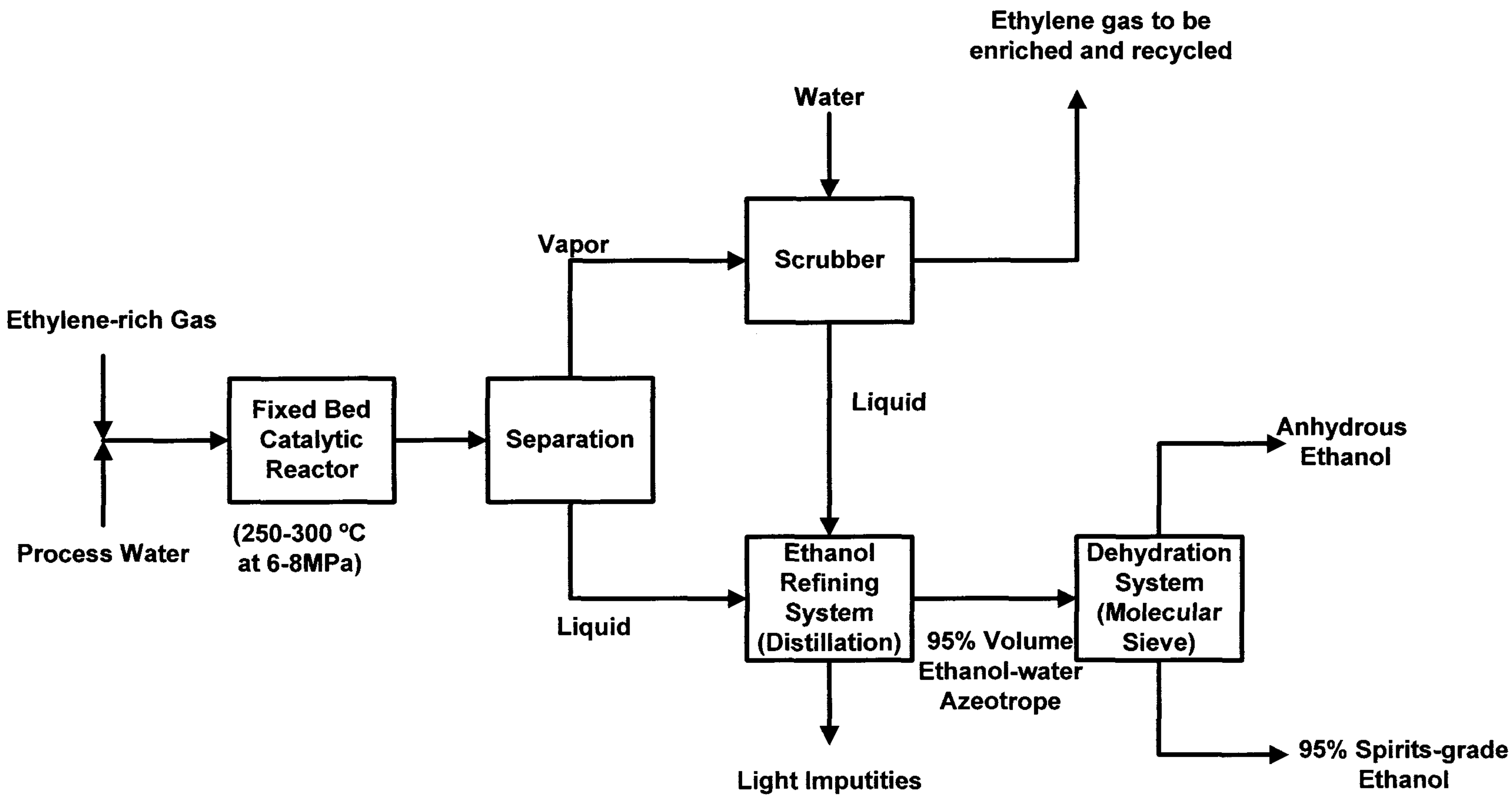


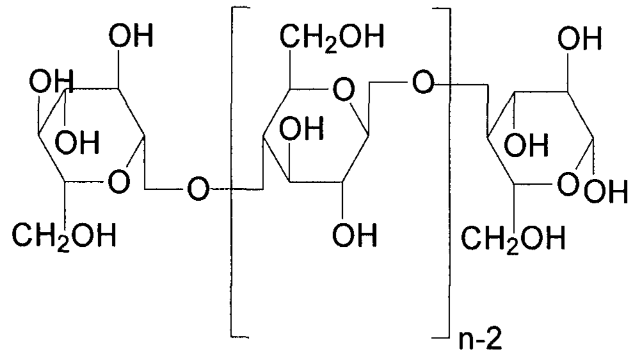
Figure 2-1. Direct Hydration of Ethylene to Ethanol

Fermentation to Ethanol

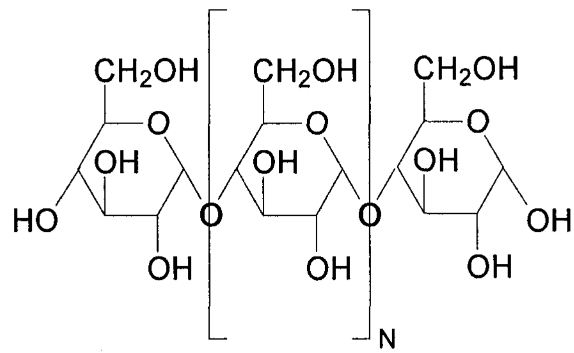
The majority of commercial ethanol plants in the United States are based upon the fermentation of either simple sugars or complex carbohydrates. Originally ethanol was produced by the fermentation of molasses. Currently in the U.S. ethanol is produced principally from corn by using both the dry milling and wet milling processes. Brazil is another major producer of ethanol and they use crops such as sugar cane and sugar beets. Several other fermentation processes are under development to utilize various forms of biomass to produce ethanol.

Corn to Ethanol

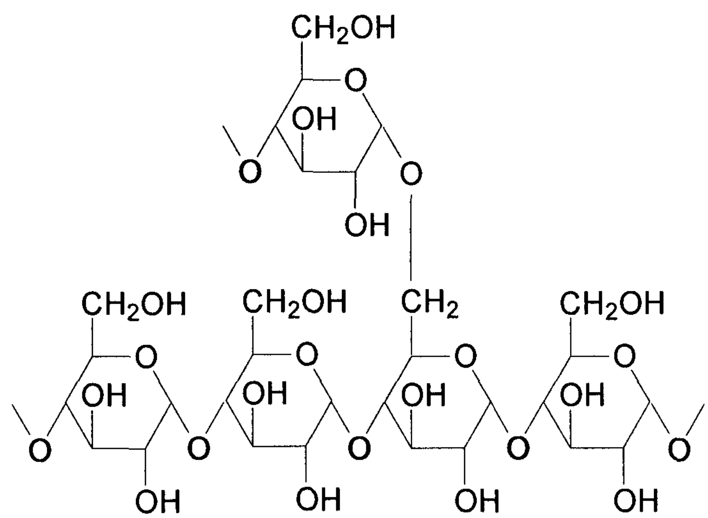
In 2005, 1.43 billion bushels of corn were used for the purpose of producing commercial ethanol (RFA, 2006). Corn is composed primarily of starch and is made up of two natural carbohydrate polymers; amylase, a linear polymer and amylopectine, a highly branched carbohydrate. Both the amylase and amylopectine can be converted to their principal sugar, α-glucopyranose. The glucose units in starch differ from the glucose units found in cellulose. Cellulose is connected by β-1,4 bonds which give a strong structure and is found in wood (Figure 2-2A). Starch on the other hand, is linked through α 1-4 bonds which give the starch molecules considerably greater flexibility. Amylose (Figure 2-2B) and the highly branched amylase pectin (Figure 2-2C) comprise the primary carbohydrates in starch.



A. Cellulose Chain



B. Amylose



C. Amylopectine

Figure 2-2. The Structures of Cellulose and Starch (Eklund and Lindstdrom, 1991)

Dry Milling. The main process being used currently is the dry milling process (Figure 2-3). In dry milling plants, the entire corn kernel is ground into what is know as "meal." The meal is then added to a liquefaction tank, which is kept at about 88 °C. Here the corn is mixed with water and alpha-amylase enzyme. The alpha-amylase converts starch polymers into maltose and higher oligomers. Caustic and lime are added as a calcium source for the enzyme and to maintain a pH of 6. Urea is also added to the tank as a source of nitrogen for future fermentation steps. The resulting slurry know as "mash" must then be brought to high temperatures (110 °C) to control bacteria before saccharification can occur (McAloon, 2000).

Saccharification is the process of breaking down complex carbohydrates in the starch (and cellulose) into simple sugars. The following reaction is a typical enzymatic hydrolysis of starch to sugar.

$$(C_6H_{10}O_5)_n + H_2O \xrightarrow{Enzymes} nC_6H_{12}O_6 \qquad (2\text{-}4)$$

Saccharification takes place in a continuous stirred tank at 60 °C. Gluco-amylase enzyme is added and is responsible for "splitting off" glucose from the maltose and higher oligomers. Sulfuric acid is used to maintain a pH of about 4.4. After 6 hours in the saccharification tank the slurry is cooled and sent to fermentors that operate at about 32-34 °C (McAloon, 2000). The addition of yeast converts the simple sugars to ethanol and carbon dioxide. The following is the basic chemical reaction for this process.

$$C_6H_{12}O_6 \xrightarrow{Yeast} 2\,CH_3CH_2OH \; + \; 2\,CO_2 \qquad (2\text{-}5)$$

(180 grams/mole) = 2x (46 grams/mole) + 2x(44 gram/mole)

The fermentation step has a residence time of about 46 hours and the product (about 9 % ethanol by weight) is known as the "beer." Theoretically, the yield of ethanol is 51.1% of the weight of the sugar fermented; that is 92 grams of ethanol can be produced from 180 grams of sugar (see equation 2-5). The gases from fermentation are scrubbed and carbon dioxide is vented into the air (McAloon, 2000). The beer is sent to distillation columns to separate the ethanol and the remaining product is referred to as "stillage", which can be treated and processed into a feed for livestock. The ethanol stream at this point is about 95% ethanol and must be dehydrated using a molecular sieve to reach virtually 100% ethanol. If the purified product is to be used for fuel purposes, then it must be mixed with about 5% gasoline to make it undrinkable. This is done to avoid beverage alcohol taxes (RFA, 2005).

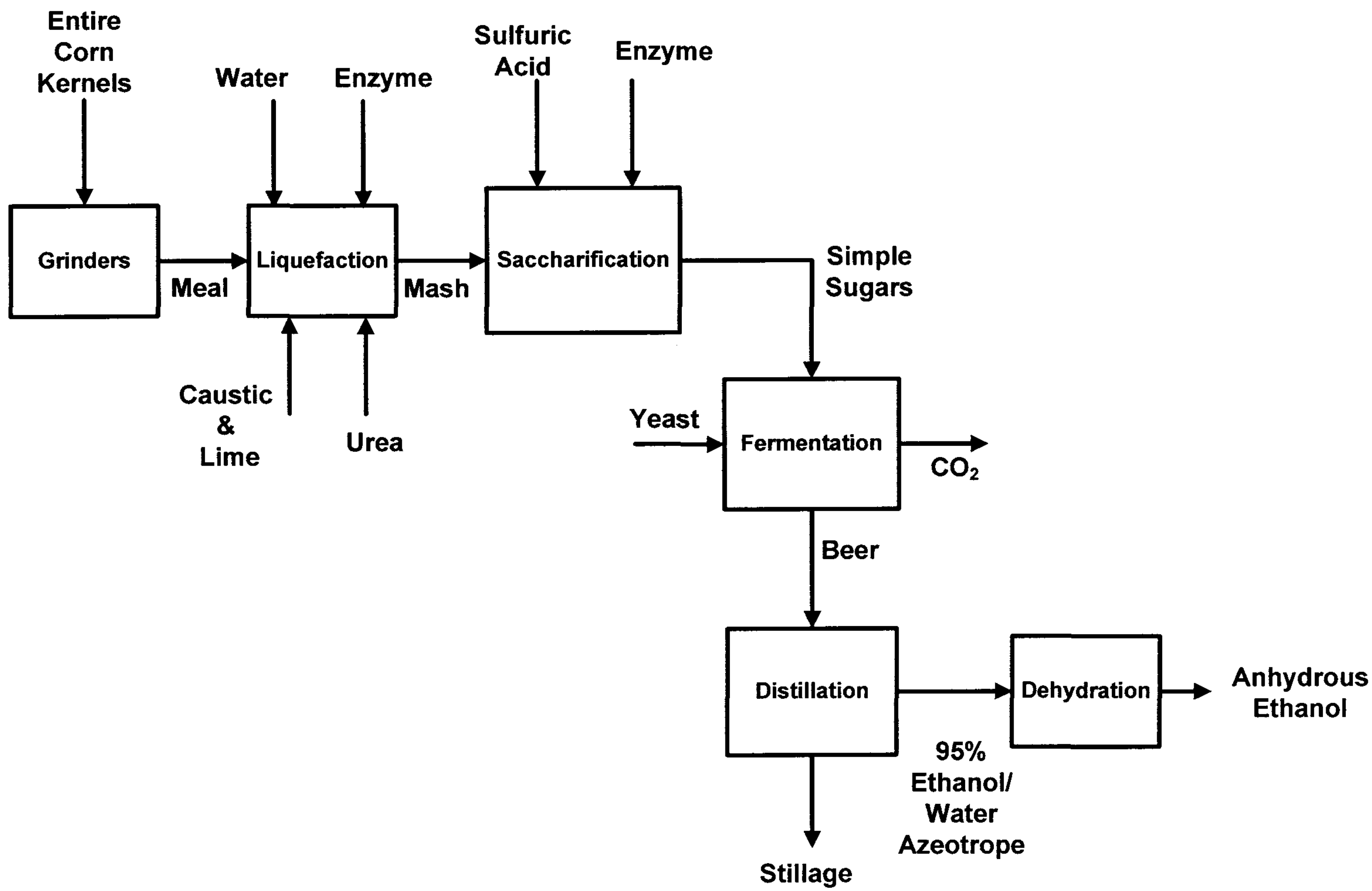


Figure 2-3. Dry Milling Process from Corn

Wet Milling. The second and less common process for fermenting corn to ethanol is called wet milling (Figure 2-4). The wet milling process consists of seven main steps: inspection and cleaning, steeping, germ separation, grinding and screening, starch separation, syrup conversion, and fermentation. When the corn arrives at a plant the first step is to clean it to remove cob, dust, and any other unwanted particles. The corn kernels are then steeped or soaked in a dilute sulfurous acid solution consisting of water and sulfur dioxide. In the steeping process about 3,000 bushels of corn are soaked at 50 °C in dilute acid for approximately 30 to 40 hours. About 0.1 percent sulfur dioxide is added to the water in order to prevent the growth of bacteria. Steeping increases the moisture of the kernels to around 45 percent. The swelling due to moisture and the acidity loosen bonds in the corn, which releases starch. After the steeping or soaking process is complete, the corn is ground in order to separate the germ from the other components and the steepwater is processed to remove nutrients for animal feed or for fermentation. The ground corn slurry is then sent to the germ separation process. Due to the low density of corn germ, it can be separated from the slurry using cyclone separators. The separated germ can be further processed into corn oil (Corn Refiners Association, 2006). The remaining corn slurry continues on to a grinding and screening stage. Here, the slurry is ground in order to separate starch and gluten from the fiber. Screens are used to catch the fiber, while allowing the gluten and starch to pass. The starch and gluten mixture goes through a starch separation process. The gluten is less dense than starch and can be removed using a centrifuge. The gluten can be used as an animal feed.

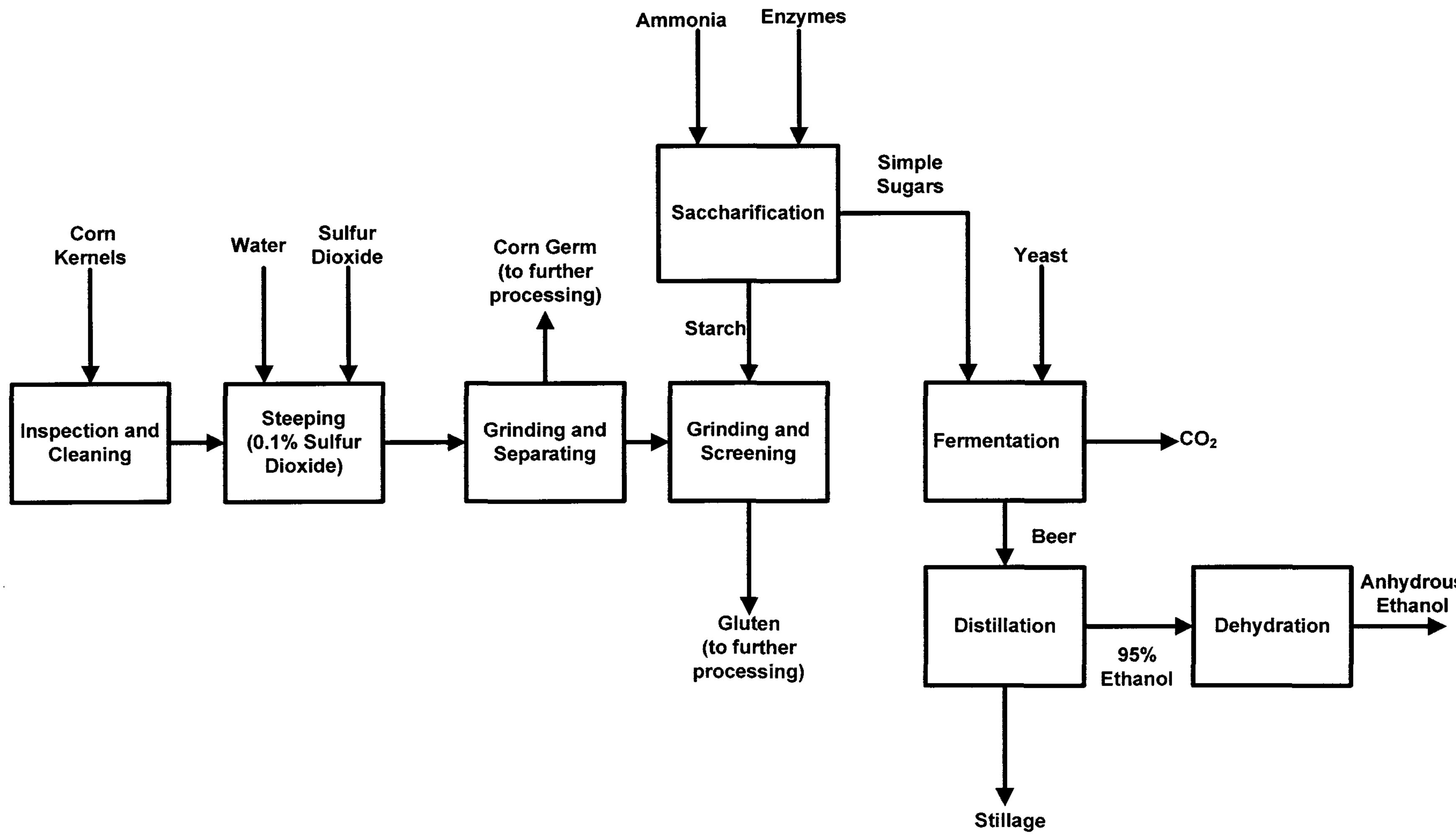


Figure 2-4. Wet Milling Process from Corn

The starch is washed, diluted, and run through hydroclones to make sure it is very pure. From this point the process is essentially the same as the dry milling process. Starch is converted to simple sugars using enzymes and the simple sugars are fermented with yeast to produce ethanol. If ethanol is not the desired product then the starch could also be alternatively processed into corn syrup (Corn Refiners Association, 2006).

Biomass to Ethanol

Companies have been developing technology to produce ethanol from biomass for quite some time now. Feedstocks include waste wood, corn stover, waste liquor, and many other lignocellulosic materials. The wood to ethanol processes currently are the most common. McCloy and O'Connor (1999) summarize five major technologies for the conversion of lignocellulosic materials into ethanol. Iogen Corporation converts agricultural and hardwood waste using an enzymatic process. BC International uses a fermentation process with a genetically modified organism that focuses on sugars common to wood waste. Arkenol Inc. utilizes a patented acid hydrolysis process on wood waste and agricultural waste. Laszlo Paszner (UBC Faculty of Forestry) has developed an organic solvent process and lastly, Bioengineering Resources Inc. employs a gasification-fermentation process (McCloy & O'Connor, 1999). Aside from these five processes, the most advanced process for the conversion of lignocellulosic biomass into ethanol has been developed by NREL. They have developed technical information applicable to the conversion of corn stover (Aden, 2002) as well as both hardwood (Wooley, 1999) and softwood (Merrick & Company, 2004). The current study focused on the conversion of hardwood to ethanol using the NREL process

<u>NREL Process Using Hardwood Chips.</u> The NREL process is the most advanced woody biomass to ethanol process currently available. The National Renewable Energy Laboratory starts their biomass to ethanol process with purchased hardwood chips (Wooley, 1999). This is a very important consideration since the price of chips is considerably higher than the price of biomass. The NREL white wood to ethanol plant consists (Figure 2-5) of nine sections: feed handling and preparation (A100), acid pretreatment (A200), fermentation (A300), cellulase generation (A400), separations/distillation (A500), waste treatment (A600), storage (A700), burner/boiler/turbo generator (A800), and utilities (A900). Figures A-1 through A-9 in Appendix A summarize the unit processes comprising the NREL white wood to ethanol plant (Wooley, 1999).

Feed Handling (A100). Referring to Figure 2-5 or for more detail Figure A-1, in the feed handling section of the plant (A100) the chips are weighed, washed, sorted and screened to remove debris and then sent to the pretreatment process (Section 200) for hydrolysis. Oversized chips are reduced in size to minimize wood losses (Wooley, 1999).

Prehydroysis and Detoxification (A200). The purpose of pretreatment of the biomass (Figure A-2) is to convert the biomass into the best possible form for simultaneous saccharification and fermentation to take place. The method used by NREL is based on high temperature (190°C) acid hydrolysis reactions using dilute sulfuric acid. This process causes a number of desirable changes in the feed composition. The hemicellulose in the biomass feed is converted to fermentable sugars including xylose, mannose, arabinose, and galactose. Also, a small percentage of the cellulose from the feedstock is converted to glucose. The high temperature acid hydrolysis also causes some of the lignin to solubilize and "expose" the cellulose for the enzymatic hydrolysis step that follows (Wooley, 1999). In the pretreatment section of the plant, the residual sulfuric acid must be neutralized using lime and thus gypsum ($CaSO_4$) is produced as a byproduct.

The pretreatment also yields some unwanted products that must be removed before simultaneous hydrolysis and fermentation can take place. Beside gypsum, some of these products include acetic acid, pentose sugars, furfural, hydroxymethyl furfural (HMF) and hexose sugars. Flash cooling (blowdown) is used to vaporize and remove some of the acetic acid and a significant amount of water, furfural, and HMF. Acetic acid is also removed from the liquid using continuous ion exchange. Before being sent to fermentation tanks, the slurry is overlimed, neutralized, and mixed with cellulose and water. Nitrogen is added as a nutrient in the form of ammonia. The fully pretreated feedstock is referred to as the detoxified hydrolyzate which goes to fermentation (Wooley, 1999).

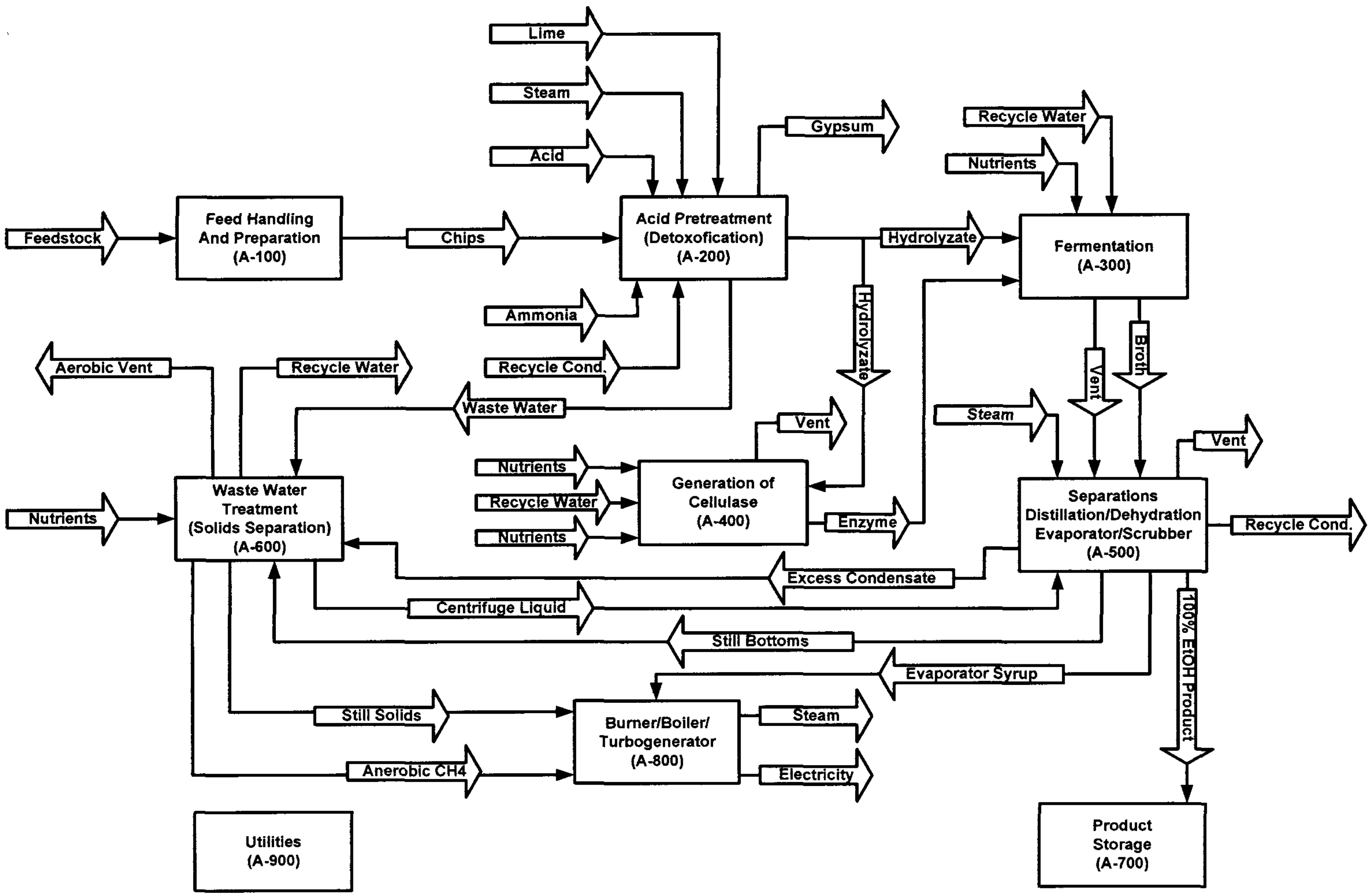


Figure 2-5. NREL Hardwood to Ethanol Process (Wooley, 1999)

Cellulase Production (A400). Cellulase enzymes are produced on site in the A400 section of the plant as shown in Figure A-4 in Appendix A. The main purpose of the cellulase enzyme is to perform the enzymatic hydrolysis of cellulose to form glucose. Cellulase is technically a compilation of three different enzymes: endoglucanases, exoglucanases, and β-glucosidase. Each of these three enzymes performs its own unique function with regards to cellulose. Endoglucanases attack cellulose fibers resulting in a quick reduction in polymer size and causing cellulose to hydrolyze to glucose and cellobiose. Exoglucanases can hydrolyze crystalline cellulose due to its propensity to attack the ends of cellulose fibers. β-glucosidase performs the important task of hydrolyzing cellobiose to glucose. The enzymes that comprise cellulase are naturally produced by numerous bacteria and fungi. NREL uses *Trichoderma reesei*, the most commonly used organism for industrial production of cellulase. Aerobic bioreactors are used to grow *Trichoderma reesei* for the production of the enzyme (Wooley, 1999).

Saccharification and Fermentation (A300). The next step in the process is known as simultaneous saccharification and fermentation (SSF). As the title clearly suggests, this step involves two operations. Saccharification is the process of breaking down (hydrolyzing) cellulose into simple sugars, xylose and glucose. The cellulase enzyme is responsible for this hydrolysis. The fermentation portion of the process is done using bacteria, known as the ethanologen, which converts xylose and glucose into ethanol. NREL uses *Zymomonas mobilis* as their ethanologen. There are several other possible yeasts and bacteria that could also be used such as *Saccharomyces cerevisiae* and *Pichia stipitis.* A step by step process is used to get a seed inoculum for the main fermentation vessels. *Zymomonas mobilis* is initially grown in a very small vessel along with nutrients and the pretreated biomass (hydrolyzate). The vessel size is then increased a series of times until there is enough ethanologen for use in the main fermentation tanks. At this point the ethanologen, diluted hydrolyzate, cellulase enzyme, and nutrients are added in a continuous fashion to the main fermentor tanks. Saccharification and fermentation simultaneously take place in these tanks resulting in the production of ethanol. The product from the tanks is then pumped to a storage tank (Wooley, 1999).

Distillation and Dehydration (A500). The product in the storage tank goes to a distillation process (Figure A-5) for purification. Distillation is used to remove dissolved carbon dioxide and the majority of the water. The bottoms product of distillation contains the insoluble solids and dissolved solids that have not been converted to ethanol. The insoluble solids are separated using a centrifuge and are burned in a fluidized bed burner/boiler system in the A800 section of the plant. The remaining liquid with dissolved solids is concentrated into syrup by evaporation and is also burned in the fluidized bed combustor. The product of distillation is a 95% ethanol and water azeotrope. The water is removed using molecular sieve adsorption and the result is a purified finished product that is almost 100% pure ethanol (Wooley, 1999).

Solids Separation and Waste Water Treatment (A600). Wastewater is always an issue with any industrial process. NREL first treats the water by anaerobic digestion, where 90% of the organic material in the water is converted to methane and carbon dioxide gas. These gases are burned for their moderate fuel value. The next step in water treatment is an aerobic digestion lagoon. Once again 90% of the organics are taken out and the sludge generated is dewatered and burned in the fluidized bed combustor. From here the water is clean enough to be recycled in some capacity back to the production process (Wooley, 1999).

Boiler, Burner and Turbogenerator (A800). NREL utilizes a fluidized bed combustor to burn by-product streams for the production of process steam and electricity. Lignin and unconverted cellulose and hemicellulose from the wood are burned in a combustion chamber along with sludge, biogas, and evaporator syrup. Boiler feed water flows through a heat exchanger in the combustor, is evaporated, and superheated steam is produced. A turbine and generator are used to produce electrical power and steam is also extracted from the turbine for use as process steam (Wooley, 1999). For more details refer to Figure A-8 in Appendix A.

Storage (A700) and Utilites (A900). 100% ethanol produced in the A500 section of the plant is stored and diluted with gasoline to produce the final denatured ethanol product, which is also stored. In addition, all of the raw materials used in the process are stored in the A700 section of the plant and distributed for use. In the utilities section (A900) river water is treated to produce fresh process and cooling water. Plant and instrument air are produced in the utilities section (A900) as well as chilled water. Heat from condensers and heat exchangers is rejected to the atmosphere in a cooling tower. Cool, fresh water is then sent back to the process for use in heat exchangers and condensers (Wooley, 1999).

Chapter 3

OBJECTIVE AND SCOPE

The objective of the work reported here was to develop a model for estimating the economics of converting hardwood biomass to ethanol. It is desired to use the model in preliminary process design and in estimating the feasibility of producing ethanol from white wood. The model developed in this study is predicated on the NREL hardwood to ethanol process, which was described by Wooley and co-workers (1999) who published process design and economic information for the conversion of white wood to ethanol.

Wooley divided the biomass to ethanol plant into separate sections or modules, and summarized design and cost information for each major section of the plant; specifically, the feed handling, acid pretreatment, cellulase production, fermentation, product purification, storage, steam and power generation, utilities and waste treatment unit processes. Process flow diagrams, material and energy balance information, equipment sizes and cost information are presented for each section of the plant.

Information on the process flow diagram and equipment sizes were used in the present study to develop cost curves for the purchased and installed capital cost for the various unit processes of a biomass to ethanol plant. The total plant would of course be equal to the sum of the sections included in the design. Operating costs for the process were estimated using the material and energy balance information; again for the various unit processes comprising the entire plant. The profitability analysis in the model was accomplished by estimating net revenue and after tax profits. In this analysis cash flow diagrams were prepared as a function of time after the purchase of land for the plant.

The profitability of the biomass to ethanol project was estimated by calculating the Net Present Value (NPV) at the end of the venture, assumed to be ten (10) years after startup. Economic barometers used to estimate profitability were the NPV and the discounted cash flow rate of return for the process.

The economic viability of three biomass to ethanol projects was considered based upon production capacity. The base case investigated was a plant that consumes 2205 dry tons of hardwood daily, produces 52 million gallons of ethanol per year, and is about the size of a typical corn to ethanol plant currently being constructed in the Midwest. For purpose of comparison the economics of two additional projects were considered; a small plant (1103 dry tons of wood per day and 26 million gallons of ethanol annually) which can be easily supplied with hardwood in most areas of the State of Maine, and a large plant (4410 dry tons of wood per day and 104 million gallons of ethanol annually) that takes advantage of economies of scale to lower the operating cost for the process. Variants of the "greenfield" or free-standing plants considered locating the biomass to ethanol plant at an existing pulp mill; biorefinery case. Siting the ethanol production plant at an existing pulp mill site has several advantages; notable of which are reductions in permitting and access to existing process equipment such as a boiler and turbo-generator, utilities and waste treatment facilities, thus lowering the capital cost for the project. Since the "greenfield" plants are very expensive, reducing the capital investment has the potential of increasing the profitability of the venture. The results of all cases were compared and analyzed in order to gain a better understanding of the feasibility of building a hardwood biomass to ethanol plant in the State of Maine.

Chapter 4

METHODOLOGY

An economic analysis has been performed on the NREL hardwood biomass to ethanol process. The study consists of an estimation of the capital and operating costs, and a profitability analysis. The work of Wooley and co-workers (1999) was used as the basis for the current study. They performed a preliminary process design for a greenfield biomass to ethanol plant that consumes 2,205 dry tons per day of white wood and corresponded to a production rate (P) of approximately 52 million gallons per year of 100% ethanol.

Wooley's report provided material and energy balance information for the process, a list of equipment for each section of the plant, and cost estimates for the fixed capital investment and the plant operating costs. Appendix B summarizes pertinent information from the NREL report used in the present study. In performing the material balance for the process, Wooley assumed a yield of 68 gallons, or 448 pounds, of 100% ethanol for each bone dry ton of white wood; that is 22.4% of the dry weight of the wood can be converted into ethanol. At 2,205 bone dry tons of white wood per day and an assumed 350.25 operating days in a year, the annual production was estimated to be 52.2 million gallons of 100% ethanol. The information presented by NREL was very complete, but unfortunately it pertains to a plant built in 1999 and the economics were limited to one plant size.

In the current analysis, the information provided by Wooley was scaled to 2005 for the base case and extrapolated to a half-size plant (1,103 tons per day) and a plant twice the size of the base case (4,410 tons per day). Investigating three plant sizes allowed for the development of equations for estimating capital and yearly operating costs for a plant of any desired size lying within the range of wood sizes 1,100 to 4,400 tons per day of biomass or 26 to 104 million gallons of 100% ethanol per year.

Estimation of Capital Cost

To estimate the capital cost of a chemical plant, the plant is divided into sections or unit processes. The unit processes are then further subdivided into unit operations and individual pieces of equipment. The installed cost of each piece of equipment comprising the unit processes is estimated and summed to get the capital cost of each area of the plant. The capital cost for each unit process is then summed to get the capital cost for the plant.

Scaling Capital Equipment Costs

The relationship between the purchased cost (C_P) and capacity or size (A) of process equipment is related by the size ratio (r) and the scaling exponent (n).

$$C_{Pb} = C_{Pa} * \left(\frac{A_b}{A_a} \right)^n = C_{Pa} * r^n \qquad (4\text{-}1)$$

The size ratio is simply the capacity (A_b) of the equipment at size (b) divided by the capacity (A_a) of the equipment at size (a). Equation (4-1) permits purchased costs of equipment of size (a) to be scaled up or down for capacity by using the scaling ratio (r) and the scaling exponent (n) (Turton et al, 2003; Peters and Timmerhaus, 1991).

The purchase cost of each piece of process equipment from time period one (C_{P1}) can be converted to 2005 by using an appropriate cost index (I_{2005}), for example the Chemical Engineering Plant Cost Index (CEPCI), the Marshall and Swift index (MS) and the Engineering News-Record Construction Index to name a few.

$$C_P(2005) = C_P(Time\ 1) * \left(\frac{I_{2005}}{I_{Time\ 1}} \right) \tag{4-2}$$

In equation (4-2), [$C_P(2005)$] is the purchased cost of the equipment in year 2005 while [C_P(Time 1)] is the purchase costs of the same equipment in the base year. The cost indices I_{2005} and $I_{Time\ 1}$ are the appropriate values for the cost index in year 2005 and the original time of purchase; which in Wooley's analysis was the mid-to-late-nineties.

The installed equipment cost (C_I) was obtained by multiplying the purchased equipment cost [$C_P(2005)$] by an installation factor ($f_{installation}$). Thus the installed cost for each piece of equipment was taken to be

$$C_I(2005) = C_P(2005) * f_{Installation} \tag{4-3}$$

Or starting with the original purchased cost of the equipment (C_{Pa}) at time one (1), the installed cost in year 2005 [$C_I(2005)$] is given by the equation

$$\begin{aligned} C_{I,b}(2005) &= C_{Pa}(Time\ 1) * \left(\frac{I_{2005}}{I_{Time\ 1}} \right) * \left(\frac{A_b}{A_a} \right)^n f_{installation.} \\ &= C_{Pa}(Time\ 1) * \left(\frac{I_{2005}}{I_{Time\ 1}} \right) * (r)^n f_{installation.} \end{aligned} \tag{4-4}$$

The total cost of installed equipment (TC) for each section of the plant, exclusive of indirect costs, was taken as the sum of the installed cost for each piece of equipment.

$$\begin{aligned} TC_j(2005) &= Total\ Installed\ Equipment\ Cost\ (Section\ J) \\ &= \sum_i C_{I,i}(2005) \end{aligned} \tag{4-5}$$

This procedure was followed for each section of the plant. The total installed equipment cost ($TIC_{Installed}$) for the plant size under consideration was taken as the sum of the installed costs for each section of the plant

$$TIC_{Installed}(2005) = Total\ Installed\ Equipment\ Cost = \sum_{i} CT_{ji}(2005) \quad (4\text{-}6)$$

Application to NREL Design. The plant described by Wooley (Figure 2-5) was broken down into the following nine sections: feed handling and preparation (A-100), acid pretreatment (A-200), fermentation (A-300), cellulase generation (A-400), separations/di85stillation (A-500), waste treatment (A-600), storage (A-700), burner/boiler/turbogenerator (A-800), and utilities (A-900). Wooley provided the following information for all equipment required in the plant: (1) equipment name, (2) the equipment number, (3) the number of units required, (4) the number of spares required, (5) the size ratio (r), (6) the cost per unit of capacity and the total cost in the base year, (7) the scaling exponent (n), (8) the installation factor, and (9) the installed cost in the base year.

Table 4-1 illustrates the methodology for scaling the capital cost estimate to the 2005 time period for the three cases investigated. Wooley estimated the original equipment cost in the base year by multiplying the cost of each piece of equipment by the number required including spares. For the base case plant in the current economic study, the purchased equipment cost for each piece of equipment was updated to 2005 from the base year in Wooley's analysis, by using the Chemical Engineering Plant Cost Index ("Economic indicators", 2006) as given by equation 4-2. The installed cost was estimated from installation factors ($f_{installation}$) provided by Wooley (1999).

Table 4-1
Explanation of Equipment Cost Estimate

Cost Factor	Calculation Method
Original Equipment Cost in Base Year	(No. Reqd. + No. Spares) X (Original Cost Per Unit)
Equipment Cost 2005	(Original Equipment Cost in Base Year) X (CEPCI 2005/CEPCI Base Year)
Scaled Cost in Base Year	(Size Ratio$^{\text{Scaling Exponent}}$) X (Original Equipment Cost in Base Year)
Original Plant Scaled 2005	(Size Ratio$^{\text{Scaling Exponent}}$) X Equipment Cost 2005
Size Ratio Double Capacity	2 X Size Ratio
Size Ratio Half Capacity	Size Ratio/2
Scaled Cost 2005 Double Capacity	(Size Ratio Double Capacity$^{\text{Scaling Exponent}}$) X Equipment Cost 2005
Scaled Cost 2005 Half Capacity	(Size Ratio Half Capacity$^{\text{Scaling Exponent}}$) X Equipment Cost 2005
Base Case Plant Installed Cost in 2005	(Scaled 2005) X (Installation Factor)
Double Capacity Plant Installed Cost 2005	(Scaled Cost 2005 Double Capacity) X (Installation Factor)
Half Capacity Plant Installed Cost 2005	(Scaled Cost 2005 Half Capacity) X (Installation Factor)

The 2005 installed costs for each plant section, A100-A900, were estimated by taking the sum of all 2005 installed equipment costs associated with a particular section (see Equation 4-5). The total installed equipment cost for the whole base case plant equals the sum of the installed costs for all nine sections (Equation 4-6). Appendix C contains tables summarizing all equipment costs by section for the base case.

Additional Cost Factors

There are a number of other capital expenses, aside from equipment costs, that were estimated by Wooley and co-workers in 1999. Additional cost factors can be divided into additional direct- and indirect- costs. Additional direct costs include a Warehouse (W) and Site Development (SD). Additional indirect costs are Pro-ratable

Costs (PC), Field Expenses (FE), Home Office and Construction (HOC), Project Contingency ($P_{Contingency}$), and Other Costs (OC). In the current analysis, the additional costs were estimated by implementing the same assumptions used by Wooley in his original work (see Table 4-2).

Table 4-2
Additional Direct and Indirect Capital Cost Factors (Wooley, 1999)

Item	Description	Amount
Additional Direct Costs		
Warehouse (W)	Storage Warehouse	1.5% of Total Equipment Costs
Site Development (SD)	Site Development: Includes fencing, curbing, parking lot, roads, well drainage, rail system, soil borings, and general paving. This factor allows for minimum site development assuming a clear site, with no unusual problems such as right-of-way, difficult land clearing, or unusual environmental problems. 9% of the installed cost of process equipment (areas A100, A200, A300, A400, and A500).	9% of the installed cost of process equipment (Areas A100-A500)
Additional Indirect Costs		
Pro-rateable Costs (PC)	This includes fringe benefits, burdens, and insurance of the construction contractor.	10% Of Total Installed Cost
Field Expenses (FE)	Consumables, small tool equipment rental, field services, temporary construction facilities, and field construction supervision.	10% Of Total Installed Cost
Home Office and Construction (HOC)	Engineering plus incidentals, purchasing, and construction.	25% of Total Installed Cost
Project Contingency ($P_{Contingency}$)	Small because of the detail included in the process design.	3% of Total Installed Cost
Other Costs (OC)	Start-up and commissioning costs. Land, rights-of-way, permits, surveys, and fees. Piling, soil compaction/dewatering, unusual foundations. Sales, use, and other taxes. Freight, insurance in transit and import duties on equipment, piping, steel instrumentation, etc. Overtime pay during construction. Field insurance. Project team. Transportation equipment, bulk shipping containers, plant vehicles, etc. Escalation or inflation of costs over time. Interest on construction loan.	10% of Total Capital Investment

Total Direct Cost. Total direct costs (TDC) include the Total Installed Equipment Costs ($TIC_{Installed}$) plus the cost of the warehouse (W) and site development (SD).

$$TDC(2005) = TC_{Installed} + W + SD = Total\ Direct\ Cost \tag{4-7}$$

Total Indirect Cost. The total indirect cost (TIC) includes all of the indirect cost items listed in Tables 4-2.

$$TIC = PC + FE + HOC + P_{Contingency} + OC \tag{4-8}$$

Land Value (L). Land was considered to be part of "Other Costs" (see Table 4-2). The cost of land was not specified by Wooley, but for this study was taken to be equal to 1.5% of the Total Capital Investment (TCI) or 15% of "Other Costs".

Total Project Investment. Lastly, the Total Project Investment (TPI) in 2005 dollars was taken as the sum of total direct (TDC) and indirect costs (TIC).

$$TPI(2005) = TDC(2005) + TIC(2005) \tag{4-9}$$

This calculation procedure is summarized in Table 4-3.

Table 4-3
Summary of Capital Cost Factors[a]

Summary of Capital Cost Factors	
Direct Costs	
2005 Total Installed Equipment Cost	Sum of 2005 Installed Equipment Costs
Warehouse	1.5% of 2005 Total Installed Equipment
Site Development	9% of 2005 Installed Equipment Cost of A100-A500
Total Installed Cost	2005 Total Installed Equipment Cost + Warehouse + Site Development
Indirect Costs	
Field Expenses & Prorateable Costs	20% of Total Installed Cost
Home Office & Construction Fee	25% of Total Installed Cost
Project Contingency	3% of Total Installed Cost
Total Costs	
Total Capital Investment	Total Installed Cost + Indirect Costs
Other Costs	10% of Total Capital Investment
Total Project Investment	Total Capital Investment + Other Costs
(a) Calculated Separately for All 3 Plant Sizes	

Scaling to Other Plant Sizes

The cost for each piece of equipment in 2005 dollars for the base case plant (2,205 Tons per day) was scaled to the other two cases (1,103 and 4,410 tons per day plant sizes). This was done by using appropriate size ratios (r) and appropriate scaling exponents (n) as given by Equation (4-1). For example, to estimate the capital cost in 2005 dollars for the 4,410 tons per day plant, the size ratio would be (2r) while for the half-size plant, the size ratio would be (r/2). The exact same procedure was used for the 1,103 and 4,410 ton per day plants as was used for the base case (2,205 ton per day) plant. The additional cost factors shown in Tables 4-2 and 4-3 were applied to the other plant sizes. Appendix C contains the capital cost estimate for equipment associated with the double capacity, and half capacity plant sizes.

Estimation of Operating Cost

The results from the capital cost analysis were used in conjunction with information provided by Wooley and co-workers in 1999 to estimate the yearly operating cost. This was done for all three plant sizes. The operating cost estimate, sometimes referred to as cost of manufacture (COM) is made up of the following components: Raw Materials (RM), Waste Disposal (WD), Total Salaries (S), Overhead for Maintenance (OM), Maintenance per se (M), Taxes and Insurance (TI), and Capital Recovery. Each of these categories of cost is explained in the following sections.

Raw Materials (RM) and Waste Disposal (WD)

The costs associated with the raw materials (RM) and waste disposal (WD) are summarized in Table 4-4. It was assumed that the raw material usage would be doubled for the double capacity plant and cut in half for the half capacity plant. Therefore the yearly cost for each material is doubled for the large plant and divided by two for the small plant. The costs of the raw materials given in the NREL report were updated from 1996 to 2005 for use in the current analysis.

Wood Costs. The most significant change to NREL's raw material table was the price of white wood feedstock. Wooley assumed a price of $25 per bone dry U.S. ton, which is no longer realistic. Price quotes were obtained from a number of local companies in the forestry and pulp and paper industries (see Figure 4-1).

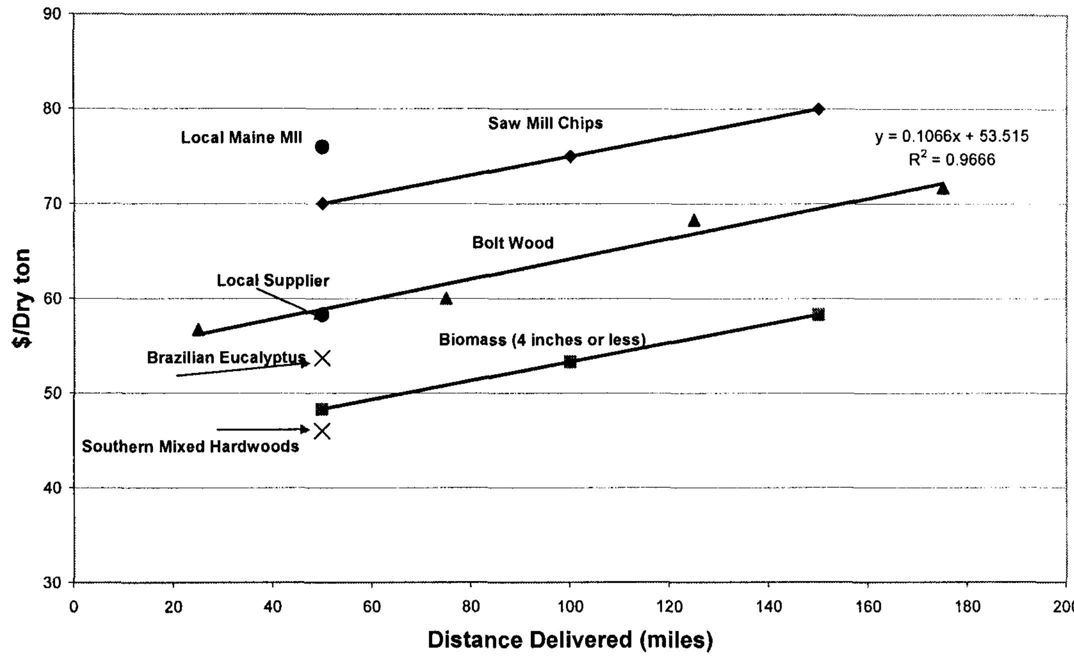


Figure 4-1. Wood Prices (Delivered) on a Bone Dry Basis ($/Ton Bone Dry)

In developing the data shown in Figure 4-1, a survey was made of delivered wood prices for saw mill chips, delivered bolt wood and biomass. The price of wood depends upon the distance of delivery. Bolt wood contains bark and is greater than 4-inches in diameter. Sawmill chips, as the name implies, come from saw mills and are derived from the outside edges of wood bolts. This is premier wood since it is derived from the outer edges of the merchantable bole. Biomass consists of wood that is 4-inch or less in diameter and consists of small trees and branches. Biomass does not contain needles or leaves, but contains considerable bark. Also included in Figure 4-1 are the cost of Brazilian Eucalyptus chips delivered in Maine and the price of mixed southern hardwood. The NREL process is applicable only to white wood and does not work on bark. In the current analysis, the cost of white wood as the feedstock was assumed to be $60 per bone dry U.S. ton. This corresponded to a delivery distance of about 60 miles.

Other Raw Materials. A majority of the other raw material costs in 2005 dollars [RM(2005)] were updated from 1996 dollars to 2005 dollars by using the inorganic chemical index and equation (4-10).

$$\mathrm{RM}(2005,\$/\mathrm{yr}) = RM(1996,\$/\mathrm{yr}) * \left[\frac{\mathrm{I}_{\mathrm{Inorganic}}(2005)}{\mathrm{I}_{\mathrm{Inorganic}}(1996)}\right] \tag{4-10}$$

Values of the Inorganic Chemical Index ($I_{Inorganic}$) were provided by Wooley et al. for 1999. The value for 2005 was a projection that was determined by extrapolation. The cost of make-up water was taken to be equal to NREL's cost for the base case, doubled for the large plant, and halved for the small plant. The price of diesel fuel in the NREL report was $0.426/gal (1998) and was taken from the DOE Energy Information Administration (EIA) in July of 1998. The current price for diesel reported by the DOE Energy Information Administration is $1.778/gallon (EIA, 2006) and was used in the current analysis.

Waste Disposal Costs (WD). The cost of disposing both ash and gypsum was $20 per metric ton in 1996. It was estimated that in 2005 the disposal charge for ash and gypsum was $40 per metric ton, hence the cost of disposal doubled for the base case plant. As with the raw materials, it was assumed that for the double capacity plant the amount of waste disposal would be doubled and for the half capacity plant the disposal would be cut in half. The total cost of all the raw materials and disposal charges will be referred to as "Raw Materials" when discussing the total yearly operating costs.

Table 4-4
Raw Materials and Waste Disposal Summary (Wooley, 1999)

Raw Material	1996 Base Case MM$/yr	Inorganic Chemical Index 1996	Inorganic Chemical Index 2005 (Projected)	2005 Base Case MM$/yr	2005 Double Capacity MM$/yr	2005 Half Capacity MM$/yr
Biomass Feedstock	19.31	1996 $25/ton	2005 $60/ton	46.34	92.69	23.17
Cellulase (a)	0		$0.0552/lb	4.60	9.20	2.30
Sulfuric Acid	0.41	119.5	131.2	0.45	0.90	0.23
Lime	0.44	119.5	131.2	0.48	0.97	0.24
Ammonia	2.2	119.5	131.2	2.17	4.34	1.08
Corn Steep Liquor	2.63	119.5	131.2	2.89	5.77	1.44
Nutrients	0.43	119.5	131.2	0.00	0.00	0.00
Ammonium Sulfate	0.16	119.5	131.2	0.18	0.35	0.09
Antifoam (Corn Oil)	1.01	119.5	131.2	0.00	0.00	0.00
WWT Nutrients	0.45	119.5	131.2	0.49	0.99	0.25
BFW Chemicals	0.01	119.5	131.2	0.01	0.02	0.01
CW Chemicals	0.1	119.5	131.2	0.11	0.22	0.05
WWT Chemicals	0.03	119.5	131.2	0.03	0.07	0.02
Make-up Water	0.45			0.45	0.90	0.23
Diesel	0.48	$0.407/gallon	$1.778/gallon	2.10	4.19	1.05
Ash Disposal	0.19	1996($20/Mt)	2005($40/Mt)	0.38	0.76	0.19
Gypsum Disposal	0.42	1996($20/Mt)	2005($40/Mt)	0.84	1.68	0.42
TOTAL	28.72			61.52	123.05	30.76
(a) The cost of the cellulase enzyme is only applicable to Co-Location Cases Band C.						

Labor Costs (LC)

The report by Wooley provided an economic summary for the employees, but it corresponds to the year 1998. Table 4-5 shows how this information was updated to a 2005 basis. The cost of salaries for the various jobs was assumed to be the same for all three plant sizes being studied. The basis for this assumption is that the equipment for the three plants varies in size, but the amount of people needed to operate the equipment remains the same.

Table 4-5
Cost of Labor and Supervision (Wooley, 1999)

Job Description	Salary	Number	Total Cost 1998	Labor Index 1998	Labor Index 2005 Projected	Total Cost 2005 All Plant Sizes
Plant Manager	$80,000	1	$80,000	17.17	19.90	$92,720
Plant Engineer	$65,000	1	$65,000	17.17	19.90	$75,335
Maintenance Supervisor	$60,000	1	$60,000	17.17	19.90	$69,540
Lab Manager	$50,000	1	$50,000	17.17	19.90	$57,950
Shift Supervisor	$37,000	5	$185,000	17.17	19.90	$214,415
Lab Technician	$25,000	2	$50,000	17.17	19.90	$57,950
Maintenance Technician	$28,000	8	$224,000	17.17	19.90	$259,616
Shift Operators	$25,000	20	$500,000	17.17	19.90	$579,499
Yard Employees	$20,000	8	$160,000	17.17	19.90	$185,440
General Manager	$100,000	1	$100,000	17.17	19.90	$115,900
Clerks & Secretaries	$20,000	5	$100,000	17.17	19.90	$115,900
Total Salaries			**$1,574,000**			**$1,824,263**

The labor cost in 1998 for each job, calculated by NREL, equals the number of people doing that particular job times the salary. The 1998 labor cost (LC_{1998}) for each job was updated to 2005 (LC_{2005}) by using the projected Labor Index (LI) values provided by Wooley. Thus for job "i" the 2005 cost is given by equation 4-11.

$$(LC_{2005})_i = (LC_{1998})_i * \left[\frac{LI_{2005}\ \text{Projected}}{LI_{1996}}\right] \qquad (4\text{-}11)$$

The total labor cost (TL) in terms of 2005 dollars was equal to the sum of individual salaries $(LC_{2005})_i$.

$$TL(2005) = \sum_i (LC_{2005})_i \tag{4-12}$$

Other Operating Costs

Other operating costs that were estimated were Overhead and Maintenance (O&M), Maintenance per se (M), and Insurance and Taxes (I&T). These costs were estimated based on percentages of other costs as summarized in Table 4-6 (Wooley, 1999).

Table 4-6
Fixed Operating Costs (Wooley, 1999)

Operating Cost	**Calculation Method**
Overhead/Maintenance	60% of Total Salaries
Maintenance	2% of 2005 Total Installed Equipment Cost
Insurance & Taxes	1.5% of Total Installed Cost

Overhead/Maintenance (O&M) refers to the following: safety, general engineering, general plant maintenance, payroll overhead including benefits, plant security, janitorial services, phone, light, heat, and plant communications. This value was assumed to be 60% of the Total Salaries. Maintenance (M) refers to annual maintenance materials and is estimated to be 2% of the 2005 Total Installed Equipment Cost. The cost of Insurance and Taxes (I&T) was estimated to be 1.5% of the Total Installed Cost.

Capital Recovery (R_C)

The last component comprising the operating cost analysis was the cost of capital and its recovery (Peters and Timmerhaus, 1991). Capital Recovery is the yearly cost of borrowing money (R_C, dollars per year) and repaying the borrowed capital (FCI, dollars) in the form of an annuity over n-years at interest rate (i).

$$R_C = FCI * \left[\frac{i\,(1+i)^n}{(1+i)^n - 1} \right] = FCI * CRF \qquad (4\text{-}13)$$

The capital recovery factor (CRF) is given as by the factor $[i(1+i)^n]/[(1+i)^n-1]$.

Recovery of capital pertains only to depreciable capital, which does not include land. To make this distinction the notation (FCI_L) was used to denote the fixed capital investment (FCI) excluding land (L). It was taken equal to the Total Project Investment minus the cost of land.

$$FCI_L = [TPC(2005) - L] \qquad (4\text{-}14)$$

For the current analysis and in NREL's document a capital recovery factor (CRF) of 0.182 was used. The yearly charge for capital was calculated using equation 4-15.

$$R_C = CRF * [FCI_L] \qquad (4\text{-}15)$$

The capital recovery factor was applied over a ten year period (n). A capital recovery factor of 0.182 is equivalent to assuming that money is borrowed at 12.7% (i) over a ten (10) year period.

Cost of Manufacturing (COM_d)

The total yearly operating cost or the cost of manufacture excluding depreciation (COM_d), was the sum of Raw Materials (RM), Waste Disposal (WD), Total Labor Cost (TL), Overhead/Maintenance (O&M), Maintenance (M), Insurance and Taxes (I&T), and Capital Recovery (R_C).

$$COM_d = RM + WD + TL + O\&M + M + I\&T + R_C \tag{4-16}$$

To determine the cost of producing a gallon of ethanol for each plant size, the COM_d for each plant size was divided by the corresponding ethanol production capacity (P).

$$C_P = \text{Pr}\,oduction\ Cost = \left[\frac{COM_d}{P}\right] = Dollars\,/\,Gallon \tag{4-17}$$

Profitability Analysis

Basis for Analysis

The basis for the profitability analyses is summarized in Table 4-7.

Table 4-7
Basis for Profitability Analysis

Item	Basis	Value
Land Value (L)	1.5% of Total Capital Investment (TCI) invested at time 0 years	1.5%*TCI
Construction Period (N_C)	Three year from purchase of Land	N_C = 3 years
Distribution of Capital Investment (FCI_L)	Distributed over construction period. Percent of FCI_L	8% Yr. 1 61% Yr. 2 31% Yr. 3
Working Capital (WC)	5% of Total Project Cost (TPC) invested in Year 3	5%TPC
Tax Rate (t)	39% of Gross Profits	39%
Project Life (N)	13 years from purchase of land or 10 years after startup	13 Years
Depreciation (d_i)	Ten year straight line (N_d = 10 years)	FCI_L/N_d
Selling Price of Ethanol (S_{EtOH})	Treated as a variable between $1.85 and $3.50/gallon. Base case was $2.50 per gallon	S_{EtOH} = $2.50/gal (Base Case)
Selling Price of Electricity (S_{Elec})	Selling into local grid at 4.3 cents per kilowatt hour	$0.043/kWhr
Recovery of Land (L) and Working Capital (WC)	Fully recovered in year 13	Year 13

The project is initiated with the purchase of land (L), which was assumed to occur at time zero. For the sake of simplicity, in the current analysis the construction was assumed to take place over three years (N_C) and startup occurs at the end of year three. The distribution of the fixed capital investment, excluding land (FCI_L) was assume to be spread over a three year period as follows: 8% of FCI_L in year one, 61% of FCI_L in year two, and 31% of FCI_L in year three. These are the same percentages used by NREL in their analysis, which assumed a slightly different startup period with a construction period of two and a half years and a startup time of 6 months. A detailed description of the tasks performed in these three years can be seen in Table B-14 of Appendix B (Wooley, 1999).

The plant was assumed to start up in year three (3). The working capital is money necessary to get the plant up and running before revenues are generated. The expense includes salaries, raw materials, and any other costs from operating the plant prior to revenues flowing to the project (Turton, 2003). In the current economic analysis and in Wooley's study, the working capital was assumed to be equal to 5% of the Total Project Investment. Revenues were assumed to be generated over a ten year plant life following startup (N). The working capital (WC) and the land (L) were assumed to be fully recovered in year 13. The selling price of ethanol was assumed to be $2.50 per gallon for the base case analysis. A sensitivity analysis was then performed to determine how the profitability of the three plants varied with the selling price of ethanol.. Discounted cash flow diagrams were constructed to show the accumulation of negative and positive cash flows over the life of each Greenfield plant as a function of plant size.

Net Profit

The net profit (NP) for any year "i" is defined as the revenue stream obtained from selling the products (R_i) minus the cost of manufacturing $(COM)_i$ minus the depreciation (d_i) multiplied by one minus the tax rate (t).

$$NP_i = \left[R - COM - d\right]_i * (1 - t) \tag{4-18}$$

In the present analysis the tax rate was assumed to be 39% or $t = 0.39$; which is the same as used by Wooley (1999). The revenue stream for the "ith" year (R_i) is equal to the annual ethanol production rate (P_{EtOH}) times the assumed selling price (S_{EtOH}) for the ethanol plus an electricity credit obtained by selling excess power (P_{Elec}) generated in the turbogenerator section of the plant times an assumed selling price (S_{Elec}).

$$R_i = \left[P_{EtOH} * S_{EtOH} + P_{Elec} * S_{Elec}\right]_i \tag{4-19}$$

Due to the fluctuating nature of the ethanol market, the cash flows were evaluated for a range of different ethanol selling prices (S_{EtOH}) for each plant size. The prices studied ranged from $1.85 per gallon to $3.50 per gallon. The selling price for electrical power (S_{Elec}) was assumed constant at ($0.04 per kilowatt hour) but the production of excess power varied with the size of the plant.

Depreciation

The depreciation is the fraction of the capital investment that the government allows companies to charge as a yearly operating expense in order to make up for the decrease in plant value over time (Turton, 2003). A straight line method of depreciation was used in the current economic analysis with a depreciation period (N_d) of 10 years.

$$d_i = \frac{FCI_L}{N_D} \tag{4-20}$$

Cash Flow

The cash flow (CF_i) for any given year (Equation 4-21) is defined as the net profit $(NP)_i$ plus the depreciation (d_i).

$$CF_i = \left[R - COM - d\right]_i * (1 - t) + d_i \tag{4-21}$$

The cash flow can be either positive or negative depending upon the cost of manufacturing and the depreciation. The depreciation invariably results in a tax credit.

$$CF_i = [R - COM]_i * (1 - t) + d_i * t \qquad (4\text{-}22)$$

Negative cash flows occur in the early years when the land is purchased, the plant is constructed and the working capital is installed. Cash flow diagrams were constructed using methods described by Turton et al, (2003). These diagrams show both negative and positive cash flows for the project. The net present value (NPV) and discounted cash flow rate of return (DCFROR) methods were used to determine profitability of the projects.

Net Present Value (NPV) Method

In the net present value method all positive and negative cash flows were discounted back to time zero, that is the point at which the land was purchased, to account for the time value of money; which was taken to be 10% (i = 10%), or the minimum rate of return acceptable for investment of capital ($i = i_C$). The yearly discounted cash flows were calculated by using appropriate discount factors. For example the discounted cash flow for the k^{th} year would be given by Equation 4-23 (Turton, 2003).

$$DCF_k = \frac{CF_k}{(1+i)^k} = \textit{Discounted Cash Flow for } k \qquad (4\text{-}23)$$

The net present worth at the end of the project (NPV) was determined by taking the sum of all negative and positive discounted cash flows for the entire 13 year project life.

$$NPV(i) = \sum_{k=0}^{k=N} \left[\frac{CF_k}{(1+i)^k} \right] \qquad (4\text{-}24)$$

When applied to the biomass to ethanol plants equation (4-24) becomes

$$NPV(i) = -\left[\frac{L}{(1+i)^0}\right] - \left[\frac{FCI_i}{(1+i)^1}\right] - \left[\frac{FCI_2}{(1+i)^2}\right] - \left[\frac{FCI_3}{(1+i)^3}\right] - \left[\frac{WC}{(1+i)^3}\right] + \sum_{k=4}^{k=13}\left[\frac{CF_k}{(1+i)^k}\right] + \left[\frac{(WC+L)}{(1+i)^{13}}\right] \tag{4-25}$$

In the net present value method, the worth of the investment is judged by the magnitude of the net present value. If the net present value is very large, the investment is judged to be very good. If the net present value is equal to zero, it would mean that the entity making the investment would just return all of the money invested. Lastly if the net present value is negative, the project would be judged to be quite deficient and not worthy of the investment of funds.

Case 1. $NPV >> 0$ Very Good Project Investment

Case 2. $NPV = 0$ Project Investment is Neutral

Case 3. $NPV < 0$ Poor Project Investment

Discounted Cash Flow Rate of Return Method

A second more rigorous method of estimating profitability was the discounted cash flow rate of return (DCFROR). The DCFROR is the discount rate at which the net present value (NPV) at the end of the project would be equal to zero (Turton, 2003). In the discounted cash flow rate of return on investment, the interest rate (i) is found so that the negative cash flows are just balanced by the positive cash flows. The DCFROR method of judging projects involves a trial and error solution. The discount rate (i) was varied until the net present value (NPV) became zero.

$$\sum_{k=0}^{k=N=Ter\min ation}\left[\frac{CF_k}{(1+i)^k}\right] = 0 \tag{4-26}$$

In the discounted cash flow rate of return method a project is again judged to be good, neutral or bad by comparing the calculated discounted rate of return (i) to the cost of capital (i_C); which in the present study was taken to be 10%.

Case 1. $i >> i_C$ Very Good Project Investment

Case 2. $i = i_C$ Project Investment is Neutral

Case 3. $i << i_C$ Poor Project Investment

Co-Location or Alternative Cases

A complete economic analysis was performed for the Greenfield hardwood biomass to ethanol plant as a function of plant size. In addition, three alternative project cases were evaluated in an effort to increase the profitability of the projects. These cases were referred to as Co-Location Case A, Co-Location Case B, and Co-Location Case C. These alternatives employed the same production process but were assumed to be located at an existing pulp mill in an effort to share utility cost and decrease the capital investment. The purpose of the co-location cases was to determine the savings that could be gained by decreasing the capital investment in some realistic manner.

Co-Location Case A. Existing Power Generation Facilities. Co-Location Case A is a hardwood to ethanol plant that is co-located at a pulp mill that already has a recovery boiler, burner, and turbogenerator with excess unused capacity that can burn residual lignin and by-products from the fermentation process. In this case, it was assumed that the company building the ethanol plant also owned the pulp mill; therefore there would not be additional charges for using the boiler system and the installed cost of the burner/boiler/turbogenerator section (A-800) would be equal to zero. This case coincides with a paper company wishing to produce additional products besides pulp.

Co-Location Case B. Existing Power Generation and Purchase of Cellulase. Co-Location Case B is the same as Co-Location Case A except that the cellulase enzyme is now purchased from a cellulase supplier rather than produced on site. The installed equipment cost for section A-400 (cellulase generation) becomes zero and cellulase is now considered a raw material.

Co-Location Case C. Existing Power Generation, Purchase of Cellulase and Existing Waste Treatment and Wood Yard Facilities. Co-Location Case C continues to build on the concept of Co-Location Case B and the facility is now sited at a pulp mill that has additional waste treatment capacity (A-600) and a wood yard (A-100) that could handle the additional wood as ethanol feedstock.

Methodology of Economic Analysis for Co-Location Cases. For the three co-location options, similar economic analyses were performed as for the Greenfield plant; that is the base case plant (2,205 tons per day of biomass), double capacity plant (4,410 tons per day), and half capacity plant (1,103 tons per day). The methodology for the co-location cases was essentially the same as that used for the Greenfield plants. The capital costs were reduced accordingly for each co-location case. Factors that were estimated by taking a percentage of a capital cost factor were reduced accordingly. For example, capital recovery charges were reduced because they are directly related to the fixed capital investment. The only other change in methodology was the addition of cellulase as a raw material for co-location cases B and C. This was done by multiplying the cellulase usage by the price of cellulase, which was assumed to be $0.0552/lb (Aden, 2002).

Chapter 5

RESULTS OF ECONOMIC ANALYSIS

Estimation of Total Project Investment

A capital cost estimate was made for the Greenfield hardwood to ethanol plant. The installed capital cost (TIC_{2005}) was estimated as a function of the nine (9) sections or unit processes comprising the plant (Table 5-1). Details of the capital cost estimate are summarized in Appendix C.

Table 5-1
Installed Equipment Cost by Plant Section: Greenfield Plant

	Total Installed Equipment Cost ($Millions)		
Plant Section	Base Case Plant 2005 (2205 Tons/Day)	Double Capacity Plant 2005 (4410 Tons/Day)	Half Capacity Plant 2005 (1102.5 Tons/Day)
Feed Handling (A100)	6.0	9.1	4.0
Pretreatment/Detox (A200)	32.0	52.0	19.9
Fermentation (A300)	16.7	32.4	8.6
Cellulase Production (A400)	18.1	32.7	10.4
Distillation (A500)	19.9	32.5	12.2
Waste Water Treatment (A600)	12.6	19.2	8.4
Storage (A700)	2.2	3.5	1.4
Boiler/Turbogenerator (A800)	54.1	89.1	33.0
Utilities (A900)	6.3	10.7	3.8
Total Installed Equipment Cost	**167.9**	**281.2**	**101.6**

Figure 5-1 illustrates the distribution of the total installed equipment for the various sections of the plant for the base case, 52 million gallon per year ethanol plant. This figure is essentially the same for the other plant sizes. The major areas of the plant that contribute to the total capital investment are the boiler and turbo-generator facility (31%), pretreatment and detoxification area (19%), the distillation area (12%), the cellulase production area (11%), fermentation area (10%) and the waste treatment area (8%). All other areas are relatively small compared to these. The large expenditure of capital on the boiler and turbo-generator facility arises because of the large amount of lignin in the wood and residual products from the ethanol fermentation.

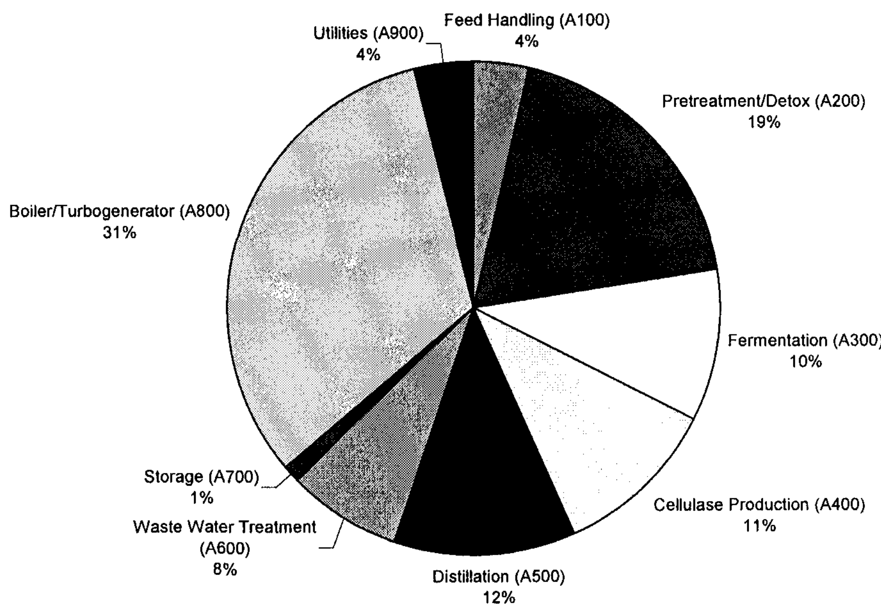


Figure 5-1 Breakdown of the Installed Equipment Cost by Sections of the Plant (Greenfield Plant at the Base Case Size, 52 Million Gallons of Ethanol per Year)

Capital costs estimates were also developed for the three Co-Location cases considered in this thesis.

Co-Location Case A. Locate Ethanol Plant at an Existing Pulp Mill with Spare Power Generation Capacity.

Co-Location Case B. Locate Ethanol Plant at an Existing Pulp Mill with Spare Power Generation Capacity and also Purchase Cellulase Enzyme.

Co-Location Case C. Locate Ethanol Plant at Existing Pulp Mill with Spare Power Generation Capacity, Waste Treatment and Wood Yard Facilities plus Purchase Cellulase Enzyme

Estimates for the cost of the installed equipment are summarized in Tables D-4, D-6, and D-8 for the three co-location cases. The zeros in Tables D-4, -6 and -8 for the total installed equipment cost result from the assumption of using existing equipment in the co-location cases. For each plant section, the total installed equipment cost was plotted versus the daily hardwood usage. This procedure permitted capital cost curves to be generated for the various sections of the plant (see Figures D-6 through D-14). These equations are summarized in Table D-2 and can be used to estimate the total installed equipment cost for white wood to ethanol plants ranging in size from 1,103 to 4,410 tons per day of hardwood consumption.

Total Project Investment (TPI_{2005}) for the Greenfield hardwood to ethanol plant includes the Total Installed Equipment Cost ($TIC_{Installed}$), plus the cost of a warehouse (W), site development (SD), and various indirect costs (IC) that were outlined in Table 4-2. Estimates for the Total Project Investment are summarized in Table D-3 for the Greenfield case and also in Tables D-5, -7, and -9 for the three (3) co-location cases. The estimated Total Project Investment for the four (4) investment scenarios is shown as a function of plant size in Figure 5-2. The trend line equations located in the upper left hand portion of the graph can be used to determine the Total Project Investment for plants of any desired hardwood feed rates within the limits of the analysis.

Building a Greenfield hardwood to ethanol plant of the type developed by NREL clearly involves a sizeable investment; regardless of plant size. Figure 5-2 illustrates that major reductions in capital can be achieved by finding a site for co-location. In essence the co-location cases involve finding an entity that will underwrite a sizeable portion of the hardwood to ethanol plant at no cost to the project. Relative to the base case plant size of a Greenfield plant, the Total Project Investment can be reduced by approximately 47%, or $153.6 million dollars, by meeting the criteria of Co-Location Case C; namely finding a plant site with existing power generation facilities, substituting purchase of cellulase for on-site production of the enzyme, and having existing wood yard and waste treatment facilities. It would, of course, be difficult to fulfill the criteria outlined in Option C. Most likely, this project could only be undertaken by some state government as a project to promote employment in a depressed region. Under Option C, an existing mill would have to be purchased and renovated to meet the needs of the ethanol production facility. Cases A and B would appear to be more realistic, especially the case where cellulase is purchased rather than produced on site. These latter two options also lead to an appreciable reduction in the capital investment required to build a white wood to ethanol plant.

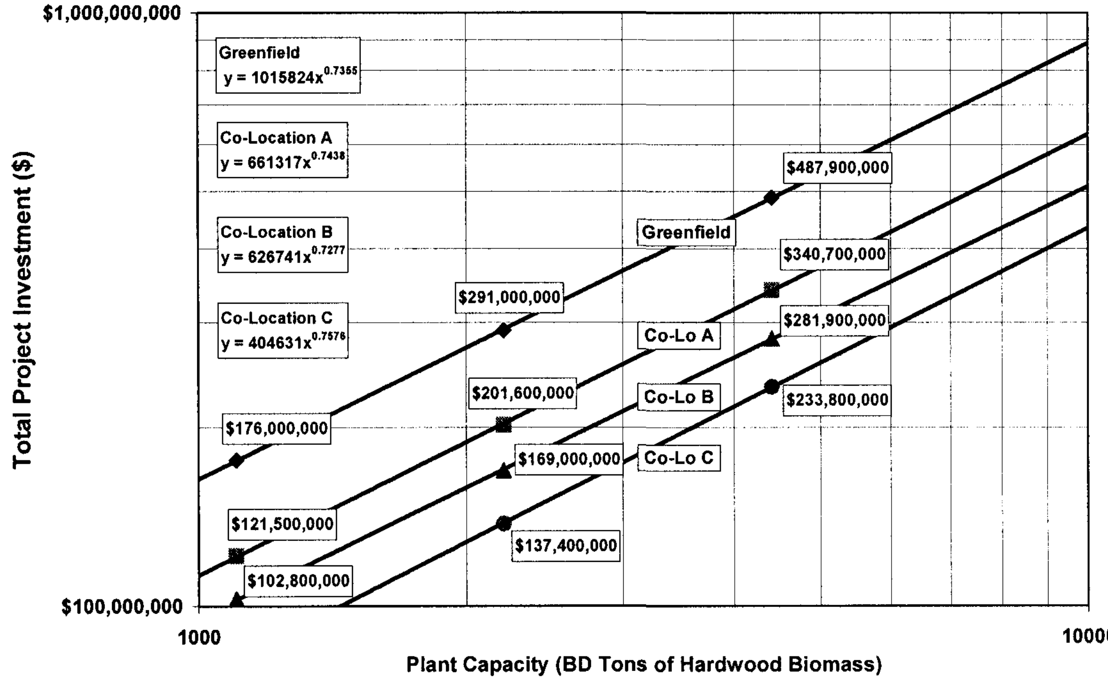


Figure 5-2. Summary of Total Project Investment for Four Investment Scenarios

Economy of Scale. The capital investment increases with an increase in plant capacity, according to a scaling exponent (n); usually about a power of 0.7 when plotted on a log-log graph. Generally, when the size of the plant is increased the cost of the installed equipment per unit of plant capacity will decrease. For example, when the Greenfield plant was doubled (100% increase) in size from 52.2 million gallons of annual ethanol production capacity to 104.4 million gallons, the cost of installed equipment only increased by about 67%; or from about $167.9 million to $281.2 million dollars. This concept is further illustrated in Figure 5-3 where the total project investment (TPI_{2005}) is given in terms of dollars per unit of capacity (UC_{2005}) for the four investment scenarios considered.

$$UC_{2005}(Dollar/Gallon/year) = \left[\frac{TPI_{2005}(Dollars)}{P(Gal.\ ETOH/year)}\right]$$

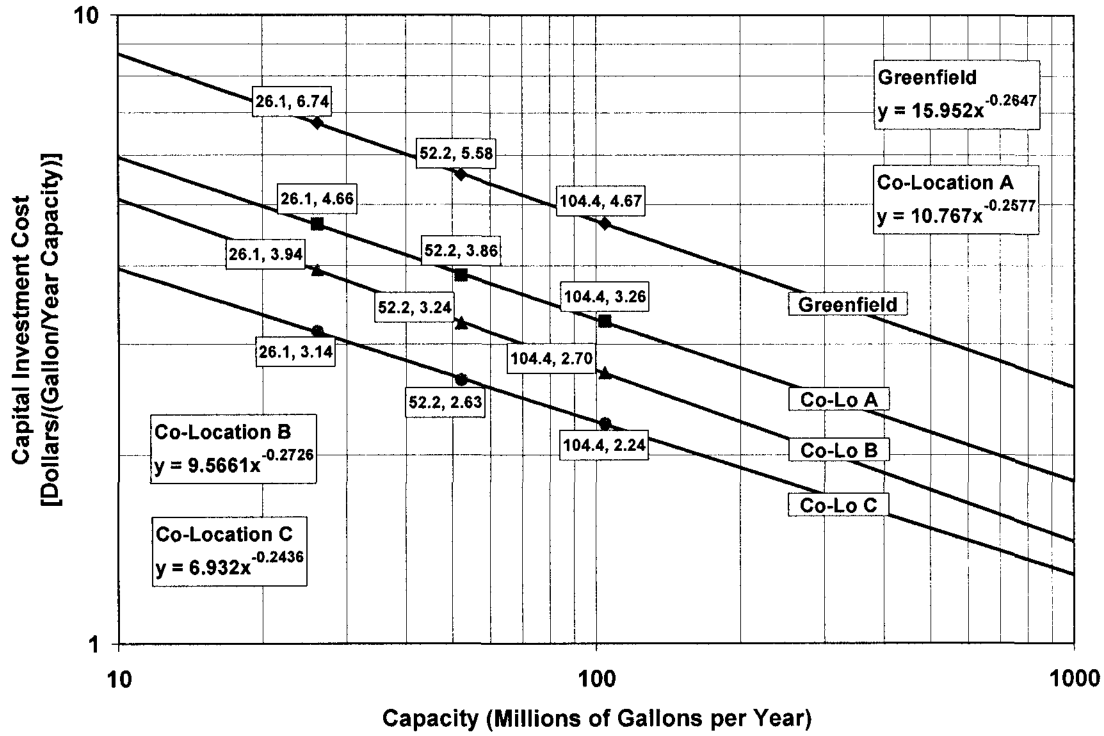


Figure 5-3. Total Project Investment per Unit of Ethanol Capacity

Table 5-2 summarizes the information from Figure 5-3 along with the unit cost associated with the installed equipment for each plant size and option considered. It is clear from the negative slope of the lines (see Figure 5-3) that the capital cost per unit of production capacity decreases with an increase in plant size; clearly illustrating economy of scale. The unit cost of the Greenfield plant will be reduced accordingly for the Co-Location Cases. The factor that limits the size of the plant, and thus the effect of economy of scale, will of course be the ability to supply white hardwood to the plant site. For any given plant location only so much wood can be economically delivered to the site on a sustainable basis. Consequently, having ever increasing plant sizes becomes unrealistic.

Table 5-2
Total Plant Cost for White Wood to Ethanol Plant (Co-Location Case B)

Plant Case	Ethanol Plant Capacity [Millions Gallons/Year]	Unit Installed Equipment Cost[a] [Dollars/(Gallon/Year)]	Unit Plant Cost [Dollars/(Gallon/Year)]
Greenfield Plant	26.1	3.89	6.74
	52.2	3.22	5.58
	104.4	2.69	4.67
Co-Location Case A	26.1	2.63	4.66
	52.2	2.18	3.86
	104.4	1.84	3.26
Co-Location Case B	26.1	2.23	3.94
	52.2	1.83	3.24
	104.4	1.53	2.70
Co-Location Case C	26.1	1.76	3.14
	52.2	1.48	2.63
	104.4	1.26	2.24
(a) Installed Equipment (Not Including Warehouse, Site Development and Indirect Costs)			

Estimation of Annual Operating Cost

Details of the estimated yearly operating costs are summarized in Appendix E. Table E-1 is a summary of the raw material usage for the Greenfield case. This information is identical for Co-Location Case A. The raw material usage for Co-Location Cases B and C are shown in Table E-2. In these cases, cellulase enzyme has been added to the raw materials table. Also, the nutrients and corn oil (antifoam) have been eliminated and the amount of ammonia has been decreased appropriately because these chemicals are no longer being used for the on site production of the cellulase enzyme. The yearly salaries for plant employees are summarized in Table E-3 and were assumed to be equal for all plant cases.

Table 5-3 summarizes the annual production of products from the white wood to ethanol plant. The major products are ethanol, electrical energy, gypsum and ash. In the analysis, the ash and gypsum were treated as by-products but assigned negative costs associated with their disposal. The electrical energy is put into the grid but contributes modestly to the revenue of the plant. Virtually all of the positive revenue originates from the sale of the ethanol.

Table 5-3
Summary of Product Products Produced at White Wood to Ethanol Plant

Item	Amount	Yearly Revenue/Cost ($Millions)
Half Capacity Plant		
White Wood at $60/BD Ton	1103 BD tons/day	-23.2
Ethanol at $2.50/Gallon	26.1 million gallons/year	65.3
Electricity at $0.043/kWhr	5,471 kW	2.1
Gypsum at $40/Ton (Disposal Fee)	5,100 tons/year	-0.2
Ash at $40/Ton (Disposal Fee)	11,300 tons/year	-0.4
Base Case Plant Size		
White Wood at $60/BD Ton	2205 BD tons/day	-46.3
Ethanol at $2.50/Gallon	52.2 million gallons/year	130.5
Electricity at $0.043/kWhr	10,942 kW	4.2
Gypsum at $40/Ton (Disposal Fee)	10,200 tons/year	-0.38
Ash at $40/Ton (Disposal Fee)	22,500 tons/year	-0.84
Double Capacity Plant		
White Wood at $60/BD Ton	2205 BD tons/day	-92.7
Ethanol at $2.50/Gallon	104.4 million gallons/year	261
Electricity at $0.043/kWhr	21,884 kW	8.3
Gypsum at $40/Ton (Disposal Fee)	20,500 tons/year	-0.76
Ash at $40/Ton (Disposal Fee)	45,100 tons/year	-1.68

Table 5-4 summarizes the yearly operating cost and the cost of producing a gallon of ethanol for the Greenfield plant case. From inspection it is clear that the yearly operating cost, or Cost of Manufacture (COM_d), of a hardwood to ethanol plant is quite high; primarily associated with the cost of wood. For the Greenfield plant it was estimated that it would range from $67.4 million dollars per year at the small plant size to $217.6 million dollars per year at the large plant size. The breakdown of yearly operating costs for the base case plant is summarized in Figure 5-4.

Table 5-4
Estimated Yearly Operating (Manufacturing) Costs for Greenfield Plant

	Yearly Operating Costs ($Millions)		
Yearly Costs	Base Case Plant 2005 (2205 Tons/Day)	Double Capacity Plant 2005 (4410 Tons/Day)	Half Capacity Plant 2005 (1102.5 Tons/Day)
Raw Materials	58.8	117.5	29.4
Total Salaries	1.8	1.8	1.8
Overhead/Maintenance	1.1	1.1	1.1
Maintenance	3.4	5.6	2.0
Insurance & Taxes	2.7	4.5	1.6
Recovery of Capital	52.2	87.6	31.6
TOTAL	**120.0**	**218.1**	**67.5**
Cost Per Gallon of Ethanol Produced	**$2.30**	**$2.09**	**$2.59**

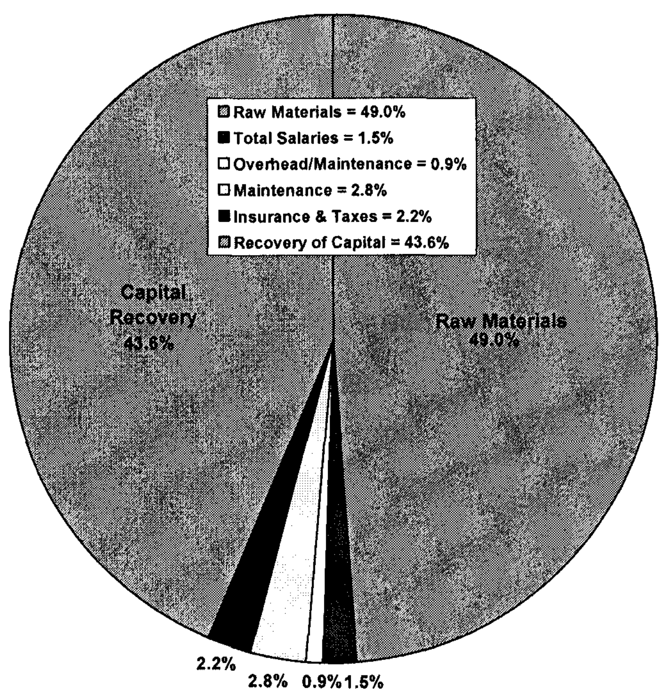


Figure 5-4. Breakdown of Yearly Operating Costs for the Greenfield Plant (Base Case Size, 52 Million Gallons of Ethanol per Year)

Two factors essentially dictate the cost of manufacturing (COM_d); namely the high cost of the raw materials, most notably the cost of wood, and secondly the cost of capital. For the Greenfield plant at base case size, the raw materials account for 49% of the yearly operating costs. The wood alone is responsible for about 79% of the raw material cost or approximately 39% of the yearly operating cost. The second factor contributing to the high operating cost is the cost of recovering capital. For the Greenfield plant case at the base case size, the cost of recovering the capital investment ($0.182*FCI_L$) accounts for nearly 44% of the COM_d.

Tables E-5, E-6, and E-7 show the yearly operating cost summary for Co-Location Cases A, B, and C respectively. For comparison purposes, the yearly operating costs for all cases are displayed graphically in Figure 5-5. Each of these lines can also be viewed separately in Figures E-1, E-2, E-3, and E-4 of Appendix E. The equations in the upper left of the graph can be used to estimate the yearly operating cost for a plant of any desired daily hardwood feed rate. The yearly operating cost is reduced significantly for the co-location cases because the capital recovery cost is significantly reduced. Capital recovery remains a significant factor even in the co-location cases. Reductions in capital expenditures are the driving force in the co-location cases considered

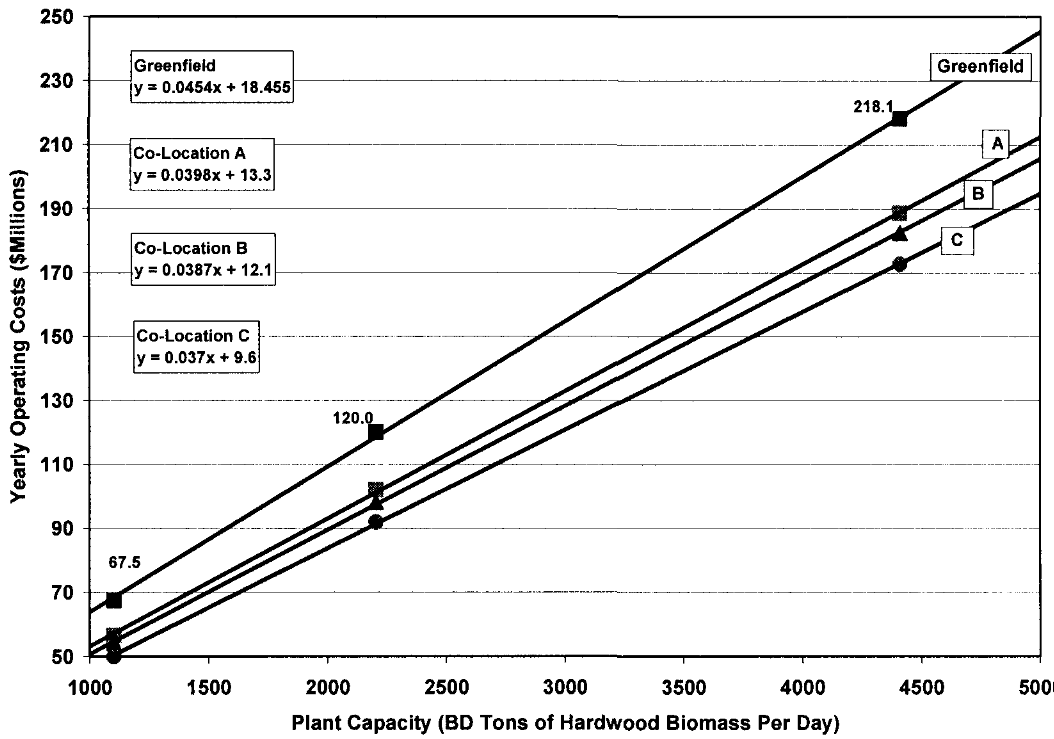


Figure 5-5. Yearly Operating Cost vs. Plant Capacity for All Economic Cases

Results of Profitability Analysis for the Greenfield Plant

Net Present Value Method (Greenfield)

Figure 5-6 summarizes the profitability for the Greenfield plant cases when analyzed using the Net Present Value (NPV) method shown in Equation (4-25). The results are summarized in Figure 5-6 and correspond to an ethanol selling price of $2.50 per gallon, a wood cost of $60 per bone dry ton, and a discount rate of 10%; which is the cost of capital to the investing organization. A positive Net Present Value at the end of the 13th year means that the project returns money over the life of the project. It should be noted that a 10% cost of capital (i_C) is not a particularly difficult hurdle to meet.

Figure 5-6 illustrates the cash flows by year over the entire life of the project. The first three years have negative cash flows and correspond to the period that the capital investment is being made. The cash flow values over the next ten years represent the after tax profits plus the value of the land and working capital, which are fully recovered in year thirteen (13) of the investment. Clearly the white wood to ethanol project will not meet the minimum criteria for a positive Net Present Value (NPV) at the end of the life of the project no matter what the plant size considered. The small- , base-case, and large-plant sizes fail the NPV criteria for the case of $2.50 per gallon of ethanol, $60 per ton white wood and a discount rate of 10%.

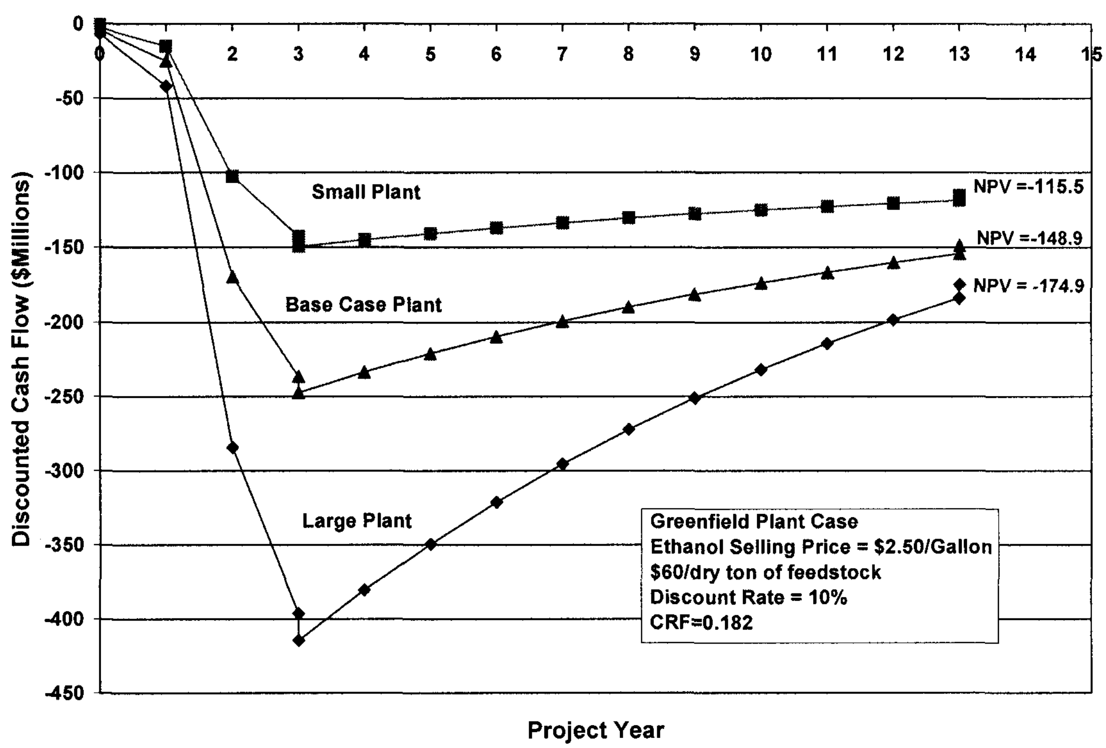


Figure 5-6. Net Present Value at a Discount Factor of 10% (Greenfield Plant)

Discounted Cash Flow Rate of Return (Greenfield)

The second economic barometer used in the profitability analysis was the Discounted Cash Flow Rate of Return (DCFROR) given in Equation (3-27). Unfortunately, the DCFROR method could not be used for the Greenfield plant case, because there is no discount rate that causes the investments to break even. The NPV for each plant size is below zero even when the discount factor is set equal to zero.

Effect of Ethanol Selling Price on Profitability of the Greenfield Plant

The selling price of ethanol has been changing erratically over the last year (see Figure 5-7). The graph shows the price of ethanol has fluctuated from $2.20 per gallon in January of 2006, to $4.00 per gallon in July of 2006, and back down as low as $1.70 per gallon in September 2006. This volatile price reflects supply and demand as well as some speculation.

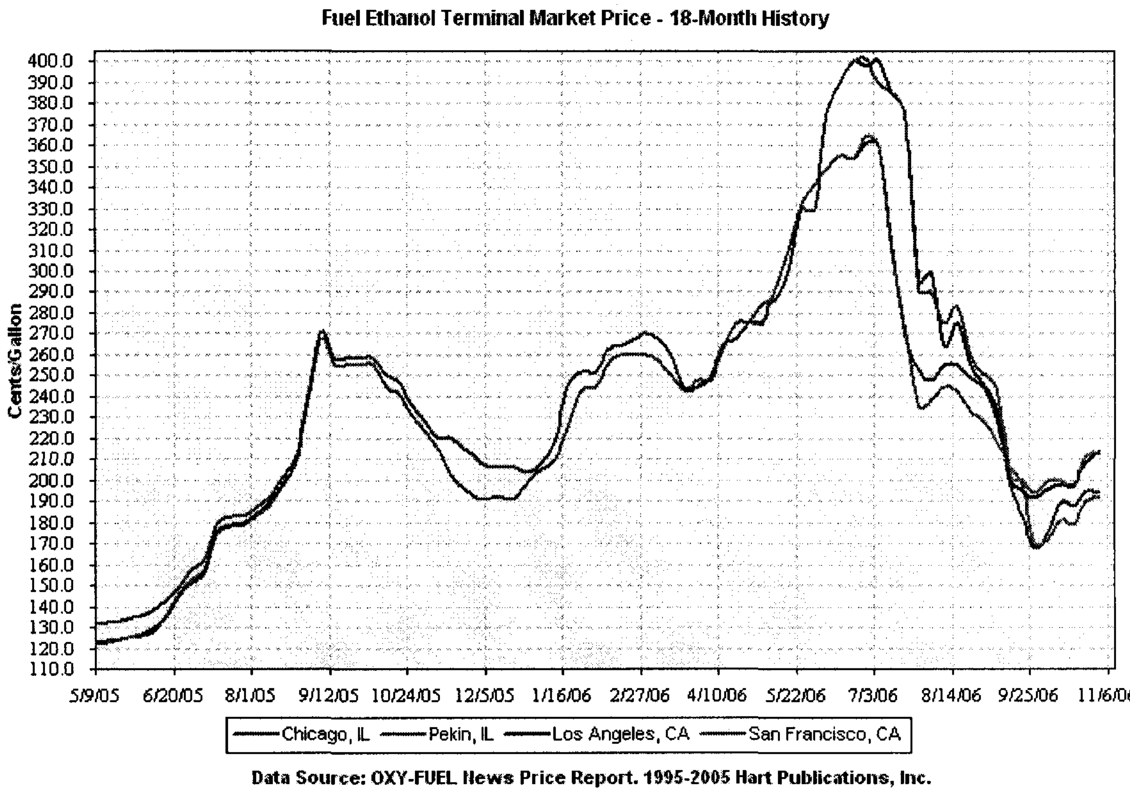


Figure 5-7. Fuel Ethanol Price Market in the Past 18 Months.
(California Energy Commission, 2006)

Figure 5-8 shows the net present value (NPV) for the Greenfield plant case at the 52 million gallon per year plant size (base case) assuming various selling prices for ethanol. The purpose of the graph is to clearly show the impact of the ethanol selling price on the profitability of a white wood to ethanol project. Similar graphs for the large plant and small plant can be seen in Figures F-1 and F-2. All of the prices examined in these graphs fall within the range of the selling price of ethanol during the period of May 2005 to November 2006 (see Figure 5-7). Since ethanol is the major product from the investment, the selling price of ethanol has a major impact on the economic viability of the Greenfield white wood to ethanol project.

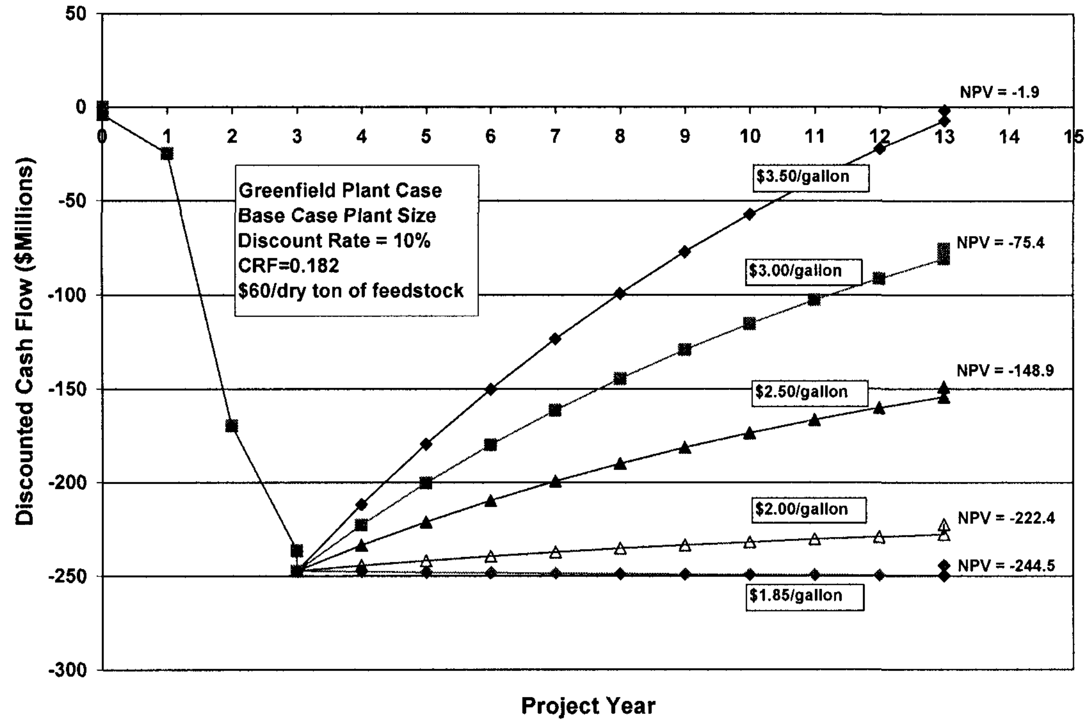


Figure 5-8. Effect of Ethanol Selling Price on Net Present Value
for the Greenfield Plant (2,205 BD Tons/Day or 52 Million Gallons Per Year Ethanol)

For the base case Greenfield plant (52 million gallons per year) the NPV ranges from negative -$244.5 million dollars at the minimum selling price of $1.85/gallon to approximately zero, actually $-1.9 million dollars, when the selling price of ethanol is assumed to be $3.50/gallon. For the Greenfield case, the only situation where altering the ethanol selling price leads to a positive NPV is for the 104 million gallon per year plant capacity at a selling price of $3.50 per gallon (see Figure F-1). For this case the net present value after the thirteen year period would be a positive $119.1 million dollars. This assumes that the selling price of ethanol averages $3.50 over the investment period and there is sufficient wood (4,410 tons per day) available to operate the plant on a sustainable basis. Similar analyses were performed for Co-Location Case B (Figures F-11, F-12, and F-13) and Case C (Figures F-19, F-20, and F-21).

Effect of Wood Cost on Profitability of the Greenfield Plant

A sensitivity analysis was performed to determine the effect of changing the cost of wood on the overall profitability of the Greenfield hardwood to ethanol plant. Figure 5-9 shows the result of changing wood cost for the base case size of a Greenfield plant. The selling price of ethanol was assumed to be \$2.50/gallon, the cost of capital used in estimating factors was set equal to 10%, and the cost of wood was evaluated in ten dollar increments over the range between \$30 and \$60 per dry ton of wood.

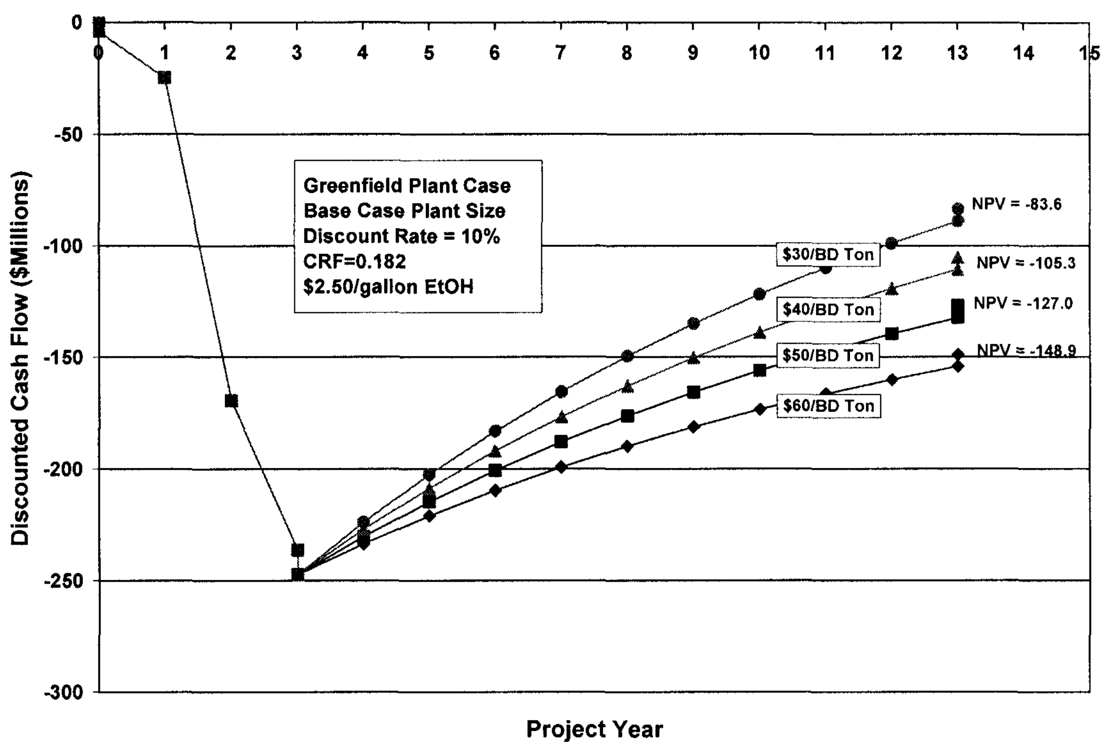


Figure 5-9. Effect of Hardwood Cost on the Net Present Value (NPV) for Greefield Plant (2,205 BD Tons/Day)

The Net Present Value for the hardwood to ethanol project is significantly affected by the price of the wood. In the case shown in Figure 5-9, the NPV at the end of the project increases by approximately $21.7 million dollars for every ten dollar reduction in the cost of wood. The increase in Net Present Value for every ten dollar reduction in wood cost ($21.7 million) approximately doubles for the large plant (see Figure F-3) and is approximately cut in half for the small plant (see Figure F-4). However, lowering the wood cost does not lead to a positive Net Present Value for the Greenfield plant no matter which size plant is selected when the selling price of ethanol is assumed to be $2.50 per gallon and the cost of capital is 10%. Graphs for additional cases can be found in Appendix F. Sensitivity analyses predicated on the cost of white wood are shown for the small and large plant sizes for a Greenfield plant (see Figures F-3 and F-4). Similarly the Co-location cases are illustrated in Figures F-14 through F-16 for Co-location Case B and Figures F-22 through F-24 for Case C.

Profitability Analysis for Co-Location Cases

Net Present Value Method

The profitability of each co-location case was measured using the same methods as the Greenfield case. Co-Location Case A involved citing the plant at a pulp mill with existing boiler and turbo-generation facilities (Section A800). Figure 5-10 illustrates the Net Present Value for hardwood to ethanol plants operating under the assumptions associated with Co-Location Case A.

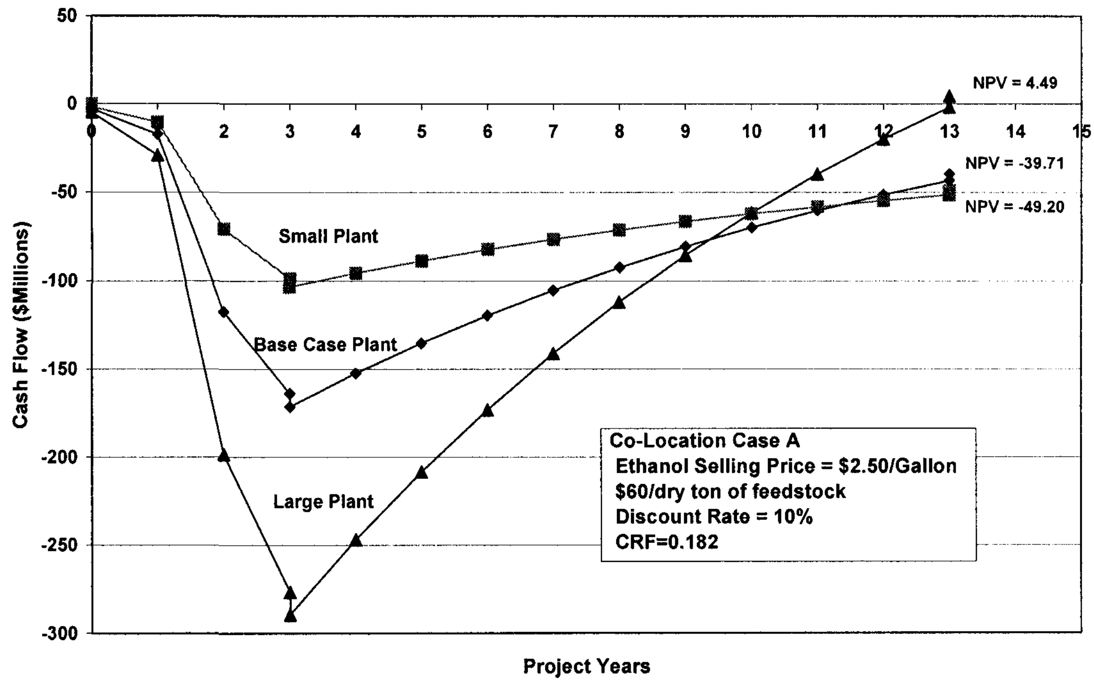


Figure 5-10. Net Present Value at a Discount Rate of 10% (Co-Location Case A)

Comparing Figure 5-10 to the Net Present Value analysis for the Greenfield plant (Figure 5-6) shows that siting the ethanol plant at an existing pulp mill (Co-Location Case A) leads to improved economics; although none is acceptable using the investment criteria set forth previously. For Co-location Case A the small plant is the least profitable venture with a Net Present Value of -$49.2 million dollars and the large plant is the most profitable with a NPV of $4.5 million dollars. The base case plant was slightly better than the small plant with a NPV of -$39.7 million dollars. These figures are still not promising, especially considering the magnitude of the capital investment, but they do show a marked improvement over the Greenfield case.

Discounted Cash Flow Rate of Return

The Discounted Cash Flow Rate of Return (DCFROR) method was also used as an economic barometer for Co-Location Case A (see Figure 5-11). For the small plant, there was no discount rate that caused the project to break even. At a discount rate of zero the small plant showed a loss of $7.6 million dollars at the end of the project. The base case had a DCFROR of approximately 5.2%. As discussed in Chapter 4, if the DCRFOR is below 10% the project is not deemed acceptable. The large plant showed the most promising results at a DCFROR of about 10.3%. This is marginally greater than the 10%, which is the cost of capital and therefore the project would be acceptable; although an investment of this magnitude would, most likely, not be considered acceptable if the substantial risk is considered.

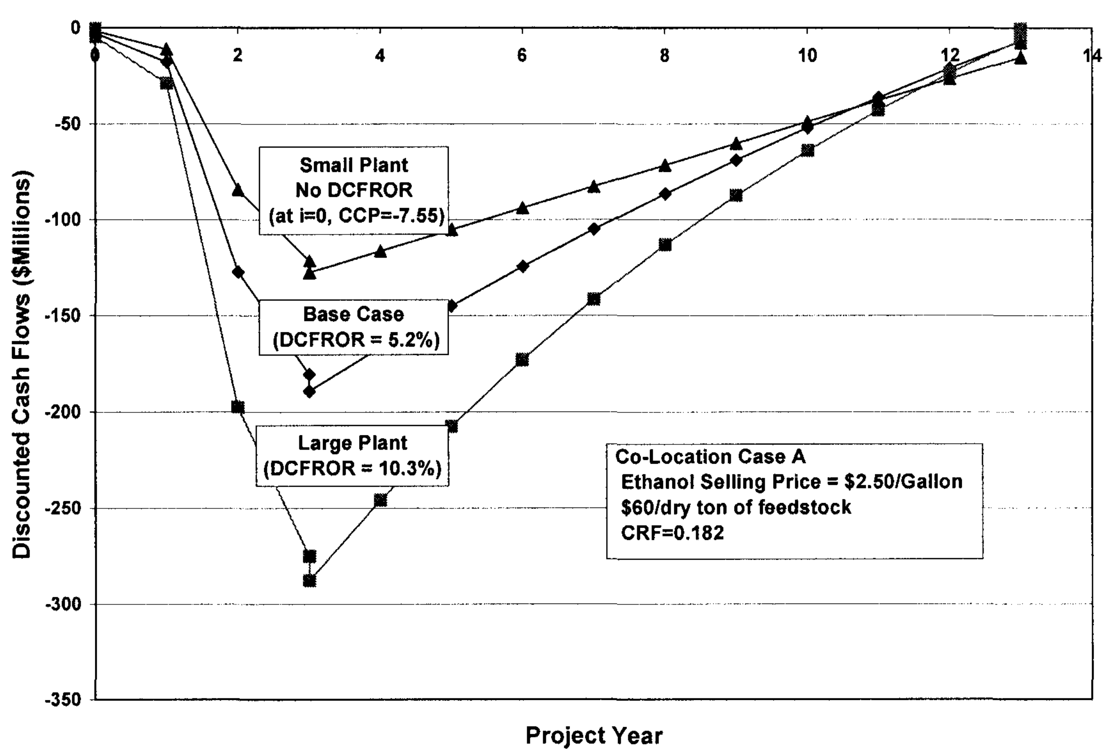


Figure 5-11. Discounted Cash Flow Rate of Return for Co-Location Case A (Citing Plant at Pulp Mill with Existing Power Generation Capability)

Effect of Ethanol Selling Price on Profitability

The effect of the ethanol selling price on the profitability of Co-Location Case A is displayed in Figure 5-12 for the 52 million gallon per year ethanol plant (base case). Analogous graphs for the large plant and small plant are shown in Figures F-5 and F-6 in Appendix F. For the base case plant size (see Figure 5-12) the NPV is negative for all prices except the cases where the ethanol selling price is $3.00 and $3.50 per gallon. The Co-location project A would break even after about five years of plant operation following startup for the 52 million gallon per year plant (base case size) assuming the ethanol selling price is $3.50 per gallon. The NPV after ten years of operation is estimated to be approximately $107.3 million dollars for Co-location Case A. For the 104 million gallon per year large plant (see Figure F-5) the project would break even just prior to the fourth year of operation assuming that ethanol sells for $3.50 per gallon. In this case the final NPV is estimated to be $298.5 million dollars. Both these cases are attractive provided the ethanol could be sold for $3.50 per gallon and the plants are sustainable with regards to wood.

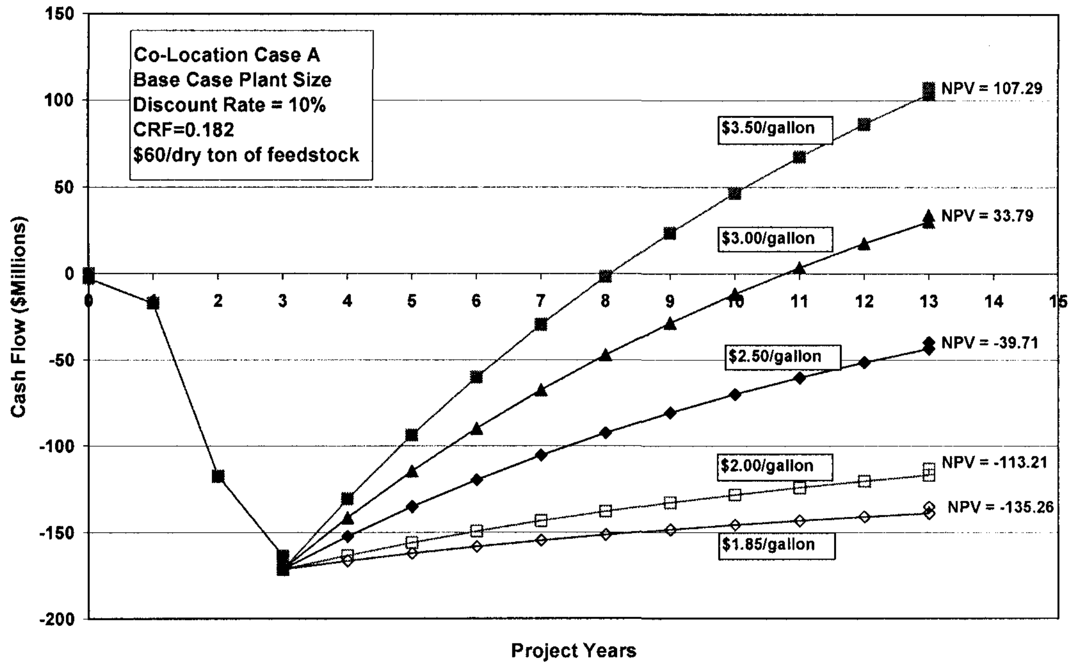


Figure 5-12. Effect of Ethanol Selling Prices on NPV for Co-Location Case A (Citing Plant at Pulp Mill with Existing Power Generation Capability)

The 26 million gallon per year plant (Figure F-6) fails to break even assuming every selling price except for the case where ethanol sells for $3.50 per gallon; in which case the NPV is estimated to be $24.3 million dollars.

Effect of Wood Cost on Profitability

The effect of the cost of wood on the profitability of Co-Location Case A at the base case size is shown in Figure 5-13. The corresponding graphs for the large and small plants are shown in Figures F-7 and F-8 in Appendix F. The base case plant size for Co-Location Case A shows a positive NPV when wood is purchased at $30 and $40 per dry ton. At $30 per ton the NPV is $25.6 million dollars, which is a significant improvement from the NPV of -$82.7 million dollars for the Greenfield plant with the same wood cost.

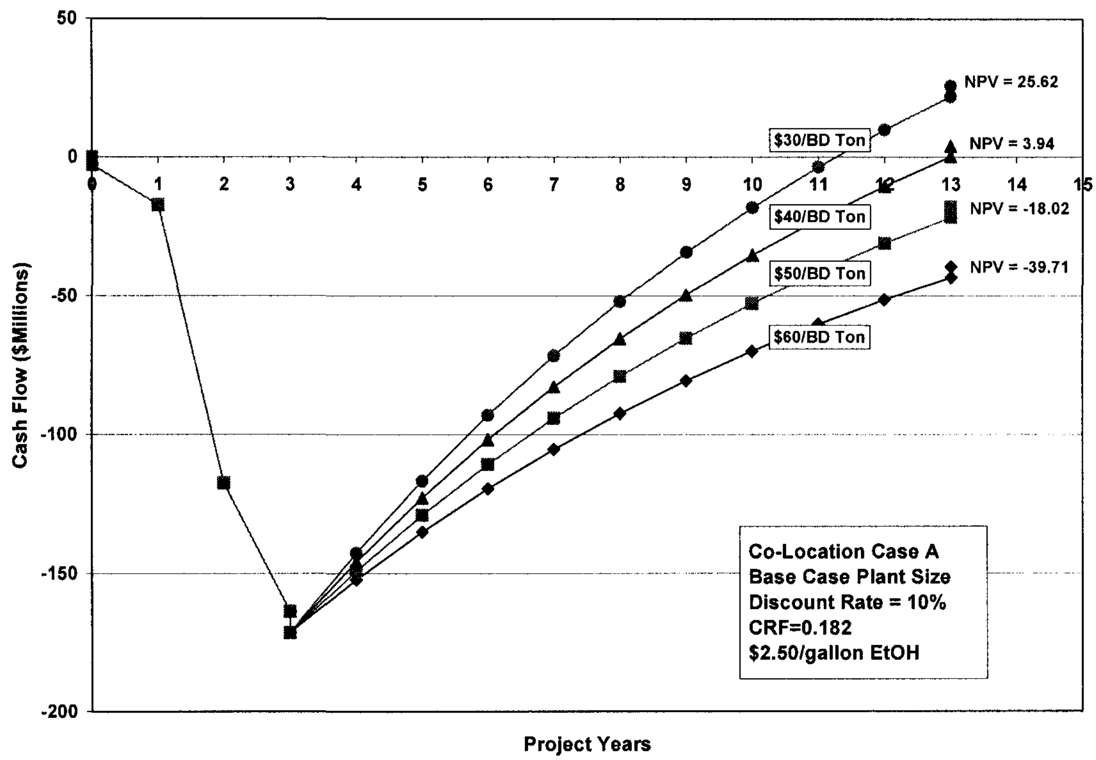


Figure 5-13. Net Present Value at Varying Hardwood Cost for Co-Location Case A (Plant Size of 2,205 BD Tons/Day or 52 Million Gallons per Year Ethanol)

The NPV for a large plant (Figure F-7) is positive for all the wood prices evaluated, with a maximum NPV of $135.2 million dollars when wood costs $30 per dry ton. There is no wood price that enables the small plant (Figure F-8) to become profitable.

Net Present Value Analysis for All Cases

The effect of co-location on the NPV of each plant case at the base case size is shown in Figure 5-14. The purpose of the graph is to illustrate the magnitude of the calculated Net Present Value for the various co-location cases. For purpose of comparison the Greenfield case is also illustrated. For the 52 million gallon per year (base case) plant size the NPV improves from -$148.9 million dollars for the Greenfield plant to $30.7 million for Co-Location Case C, a difference of $178.5 million dollars.

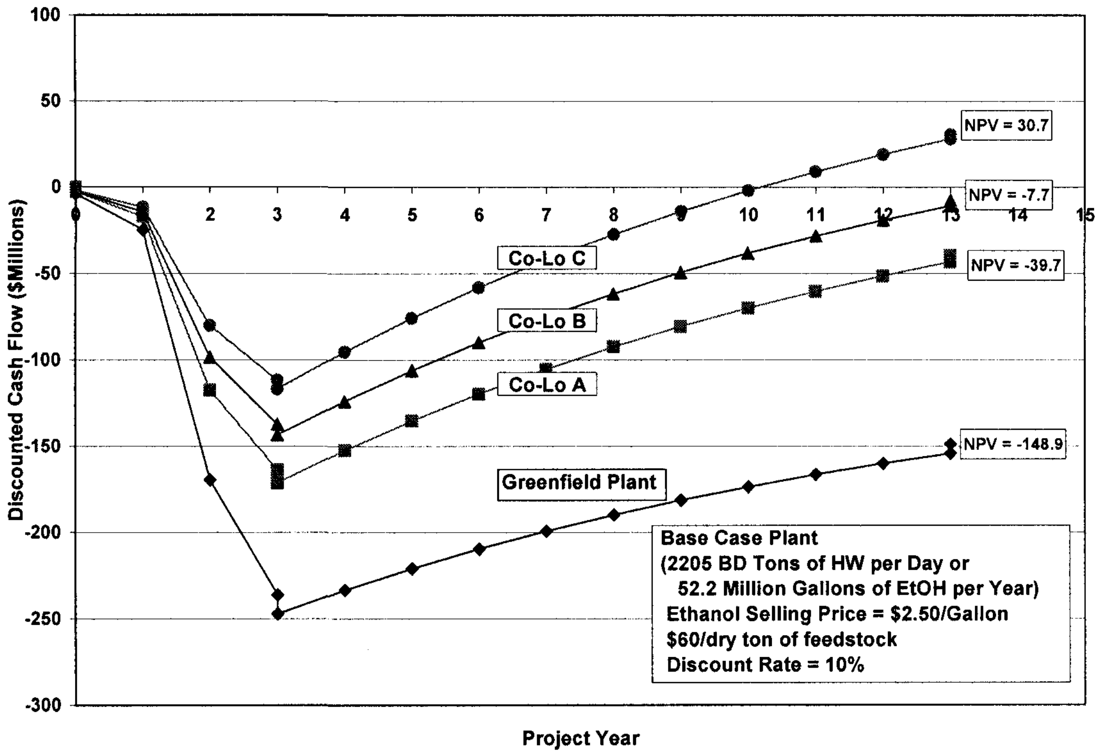


Figure 5-14. Effect of Co-Location on the NPV for All Plant Cases at Base Case Size (2,205 BD Tons/Day and 52 Million Gallon per Year Ethanol)

The NPV analysis for all plant sizes can be seen in Figure 5-4 for the Greenfield plant, Figure 5-8 for Co-Location Case A, Figure F-9 for Co-Location Case B, and Figure F-17 for Co-Location Case C. Perhaps a better way to quantify the effects of co-location is a summary of the Net Present Value at the end of year 13 for all plant cases as a function of plant size (see Figure 5-15). The results summarized in Figure 5-15 correspond to an ethanol selling price of $2.50 per gallon, a wood cost of $60 per bone dry ton, and a discount rate of 10%. When looking at this graph it is also important to remember that the capital investment for all cases is quite large. The Greenfield plant and Co-Location Case A do not show any potential to be profitable for the conditions investigated in the current study. Co-Location Case B only begins to approach an acceptable investment for the 104 million gallon per year plant size; where the NPV is $60.4 million dollars. Co-Location Case C is by far the most attractive option due to the significant reduction in the capital requirements. This is the only case in which the 52 million gallons per year plant size results in a positive net present value; NPV equal to $30.7 million dollars. The 104 million gallon per year plant size for Case C is the most profitable case with a NPV of $119.2 million dollars.

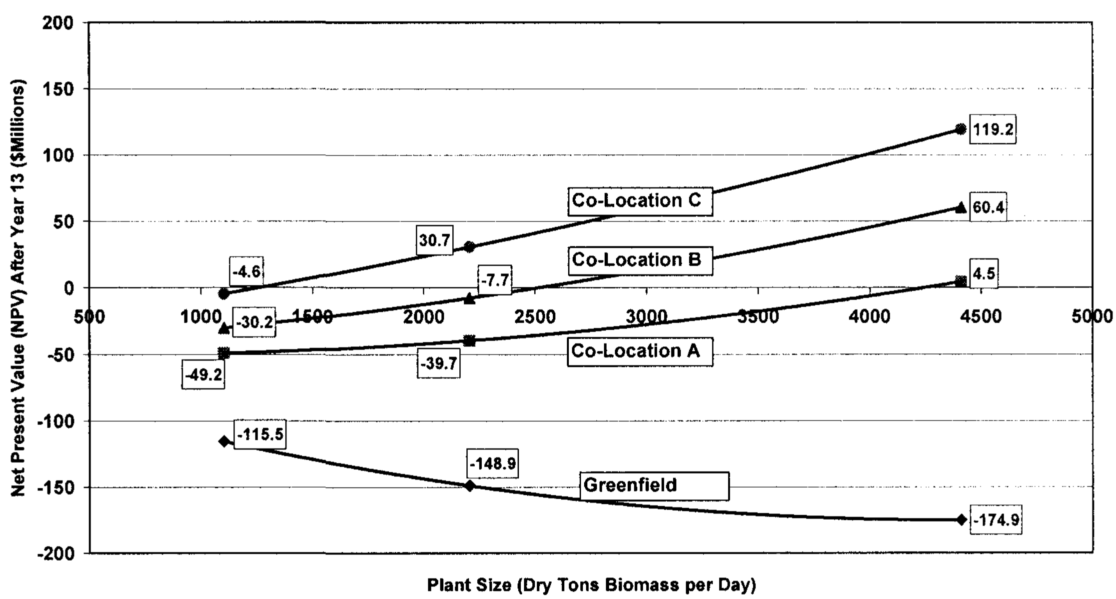


Figure 5-15. Net Present Value as a Function of Plant Size for All Cases (EtOH Selling Price = $2.50/gallon, Wood Cost = $60/BD Ton, Discount Factor =10%)

Discounted Cash Flow Rate of Return for All Cases

The Discounted Cash Flow Rate of Return (DCFROR) was compared for all co-location cases. This analysis is presented in Figure 5-16 as a function of the biomass input to the plant. The shaded region of the graph represents all DCFROR values that fall below the minimum acceptable rate of return, which is 10%. All cases within the shaded region are unacceptable.

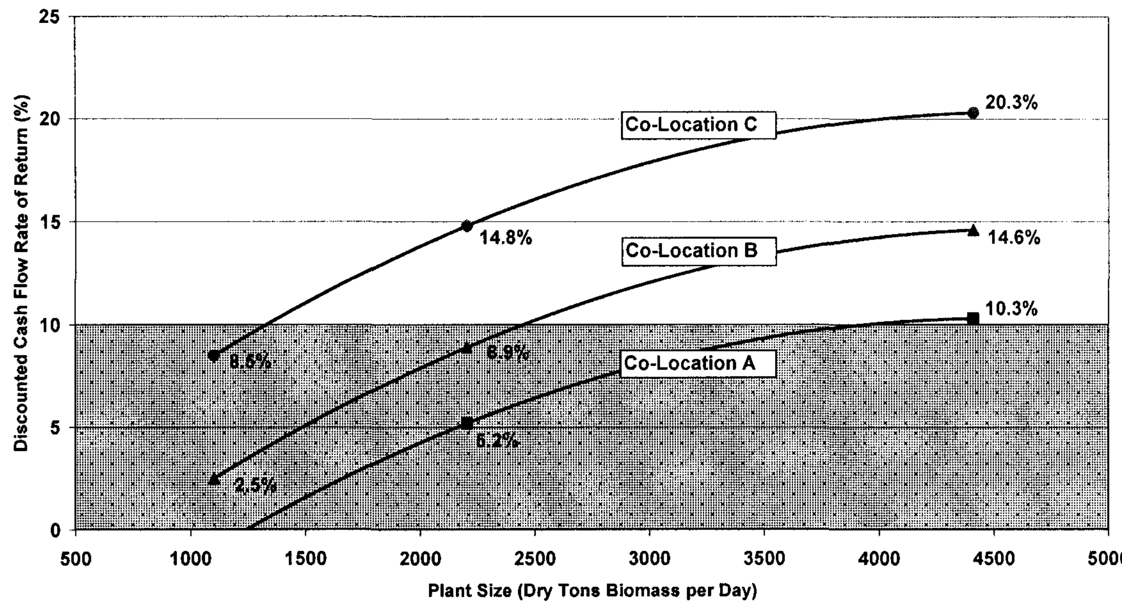


Figure 5-16. DCFROR as a Function of Plant Size for All Cases
(EtOH Selling Price = $2.50/gallon, Wood Cost = $60/BD Ton, Discount Factor =10%)

Co-Location Case A reaches an acceptable DCFROR above 10% only for the 104 million gallon per year plant size, and the value of 10.3% as mentioned previously is probably not worth the risk associated with the capital investment. The graph shows that in order for Co-Location Case B to be acceptable, the plant would have to be larger than the 52 million gallons per year (base case), where the DCFROR is 8.9%. Co-Location Case C showed much more attractive results, but the small plant still failed to meet the conditions of an acceptable investment. The DCFROR for the 52 million gallon per year plant size (base case) for Co-location Case C is approximately 14.8%

and meets the criteria of an acceptable investment. The 104 million gallon per year plant size resulted in a DCFROR of 20.3%; which clearly meets the investment criteria.

<u>Effect of Capital Recovery Factor on Profitability</u>

The profitability of a 52 million gallon per year plant was analyzed over a range of capital recovery factors (CRF) from 0.1 to 0.22.

CRF = 0.1 Equivalent to money borrowed at 0% (i) over a ten year period.

CRF = 0.14 Equivalent to money borrowed at 6.6% (i) over a ten year period.

CRF = 0.182 Equivalent to money borrowed at 12.7% (i) over a ten year period.

CRF = 0.22 Equivalent to money borrowed at 17.7% (i) over a ten year period.

Figure 5-17 summarizes the effect of altering the capital recovery factor on the Discounted Cash Flow Rate of Return for the base case size of all plant cases studied. The results correspond to an ethanol selling price of $2.50 per gallon, wood cost of $60 per dry ton and a discount factor of 10%. The shaded area in Figure 5-17 represents all DCFROR values that fall below the minimum acceptable rate of return (10%). The equations on the graph for each co-location case can be used to estimate the DCFROR for any given capital recovery factor within the range evaluated.

The CRF in all previous cases was assumed to be 0.182; and only Co-Location Case C, the 52 million gallon per year plant size, was able to meet the DCFROR profitability criteria of 10% (i_C). All co-location cases meet the DCFROR profitability criteria when the CRF is equal to 0.1. However, a CRF of 0.1 over ten years corresponds to an interest rate of zero (0%) and is highly unlikely. With a CRF of 0.14, Case B and Case C were the only profitable projects and at a CRF of 0.22, Case C was the only project able to exceed the DCFROR criteria of 10%. The Greenfield plant did not meet profitability criteria for any CRF.

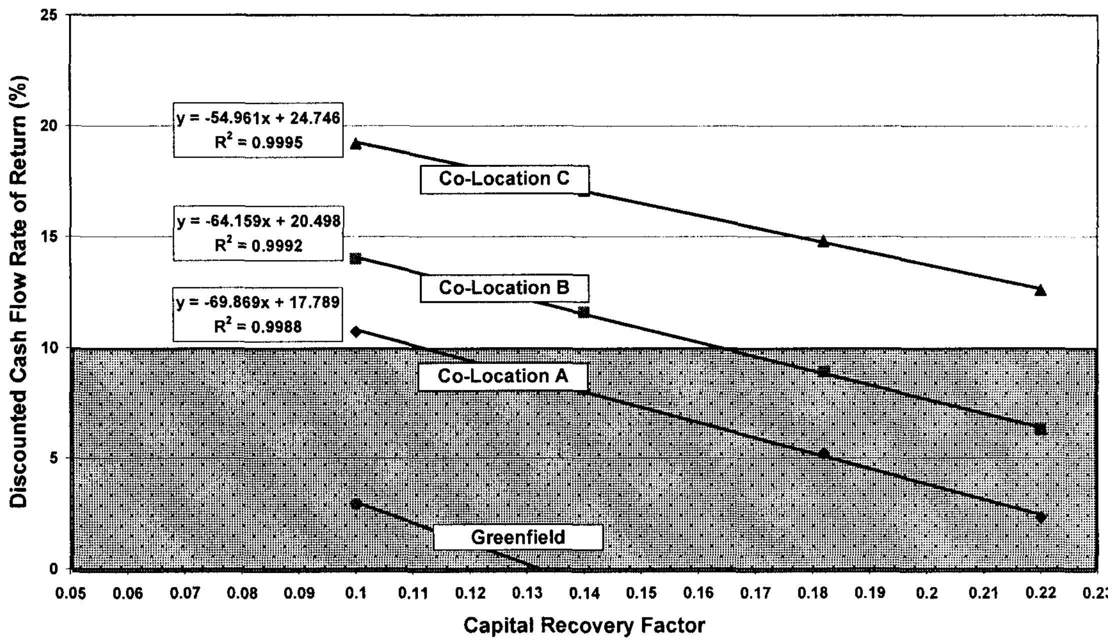


Figure 5-17. Effect of Capital Recovery Factor on the DCFROR for all Plant Cases (Base Case Plant Size, 52 Million Gallons per Year)

Break Even Analysis

The results from nearly every case studied showed that the 104 million gallons per year plant size has a much better chance of being economically viable than either the 52 million gallon per year (base case) or the 26 million gallon per year (small size) plant sizes. With each step in the Co-Location study, the capital investment is decreased and therefore the likelihood of meeting criteria for an acceptable investment increases. Figure 5-18 shows the ethanol selling price required for the net present value to go to zero or break even at the end of the investment, for the four cases under consideration. The price of wood for all cases is $60 per dry ton and the discount factor is equal to 10%. The graph provides a graphical representation of what happens to the economic feasibility when the plant size or capital investment changes. The best case scenario is a 104 million gallon per year plant size (large plant) under the conditions of Co-Location Case C which requires a selling price of $2.10 per gallon. The worst case is the 26 million gallon per year (small size) Greenfield plant which requires a selling price of $4.07 per gallon. The best Greenfield case is the 104 million gallon per year plant size (large plant) which requires a selling price of $3.10 per gallon.

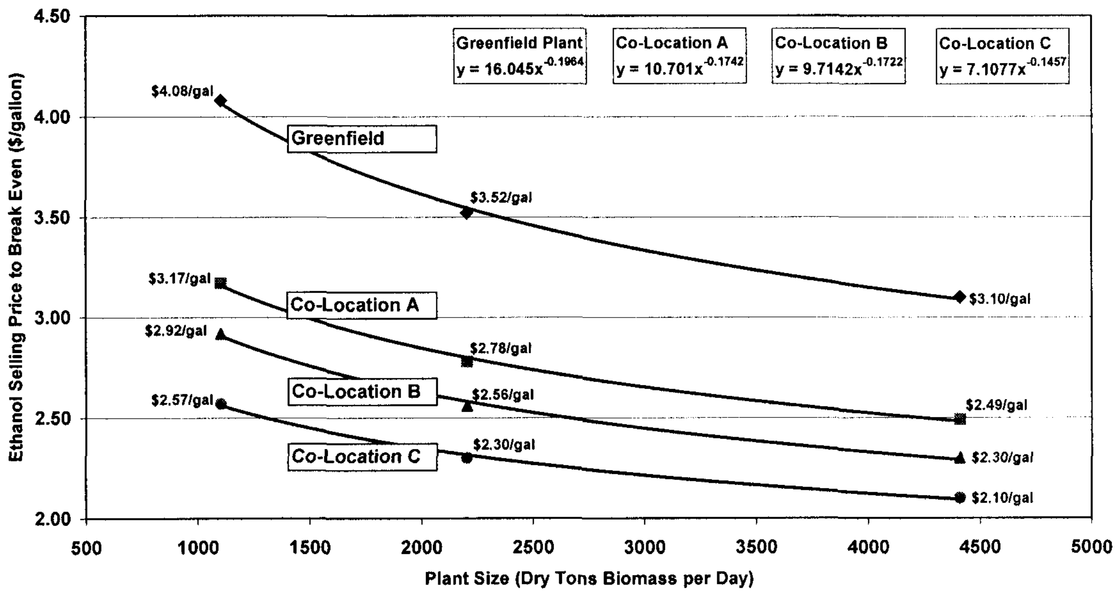


Figure 5-18. Breakeven Selling Price vs. Plant Size for All Economic Cases
(Wood Cost = $60/BD Ton, Discount Factor =10%)

Thermal Efficiency

The thermal efficiency ($\eta_{Thermal}$) of a white wood to ethanol plant of the base case size was estimated from the wood use (m_{wood}), the higher heating value for the fuel oil (HHV_{FO}), the wood (HHV_{wood}) and ethanol (HHV_{EtOH}) and the production rate for ethanol (P_{EtOH}), the amount of electricity sold into the grid (P_{Elec}) and the mass flow rate of fuel oil (m_{FO}) consumed in the process.

$$\eta_{Thermal}(\%) = \left[\frac{P_{EtOH} * HHV_{EtOH} + P_{Elec}}{m_{Wood} * HHV_{wood} + m_{FO} * HHV_{FO}} \right] * 100$$

This calculation is summarized in Table 5-5. Thermal efficiency is a measure of how much internal energy from the raw materials is retained in the final products. The ethanol accounts for approximately 93.4% of the energy content of the products and electricity accounts for the remaining 6.6%. The thermal efficiency was calculated by dividing the total energy content of the products (ethanol and electricity) by the total energy content of the raw material fuels (white wood and diesel). The NREL process results in a thermal efficiency of approximately 36%. In other words, 64% of the stored energy in the raw material fuels is lost. This thermal efficiency is approximately the same as what is obtained in a large coal burning power station where coal is burned and the energy in the coal is converted to electrical energy. In the current analysis, the thermal efficiency would also be approximately 36% both for the large plant and small plant since the amount of raw materials used in the process were scaled linearly.

Table 5-5
Thermal Efficiency Summary for the Base Case Plant Size

	Item	**Ammount**	**HHV (BTU/lb)**	**Total Energy Content (BTU) (One year of operation or 350.25 days)**
Raw Materials	White Wood	2205 dry tons/day	8,384	1.295 x 10^13
	Diesel	443 kg/day	19,676	1.6153 x 10^11
Products	Ethanol	52.2 million gallons per year	12,853	4.4176 x 10^12
	Electricity	10,942 kW	----------	3.1384 x 10^11
$\text{Thermal Efficiency} = \frac{\text{Total Energy Content of Products}}{\text{Total Energy Content of Raw Materials}} = 0.36 \text{ or } 36\%$				

Chapter 6

CONCLUSIONS AND RECOMMENDATIONS

Profitability Indicators

The profitability of the various investment scenarios was evaluated by using the discounted cash flow rate of return on investment (DCFROR) and the net present value method (NPV). Both of these methods essentially gave the same results and proved extremely useful for evaluating the investments under study. It is recommended that in future studies the discounted cash flow rate of return be used as the economic barometer. Essentially both methods are the same except that the discounted cash flow rate of return finds the interest rate at which the net present value goes to zero at the end of the investment. This interest rate can then be simply compared to the cost of capital (i_C) and a decision made.

High Capital Investments for Greenfield Plants

Capital costs for the Greenfield plant for the NREL white wood to ethanol process are quite high. The total cost of installed equipment varied between $102 million dollars for a 26 million gallon per year ethanol plant and $281 million dollars for a 104 million gallon per year ethanol plant. This amounts to a capital investment of 3 and 4 dollars for every one gallon of ethanol capacity. The total capital investment is reduced as the plant size increases and results from economy of scale.

Investment in Boiler and Turbo-generator Facilities

Approximately 31% of the total capital investment results from expenditures on boiler and turbo-generation facilities arising because of the large amount of lignin in the wood and residual products from the ethanol fermentation. This is a major drawback to the NREL process since a sizable fraction of the raw material is going to produce electricity rather than ethanol.

On Site Production of Cellulase Enzyme

Approximately 11% of the total capital investment goes for the production of the cellulase enzyme. It does not appear to be economically attractive for the ethanol producer to also produce the enzyme on site as originally shown in the NREL design. It would be prudent to avoid this capital investment and purchase the enzyme directly. This will decrease capital costs and most likely lead to a more efficient enzyme that will produce higher yields.

Profitability of the Greenfield Plant

The Greenfield case was not profitable, did not meet the investment criteria, at any plant size for ethanol selling prices up to $3.50 per gallon. The cost of capital, the price of ethanol and the cost of wood are the economic drivers that determine the economic viability of a white wood to ethanol investment.

Limiting Factors

An ethanol from white wood plant of the base case size, 52 million gallons per year, under the conditions discussed in this analysis can only be successful under very limited circumstances. There must be large reductions in the capital investment if the process is to be economically attractive; which can be brought about by co-locating the site at an existing pulp mill where portions of the investment are carried by the manufacture of pulp. Also as the cost of wood increases, it becomes harder to meet the investment criteria. Similarly as the price of ethanol decreases and the capital recovery factor increases, the white wood to ethanol process becomes less and less attractive.

Capital Investment for Co-Location Cases

The capital cost can be significantly reduced if the plant can be co-located at an existing mill site where some of the capital charges can be off-loaded by dual usage This concept is limited by the number of sites that are available; and finding partners that are willing to commit their facilities to the production of ethanol. The order of increasing profitability will occur in the order at which the capital investment can be off-loaded; specifically:

Co-Location Case A. Locate Ethanol Plant at an Existing Pulp Mill with Spare Power Generation Capacity.

Co-Location Case B. Locate Ethanol Plant at an Existing Pulp Mill with Spare Power Generation Capacity and also Purchase Cellulase Enzyme.

Co-Location Case C. Locate Ethanol Plant at Existing Pulp Mill with Spare Power Generation Capacity, Waste Treatment and Wood Yard Facilities plus Purchase Cellulase Enzyme

Profitability of the Co-Location Plants

The co-location cases were evaluated based on the investment criteria of a 10% after tax rate of return when measured by the discounted cash flow rate of return method for evaluating potential projects. For each case the ethanol selling price was $2.50 per gallon and the wood cost was $60 per dry ton when determining the DCFROR.

For Co-Location Case A the small plant (26 million gallons per year) and base case plant (52 million gallons per year) did not meet the criteria of a DCFROR of 10%. The large plant (104 million gallons per year) for Case A resulted in a DCFROR of 10.3%, which barely meets the criteria and is most likely not worth the risk.

For Co-Location Case B no case met the investment criteria except for the 104 million gallon per year plant, with a DCFROR of 14.6%.

Co-location C showed the best results. The best small plant investment was in Case C, but the DCFROR still was only 8.5%, which of course did not meet the investment criteria. The base case plant was able to meet the investment criteria and had a DCFROR of 14.8%. The large plant for Case C was the most profitable option with a DCFROR of 20.3%.

The only plant size meeting the criteria for all three co-location cases was the 104 million gallon per year ethanol plant. The large plant size however is subject to the caveat that there is sufficient white wood to meet the demands of the plant. Wood sustainability was not considered in the work reported here.

Economy of Scale

Economy of scale reduces the capital cost per unit of production as the plant size increases. A large plant shows potential for meeting the profitability criteria provided the area surrounding the cite can sustain the wood demands for the plant and if there is sufficient market demand for ethanol. However, this is currently only true when significant capital reduction occurs due to co-location at an existing pulp mill. A small plant would not be feasible under the assumptions of the analysis. The success of the project would rely on major capital reduction, high ethanol prices, and low wood prices. It would be nearly impossible to meet all of these conditions, so the project would most likely be very risky.

Wood Costs

One significant factor that improves the process economics is reduction in the price of wood. As the cost of wood is reduced, then there is a concomitant reduction in the cost of manufacturing ethanol. However, the NREL process is predicated on white wood being used; the highest value raw material. Consequent, it would be very beneficial if the process could be made to work on wood containing bark which would reduce wood cost. One advantage of gasification to syngas and conversion to alcohols via the Fisher- Tropsch process is that low cost wood can be used, biomass 4-inches and less, and potentially some of the lignin in the wood can be used to produce alcohols.

Government Subsidies

If the Federal government thought it was in the best interests of the country to begin producing ethanol from wood rather than corn, then subsidies to ethanol producers would reduce the cost of manufacture and improve the economic viability of the process. However, under this scenario the ethanol producer is at the mercy of the government; a risky business at best.

Wood Supply

One shortcoming of the current study relates to the wood supply and the ability to supply wood to the site on a sustainable basis. It was assumed that the wood required for the white wood to ethanol plant would be available on a sustainable basis and did not affect the process economics. This assumption may not be true. For future analysis it is recommended that representatives from the Maine Forest Service work with the individuals performing the process economics to determine what a realistic wood supply would be for the site of interest.

Updated Process and Cost Data

The current study is predicated on yield and cost information provided in the report by Wooley (1999). In future analysis, it is recommended that a survey be performed, with site visits if necessary, to obtain more accurate information upon which to base the design and technical economic analysis.

Advances in Technology

Advancements are currently being made in the area of genetic engineering that impact the current problem (Steeves, 2006). These researchers are genetically modifying lignin in hybrid poplar, a species of wood which grows quickly in many climates. The goal of this research is to reduce the amount of lignin present in the hybrid poplar and to modify the lignin thus making the cellulose and other carbohydrates more accessible for hydrolysis and fermentation. If research of this type comes to fruition, ethanol yields would improve and less wood would be utilized in the generation of by-product electricity. This would improve the thermal efficiency of the process (estimated to be 36% for the NREL process) and thus improve the process economics.

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APPENDICIES

Appendix A

NREL HARDWOOD TO ETHANOL PLANT SECTION DIAGRAMS

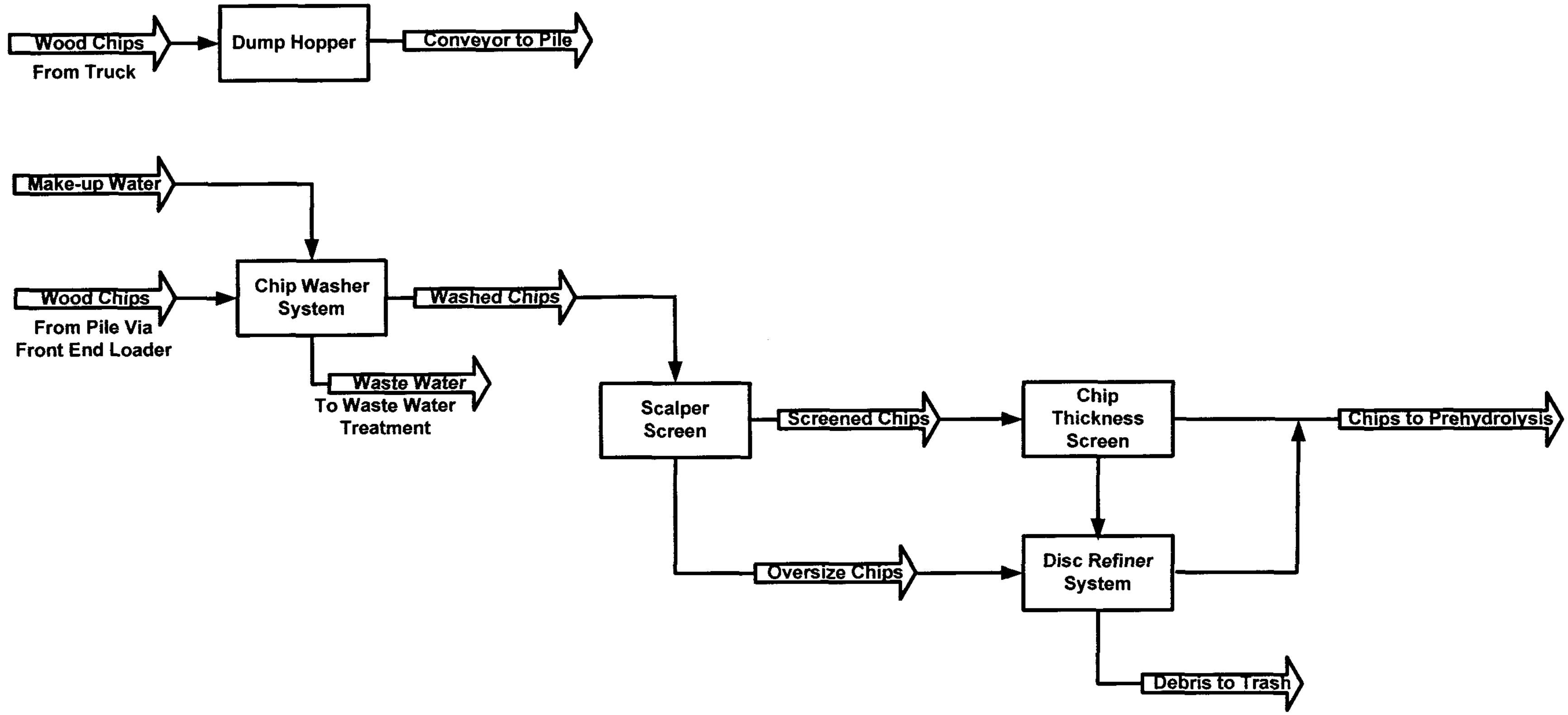

Figure A-1. Feed Handling Flow Diagram (A100)

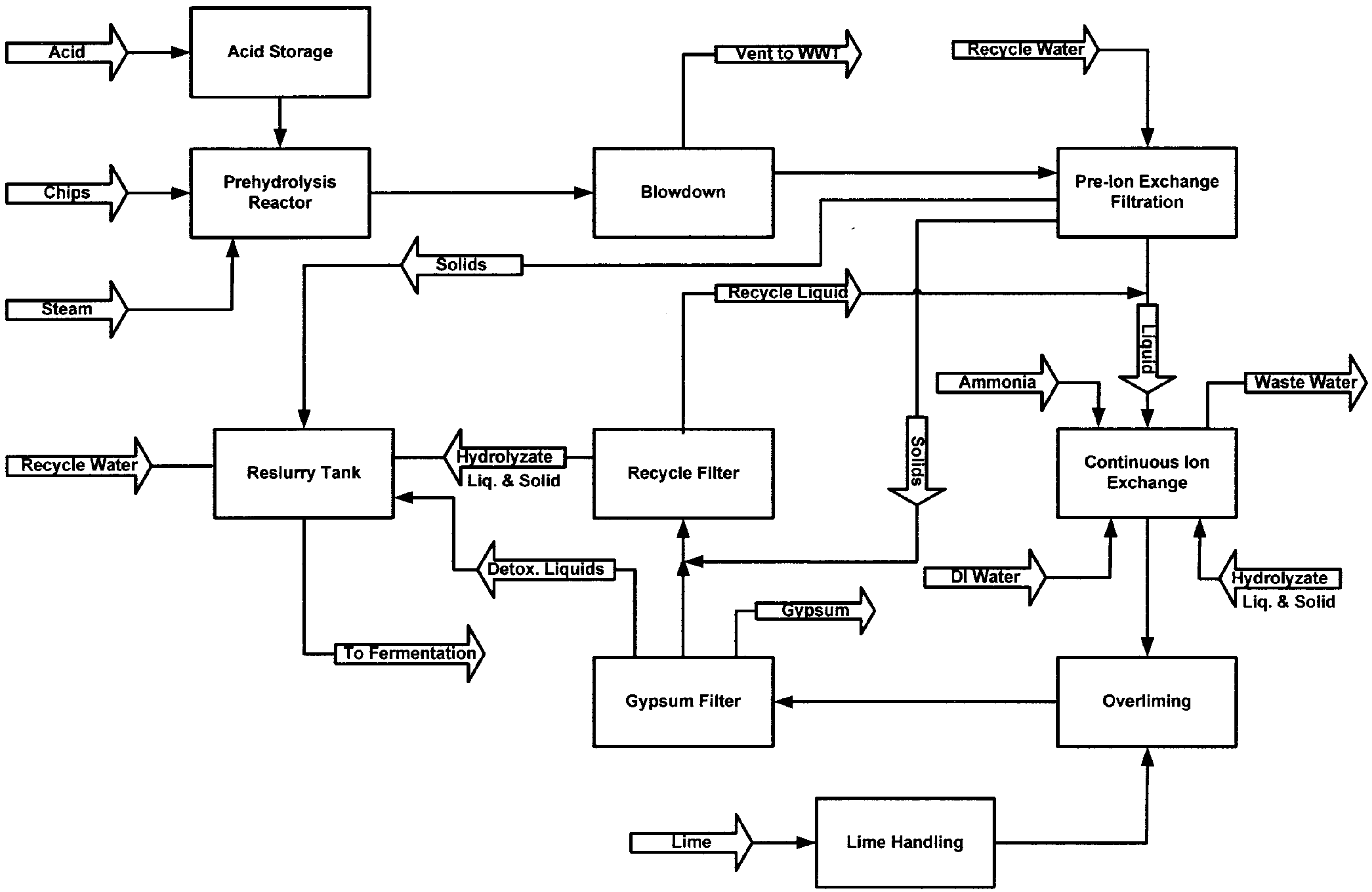

Figure A-2. Pretreatment Flow Diagram (A200)

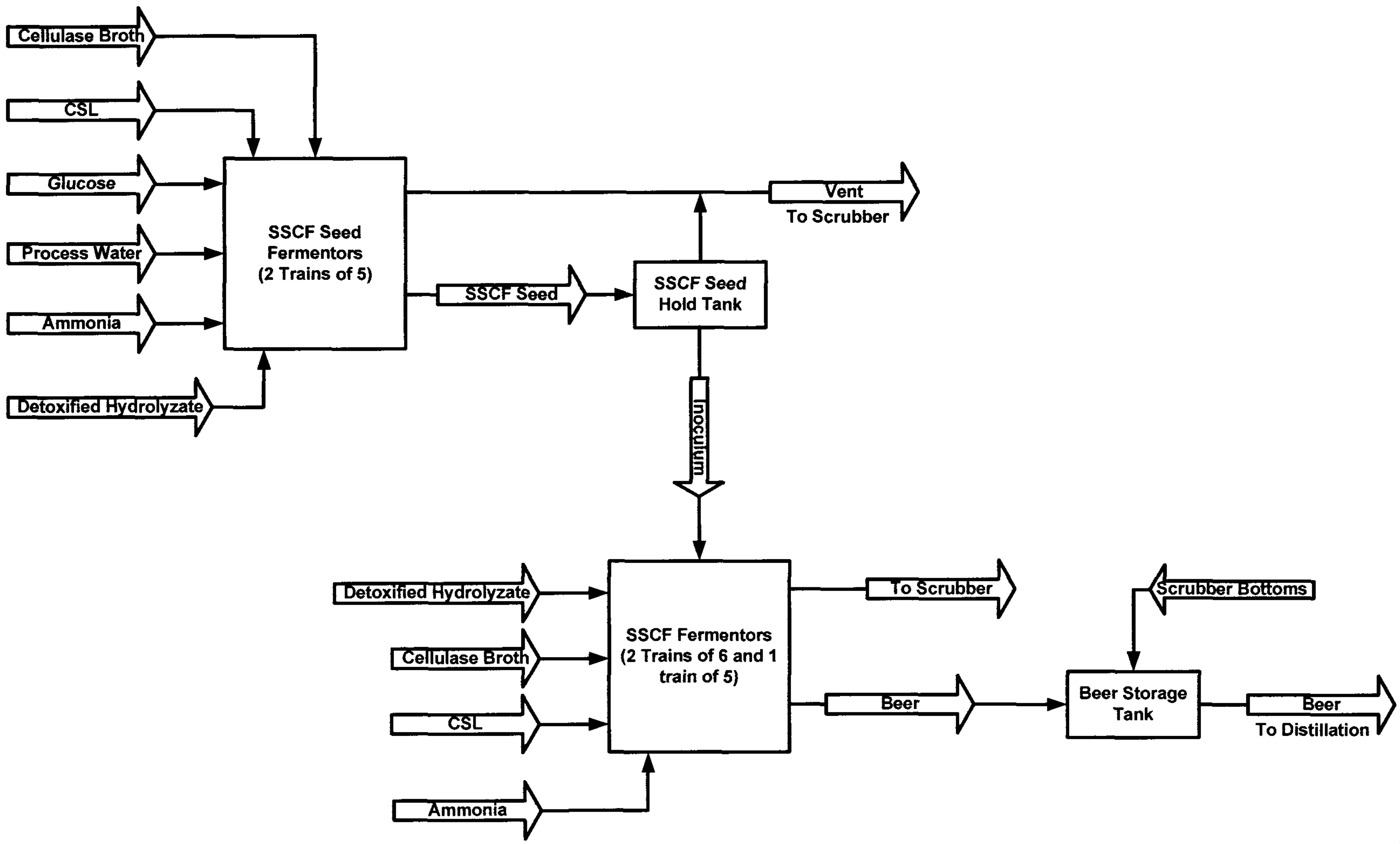

Figure A-3. Fermentation Flow Diagram (A300)

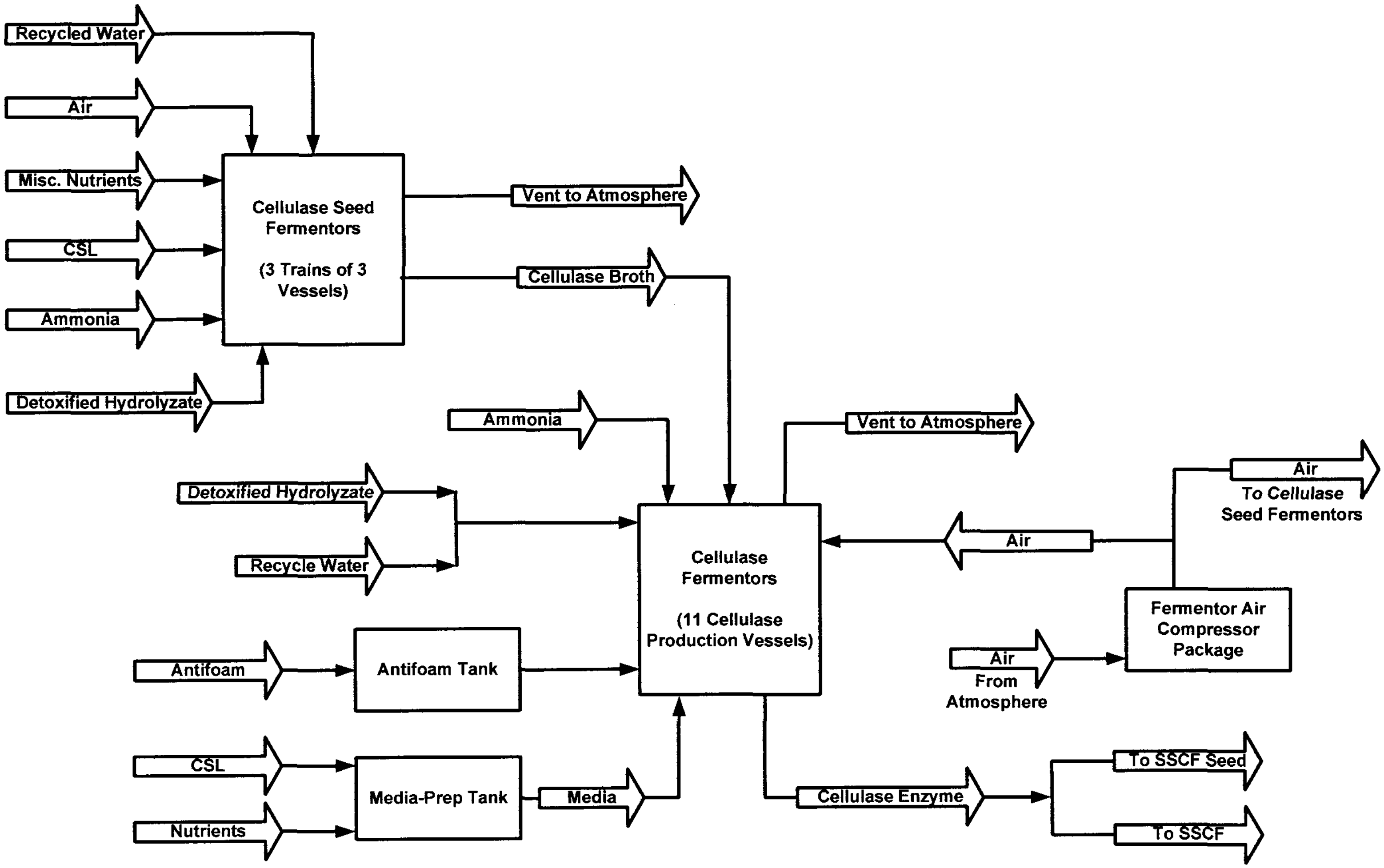

Figure A-4. Cellulase Production Flow Diagram (A400)

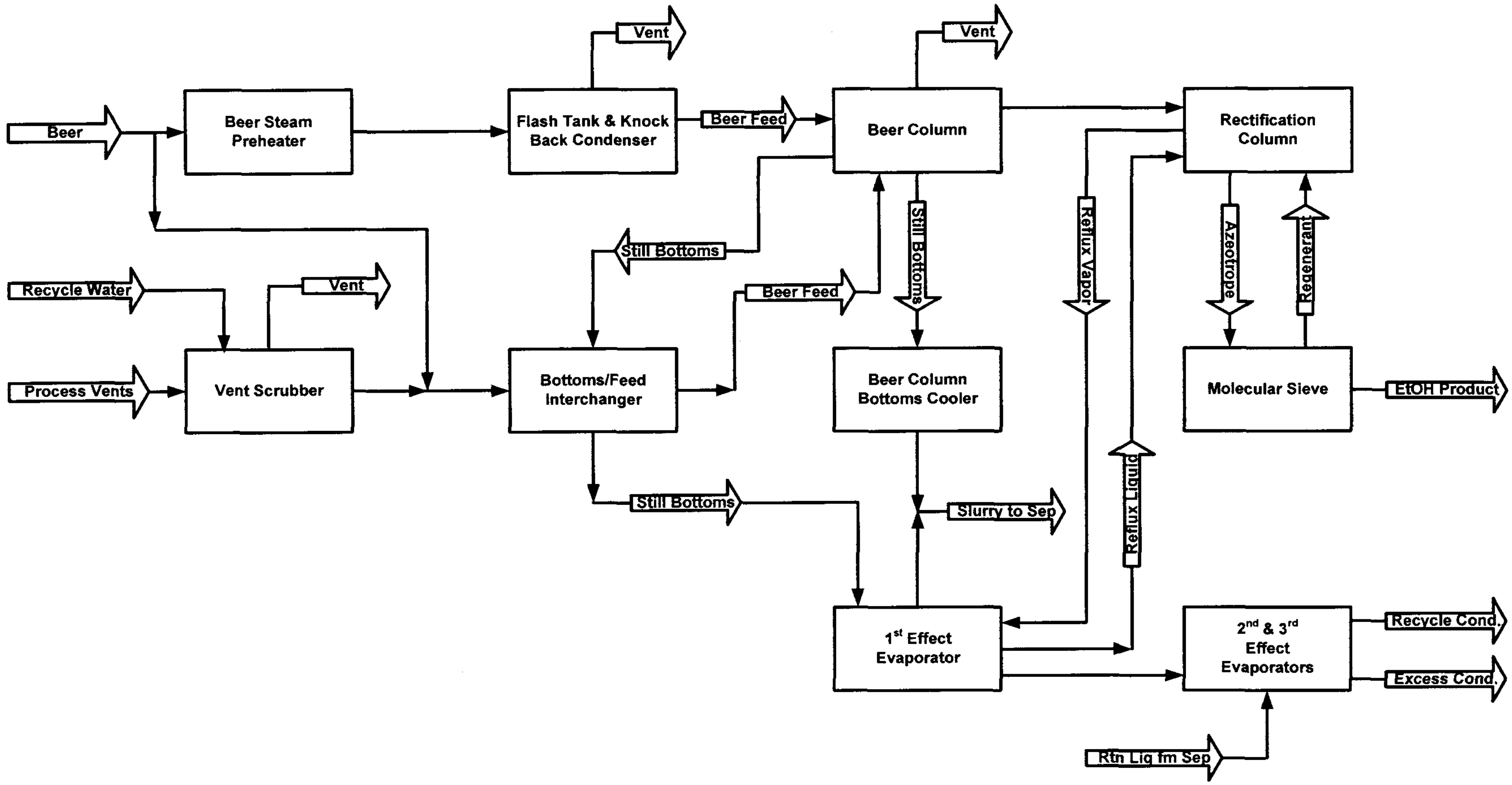

Figure A-5. Distillation, Dehydration, Evaporator and Scrubber Flow Diagram (A500)

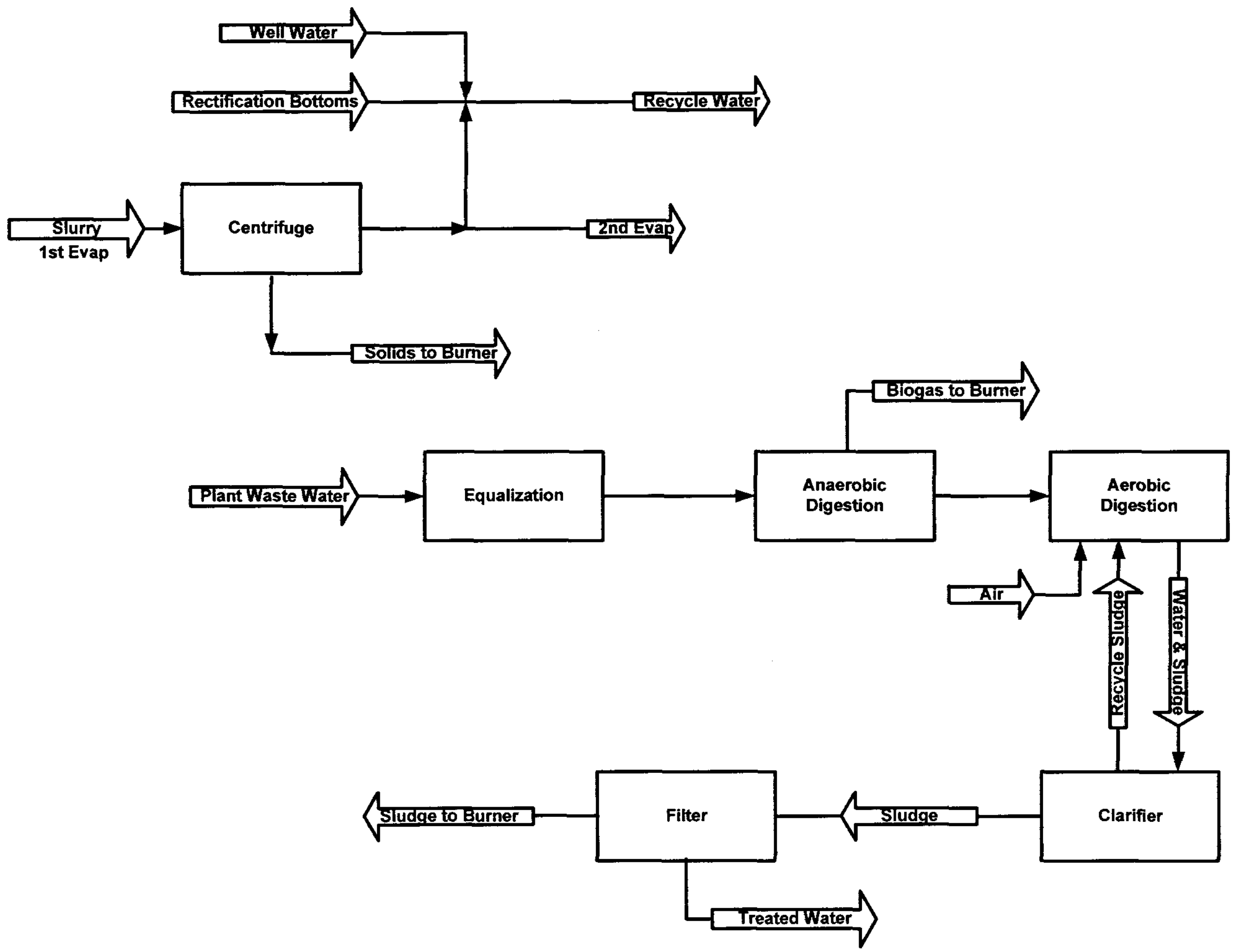

Figure A-6. Waste Water Treatment Flow Diagram (A600)

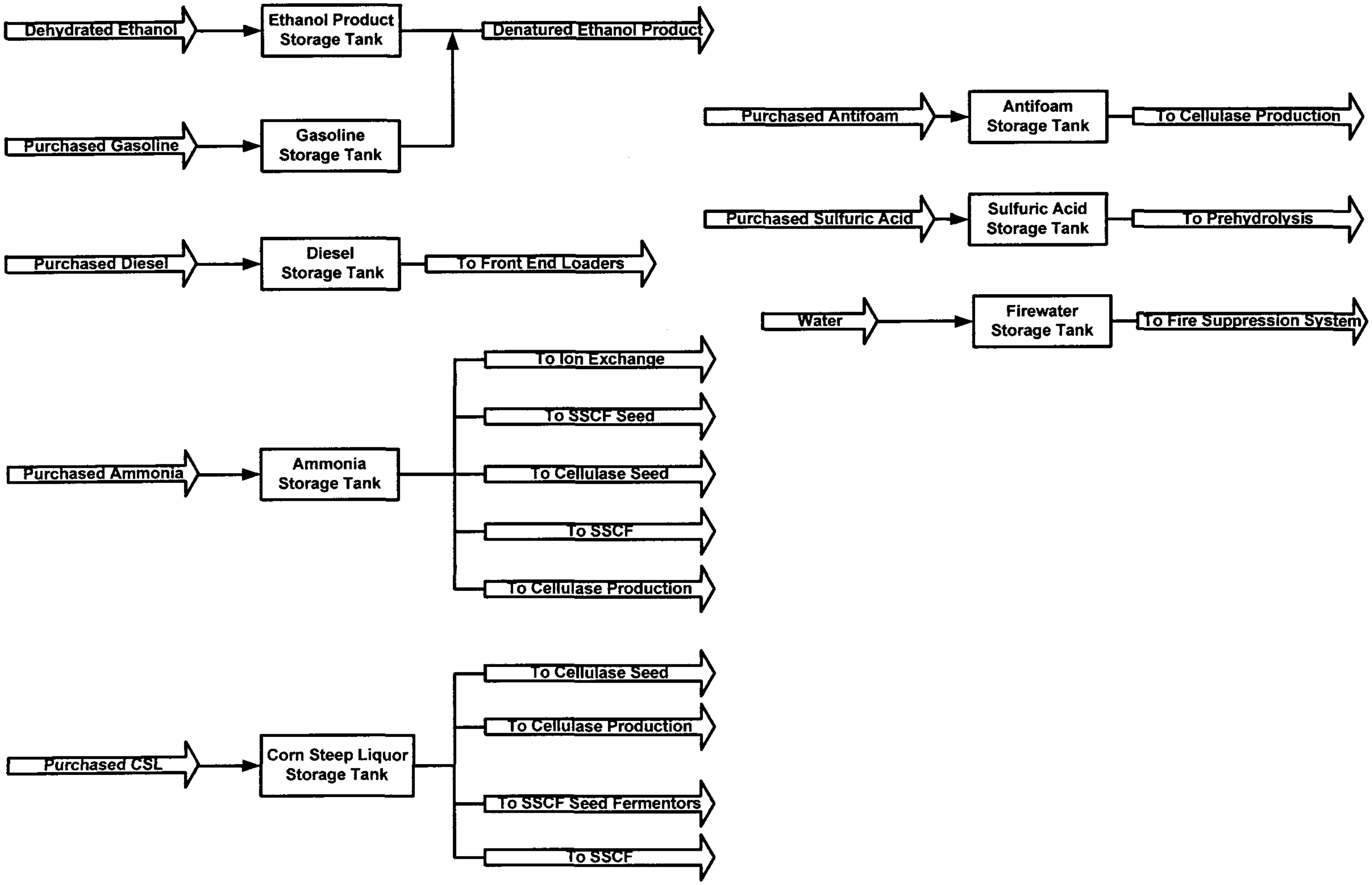

Figure A-7. Storage Flow Diagram (A700)

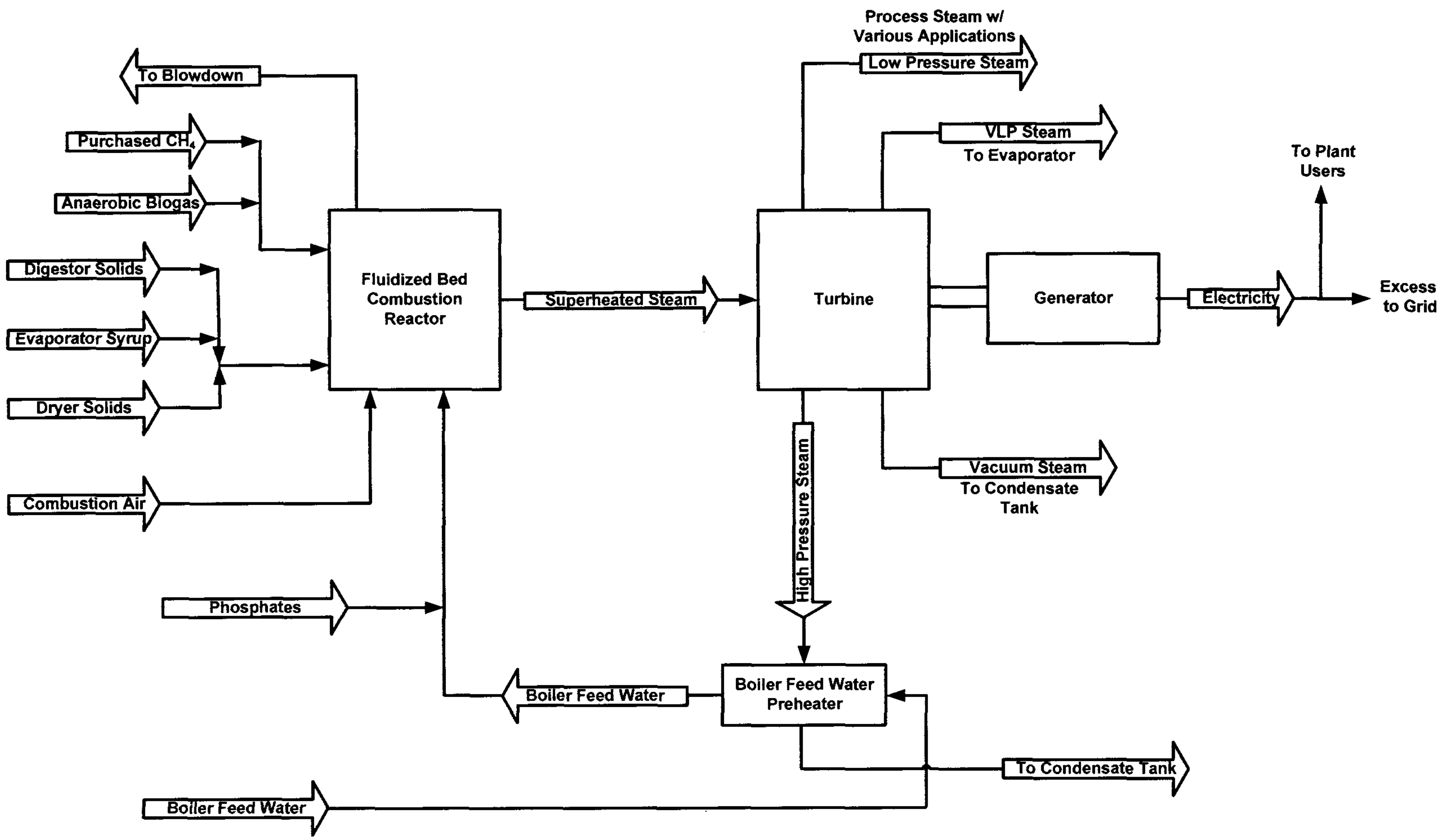

Figure A-8. Boiler, Burner, and Turbogenerator Flow Diagram (A800)

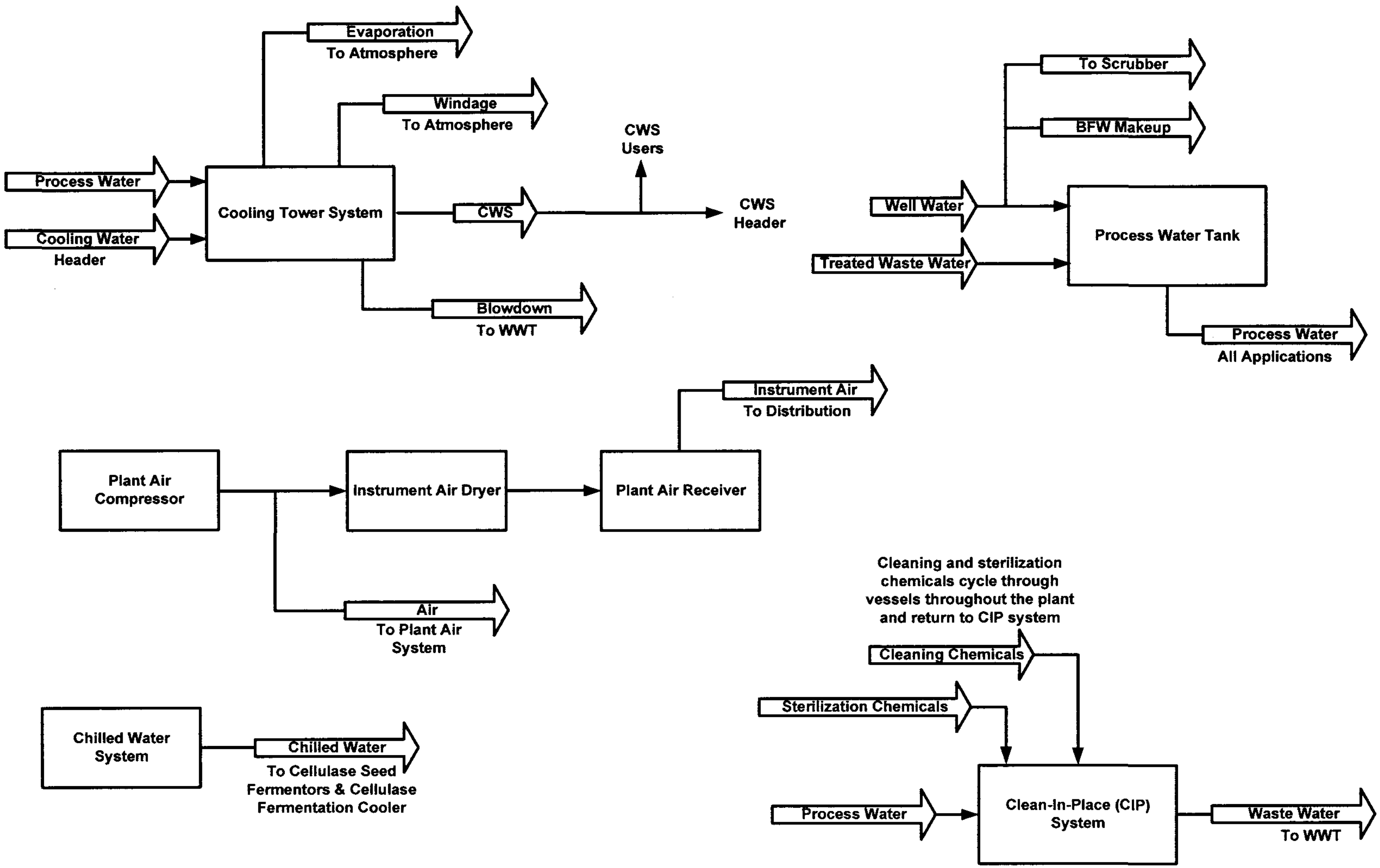

Figure A-9. Utilities Flow Diagram (A900)

Appendix B

ADDITIONAL INFORMATION FROM NREL (Wooley, 1999)

Table B-1. Feedstock Composition (Yellow Poplar Hardwood)

Component	% Dry Basis
Cellulose	42.67
Xylan	19.05
Arabinan	0.79
Mannan	3.93
Galactan	0.24
Acetate	4.64
Lignin	27.68
Ash	1
Moisture	47.9

Table B-2. Pretreatment (A200) Reactor Conditions

Acid Concentration	0.50%
Residence Time	10 minutes
Temperature	190°C
Solids in the Reactor	22%

Table B-3. Pretreatment Hydrolyzer Reactions and Conversions

Reaction	Conversion
$(Cellulose)_n$ + n H_2O → n Glucose	Cellulose 0.065
$(Cellulose)_n$ + m H_2O → m Glucose Olig	Cellulose 0.007
$(Cellulose)_n$ + 1/2 n H_2O → 1/2 n Cellobiose	Cellulose 0.007
$(Xylan)_n$ + n H_2O → n Xylose	Xylan 0.75
$(Xylan)_n$ + m H_2O → m Xylose Olig	Xylan 0.05
$(Xylan)_n$ → n Furfural + 2n H_2O	Xylan 0.1
$(Xylan)_n$ + n H_2O → $(Tar)_n$	Xylan 0.05
$(Mannan)_n$ + n H_2O → n Mannose	Mannan 0.75
$(Mannan)_n$ + m H_2O → m Mannose Olig	Mannan 0.05
$(Mannan)_n$ → n HMF + 2n H_2O	Mannan 0.15
$(Galactan)_n$ + n H_2O → n Galactose	Galactan 0.75
$(Galactan)_n$ + m H_2O → m Galactose Olig	Galactan 0.05
$(Galactan)_n$ + n H_2O → n HMF + 2n H_2O	Galactan 0.15
(Arabinan)n + n H_20 → n Arabinose	Arabinan 0.75
(Arabinan)n + m H_20 → m Arabinose Olig	Arabinan 0.05
(Arabinan)n → Furfural + 2n H_2O	Arabinan 0.1
(Arabinan)n + n H_20 → $(Tar)_n$	Arabinan 0.05
Acetate → Acetic Acid	Acetate 1.0
n Furfural + 3n H_2O → $(Tar)_n$	Furfural 1.0
n HMF + 3n H_2O → 1.2 $(Tar)_n$	HMF 1.0
Summarizes hydrolysis and side reactions taking place during pretreatment	

Table B-4. Seed Train Specifications

Inoculum Level	10% of total
Batch Time	24 hr
Fermenter Turn-Around Time	12 hr
Number of Trains	2
Number of Fermenter Stages	5
Maximum Fermenter Volume	655 m^3 (173000 gal)
Cultivation of fermentation organism (ethanalogen)	

Table B-5. SSCF Seed Train Reactions and Conversion (A300)

Saccharification Reaction	Conversion
$Cellulose_n$ + n H_2O → n Glucose	Cellulose 0.2
Fermentation Reactions	**Conversion**
Glucose + 2 Ethanol → 2 CO_2	Glucose 0.85
Glucose + 1.2 NH_3 → 6 *Z. mobilis* + 2.4 H_2O + 0.3 O_2	Glucose 0.04
Glucose + 2 H_2O → 2 Glycerol + O_2	Glucose 0.002
Glucose + 2 CO_2 → 2 Succinic Acid + O_2	Glucose 0.008
Glucose → 3 Acetic Acid	Glucose 0.022
Glucose → 2 Lactic Acid	Glucose 0.013
3 Xylose → 5 Ethanol +5 CO_2	Xylose 0.8
Xylose + NH_3 → 5 Z. mobilis + 2 H_2O + 0.25 O_2	Xylose 0.03
3 Xylose + 5 H_2O → 5 Glycerol + 2.5 O_2	Xylose 0.02
Xylose + H_2O → Xylitol + 9.5 O_2	Xylose 0.02
3 Xylose + 5 CO_2 → 5 Succinic Acid + 2.5 O_2	Xylose 0.01
2 Xylose → 5 Acetic Acid	Xylose 0.01
3 Xylose → 5 Lactic Acid	Xylose 0.01
Hydrolysis (saccharification) and microorganism produciton	

Table B-6. SSCF Production Specifications (A300)

Temperature	30°C
Initial Fermentation Solids Level	20%
Residence Time	7 days
Size of Vessels	3596 m^3 (950,000 gal) each
Number of Vessels	18
Number of Continuous Trains	3
Inoculum Level	10%
Cellulase Loading	15 FPU/g cellulase
Corn Steep Liquor Level	0.25%
SSCF design specifications	

Table B-7. Production SSCF Saccharification Reactions and Conversions (A300)

Reaction	Conversion
$(Cellulose)_n$ + m H_2O → m Glucolse Olig	Cellulose 0.068
$(Cellulose)_n$ + 1/2n H_2O → 1/2n Cellobiose	Cellulose 0.012
$(Cellulose)_n$ + n H_2O → n Glucose	Cellulose 0.8
Cellobiose + 2 H_2O → 2 Glucose	Cellobiose 1.0
Hydrolysis reactions taking place simultaneously with fermentation	

Table B-8. SSCF Fermentation Reactions and Conversions (A300)

Reaction	**Conversion**
Glucose → Ethanol + 2 CO2	Glucose 0.92
Glucose + 1.2 NH_3 → 6 *Z. mobilis* + 2.4 H_2O + 0.3 O_2	Glucose 0.027
Glucose + 2 H_2O → 2 Glycerol + O_2	Glucose 0.002
Glucose + 2 CO_2 → 2 Succinic Acid + O_2	Glucose 0.008
Glucose → 3 Acetic Acid	Glucose 0.022
Glucose → 2 Lactic Acid	Glucose 0.013
3 Xylose → 5 Ethanol +5 CO_2	Xylose 0.85
Xylose + NH_3 → 5 Z. mobilis + 2 H_2O + 0.25 O_2	Xylose 0.029
3 Xylose + 5 H_2O → 5 Glycerol + 2.5 O_2	Xylose 0.002
Xylose + H_2O → Xylitol + 0.5 O_2	Xylose 0.006
3 Xylose + 5 CO_2 → 5 Succinic Acid + 2.5 O_2	Xylose 0.009
2 Xylose → 5 Acetic Acid	Xylose 0.024
3 Xylose → 5 Lactic Acid	Xylose 0.014

Table B-9. Production SSCF Contamination Loss Reactions

Reaction		Conversion	
Glucose	→ 2 Lactic Acid	Glucose	1.0
3 Xylose	→ 5 Lactic Acid	Xylose	1.0
3 Arabinose	→ 5 Lactic Acid	Arabinose	1.0
Galactose	→ 2 Lactic Acid	Galactose	1.0
Mannose	→ 2 Lactic Acid	Mannose	1.0

Table B-10. Cellulase Production Parameters

Cellulase Requirement for SSCF	15 FPU/g cellulose
Yield	200 FPU/ (g cellulose + xylose)
Productivity	75 FPU/ (L* hr)
Initial Cellulose Concentration	4%

Table B-11. Cellulase Production Nutrient Requirements

Component	**Amount (g/L)**
$(NH_4)_2SO_4$	1.4
KH_2PO_4	2
$MgSO_4 * 7H_2O$	0.3
$CaCl_2 * 2H_2O$	0.4
Tween 80	0.2

Table B-12. Boiler Costs

Vendor/ Requestor	**Year**	**Steam Conditions Pressure/Temp**	**Steam Production (1000 lb/hr)**	**Total Cost ($MM)**	**Cost ($98/lb steam)**	**Scope**
FWEC/REI	1998	915-1265 psia/ 950°F	752	24.9	33	CFB
FWEC/NREL	1994	1515 psia/ 950°F	694	22.9	34.5	CFBC
Ahlstrom Pyropower/ Radian	1991	1515 psia/ 950°F	279-385	18-24	70-68	FBC
ABB/Chem Systems	1990	1100 psia/ 875°F	434	19.8	50	Dryer/ FBC

Table B-13. Equipment Installation Factors

Equipment	Multiplier
Agitators- Carbon Steel	1.3
Agitators- Stainless Steel	1.2
Boilers	1.3
Compressors (motor driven)	1.3
Cooling Towers	1.2
Distillation columns - Carbon Steel	3
Distillation columns - Stainless Steel	2.1
Filters	1.4
Heat Exchangers (S&T) - CS/SS	2.1
Pumps - Lobe	1.4
Pumps - Centrifugal, Carbon Steel	2.8
Pumps - Centrifugal, Stainless Steel	2
Pressure Vessels - Carbon Steel	2.8
Pressure Vessels -Stainless Steel	1.7
Tanks - Field Erected, Carbon Steel	1.4
Tanks - Field Erected, Carbon Steel with Lining	1.6
Tanks - Field Erected, Stainless Steel	1.2
Solids Handling Equipment	1.2-1.4
Rotary Dryer	1.6
Turbogenerator	1.5

Table B-14. Breakdown of Construction Costs

Project Year	Description	% of Project Cost
1	Establish project plan and schedule, complete P& ID's, and make arrangements for equipment.	8%
2	All site preparation and plant structure completed including sewer, foundations, electrical and piping. All equipment purchased and delivered. 80% of major process equipment set.	61%
3	Completion of process equipment installation, all buildings and landscaping completed, and commissioning completed. Start-up. Initial performance testing completed	31%

Appendix C

EQUIPMENT SUMMARIES BY SECTION

Table C-1
Basis for Capital Cost Estimate

Cost Factor	Calculation Method
Original Equipment Cost in Base Year	(No. Reqd. + No. Spares) X (Original Cost Per Unit)
Equipment Cost 2005	(Original Equipment Cost in Base Year) X (CEPCI 2005/CEPCI Base Year)
Scaled Cost in Base Year	(Size Ratio$^{\text{Scaling Exponent}}$) X (Original Equipment Cost in Base Year)
Original Plant Scaled 2005	(Size Ratio$^{\text{Scaling Exponent}}$) X Equipment Cost 2005
Size Ratio Double Capacity	2 X Size Ratio
Size Ratio Half Capacity	Size Ratio/2
Scaled Cost 2005 Double Capacity	(Size Ratio Double Capacity$^{\text{Scaling Exponent}}$) X Equipment Cost 2005
Scaled Cost 2005 Half Capacity	(Size Ratio Half Capacity$^{\text{Scaling Exponent}}$) X Equipment Cost 2005
Base Case Plant[a] Installed Cost in 2005	(Scaled 2005) X (Installation Factor)
Double Capacity Plant[b] Installed Cost 2005	(Scaled Cost 2005 Double Capacity) X (Installation Factor)
Half Capacity Plant[c] Installed Cost 2005	(Scaled Cost 2005 Half Capacity) X (Installation Factor)

(a) 2205 BD tons of biomass per day, 52 million gallons of ethanol produced per year
(b) 4410 BD tons of biomass per day, 104 million gallons of ethanol produced per year
(c) 1103 BD tons of biomass per day, 26 million gallons of ethanol produced per year

Table C-2
Installed Equipment Cost Summary: Feed Handling (A100)
Base Case Capacity Plant (2205 BD tons of biomass per day, 52 million gallons of ethanol produced per year)

A100	Feed Handling													
Equip. No.		No. Reqd. + No Spares	Size Ratio	Original Cost (Per Unit)	Base Year	CEPCI	2005 CEPCI	Total Original Equip. Cost in Base Year	Equip. Cost 2005	Scaling Exponent	Scaled Cost in Base Year	Base Case Plant Scaled 2005	Inst. Factor (f)	Base Case Plant Installed Cost in 2005$
C-101	Hopper Feeder	4	1.00	$8,000	1999	390.6	468.2	$32,000	$38,357	0.76	$32,000	$38,357	1.3	$49,865
C-102	Transfer Belt Conveyor	1	1.00	$78,120	1999	390.6	468.2	$78,120	$93,640	0.76	$78,120	$93,640	1.3	$121,732
C-103	Radial Stacker Conveyor	1	1.00	$200,100	1999	390.6	468.2	$200,100	$239,854	0.76	$200,100	$239,854	1.3	$311,810
C-104	Reclaim Hopper Feeder	2	1.00	$8,000	1999	390.6	468.2	$16,000	$19,179	0.76	$16,000	$19,179	1.3	$24,932
C-105	Reclaim Hopper Conveyor	1	1.00	$172,976	1999	390.6	468.2	$172,976	$207,341	0.76	$172,976	$207,341	1.3	$269,543
C-106	Chip Washer Feeder	4	1.00	$5,500	1999	390.6	468.2	$22,000	$26,371	0.76	$22,000	$26,371	1.3	$34,282
C-107	Scalper Screen Feeder	2	1.00	$13,392	1998	389.5	468.2	$26,784	$32,196	0.76	$26,784	$32,196	1.3	$41,855
C-108	Pretreatment Feeder	1	1.00	$95,255	1998	389.5	468.2	$95,255	$114,502	0.76	$95,255	$114,502	1.3	$148,852
M-101	Hydraulic Truck Dump	4	1.00	$80,000	1998	389.5	468.2	$320,000	$384,657	0.6	$320,000	$384,657	1.3	$500,054
M-104	Disk Refiner System	1	1.00	$382,500	1997	386.5	468.2	$382,500	$463,354	0.62	$382,500	$463,354	1.3	$602,361
S-101	Magnetic Separator	1	1.00	$13,863	1998	389.5	468.2	$13,863	$16,664	0.6	$13,863	$16,664	1.3	$21,663
S-102	Scalper Screener	2	1.00	$29,554	1998	389.5	468.2	$59,108	$71,051	0.75	$59,108	$71,051	1.3	$92,366
S-103	Chip Thickness Screen	1	1.00	$218,699	1998	389.5	468.2	$218,699	$262,888	0.75	$218,699	$262,888	1.3	$341,754
T-101	Dump Hopper	4	1.00	$28,327	1998	389.5	468.2	$113,308	$136,202	0.71	$113,308	$136,202	1.4	$190,683
T-102	Reclaim Hipper	2	1.00	$28,327	1998	389.5	468.2	$56,654	$68,101	0.51	$56,654	$68,101	1.4	$95,342
T-103	Washing/Refining Surge Bin	4	1.00	$36,103	1998	389.5	468.2	$144,412	$173,591	0.51	$144,412	$173,591	1.4	$243,027
W-101	Chip Washer System	4	1.00	$400,000	1998	389.5	468.2	$1,600,000	$1,923,286	0.6	$1,600,000	$1,923,286	1.3	$2,500,272
M-103	Front End Loaders	2	1.00	$156,000	1998	389.5	468.2	$312,000	$375,041	1	$312,000	$375,041	1	$375,041
TOTAL								**$3,863,779**	**$4,646,275**			**$4,646,275**		**$5,965,435**

Table C-3
Installed Equipment Cost Summary: Pretreatment (A200)
Base Case Capacity Plant (2205 BD tons of biomass per day, 52 million gallons of ethanol produced per year)

A200	Pretreatment (Detoxification)													
Equip. No.	Equipment Description	No. Reqd. + No Spares	Size Ratio	Original Cost (Per Unit)	Base Year	CEPCI	2005 CEPCI	Total Original Equip. Cost in Base Year	Equip. Cost 2005	Scaling Exponent	Scaled Cost in Base Year	Base Case Plant Scaled 2005	Inst. Factor (f)	Base Case Plant Installed Cost in 2005$
A201	In-Line Sulfuric Acid Mixer	1	0.88	$1,900	1997	386.5	468.2	$1,900	$2,302	0.48	$1,787	$2,165	1	$2,165
A202	In-Line NH3 Mixer	1	1.25	$1,500	1997	386.5	468.2	$1,500	$1,817	0.48	$1,670	$2,023	1	$2,023
A209	Overliming Tank Agitator	1	1.30	$19,800	1997	386.5	468.2	$19,800	$23,985	0.51	$22,635	$27,419	1.3	$35,645
A224	Reacidification Tank Agitator	1	1.30	$65,200	1997	386.5	468.2	$65,200	$78,982	0.51	$74,535	$90,290	1.2	$108,348
A232	Reslurrying Tank Agitator	1	1.01	$36,000	1997	386.5	468.2	$36,000	$43,610	0.51	$36,183	$43,832	1.2	$52,598
A235	In-Line Acidification Mixer	1	1.30	$2,600	1997	386.5	468.2	$2,600	$3,150	0.48	$2,949	$3,572	1	$3,572
C201	Hydrolyzate Screw Conveyor	1	1.00	$59,400	1997	386.5	468.2	$59,400	$71,956	0.78	$59,400	$71,956	1.3	$93,543
C202	Wash Solids Screw Conveyor	4	0.75	$23,700	1997	386.5	468.2	$94,800	$114,839	1	$71,100	$86,129	1.3	$111,968
H200	Hydrolyzate Cooler	1	1.31	$45,000	1997	386.5	468.2	$45,000	$54,512	0.51	$51,644	$62,561	2.1	$131,378
H201	Beer Column Feed Economizer	3	1.02	$132,800	1997	386.5	468.2	$398,400	$482,615	0.68	$403,801	$489,158	2.1	$1,027,232
M202	Prehydrolysis Reactor System	1	1.00	$12,461,841	1998	389.5	468.2	$12,461,841	$14,979,805	0.78	$12,461,841	$14,979,805	1.5	$22,469,707
P201	Sulfuric Acid Pump	2	1.13	$4,800	1997	386.5	468.2	$9,600	$11,629	0.79	$10,573	$12,808	2.8	$35,863
P209	Overlimed Hydrolyzate Pump	2	1.30	$10,700	1997	386.5	468.2	$21,400	$25,924	0.79	$26,329	$31,894	2.8	$89,304
P222	Filtered Hydrolyzate Pump	2	1.33	$10,800	1997	386.5	468.2	$21,600	$26,166	0.79	$27,058	$32,778	2.8	$91,778
P223	Lime Unloading Blower	1	1.31	$47,600	1998	389.5	468.2	$47,600	$57,218	0.5	$54,481	$65,489	1.4	$91,684
P224	Fermentation Feed Pump	3	1.01	$61,368	1998	389.5	468.2	$184,104	$221,303	0.7	$185,391	$222,850	2.8	$623,979
P225	ISEP Elution Pump	2	1.25	$7,900	1997	386.5	468.2	$15,800	$19,140	0.79	$18,846	$22,830	2.8	$63,923
P226	ISEP Reload Pump	2	1.30	$8,700	1997	386.5	468.2	$17,400	$21,078	0.79	$21,407	$25,933	2.8	$72,611
P227	ISEP Hydrolyzate Feed Pump	2	1.31	$10,700	1997	386.5	468.2	$21,400	$25,924	0.79	$26,489	$32,088	2.8	$89,846
P239	Readcidified Liquor Pump	2	1.30	$10,800	1997	386.5	468.2	$21,600	$26,166	0.79	$26,575	$32,192	2.8	$90,138
S202	Pre-IX Belt Filter Press	8	1.03	$200,000	1998	389.5	468.2	$1,600,000	$1,923,286	0.39	$1,618,551	$1,945,586	1.4	$2,723,820
S221	ISEP	1	1.00	$2,058,000	1997	386.5	468.2	$2,058,000	$2,493,029	0.33	$2,058,000	$2,493,029	1.2	$2,991,634
S222	Hydroclone & Rotary Drum Filter	1	0.47	$165,000	1998	389.5	468.2	$165,000	$198,339	0.39	$122,914	$147,750	1.4	$206,849
S227	LimeDust Vent Baghouse	1	1.30	$32,200	1997	386.5	468.2	$32,200	$39,007	1	$41,860	$50,709	1.5	$76,063
T201	Sulfuric Acid Storage	1	1.13	$5,760	1996	281.7	468.2	$5,760	$9,573	0.71	$6,282	$10,441	1.4	$14,618
T203	Blowdown Tank	1	1.00	$64,100	1997	386.5	468.2	$64,100	$77,650	0.93	$64,100	$77,650	1.2	$93,180
T209	Overliming Tank Agitator	1	1.30	$71,000	1997	386.5	468.2	$71,000	$86,008	0.71	$85,538	$103,619	1.4	$145,067
T220	Lime Storage Bin	1	1.30	$69,200	1997	386.5	468.2	$69,200	$83,828	0.46	$78,076	$94,581	1.3	$122,955
T224	Reacidification Tank	1	1.30	$147,800	1997	386.5	468.2	$147,800	$179,043	0.51	$168,961	$204,676	1.2	$245,612
T232	Slurrying Tank	1	1.01	$44,800	1997	386.5	468.2	$44,800	$54,270	0.71	$45,118	$54,655	1.2	$65,586
C225	Lime Solids Feeder	1	1.00	$3,900	1997	386.5	468.2	$3,900	$4,724		$3,900	$4,724	1.3	$6,142
TOTAL								$17,808,705	$21,440,877			$21,525,190		$31,978,830

Table C-4
Installed Equipment Cost Summary: Fermentation (A300)
Base Case Capacity Plant (2205 BD tons of biomass per day, 52 million gallons of ethanol produced per year)

A300	**Fermentation**													
Equip. No.		**No. Reqd. + No Spares**	**Size Ratio**	**Original Cost (Per Unit)**	**Base Year**	**CEPCI**	**2005 CEPCI**	**Total Original Equip. Cost in Base Year**	**Equip. Cost 2005**	**Scaling Exponent**	**Scaled Cost in Base Year**	**Base Case Plant Scaled 2005**	**Inst. Factor (f)**	**Base Case Plant Installed Cost in 2005$**
A301	Seed Hold Tank Agitator	1	0.91	$12,551	1996	281.7	468.2	$12,551	$20,860	0.51	$11,962	$19,881	1.2	$23,857
A304	4th Seed Vessel Agitator	2	0.91	$11,700	1997	386.5	468.2	$23,400	$28,346	0.51	$22,301	$27,015	1.2	$32,418
A305	5th Seed Vessel Agitator	2	0.91	$10,340	1996	281.7	468.2	$20,680	$34,371	0.51	$19,709	$32,757	1.2	$39,309
A306	Beer Surge Tank Agitator	1	1.00	$10,100	1997	386.5	468.2	$10,100	$12,235	0.51	$10,100	$12,235	1.2	$14,682
F304	4th SSCF Seed Fermentor	2	0.91	$39,500	1997	386.5	468.2	$79,000	$95,699	0.93	$72,366	$87,663	1.2	$105,196
F305	5th SSCF Seed Fermentor	2	0.91	$147,245	1998	389.5	468.2	$294,490	$353,993	0.51	$280,661	$337,369	1.2	$404,843
H300	Fermentation Cooler	18	1.33	$4,000	1997	386.5	468.2	$72,000	$87,220	0.78	$89,937	$108,948	2.1	$228,790
H301	SSCF Seed Hydrolyzate Cooler	1	0.91	$15,539	1998	389.5	468.2	$15,539	$18,679	0.78	$14,437	$17,354	2.1	$36,443
H302	SSCF Hydrolyzate Cooler	3	0.98	$25,409	1998	389.5	468.2	$76,227	$91,629	0.78	$75,035	$90,196	2.1	$189,412
H304	4th Seed Fermentor Coils	1	0.92	$3,300	1997	386.5	468.2	$3,300	$3,998	0.83	$3,079	$3,730	1.2	$4,476
H305	5th Seed Fermentor Coils	1	0.92	$18,800	1997	386.5	468.2	$18,800	$22,774	0.98	$17,325	$20,987	1.2	$25,184
P300	SSCF Recirculation and Transfer Pump	18	1.33	$8,000	1997	386.5	468.2	$144,000	$174,439	0.79	$180,387	$218,518	2.8	$611,850
P301	SSCF Seed Transfer Pump	2	0.91	$22,194	1998	389.5	468.2	$44,388	$53,357	0.7	$41,552	$49,948	1.4	$69,927
P302	Seed Transfer Pump	2	0.91	$54,088	1998	389.5	468.2	$108,176	$130,033	0.7	$101,265	$121,726	1.4	$170,417
P306	Beer Transfer Pump	2	1.00	$17,300	1997	386.5	468.2	$34,600	$41,914	0.79	$34,600	$41,914	2.8	$117,359
T301	SSCF Seed Hold Tank	1	0.91	$161,593	1998	389.5	468.2	$161,593	$194,243	0.51	$154,005	$185,122	1.2	$222,146
T306	Beer Storage Tank	1	1.00	$34,900	1997	386.5	468.2	$34,900	$42,277	0.71	$34,900	$42,277	1.2	$50,733
A300	SSCF Fermentor Agitators	34	1.00	$19,676	1996	281.7	468.2	$668,984	$1,111,886	1	$668,984	$1,111,886	1.2	$1,334,263
F300	SSCF Fermentors	17	1.00	$493,391	1998	389.5	468.2	$8,387,647	$10,082,404	1	$8,387,647	$10,082,404	1.2	$12,098,885
F301	1st SSCF Seed Fermentor	2	1.00	$14,700	1997	386.5	468.2	$29,400	$35,615	1	$29,400	$35,615	2.8	$99,721
F302	2nd SSCF Seed Fermentor	2	1.00	$32,600	1997	386.5	468.2	$65,200	$78,982	1	$65,200	$78,982	2.8	$221,150
F303	3rd SSCF Seed Fermentor	2	1.00	$81,100	1997	386.5	468.2	$162,200	$196,487	1	$162,200	$196,487	2.8	$550,162
TOTAL								**$10,467,175**	**$12,911,442**			**$12,923,015**		**$16,651,226**

Table C-5
Installed Equipment Cost Summary: Cellulase Production (A400)
Base Case Capacity Plant (2205 BD tons of biomass per day, 52 million gallons of ethanol produced per year)

A400	Cellulase (Enzyme Production)													
Equip. No.		No. Reqd. + No Spares	Size Ratio	Original Cost (Per Unit)	Base Year	CEPCI	2005 CEPCI	Total Original Equip. Cost in Base Year	Equip. Cost 2005	Scaling Exponent	Scaled Cost in Base Year	Base Case Plant Scaled 2005	Inst. Factor (f)	Base Case Plant Installed Cost in 2005$
F401	1st cellulase seed fermentor	3	0.92	$22,500	1997	386.5	468.2	$67,500	$81,768	0.93	$62,464	$75,667	2	$151,335
F402	2nd cellulase seed fermentor	3	0.92	$54,100	1997	386.5	468.2	$162,300	$196,608	0.93	$150,190	$181,938	2	$363,876
F403	3rd cellulase seed fermentor	3	0.92	$282,100	1997	386.5	468.2	$846,300	$1,025,194	0.93	$783,154	$948,700	2	$1,897,400
H400	Cellulase fermentation cooler	11	1.00	$34,400	1997	386.5	468.2	$378,400	$458,388	0.78	$378,400	$458,388	2.1	$962,614
M401	Fermentor Air Compressor Package	3	3.10	$596,342	1998	389.5	468.2	$1,789,026	$2,150,506	0.34	$2,628,328	$3,159,392	1.3	$4,107,210
P400	Cellulase Transfer Pump	2	0.97	$9,300	1997	386.5	468.2	$18,600	$22,532	0.79	$18,158	$21,996	2.8	$61,589
P401	Cellulase Seed Pump	2	0.92	$12,105	1998	389.5	468.2	$24,210	$29,102	0.7	$22,837	$27,452	1.2	$32,942
P405	Media Pump	2	0.99	$8,300	1997	386.5	468.2	$16,600	$20,109	0.79	$16,469	$19,950	2.8	$55,860
P420	Anti-foam Pump	2	1.00	$5,500	1997	386.5	468.2	$11,000	$13,325	0.79	$11,000	$13,325	2.8	$37,311
T405	Media-Prep Tank	1	0.99	$64,600	1997	386.5	468.2	$64,600	$78,255	0.71	$64,141	$77,699	1.2	$93,239
T420	Anti-foam Tank	1	1.00	$402	1998	389.5	468.2	$402	$483	0.71	$402	$483	1.2	$580
A00	Cellulase Fermentors	11	1.00	$550,000	1998	389.5	468.2	$6,050,000	$7,272,426	1	$6,050,000	$7,272,426	1.1	$7,999,669
F400	Cellulase Fermentor Agitators	11	1.00	$179,952	1998	389.5	468.2	$1,979,472	$2,379,432	1	$1,979,472	$2,379,432	1	$2,379,432
TOTAL								**$11,408,410**	**$13,728,129**			**$14,636,849**		**$18,143,056**

Table C-6
Installed Equipment Cost Summary: Cellulase Enzyme Production (A500)
Base Case Capacity Plant (2205 BD tons of biomass per day, 52 million gallons of ethanol produced per year)

Equip. No.		No. Reqd. + No Spares	Size Ratio	Original Cost (Per Unit)	Base Year	CEPCI	2005 CEPCI	Total Original Equip. Cost in Base Year	Equip. Cost 2005	Scaling Exponent	Scaled Cost in Base Year	Base Case Plant Scaled 2005	Inst. Factor (f)	Base Case Plant Installed Cost in 2005$
A500	**Distillation/Dehydration/Evaporator/Scrubber**													
D501	Beer Column	1	0.94	$636,976	1996	281.7	468.2	$636,976	$1,058,687	0.78	$606,964	$1,008,805	2.1	$2,118,491
D502	Rectification column	1	0.99	$525,800	1996	281.7	468.2	$525,800	$873,907	0.78	$521,694	$867,083	2.1	$1,820,874
E501	1st Effect Evaporation	2	1.23	$544,595	1996	281.7	468.2	$1,089,190	$1,810,290	0.68	$1,253,831	$2,083,933	2.1	$4,376,259
E502	2nd Effect Evaporation	1	1.23	$435,650	1996	281.7	468.2	$435,650	$724,073	0.68	$501,503	$833,523	2.1	$1,750,399
E503	3rd Effect Evaporation	2	1.23	$435,650	1996	281.7	468.2	$871,300	$1,448,146	0.68	$1,003,005	$1,667,047	2.1	$3,500,798
H501	Beer Column Reboiler	1	0.99	$158,374	1996	281.7	468.2	$158,374	$263,226	0.68	$157,295	$261,433	2.1	$549,009
H502	Rectification Column Reboiler	1	0.99	$29,600	1997	386.5	468.2	$29,600	$35,857	0.68	$29,398	$35,613	2.1	$74,787
H504	Beer Column Condenser	1	0.89	$29,544	1996	281.7	468.2	$29,544	$49,104	0.68	$27,293	$45,363	2.1	$95,262
H505	*Rectification Column Condenser*	1	0.99	$86,174	1996	281.7	468.2	$86,174	$143,226	0.68	$85,587	$142,250	2.1	$298,725
H512	Beer Column Feed Interchange	2	1.00	$19,040	1996	281.7	468.2	$38,080	$63,291	0.68	$38,080	$63,291	2.1	$132,911
H517	Evaporator Condenser	2	1.18	$121,576	1996	281.7	468.2	$243,152	$404,131	0.68	$272,118	$452,275	2.1	$949,777
M503	Molecular Sieve (9 pieces)	1	0.91	$2,700,000	1998	389.5	468.2	$2,700,000	$3,245,546	0.7	$2,527,509	$3,038,202	1	$3,038,202
P501	Beer Column Bottoms Pump	2	1.00	$42,300	1997	386.5	468.2	$84,600	$102,483	0.79	$84,600	$102,483	2.8	$286,953
P503	Beer Column Reflux Pump	2	0.89	$1,357	1998	389.5	468.2	$2,714	$3,262	0.79	$2,475	$2,975	2.8	$8,331
P504	Rectification Column Bottoms Pump	2	0.98	$4,916	1998	389.5	468.2	$9,832	$11,819	0.79	$9,676	$11,631	2.8	$32,568
P505	Rectification Column Reflux Pump	2	0.99	$4,782	1998	389.5	468.2	$9,564	$11,496	0.79	$9,488	$11,406	2.8	$31,935
P511	1st Effect Pump	3	0.97	$19,700	1997	386.5	468.2	$59,100	$71,593	0.79	$57,695	$69,891	2.8	$195,694
P512	2nd Effect Pump	2	0.83	$13,900	1997	386.5	468.2	$27,800	$33,676	0.79	$23,995	$29,067	2.8	$81,387
P513	3rd Effect Pump	3	0.51	$8,000	1997	386.5	468.2	$24,000	$29,073	0.79	$14,099	$17,079	2.8	$47,822
P514	Evaporator Condensate Pump	2	1.18	$12,300	1997	386.5	468.2	$24,600	$29,800	0.79	$28,036	$33,963	2.8	$95,096
P515	Scrubber Bottoms Pump	1	0.88	$2,793	1998	389.5	468.2	$2,793	$3,357	0.79	$2,525	$3,035	2.8	$8,498
T503	Beer Column Reflux Drum	1	0.89	$11,900	1997	386.5	468.2	$11,900	$14,415	0.93	$10,678	$12,935	2.1	$27,163
T505	Rectification Column Reflux Drum	1	0.99	$45,600	1997	386.5	468.2	$45,600	$55,239	0.72	$45,271	$54,841	2.1	$115,166
T512	Vent Scrubber	1	1.00	$99,000	1998	389.5	468.2	$99,000	$119,003	0.78	$99,000	$119,003	2.1	$249,907
TOTAL								**$7,245,343**	**$10,604,701**			**$10,967,126**		**$19,886,014**

Table C-7
Installed Equipment Cost Summary: Waste Water Treatment (A600)
Base Case Capacity Plant (2205 BD tons of biomass per day, 52 million gallons of ethanol produced per year)

A600	Waste Water Treatment (SOLIDS SEPARATION)													
Equip. No.		No. Reqd. + No Spares	Size Ratio	Original Cost (Per Unit)	Base Year	CEPCI	2005 CEPCI	Total Original Equip. Cost in Base Year	Equip. Cost 2005	Scaling Exponent	Scaled Cost in Base Year	Base Case Plant Scaled 2005	Inst. Factor (f)	Base Case Plant Installed Cost in 2005$
A602	Equalization Basin Agitator	1	0.95	$28,400	1997	386.5	468.2	$28,400	$34,403	0.51	$27,667	$33,515	1.2	$40,218
A606	Anaerobic Agitator	4	1.02	$30,300	1997	386.5	468.2	$121,200	$146,820	0.51	$122,430	$148,310	1.2	$177,972
A608	Aerobic Lagoon Agitators	16	1.02	$31,250	1998	389.5	468.2	$500,000	$601,027	0.51	$505,075	$607,128	1.4	$849,979
A630	Recycled Water Tank Agitator	1	0.95	$5,963	1998	389.5	468.2	$5,963	$7,168	0.51	$5,809	$6,983	1.3	$9,078
C601	Lignin Wet Cake Screw	1	0.99	$31,700	1997	386.5	468.2	$31,700	$38,401	0.78	$31,452	$38,101	1.4	$53,341
C614	Aerobic Sludge Screw	1	0.94	$5,700	1997	386.5	468.2	$5,700	$6,905	0.78	$5,431	$6,580	1.4	$9,211
H602	Anaerobic Digestor Feed Cooler	1	0.98	$128,600	1997	386.5	468.2	$128,600	$155,784	0.74	$126,692	$153,472	2.1	$322,292
M606	Biogas Emergency Flare	1	1.02	$20,793	1998	389.5	468.2	$20,793	$24,994	0.6	$21,042	$25,293	1.68	$42,492
P602	Anaerobic Reactor Feed Pump	2	0.95	$11,400	1997	386.5	468.2	$22,800	$27,620	0.79	$21,895	$26,523	2.8	$74,264
P606	Aerobic Digestor Feed Pump	2	0.95	$10,700	1997	386.5	468.2	$21,400	$25,924	0.79	$20,550	$24,894	2.8	$69,704
P608	Aerobic Sludge Recycle Pump	1	0.94	$11,100	1997	386.5	468.2	$11,100	$13,446	0.79	$10,570	$12,805	1.4	$17,927
P610	Aerobic Sludge Pump	1	0.94	$11,100	1997	386.5	468.2	$11,100	$13,446	0.79	$10,570	$12,805	1.4	$17,927
P611	Aerobic Digestion Outlet Pump	2	0.95	$10,700	1997	386.5	468.2	$21,400	$25,924	0.79	$20,550	$24,894	2.8	$69,704
P614	Sludge Filtrate Recycle Pump	2	0.94	$6,100	1997	386.5	468.2	$12,200	$14,779	0.79	$11,618	$14,074	2.8	$39,407
P616	Treated Water Pump	2	0.95	$10,600	1997	386.5	468.2	$21,200	$25,681	0.79	$20,358	$24,661	2.8	$69,052
P630	Recycled Water Pump	2	0.95	$10,600	1997	386.5	468.2	$21,200	$25,681	0.79	$20,358	$24,661	2.8	$69,052
S600	Bar Screen	1	0.95	$117,818	1991	361.3	468.2	$117,818	$152,678	0.3	$116,019	$150,346	1.2	$180,415
S601	Beer Column Bottoms Centrifuge	3	0.96	$659,550	1998	389.5	468.2	$1,978,650	$2,378,444	0.6	$1,930,775	$2,320,896	1.2	$2,785,075
S614	Belt Filter Press	1	1.02	$650,223	1998	389.5	468.2	$650,223	$781,603	0.72	$659,560	$792,827	1.8	$1,427,088
T602	Equalization Basin Agitator	1	0.95	$350,800	1998	389.5	468.2	$350,800	$421,681	0.51	$341,742	$410,793	1.42	$583,325
T606	Anaerobic Digestor	4	1.02	$881,081	1998	389.5	468.2	$3,524,324	$4,236,427	0.51	$3,560,098	$4,279,429	1.04	$4,450,607
T608	Aerobic Digestor	1	0.95	$635,173	1998	389.5	468.2	$635,173	$763,512	1	$603,414	$725,337	1	$725,337
T610	Clarifier	1	0.95	$174,385	1998	389.5	468.2	$174,385	$209,620	0.51	$169,882	$204,208	1.96	$400,247
T630	Recycled Water Tank	1	0.95	$14,515	1998	389.5	468.2	$14,515	$17,448	0.745	$13,971	$16,794	1.4	$23,511
M604	Nutrient Feed System	1	1.00	$31,400	1998	389.5	468.2	$31,400	$37,744	1	$31,400	$37,744	2.58	$97,381
M612	Filter Precoat System	1	1.00	$3,000	1998	389.5	468.2	$3,000	$3,606	1	$3,000	$3,606	1.4	$5,049
TOTAL								$8,465,044	$10,190,766			$10,126,679		$12,609,654

Table C-8
Installed Equipment Cost Summary: Storage (A700)
Base Case Capacity Plant (2205 BD tons of biomass per day, 52 million gallons of ethanol produced per year)

A700	Storage													
Equip. No.		No. Reqd. + No Spares	Size Ratio	Original Cost (Per Unit)	Base Year	CEPCI	2005 CEPCI	Total Original Equip. Cost in Base Year	Equip. Cost 2005	Scaling Exponent	Scaled Cost in Base Year	Base Case Plant Scaled 2005	Inst. Factor (f)	Base Case Plant Installed Cost in 2005$
A701	Denaturant In-Line Mixer	1	1.00	$1,900	1997	386.5	468.2	$1,900	$2,302	0.48	$1,900	$2,302	1	$2,302
P701	Ethanol Product Pump	3	1.00	$7,500	1997	386.5	468.2	$22,500	$27,256	0.79	$22,500	$27,256	2.8	$76,317
P703	Sulfuric Acid Pump	2	1.13	$8,000	1997	386.5	468.2	$16,000	$19,382	0.79	$17,622	$21,347	2.8	$59,771
P704	Firewater Pump	2	1.00	$18,400	1997	386.5	468.2	$36,800	$44,579	0.79	$36,800	$44,579	2.8	$124,821
P706	Ammonia Pump	2	1.20	$5,000	1997	386.5	468.2	$10,000	$12,114	0.79	$11,549	$13,991	2.8	$39,174
P707	Antifoam Store Pump	2	1.00	$5,700	1997	386.5	468.2	$11,400	$13,810	0.79	$11,400	$13,810	2.8	$38,667
P708	Diesel Pump	2	1.00	$6,100	1997	386.5	468.2	$12,200	$14,779	0.79	$12,200	$14,779	2.8	$41,381
P710	Gasoline Pump	2	0.98	$4,500	1997	386.5	468.2	$9,000	$10,902	0.79	$8,857	$10,730	2.8	$30,044
P720	CSL Pump	2	0.97	$8,800	1997	386.5	468.2	$17,600	$21,320	0.79	$17,182	$20,813	2.8	$58,278
T701	Ethanol Product Storage Tank	2	1.00	$165,800	1997	386.5	468.2	$331,600	$401,695	0.51	$331,600	$401,695	1.4	$562,373
T703	Sulfuric Acid Storage Tank	1	1.13	$42,500	1997	386.5	468.2	$42,500	$51,484	0.51	$45,233	$54,795	1.2	$65,754
T704	Firewater Storage Tank	1	1.00	$166,100	1997	386.5	468.2	$166,100	$201,211	0.51	$166,100	$201,211	1.4	$281,695
T706	Ammonia Storage Tank	1	1.20	$287,300	1997	386.5	468.2	$287,300	$348,031	0.72	$327,602	$396,852	1.4	$555,592
T707	Antifoam Storage Tank	1	1.00	$14,400	1997	386.5	468.2	$14,400	$17,444	0.71	$14,400	$17,444	1.4	$24,422
T708	Diesel Storage Tank	1	1.00	$14,400	1997	386.5	468.2	$14,400	$17,444	0.51	$14,400	$17,444	1.4	$24,422
T710	Gasoline Storage Tank	1	0.98	$43,500	1997	386.5	468.2	$43,500	$52,695	0.51	$43,054	$52,155	1.4	$73,017
T720	CSL Storage Pump	1	0.97	$88,100	1997	386.5	468.2	$88,100	$106,723	0.79	$86,005	$104,186	1.4	$145,860
TOTAL								$1,125,300	$1,363,171		$1,168,405	$1,415,387		$2,203,888

Table C-9

Installed Equipment Cost Summary: Burner, Boiler, and Turbogenerator (A800)

Base Case Capacity Plant (2205 BD tons of biomass per day, 52 million gallons of ethanol produced per year)

A800	Burner/Boiler Turbogenerator													
Equip. No.		No. Reqd. + No Spares	Size Ratio	Original Cost (Per Unit)	Base Year	CEPCI	2005 CEPCI	Total Original Equip. Cost in Base Year	Equip. Cost 2005	Scaling Exponent	Scaled Cost in Base Year	Base Case Plant Scaled 2005	Inst. Factor (f)	Base Case Plant Installed Cost in 2005$
H811	BFW Preheater	1	1.03	$58,400	1997	386.5	468.2	$58,400	$70,745	0.68	$59,586	$72,181	2.1	$151,581
M801	Solid Feed Rotary Dryer	1	1.00	$1,620,000	1998	389.5	468.2	$1,620,000	$1,947,327	0.45	$1,620,000	$1,947,327	1.6	$3,115,724
M803	Fluidized Bed Combustion Reactor	1	0.69	$24,900,000	1998	389.5	468.2	$24,900,000	$29,931,142	0.75	$18,851,077	$22,660,011	1.3	$29,458,014
M804	Combustion Gas Baghouse	1	0.22	$2,536,300	1998	389.5	468.2	$2,536,300	$3,048,769	0.58	$1,053,915	$1,266,863	1.5	$1,900,295
M811	Turbine/Generator	1	0.84	$10,000,000	1998	389.5	468.2	$10,000,000	$12,020,539	0.71	$8,835,646	$10,620,923	1.5	$15,931,384
M820	Hot Process Water Softener System	1	0.96	$1,381,300	1999	390.6	468.2	$1,381,300	$1,655,721	0.82	$1,335,828	$1,601,215	1.3	$2,081,579
M830	Hydrazine Addition Pkg.	1	1.06	$19,000	1994	368.1	468.2	$19,000	$24,167	0.6	$19,676	$25,027	1	$25,027
M832	Ammonia Addition Pkg.	1	1.06	$19,000	1994	368.1	468.2	$19,000	$24,167	0.6	$19,676	$25,027	1	$25,027
M834	Phosphate Addition Pkg.	1	1.06	$19,000	1994	368.1	468.2	$19,000	$24,167	0.6	$19,676	$25,027	1	$25,027
P804	Condensate Pump	2	2.36	$7,100	1997	386.5	468.2	$14,200	$17,202	0.79	$27,983	$33,898	2.8	$94,914
P811	Turbine Condensate Pump	2	1.40	$7,800	1997	386.5	468.2	$15,600	$18,898	0.79	$20,350	$24,652	2.8	$69,025
P824	Deaerator Feed Pump	2	0.74	$9,500	1997	386.5	468.2	$19,000	$23,016	0.79	$14,978	$18,144	2.8	$50,803
P826	BFW Pump	5	0.43	$52,501	1998	389.5	468.2	$262,505	$315,545	0.79	$134,765	$161,995	2.8	$453,586
P828	Blowdown Pump	2	1.10	$5,100	1997	386.5	468.2	$10,200	$12,356	0.79	$10,998	$13,322	2.8	$37,303
P830	Hydrazine Transfer Pump	1	1.06	$5,500	1997	386.5	468.2	$5,500	$6,663	0.79	$5,759	$6,976	2.8	$19,534
T804	Condensate Collection Tank	1	0.63	$7,100	1997	386.5	468.2	$7,100	$8,601	0.71	$5,114	$6,195	1.4	$8,674
T824	Condensate Surge Drum	1	0.97	$49,600	1997	386.5	468.2	$49,600	$60,085	0.72	$48,524	$58,781	1.7	$99,928
T826	Deaerator	1	0.91	$165,000	1998	389.5	468.2	$165,000	$198,339	0.72	$154,168	$185,318	2.8	$518,891
T828	Blowdown Flash Drum	1	1.11	$9,200	1997	386.5	468.2	$9,200	$11,145	0.72	$9,918	$12,014	2.8	$33,640
T830	Hydrazine Drum	1	1.06	$12,400	1997	386.5	468.2	$12,400	$15,021	0.93	$13,090	$15,858	1.7	$26,958
TOTAL								$41,123,305	$49,433,614			$38,780,754		$54,126,911

Table C-10

Installed Equipment Cost Summary: Utilities (A900)

Base Case Capacity Plant (2205 BD tons of biomass per day, 52 million gallons of ethanol produced per year)

A900	Utilities													
Equip. No.		No. Reqd. + No Spares	Size Ratio	Original Cost (Per Unit)	Base Year	CEPCI	2005 CEPCI	Total Original Equip. Cost in Base Year	Equip. Cost 2005	Scaling Exponent	Scaled Cost in Base Year	Base Case Plant Scaled 2005	Inst. Factor (f)	Base Case Plant Installed Cost in 2005$
M902	Cooling Tower System	1	0.79	$1,659,000	1998	389.5	468.2	$1,659,000	$1,994,207	0.78	$1,380,370	$1,659,279	1.2	$1,991,135
M904	Plant Air Compressor	3	1.00	$60,100	1997	386.5	468.2	$180,300	$218,413	0.34	$180,300	$218,413	1.3	$283,936
M908	Chilled Water Package	3	0.96	$380,000	1997	386.5	468.2	$1,140,000	$1,380,978	0.8	$1,103,372	$1,336,607	1.2	$1,603,928
M910	CIP System	1	1.00	$95,000	1995	381.1	468.2	$95,000	$116,712	0.6	$95,000	$116,712	1.2	$140,055
P902	cooling Water Pumps	2	0.76	$332,300	1997	386.5	468.2	$664,600	$805,086	0.79	$535,061	$648,164	2.8	$1,814,860
P912	Make-up Water Pump	2	0.76	$10,800	1997	386.5	468.2	$21,600	$26,166	0.79	$17,390	$21,066	2.8	$58,984
P914	Process Water Circulating Pump	3	0.78	$11,100	1997	386.5	468.2	$33,300	$40,339	0.79	$27,365	$33,150	2.8	$92,819
S904	Instrument Air Dryer	2	1.00	$15,498	1999	390.6	468.2	$30,996	$37,154	0.6	$30,996	$37,154	1.3	$48,300
T904	Plant Air Receiver	1	1.00	$13,000	1997	386.5	468.2	$13,000	$15,748	0.72	$13,000	$15,748	1.3	$20,472
T914	Process Water Tank	1	0.78	$195,500	1997	386.5	468.2	$195,500	$236,826	0.51	$172,232	$208,640	1.4	$292,096
TOTAL								$4,033,296	$4,871,629			$4,294,932		$6,346,586

Table C-11
Installed Equipment Cost Summary: Feed Handling (A100)
Large Plant (4410 BD tons of biomass per day, 104 million gallons of ethanol produced per year)

A100	Feed Handling												
Equip. No.		No. Reqd. + No Spares	Size Ratio Doubled Capacity	Original Cost (Per Unit)	Base Year	CEPCI	2005 CEPCI	Total Original Equip. Cost in Base Year	Equip. Cost 2005	Scaling Exponent	Scaled Cost 2005 Double Capacity	Inst. Factor (f)	Double Capacity Plant Installed Cost in 2005$
C-101	Hopper Feeder	4	2.00	$8,000	1999	390.6	468.2	$32,000	$38,357	0.76	$64,958	1.3	$84,445
C-102	Transfer Belt Conveyor	1	2.00	$78,120	1999	390.6	468.2	$78,120	$93,640	0.76	$158,578	1.3	$206,152
C-103	Radial Stacker Conveyor	1	2.00	$200,100	1999	390.6	468.2	$200,100	$239,854	0.76	$406,190	1.3	$528,047
C-104	Reclaim Hopper Feeder	2	2.00	$8,000	1999	390.6	468.2	$16,000	$19,179	0.76	$32,479	1.3	$42,223
C-105	Reclaim Hopper Conveyor	1	2.00	$172,976	1999	390.6	468.2	$172,976	$207,341	0.76	$351,130	1.3	$456,469
C-106	Chip Washer Feeder	4	2.00	$5,500	1999	390.6	468.2	$22,000	$26,371	0.76	$44,659	1.3	$58,056
C-107	Scalper Screen Feeder	2	2.00	$13,392	1998	389.5	468.2	$26,784	$32,196	0.76	$54,523	1.3	$70,880
C-108	Pretreatment Feeder	1	2.00	$95,255	1998	389.5	468.2	$95,255	$114,502	0.76	$193,907	1.3	$252,080
M-101	Hydraulic Truck Dump	4	2.00	$80,000	1998	389.5	468.2	$320,000	$384,657	0.6	$583,031	1.3	$757,941
M-104	Disk Refiner System	1	2.00	$382,500	1997	386.5	468.2	$382,500	$463,354	0.62	$712,118	1.3	$925,753
S-101	Magnetic Separator	1	2.00	$13,863	1998	389.5	468.2	$13,863	$16,664	0.6	$25,258	1.3	$32,835
S-102	Scalper Screener	2	2.00	$29,554	1998	389.5	468.2	$59,108	$71,051	0.75	$119,493	1.3	$155,341
S-103	Chip Thickness Screen	1	2.00	$218,699	1998	389.5	468.2	$218,699	$262,888	0.75	$442,123	1.3	$574,760
T-101	Dump Hopper	4	2.00	$28,327	1998	389.5	468.2	$113,308	$136,202	0.71	$222,800	1.4	$311,920
T-102	Reclaim Hipper	2	2.00	$28,327	1998	389.5	468.2	$56,654	$68,101	0.51	$96,979	1.4	$135,771
T-103	Washing/Refining Surge Bin	4	2.00	$36,103	1998	389.5	468.2	$144,412	$173,591	0.51	$247,202	1.4	$346,083
W-101	Chip Washer System	4	2.00	$400,000	1998	389.5	468.2	$1,600,000	$1,923,286	0.6	$2,915,157	1.3	$3,789,704
M-103	Front End Loaders	2	1.00	$156,000	1998	389.5	468.2	$312,000	$375,041	1	$375,041	1	$375,041
TOTAL								**$3,863,779**	**$4,646,275**		**$7,045,628**		**$9,103,502**

Table C-12
Installed Equipment Cost Summary: Pretreatment (A200)
Large Plant (4410 BD tons of biomass per day, 104 million gallons of ethanol produced per year)

A200	Pretreatment (Detoxification)												
Equip. No.	Equipment Description	No. Reqd. + No Spares	Size Ratio Doubled Capacity	Original Cost (Per Unit)	Base Year	CEPCI	2005 CEPCI	Total Original Equip. Cost in Base Year	Equip. Cost 2005	Scaling Exponent	Scaled Cost 2005 Double Capacity	Inst. Factor (f)	Double Capacity Plant Installed Cost in 2005$
A201	In-Line Sulfuric Acid Mixer	1	1.76	$1,900	1997	386.5	468.2	$1,900	$2,302	0.48	$3,019	1	$3,019
A202	In-Line NH3 Mixer	1	2.50	$1,500	1997	386.5	468.2	$1,500	$1,817	0.48	$2,821	1	$2,821
A209	Overliming Tank Agitator	1	2.60	$19,800	1997	386.5	468.2	$19,800	$23,985	0.51	$39,047	1.3	$50,761
A224	Reacidification Tank Agitator	1	2.60	$65,200	1997	386.5	468.2	$65,200	$78,982	0.51	$128,578	1.2	$154,293
A232	Reslurrying Tank Agitator	1	2.02	$36,000	1997	386.5	468.2	$36,000	$43,610	0.51	$62,419	1.2	$74,902
A235	In-Line Acidification Mixer	1	2.60	$2,600	1997	386.5	468.2	$2,600	$3,150	0.48	$4,982	1	$4,982
C201	Hydrolyzate Screw Conveyor	1	2.00	$59,400	1997	386.5	468.2	$59,400	$71,956	0.78	$123,558	1.3	$160,626
C202	Wash Solids Screw Conveyor	4	1.50	$23,700	1997	386.5	468.2	$94,800	$114,839	1	$172,259	1.3	$223,936
H200	Hydrolyzate Cooler	1	2.62	$45,000	1997	386.5	468.2	$45,000	$54,512	0.51	$89,090	2.1	$187,089
H201	Beer Column Feed Economizer	3	2.04	$132,800	1997	386.5	468.2	$398,400	$482,615	0.68	$783,700	2.1	$1,645,770
M202	Prehydrolysis Reactor System	1	2.00	$12,461,841	1998	389.5	468.2	$12,461,841	$14,979,805	0.78	$25,722,285	1.5	$38,583,428
P201	Sulfuric Acid Pump	2	2.26	$4,800	1997	386.5	468.2	$9,600	$11,629	0.79	$22,146	2.8	$62,009
P209	Overlimed Hydrolyzate Pump	2	2.60	$10,700	1997	386.5	468.2	$21,400	$25,924	0.79	$55,147	2.8	$154,413
P222	Filtered Hydrolyzate Pump	2	2.66	$10,800	1997	386.5	468.2	$21,600	$26,166	0.79	$56,675	2.8	$158,690
P223	Lime Unloading Blower	1	2.62	$47,600	1998	389.5	468.2	$47,600	$57,218	0.5	$92,615	1.4	$129,661
P224	Fermentation Feed Pump	3	2.02	$61,368	1998	389.5	468.2	$184,104	$221,303	0.7	$362,020	2.8	$1,013,657
P225	ISEP Elution Pump	2	2.50	$7,900	1997	386.5	468.2	$15,800	$19,140	0.79	$39,474	2.8	$110,527
P226	ISEP Reload Pump	2	2.60	$8,700	1997	386.5	468.2	$17,400	$21,078	0.79	$44,839	2.8	$125,550
P227	ISEP Hydrolyzate Feed Pump	2	2.62	$10,700	1997	386.5	468.2	$21,400	$25,924	0.79	$55,482	2.8	$155,350
P239	Readcidified Liquor Pump	2	2.60	$10,800	1997	386.5	468.2	$21,600	$26,166	0.79	$55,663	2.8	$155,856
S202	Pre-IX Belt Filter Press	8	2.06	$200,000	1998	389.5	468.2	$1,600,000	$1,923,286	0.39	$2,549,483	1.4	$3,569,276
S221	ISEP	1	2.00	$2,058,000	1997	386.5	468.2	$2,058,000	$2,493,029	0.33	$3,133,770	1.2	$3,760,525
S222	Hydroclone & Rotary Drum Filter	1	0.94	$165,000	1998	389.5	468.2	$165,000	$198,339	0.39	$193,610	1.4	$271,054
S227	LimeDust Vent Baghouse	1	2.60	$32,200	1997	386.5	468.2	$32,200	$39,007	1	$101,417	1.5	$152,126
T201	Sulfuric Acid Storage	1	2.26	$5,760	1996	281.7	468.2	$5,760	$9,573	0.71	$17,080	1.4	$23,912
T203	Blowdown Tank	1	2.00	$64,100	1997	386.5	468.2	$64,100	$77,650	0.93	$147,944	1.2	$177,533
T209	Overliming Tank Agitator	1	2.60	$71,000	1997	386.5	468.2	$71,000	$86,008	0.71	$169,501	1.4	$237,301
T220	Lime Storage Bin	1	2.60	$69,200	1997	386.5	468.2	$69,200	$83,828	0.46	$130,100	1.3	$169,129
T224	Reacidification Tank	1	2.60	$147,800	1997	386.5	468.2	$147,800	$179,043	0.51	$291,469	1.2	$349,763
T232	Slurrying Tank	1	2.02	$44,800	1997	386.5	468.2	$44,800	$54,270	0.71	$89,405	1.2	$107,285
C225	Lime Solids Feeder	1	2.00	$3,900	1997	386.5	468.2	$3,900	$4,724		$4,724	1.3	$6,142
TOTAL								$17,808,705	$21,440,877		$34,744,323		$51,981,387

Table C-13
Installed Equipment Cost Summary: Fermentation (A300)
Large Plant (4410 BD tons of biomass per day, 104 million gallons of ethanol produced per year)

A300	**Fermentation**												
Equip. No.		**No. Reqd. + No Spares**	**Size Ratio Doubled Capacity**	**Original Cost (Per Unit)**	**Base Year**	**CEPCI**	**2005 CEPCI**	**Total Original Equip. Cost in Base Year**	**Equip. Cost 2005**	**Scaling Exponent**	**Scaled Cost 2005 Double Capacity**	**Inst. Factor (f)**	**Double Capacity Plant Installed Cost in 2005$**
A301	Seed Hold Tank Agitator	1	1.82	$12,551	1996	281.7	468.2	$12,551	$20,860	0.51	$28,311	1.2	$33,974
A304	4th Seed Vessel Agitator	2	1.82	$11,700	1997	386.5	468.2	$23,400	$28,346	0.51	$38,471	1.2	$46,165
A305	5th Seed Vessel Agitator	2	1.82	$10,340	1996	281.7	468.2	$20,680	$34,371	0.51	$46,648	1.2	$55,977
A306	Beer Surge Tank Agitator	1	2.00	$10,100	1997	386.5	468.2	$10,100	$12,235	0.51	$17,423	1.2	$20,908
F304	4th SSCF Seed Fermentor	2	1.82	$39,500	1997	386.5	468.2	$79,000	$95,699	0.93	$167,023	1.2	$200,427
F305	5th SSCF Seed Fermentor	2	1.82	$147,245	1998	389.5	468.2	$294,490	$353,993	0.51	$480,431	1.2	$576,517
H300	Fermentation Cooler	18	2.66	$4,000	1997	386.5	468.2	$72,000	$87,220	0.78	$187,078	2.1	$392,863
H301	SSCF Seed Hydrolyzate Cooler	1	1.82	$15,539	1998	389.5	468.2	$15,539	$18,679	0.78	$29,799	2.1	$62,578
H302	SSCF Hydrolyzate Cooler	3	1.96	$25,409	1998	389.5	468.2	$76,227	$91,629	0.78	$154,879	2.1	$325,246
H304	4th Seed Fermentor Coils	1	1.84	$3,300	1997	386.5	468.2	$3,300	$3,998	0.83	$6,631	1.2	$7,957
H305	5th Seed Fermentor Coils	1	1.84	$18,800	1997	386.5	468.2	$18,800	$22,774	0.98	$41,396	1.2	$49,676
P300	SSCF Recirculation and Transfer Pump	18	2.66	$8,000	1997	386.5	468.2	$144,000	$174,439	0.79	$377,834	2.8	$1,057,935
P301	SSCF Seed Transfer Pump	2	1.82	$22,194	1998	389.5	468.2	$44,388	$53,357	0.7	$81,141	1.4	$113,597
P302	Seed Transfer Pump	2	1.82	$54,088	1998	389.5	468.2	$108,176	$130,033	0.7	$197,745	1.4	$276,843
P306	Beer Transfer Pump	2	2.00	$17,300	1997	386.5	468.2	$34,600	$41,914	0.79	$72,472	2.8	$202,922
T301	SSCF Seed Hold Tank	1	1.82	$161,593	1998	389.5	468.2	$161,593	$194,243	0.51	$263,623	1.2	$316,347
T306	Beer Storage Tank	1	2.00	$34,900	1997	386.5	468.2	$34,900	$42,277	0.71	$69,157	1.2	$82,989
A300	SSCF Fermentor Agitators	34	2.00	$19,676	1996	281.7	468.2	$668,984	$1,111,886	1	$2,223,772	1.2	$2,668,527
F300	SSCF Fermentors	17	2.00	$493,391	1998	389.5	468.2	$8,387,647	$10,082,404	1	$20,164,808	1.2	$24,197,769
F301	1st SSCF Seed Fermentor	2	2.00	$14,700	1997	386.5	468.2	$29,400	$35,615	1	$71,229	2.8	$199,442
F302	2nd SSCF Seed Fermentor	2	2.00	$32,600	1997	386.5	468.2	$65,200	$78,982	1	$157,965	2.8	$442,301
F303	3rd SSCF Seed Fermentor	2	2.00	$81,100	1997	386.5	468.2	$162,200	$196,487	1	$392,973	2.8	$1,100,325
TOTAL								**$10,467,175**	**$12,911,442**		**$25,270,809**		**$32,431,285**

Table C-14
Installed Equipment Cost Summary: Cellulase Production (A400)
Large Plant (4410 BD tons of biomass per day, 104 million gallons of ethanol produced per year)

A400	**Cellulase (Enzyme Production)**												
Equip. No.		**No. Reqd. + No Spares**	**Size Ratio Doubled Capacity**	**Original Cost (Per Unit)**	**Base Year**	**CEPCI**	**2005 CEPCI**	**Total Original Equip. Cost in Base Year**	**Equip. Cost 2005**	**Scaling Exponent**	**Scaled Cost 2005 Double Capacity**	**Inst. Factor (f)**	**Double Capacity Plant Installed Cost in 2005$**
F401	1st cellulase seed fermentor	3	1.84	$22,500	1997	386.5	468.2	$67,500	$81,768	0.93	$144,167	2	$288,334
F402	2nd cellulase seed fermentor	3	1.84	$54,100	1997	386.5	468.2	$162,300	$196,608	0.93	$346,642	2	$693,284
F403	3rd cellulase seed fermentor	3	1.84	$282,100	1997	386.5	468.2	$846,300	$1,025,194	0.93	$1,807,535	2	$3,615,071
H400	Cellulase fermentation cooler	11	2.00	$34,400	1997	386.5	468.2	$378,400	$458,388	0.78	$787,112	2.1	$1,652,935
M401	Fermentor Air Compressor Package	3	6.20	$596,342	1998	389.5	468.2	$1,789,026	$2,150,506	0.34	$3,999,022	1.3	$5,198,728
P400	Cellulase Transfer Pump	2	1.94	$9,300	1997	386.5	468.2	$18,600	$22,532	0.79	$38,033	2.8	$106,492
P401	Cellulase Seed Pump	2	1.84	$12,105	1998	389.5	468.2	$24,210	$29,102	0.7	$44,596	1.2	$53,515
P405	Media Pump	2	1.98	$8,300	1997	386.5	468.2	$16,600	$20,109	0.79	$34,495	2.8	$96,586
P420	Anti-foam Pump	2	2.00	$5,500	1997	386.5	468.2	$11,000	$13,325	0.79	$23,040	2.8	$64,513
T405	Media-Prep Tank	1	1.98	$64,600	1997	386.5	468.2	$64,600	$78,255	0.71	$127,100	1.2	$152,520
T420	Anti-foam Tank	1	2.00	$402	1998	389.5	468.2	$402	$483	0.71	$790	1.2	$949
A00	Cellulase Fermentors	11	2.00	$550,000	1998	389.5	468.2	$6,050,000	$7,272,426	1	$14,544,852	1.1	$15,999,338
F400	Cellulase Fermentor Agitators	11	2.00	$179,952	1998	389.5	468.2	$1,979,472	$2,379,432	1	$4,758,864	1	$4,758,864
TOTAL								**$11,408,410**	**$13,728,129**		**$26,656,249**		**$32,681,128**

Table C-15
Installed Equipment Cost Summary: Cellulase Enzyme Production (A500)
Large Plant (4410 BD tons of biomass per day, 104 million gallons of ethanol produced per year)

A500	Distillation/Dehydration/Evaporator/Scrubber												
Equip. No.		No. Reqd. + No Spares	Size Ratio Doubled Capacity	Original Cost *(Per Unit)*	Base Year	CEPCI	2005 CEPCI	Total Original Equip. Cost in Base Year	Equip. Cost 2005	Scaling Exponent	Scaled Cost 2005 Double Capacity	Inst. Factor *(f)*	Double Capacity Plant Installed Cost *in 2005$*
D501	Beer Column	1	1.88	$636,976	1996	281.7	468.2	$636,976	$1,058,687	0.78	$1,732,251	2.1	$3,637,727
D502	Rectification column	1	1.98	$525,800	1996	281.7	468.2	$525,800	$873,907	0.78	$1,488,895	2.1	$3,126,679
E501	1st Effect Evaporation	2	2.46	$544,595	1996	281.7	468.2	$1,089,190	$1,810,290	0.68	$3,338,751	2.1	$7,011,378
E502	2nd Effect Evaporation	1	2.46	$435,650	1996	281.7	468.2	$435,650	$724,073	0.68	$1,335,421	2.1	$2,804,384
E503	3rd Effect Evaporation	2	2.46	$435,650	1996	281.7	468.2	$871,300	$1,448,146	0.68	$2,670,842	2.1	$5,608,768
H501	Beer Column Reboiler	1	1.98	$158,374	1996	281.7	468.2	$158,374	$263,226	0.68	$418,852	2.1	$879,590
H502	Rectification Column Reboiler	1	1.98	$29,600	1997	386.5	468.2	$29,600	$35,857	0.68	$57,057	2.1	$119,819
H504	Beer Column Condenser	1	1.78	$29,544	1996	281.7	468.2	$29,544	$49,104	0.68	$72,677	2.1	$152,623
H505	Rectification Column Condenser	1	1.98	$86,174	1996	281.7	468.2	$86,174	$143,226	0.68	$227,905	2.1	$478,600
H512	Beer Column Feed Interchange	2	2.00	$19,040	1996	281.7	468.2	$38,080	$63,291	0.68	$101,401	2.1	$212,942
H517	Evaporator Condenser	2	2.36	$121,576	1996	281.7	468.2	$243,152	$404,131	0.68	$724,607	2.1	$1,521,675
M503	Molecular Sieve (9 pieces)	1	1.82	$2,700,000	1998	389.5	468.2	$2,700,000	$3,245,546	0.7	$4,935,574	1	$4,935,574
P501	Beer Column Bottoms Pump	2	2.00	$42,300	1997	386.5	468.2	$84,600	$102,483	0.79	$177,201	2.8	$496,163
P503	Beer Column Reflux Pump	2	1.78	$1,357	1998	389.5	468.2	$2,714	$3,262	0.79	$5,145	2.8	$14,405
P504	Rectification Column Bottoms Pump	2	1.96	$4,916	1998	389.5	468.2	$9,832	$11,819	0.79	$20,112	2.8	$56,313
P505	Rectification Column Reflux Pump	2	1.98	$4,782	1998	389.5	468.2	$9,564	$11,496	0.79	$19,721	2.8	$55,219
P511	1st Effect Pump	3	1.94	$19,700	1997	386.5	468.2	$59,100	$71,593	0.79	$120,846	2.8	$338,369
P512	2nd Effect Pump	2	1.66	$13,900	1997	386.5	468.2	$27,800	$33,676	0.79	$50,259	2.8	$140,725
P513	3rd Effect Pump	3	1.02	$8,000	1997	386.5	468.2	$24,000	$29,073	0.79	$29,532	2.8	$82,689
P514	Evaporator Condensate Pump	2	2.36	$12,300	1997	386.5	468.2	$24,600	$29,800	0.79	$58,724	2.8	$164,428
P515	Scrubber Bottoms Pump	1	1.76	$2,793	1998	389.5	468.2	$2,793	$3,357	0.79	$5,247	2.8	$14,693
T503	Beer Column Reflux Drum	1	1.78	$11,900	1997	386.5	468.2	$11,900	$14,415	0.93	$24,644	2.1	$51,753
T505	Rectification Column Reflux Drum	1	1.98	$45,600	1997	386.5	468.2	$45,600	$55,239	0.72	$90,333	2.1	$189,699
T512	Vent Scrubber	1	2.00	$99,000	1998	389.5	468.2	$99,000	$119,003	0.78	$204,344	2.1	$429,123
TOTAL								$7,245,343	$10,604,701		$17,910,341		$32,523,334

Table C-16
Installed Equipment Cost Summary: Waste Water Treatment (A600)
Large Plant (4410 BD tons of biomass per day, 104 million gallons of ethanol produced per year)

A600	Waste Water Treatment (SOLIDS SEPARATION)												
Equip. No.		No. Reqd. + No Spares	Size Ratio Doubled Capacity	Original Cost (Per Unit)	Base Year	CEPCI	2005 CEPCI	Total Original Equip. Cost in Base Year	Equip. Cost 2005	Scaling Exponent	Scaled Cost 2005 Double Capacity	Inst. Factor (f)	Double Capacity Plant Installed Cost in 2005$
A602	Equalization Basin Agitator	1	1.90	$28,400	1997	386.5	468.2	$28,400	$34,403	0.51	$47,727	1.2	$57,272
A606	Anaerobic Agitator	4	2.04	$30,300	1997	386.5	468.2	$121,200	$146,820	0.51	$211,201	1.2	$253,441
A608	Aerobic Lagoon Agitators	16	2.04	$31,250	1998	389.5	468.2	$500,000	$601,027	0.51	$864,580	1.4	$1,210,412
A630	Recycled Water Tank Agitator	1	1.90	$5,963	1998	389.5	468.2	$5,963	$7,168	0.51	$9,944	1.3	$12,927
C601	Lignin Wet Cake Screw	1	1.98	$31,700	1997	386.5	468.2	$31,700	$38,401	0.78	$65,424	1.4	$91,594
C614	Aerobic Sludge Screw	1	1.88	$5,700	1997	386.5	468.2	$5,700	$6,905	0.78	$11,298	1.4	$15,817
H602	Anaerobic Digestor Feed Cooler	1	1.96	$128,600	1997	386.5	468.2	$128,600	$155,784	0.74	$256,326	2.1	$538,284
M606	Biogas Emergency Flare	1	2.04	$20,793	1998	389.5	468.2	$20,793	$24,994	0.6	$38,337	1.68	$64,406
P602	Anaerobic Reactor Feed Pump	2	1.90	$11,400	1997	386.5	468.2	$22,800	$27,620	0.79	$45,860	2.8	$128,407
P606	Aerobic Digestor Feed Pump	2	1.90	$10,700	1997	386.5	468.2	$21,400	$25,924	0.79	$43,044	2.8	$120,523
P608	Aerobic Sludge Recycle Pump	1	1.88	$11,100	1997	386.5	468.2	$11,100	$13,446	0.79	$22,141	1.4	$30,997
P610	Aerobic Sludge Pump	1	1.88	$11,100	1997	386.5	468.2	$11,100	$13,446	0.79	$22,141	1.4	$30,997
P611	Aerobic Digestion Outlet Pump	2	1.90	$10,700	1997	386.5	468.2	$21,400	$25,924	0.79	$43,044	2.8	$120,523
P614	Sludge Filtrate Recycle Pump	2	1.88	$6,100	1997	386.5	468.2	$12,200	$14,779	0.79	$24,335	2.8	$68,137
P616	Treated Water Pump	2	1.90	$10,600	1997	386.5	468.2	$21,200	$25,681	0.79	$42,642	2.8	$119,396
P630	Recycled Water Pump	2	1.90	$10,600	1997	386.5	468.2	$21,200	$25,681	0.79	$42,642	2.8	$119,396
S600	Bar Screen	1	1.90	$117,818	1991	361.3	468.2	$117,818	$152,678	0.3	$185,098	1.2	$222,117
S601	Beer Column Bottoms Centrifuge	3	1.92	$659,550	1998	389.5	468.2	$1,978,650	$2,378,444	0.6	$3,517,820	1.2	$4,221,384
S614	Belt Filter Press	1	2.04	$650,223	1998	389.5	468.2	$650,223	$781,603	0.72	$1,305,930	1.8	$2,350,675
T602	Equalization Basin Agitator	1	1.90	$350,800	1998	389.5	468.2	$350,800	$421,681	0.51	$584,989	1.42	$830,685
T606	Anaerobic Digestor	4	2.04	$881,081	1998	389.5	468.2	$3,524,324	$4,236,427	0.51	$6,094,122	1.04	$6,337,887
T608	Aerobic Digestor	1	1.90	$635,173	1998	389.5	468.2	$635,173	$763,512	1	$1,450,673	1	$1,450,673
T610	Clarifier	1	1.90	$174,385	1998	389.5	468.2	$174,385	$209,620	0.51	$290,802	1.96	$569,972
T630	Recycled Water Tank	1	1.90	$14,515	1998	389.5	468.2	$14,515	$17,448	0.745	$28,146	1.4	$39,404
M604	Nutrient Feed System	1	2.00	$31,400	1998	389.5	468.2	$31,400	$37,744	1	$75,489	2.58	$194,762
M612	Filter Precoat System	1	2.00	$3,000	1998	389.5	468.2	$3,000	$3,606	1	$7,212	1.4	$10,097
TOTAL								$8,465,044	$10,190,766		$15,330,966		$19,210,188

Table C-17
Installed Equipment Cost Summary: Storage (A700)
Large Plant (4410 BD tons of biomass per day, 104 million gallons of ethanol produced per year)

A700	Storage												
Equip. No.		No. Reqd. + No Spares	Size Ratio Doubled Capacity	Original Cost (Per Unit)	Base Year	CEPCI	2005 CEPCI	Total Original Equip. Cost in Base Year	Equip. Cost 2005	Scaling Exponent	Scaled Cost 2005 Double Capacity	Inst. Factor (f)	Double Capacity Plant Installed Cost in 2005$
A701	Denaturant In-Line Mixer	1	2.00	$1,900	1997	386.5	468.2	$1,900	$2,302	0.48	$3,210	1	$3,210
P701	Ethanol Product Pump	3	2.00	$7,500	1997	386.5	468.2	$22,500	$27,256	0.79	$47,128	2.8	$131,958
P703	Sulfuric Acid Pump	2	2.26	$8,000	1997	386.5	468.2	$16,000	$19,382	0.79	$36,910	2.8	$103,349
P704	Firewater Pump	2	2.00	$18,400	1997	386.5	468.2	$36,800	$44,579	0.79	$77,080	2.8	$215,825
P706	Ammonia Pump	2	2.40	$5,000	1997	386.5	468.2	$10,000	$12,114	0.79	$24,191	2.8	$67,734
P707	Antifoam Store Pump	2	2.00	$5,700	1997	386.5	468.2	$11,400	$13,810	0.79	$23,878	2.8	$66,859
P708	Diesel Pump	2	2.00	$6,100	1997	386.5	468.2	$12,200	$14,779	0.79	$25,554	2.8	$71,551
P710	Gasoline Pump	2	1.96	$4,500	1997	386.5	468.2	$9,000	$10,902	0.79	$18,553	2.8	$51,948
P720	CSL Pump	2	1.94	$8,800	1997	386.5	468.2	$17,600	$21,320	0.79	$35,988	2.8	$100,766
T701	Ethanol Product Storage Tank	2	2.00	$165,800	1997	386.5	468.2	$331,600	$401,695	0.51	$572,034	1.4	$800,847
T703	Sulfuric Acid Storage Tank	1	2.26	$42,500	1997	386.5	468.2	$42,500	$51,484	0.51	$78,031	1.2	$93,637
T704	Firewater Storage Tank	1	2.00	$166,100	1997	386.5	468.2	$166,100	$201,211	0.51	$286,534	1.4	$401,148
T706	Ammonia Storage Tank	1	2.40	$287,300	1997	386.5	468.2	$287,300	$348,031	0.72	$653,687	1.4	$915,161
T707	Antifoam Storage Tank	1	2.00	$14,400	1997	386.5	468.2	$14,400	$17,444	0.71	$28,535	1.4	$39,949
T708	Diesel Storage Tank	1	2.00	$14,400	1997	386.5	468.2	$14,400	$17,444	0.51	$24,841	1.4	$34,777
T710	Gasoline Storage Tank	1	1.96	$43,500	1997	386.5	468.2	$43,500	$52,695	0.51	$74,271	1.4	$103,980
T720	CSL Storage Pump	1	1.94	$88,100	1997	386.5	468.2	$88,100	$106,723	0.79	$180,145	1.4	$252,202
TOTAL								$1,125,300	$1,363,171		$2,190,570		$3,454,902

Table C-18

Installed Equipment Cost Summary: Burner, Boiler, and Turbogenerator (A800)

Large Plant (4410 BD tons of biomass per day, 104 million gallons of ethanol produced per year)

A800	Burner/Boiler Turbogenerator												
Equip. No.		No. Reqd. + No Spares	Size Ratio Doubled Capacity	Original Cost (Per Unit)	Base Year	CEPCI	2005 CEPCI	Total Original Equip. Cost in Base Year	Equip. Cost 2005	Scaling Exponent	Scaled Cost 2005 Double Capacity	Inst. Factor (f)	Double Capacity Plant Installed Cost in 2005$
H811	BFW Preheater	1	2.06	$58,400	1997	386.5	468.2	$58,400	$70,745	0.68	$115,644	2.1	$242,853
M801	Solid Feed Rotary Dryer	1	2.00	$1,620,000	1998	389.5	468.2	$1,620,000	$1,947,327	0.45	$2,660,128	1.6	$4,256,204
M803	Fluidized Bed Combustion Reactor	1	1.38	$24,900,000	1998	389.5	468.2	$24,900,000	$29,931,142	0.75	$38,109,444	1.3	$49,542,277
M804	Combustion Gas Baghouse	1	0.44	$2,536,300	1998	389.5	468.2	$2,536,300	$3,048,769	0.58	$1,893,770	1.5	$2,840,654
M811	Turbine/Generator	1	1.68	$10,000,000	1998	389.5	468.2	$10,000,000	$12,020,539	0.71	$17,373,749	1.5	$26,060,623
M820	Hot Process Water Softener System	1	1.92	$1,381,300	1999	390.6	468.2	$1,381,300	$1,655,721	0.82	$2,826,794	1.3	$3,674,832
M830	Hydrazine Addition Pkg.	1	2.12	$19,000	1994	368.1	468.2	$19,000	$24,167	0.6	$37,933	1	$37,933
M832	Ammonia Addition Pkg.	1	2.12	$19,000	1994	368.1	468.2	$19,000	$24,167	0.6	$37,933	1	$37,933
M834	Phosphate Addition Pkg.	1	2.12	$19,000	1994	368.1	468.2	$19,000	$24,167	0.6	$37,933	1	$37,933
P804	Condensate Pump	2	4.72	$7,100	1997	386.5	468.2	$14,200	$17,202	0.79	$58,612	2.8	$164,113
P811	Turbine Condensate Pump	2	2.80	$7,800	1997	386.5	468.2	$15,600	$18,898	0.79	$42,625	2.8	$119,349
P824	Deaerator Feed Pump	2	1.48	$9,500	1997	386.5	468.2	$19,000	$23,016	0.79	$31,372	2.8	$87,842
P826	BFW Pump	5	0.86	$52,501	1998	389.5	468.2	$262,505	$315,545	0.79	$280,101	2.8	$784,284
P828	Blowdown Pump	2	2.20	$5,100	1997	386.5	468.2	$10,200	$12,356	0.79	$23,035	2.8	$64,499
P830	Hydrazine Transfer Pump	1	2.12	$5,500	1997	386.5	468.2	$5,500	$6,663	0.79	$12,063	2.8	$33,776
T804	Condensate Collection Tank	1	1.26	$7,100	1997	386.5	468.2	$7,100	$8,601	0.71	$10,135	1.4	$14,188
T824	Condensate Surge Drum	1	1.94	$49,600	1997	386.5	468.2	$49,600	$60,085	0.72	$96,824	1.7	$164,600
T826	Deaerator	1	1.82	$165,000	1998	389.5	468.2	$165,000	$198,339	0.72	$305,253	2.8	$854,707
T828	Blowdown Flash Drum	1	2.22	$9,200	1997	386.5	468.2	$9,200	$11,145	0.72	$19,790	2.8	$55,412
T830	Hydrazine Drum	1	2.12	$12,400	1997	386.5	468.2	$12,400	$15,021	0.93	$30,213	1.7	$51,362
TOTAL								$41,123,305	$49,433,614		$64,003,350		$89,125,376

Table C-19
Installed Equipment Cost Summary: Utilities (A900)
Large Plant (4410 BD tons of biomass per day, 104 million gallons of ethanol produced per year)

A900	**Utilities**												
Equip. No.		**No. Reqd. + No Spares**	**Size Ratio Doubled Capacity**	**Original Cost (Per Unit)**	**Base Year**	**CEPCI**	**2005 CEPCI**	**Total Original Equip. Cost in Base Year**	**Equip. Cost 2005**	**Scaling Exponent**	**Scaled Cost 2005 Double Capacity**	**Inst. Factor (f)**	**Double Capacity Plant Installed Cost in 2005$**
M902	Cooling Tower System	1	1.58	$1,659,000	1998	389.5	468.2	$1,659,000	$1,994,207	0.78	$2,849,200	1.2	$3,419,039
M904	Plant Air Compressor	3	2.00	$60,100	1997	386.5	468.2	$180,300	$218,413	0.34	$276,457	1.3	$359,394
M908	Chilled Water Package	3	1.92	$380,000	1997	386.5	468.2	$1,140,000	$1,380,978	0.8	$2,327,168	1.2	$2,792,602
M910	CIP System	1	2.00	$95,000	1995	381.1	468.2	$95,000	$116,712	0.6	$176,903	1.2	$212,283
P902	cooling Water Pumps	2	1.52	$332,300	1997	386.5	468.2	$664,600	$805,086	0.79	$1,120,724	2.8	$3,138,027
P912	Make-up Water Pump	2	1.52	$10,800	1997	386.5	468.2	$21,600	$26,166	0.79	$36,424	2.8	$101,988
P914	Process Water Circulating Pump	3	1.56	$11,100	1997	386.5	468.2	$33,300	$40,339	0.79	$57,318	2.8	$160,492
S904	Instrument Air Dryer	2	2.00	$15,498	1999	390.6	468.2	$30,996	$37,154	0.6	$56,315	1.3	$73,209
T904	Plant Air Receiver	1	2.00	$13,000	1997	386.5	468.2	$13,000	$15,748	0.72	$25,940	1.3	$33,722
T914	Process Water Tank	1	1.56	$195,500	1997	386.5	468.2	$195,500	$236,826	0.51	$297,113	1.4	$415,959
TOTAL								**$4,033,296**	**$4,871,629**		**$7,223,562**		**$10,706,715**

Table C-20
Installed Equipment Cost Summary: Feed Handling (A100)
Small Plant (1103 BD tons of biomass per day, 26 million gallons of ethanol produced per year)

A100	Feed Handling												
Equip. No.		**No. Reqd. + No Spares**	**Size Ratio Half Capacity**	**Original Cost (Per Unit)**	**Base Year**	**CEPCI**	**2005 CEPCI**	**Total Original Equip. Cost in Base Year**	**Equip. Cost 2005**	**Scaling Exponent**	**Scaled Cost 2005 Half Capacity**	**Inst. Factor (f)**	**Half Capacity Plant Installed Cost in 2005$**
C-101	Hopper Feeder	4	0.50	$8,000	1999	390.6	468.2	$32,000	$38,357	0.76	$22,650	1.3	$29,445
C-102	Transfer Belt Conveyor	1	0.50	$78,120	1999	390.6	468.2	$78,120	$93,640	0.76	$55,294	1.3	$71,882
C-103	Radial Stacker Conveyor	1	0.50	$200,100	1999	390.6	468.2	$200,100	$239,854	0.76	$141,633	1.3	$184,122
C-104	Reclaim Hopper Feeder	2	0.50	$8,000	1999	390.6	468.2	$16,000	$19,179	0.76	$11,325	1.3	$14,722
C-105	Reclaim Hopper Conveyor	1	0.50	$172,976	1999	390.6	468.2	$172,976	$207,341	0.76	$122,434	1.3	$159,164
C-106	Chip Washer Feeder	4	0.50	$5,500	1999	390.6	468.2	$22,000	$26,371	0.76	$15,572	1.3	$20,243
C-107	Scalper Screen Feeder	2	0.50	$13,392	1998	389.5	468.2	$26,784	$32,196	0.76	$19,012	1.3	$24,715
C-108	Pretreatment Feeder	1	0.50	$95,255	1998	389.5	468.2	$95,255	$114,502	0.76	$67,613	1.3	$87,897
M-101	Hydraulic Truck Dump	4	0.50	$80,000	1998	389.5	468.2	$320,000	$384,657	0.6	$253,779	1.3	$329,913
M-104	Disk Refiner System	1	0.50	$382,500	1997	386.5	468.2	$382,500	$463,354	0.62	$301,491	1.3	$391,939
S-101	Magnetic Separator	1	0.50	$13,863	1998	389.5	468.2	$13,863	$16,664	0.6	$10,994	1.3	$14,292
S-102	Scalper Screener	2	0.50	$29,554	1998	389.5	468.2	$59,108	$71,051	0.75	$42,247	1.3	$54,921
S-103	Chip Thickness Screen	1	0.50	$218,699	1998	389.5	468.2	$218,699	$262,888	0.75	$156,314	1.3	$203,208
T-101	Dump Hopper	4	0.50	$28,327	1998	389.5	468.2	$113,308	$136,202	0.71	$83,263	1.4	$116,569
T-102	Reclaim Hopper	2	0.50	$28,327	1998	389.5	468.2	$56,654	$68,101	0.51	$47,822	1.4	$66,951
T-103	Refining Surge Bin	4	0.50	$36,103	1998	389.5	468.2	$144,412	$173,591	0.51	$121,900	1.4	$170,659
W-101	Chip Washer System	4	0.50	$400,000	1998	389.5	468.2	$1,600,000	$1,923,286	0.6	$1,268,896	1.3	$1,649,564
M-103	Front End Loaders	2	1.00	$156,000	1998	389.5	468.2	$312,000	$375,041	1	$375,041	1	$375,041
TOTAL								**$3,863,779**	**$4,646,275**		**$3,117,279**		**$3,965,249**

Table C-21
Installed Equipment Cost Summary: Pretreatment (A200)
Small Plant (1103 BD tons of biomass per day, 26 million gallons of ethanol produced per year)

A200	Pretreatment (Detoxification)												
Equip. No.	Equipment Description	No. Reqd. + No Spares	Size Ratio Half Capacity	Original Cost (Per Unit)	Base Year	CEPCI	2005 CEPCI	Total Original Equip. Cost in Base Year	Equip. Cost 2005	Scaling Exponent	Scaled Cost 2005 Half Capacity	Inst. Factor (f)	Half Capacity Plant Installed Cost in 2005$
A201	In-Line Sulfuric Acid Mixer	1	0.44	$1,900	1997	386.5	468.2	$1,900	$2,302	0.48	$1,552	1	$1,552
A202	In-Line NH3 Mixer	1	0.63	$1,500	1997	386.5	468.2	$1,500	$1,817	0.48	$1,450	1	$1,450
A209	Overliming Tank Agitator	1	0.65	$19,800	1997	386.5	468.2	$19,800	$23,985	0.51	$19,255	1.3	$25,031
A224	Reacidification Tank Agitator	1	0.65	$65,200	1997	386.5	468.2	$65,200	$78,982	0.51	$63,404	1.2	$76,085
A232	Reslurrying Tank Agitator	1	0.51	$36,000	1997	386.5	468.2	$36,000	$43,610	0.51	$30,780	1.2	$36,936
A235	In-Line Acidification Mixer	1	0.65	$2,600	1997	386.5	468.2	$2,600	$3,150	0.48	$2,561	1	$2,561
C201	Hydrolyzate Screw Conveyor	1	0.50	$59,400	1997	386.5	468.2	$59,400	$71,956	0.78	$41,905	1.3	$54,476
C202	Wash Solids Screw Conveyor	4	0.38	$23,700	1997	386.5	468.2	$94,800	$114,839	1	$43,065	1.3	$55,984
H200	Hydrolyzate Cooler	1	0.66	$45,000	1997	386.5	468.2	$45,000	$54,512	0.51	$43,932	2.1	$92,256
H201	Beer Column Feed Economizer	3	0.51	$132,800	1997	386.5	468.2	$398,400	$482,615	0.68	$305,316	2.1	$641,163
M202	Prehydrolysis Reactor System	1	0.50	$12,461,841	1998	389.5	468.2	$12,461,841	$14,979,805	0.78	$8,723,741	1.5	$13,085,611
P201	Sulfuric Acid Pump	2	0.57	$4,800	1997	386.5	468.2	$9,600	$11,629	0.79	$7,407	2.8	$20,741
P209	Overlimed Hydrolyzate Pump	2	0.65	$10,700	1997	386.5	468.2	$21,400	$25,924	0.79	$18,446	2.8	$51,648
P222	Filtered Hydrolyzate Pump	2	0.67	$10,800	1997	386.5	468.2	$21,600	$26,166	0.79	$18,957	2.8	$53,079
P223	Lime Unloading Blower	1	0.66	$47,600	1998	389.5	468.2	$47,600	$57,218	0.5	$46,308	1.4	$64,831
P224	Fermentation Feed Pump	3	0.51	$61,368	1998	389.5	468.2	$184,104	$221,303	0.7	$137,180	2.8	$384,104
P225	ISEP Elution Pump	2	0.63	$7,900	1997	386.5	468.2	$15,800	$19,140	0.79	$13,203	2.8	$36,969
P226	ISEP Reload Pump	2	0.65	$8,700	1997	386.5	468.2	$17,400	$21,078	0.79	$14,998	2.8	$41,994
P227	ISEP Hydrolyzate Feed Pump	2	0.66	$10,700	1997	386.5	468.2	$21,400	$25,924	0.79	$18,558	2.8	$51,962
P239	Readcidified Liquor Pump	2	0.65	$10,800	1997	386.5	468.2	$21,600	$26,166	0.79	$18,618	2.8	$52,131
S202	Pre-IX Belt Filter Press	8	0.52	$200,000	1998	389.5	468.2	$1,600,000	$1,923,286	0.39	$1,484,734	1.4	$2,078,628
S221	ISEP	1	0.50	$2,058,000	1997	386.5	468.2	$2,058,000	$2,493,029	0.33	$1,983,295	1.2	$2,379,954
S222	Hydroclone & Rotary Drum Filter	1	0.24	$165,000	1998	389.5	468.2	$165,000	$198,339	0.39	$112,752	1.4	$157,853
S227	LimeDust Vent Baghouse	1	0.65	$32,200	1997	386.5	468.2	$32,200	$39,007	1	$25,354	1.5	$38,031
T201	Sulfuric Acid Storage	1	0.57	$5,760	1996	281.7	468.2	$5,760	$9,573	0.71	$6,383	1.4	$8,936
T203	Blowdown Tank	1	0.50	$64,100	1997	386.5	468.2	$64,100	$77,650	0.93	$40,755	1.2	$48,906
T209	Overliming Tank Agitator	1	0.65	$71,000	1997	386.5	468.2	$71,000	$86,008	0.71	$63,344	1.4	$88,682
T220	Lime Storage Bin	1	0.65	$69,200	1997	386.5	468.2	$69,200	$83,828	0.46	$68,759	1.3	$89,386
T224	Reacidification Tank	1	0.65	$147,800	1997	386.5	468.2	$147,800	$179,043	0.51	$143,728	1.2	$172,474
T232	Slurrying Tank	1	0.51	$44,800	1997	386.5	468.2	$44,800	$54,270	0.71	$33,412	1.2	$40,094
C225	Lime Solids Feeder	1	0.50	$3,900	1997	386.5	468.2	$3,900	$4,724		$4,724	1.3	$6,142
TOTAL								$17,808,705	$21,440,877		$13,537,875		$19,939,652

Table C-22

Installed Equipment Cost Summary: Fermentation (A300)

Small Plant (1103 BD tons of biomass per day, 26 million gallons of ethanol produced per year)

A300	Fermentation												
Equip. No.		No. Reqd. + No Spares	Size Ratio Half Capacity	Original Cost (Per Unit)	Base Year	CEPCI	2005 CEPCI	Total Original Equip. Cost in Base Year	Equip. Cost 2005	Scaling Exponent	Scaled Cost 2005 Half Capacity	Inst. Factor (f)	Half Capacity Plant Installed Cost in 2005$
A301	Seed Hold Tank Agitator	1	0.46	$12,551	1996	281.7	468.2	$12,551	$20,860	0.51	$13,961	1.2	$16,753
A304	4th Seed Vessel Agitator	2	0.46	$11,700	1997	386.5	468.2	$23,400	$28,346	0.51	$18,971	1.2	$22,765
A305	5th Seed Vessel Agitator	2	0.46	$10,340	1996	281.7	468.2	$20,680	$34,371	0.51	$23,003	1.2	$27,603
A306	Beer Surge Tank Agitator	1	0.50	$10,100	1997	386.5	468.2	$10,100	$12,235	0.51	$8,592	1.2	$10,310
F304	4th SSCF Seed Fermentor	2	0.46	$39,500	1997	386.5	468.2	$79,000	$95,699	0.93	$46,011	1.2	$55,213
F305	5th SSCF Seed Fermentor	2	0.46	$147,245	1998	389.5	468.2	$294,490	$353,993	0.51	$236,908	1.2	$284,290
H300	Fermentation Cooler	18	0.67	$4,000	1997	386.5	468.2	$72,000	$87,220	0.78	$63,448	2.1	$133,240
H301	SSCF Seed Hydrolyzate Cooler	1	0.46	$15,539	1998	389.5	468.2	$15,539	$18,679	0.78	$10,106	2.1	$21,223
H302	SSCF Hydrolyzate Cooler	3	0.49	$25,409	1998	389.5	468.2	$76,227	$91,629	0.78	$52,527	2.1	$110,307
H304	4th Seed Fermentor Coils	1	0.46	$3,300	1997	386.5	468.2	$3,300	$3,998	0.83	$2,098	1.2	$2,518
H305	5th Seed Fermentor Coils	1	0.46	$18,800	1997	386.5	468.2	$18,800	$22,774	0.98	$10,640	1.2	$12,768
P300	SSCF Recirculation and Transfer Pump	18	0.67	$8,000	1997	386.5	468.2	$144,000	$174,439	0.79	$126,379	2.8	$353,860
P301	SSCF Seed Transfer Pump	2	0.46	$22,194	1998	389.5	468.2	$44,388	$53,357	0.7	$30,747	1.4	$43,045
P302	Seed Transfer Pump	2	0.46	$54,088	1998	389.5	468.2	$108,176	$130,033	0.7	$74,931	1.4	$104,904
P306	Beer Transfer Pump	2	0.50	$17,300	1997	386.5	468.2	$34,600	$41,914	0.79	$24,241	2.8	$67,874
T301	SSCF Seed Hold Tank	1	0.46	$161,593	1998	389.5	468.2	$161,593	$194,243	0.51	$129,997	1.2	$155,996
T306	Beer Storage Tank	1	0.50	$34,900	1997	386.5	468.2	$34,900	$42,277	0.71	$25,845	1.2	$31,014
A300	SSCF Fermentor Agitators	34	0.50	$19,676	1996	281.7	468.2	$668,984	$1,111,886	1	$555,943	1.2	$667,132
F300	SSCF Fermentors	17	0.50	$493,391	1998	389.5	468.2	$8,387,647	$10,082,404	1	$5,041,202	1.2	$6,049,442
F301	1st SSCF Seed Fermentor	2	0.50	$14,700	1997	386.5	468.2	$29,400	$35,615	1	$17,807	2.8	$49,861
F302	2nd SSCF Seed Fermentor	2	0.50	$32,600	1997	386.5	468.2	$65,200	$78,982	1	$39,491	2.8	$110,575
F303	3rd SSCF Seed Fermentor	2	0.50	$81,100	1997	386.5	468.2	$162,200	$196,487	1	$98,243	2.8	$275,081
TOTAL								$10,467,175	$12,911,442		$6,651,090		$8,605,775

Table C-23
Installed Equipment Cost Summary: Cellulase Production (A400)
Small Plant (1103 BD tons of biomass per day, 26 million gallons of ethanol produced per year)

A400	Cellulase (Enzyme Production)												
Equip. No.		**No. Reqd. + No Spares**	**Size Ratio Half Capacity**	**Original Cost *(Per Unit)***	**Base Year**	**CEPCI**	**2005 CEPCI**	**Total Original Equip. Cost in Base Year**	**Equip. Cost 2005**	**Scaling Exponent**	**Scaled Cost 2005 Half Capacity**	**Inst. Factor *(f)***	**Half Capacity Plant Installed *Cost in 2005$***
F401	1st cellulase seed fermentor	3	0.46	$22,500	1997	386.5	468.2	$67,500	$81,768	0.93	$39,715	2	$79,429
F402	2nd cellulase seed fermentor	3	0.46	$54,100	1997	386.5	468.2	$162,300	$196,608	0.93	$95,492	2	$190,983
F403	3rd cellulase seed fermentor	3	0.46	$282,100	1997	386.5	468.2	$846,300	$1,025,194	0.93	$497,933	2	$995,866
H400	Cellulase fermentation cooler	11	0.5	$34,400	1997	386.5	468.2	$378,400	$458,388	0.78	$266,950	2.1	$560,595
M401	Fermentor Air Compressor Package	3	1.55	$596,342	1998	389.5	468.2	$1,789,026	$2,150,506	0.34	$2,496,051	1.3	$3,244,866
P400	Cellulase Transfer Pump	2	0.485	$9,300	1997	386.5	468.2	$18,600	$22,532	0.79	$12,721	2.8	$35,620
P401	Cellulase Seed Pump	2	0.46	$12,105	1998	389.5	468.2	$24,210	$29,102	0.7	$16,899	1.2	$20,278
P405	Media Pump	2	0.495	$8,300	1997	386.5	468.2	$16,600	$20,109	0.79	$11,538	2.8	$32,306
P420	Anti-foam Pump	2	0.5	$5,500	1997	386.5	468.2	$11,000	$13,325	0.79	$7,707	2.8	$21,578
T405	Media-Prep Tank	1	0.495	$64,600	1997	386.5	468.2	$64,600	$78,255	0.71	$47,499	1.2	$56,999
T420	Anti-foam Tank	1	0.5	$402	1998	389.5	468.2	$402	$483	0.71	$295	1.2	$354
A00	Cellulase Fermentors	11	0.5	$550,000	1998	389.5	468.2	$6,050,000	$7,272,426	1	$3,636,213	1.1	$3,999,834
F400	Cellulase Fermentor Agitators	11	0.5	$179,952	1998	389.5	468.2	$1,979,472	$2,379,432	1	$1,189,716	1	$1,189,716
TOTAL								**$11,408,410**	**$13,728,129**		**$8,318,728**		**$10,428,425**

Table C-24
Installed Equipment Cost Summary: Cellulase Enzyme Production (A500)
Small Plant (1103 BD tons of biomass per day, 26 million gallons of ethanol produced per year)

A500	Distillation/Dehydration/Evaporator/Scrubber												
Equip. No.		No. Reqd. + No Spares	Size Ratio Half Capacity	Original Cost (Per Unit)	Base Year	CEPCI	2005 CEPCI	Total Original Equip. Cost in Base Year	Equip. Cost 2005	Scaling Exponent	Scaled Cost 2005 Half Capacity	Inst. Factor (f)	Half Capacity Plant Installed Cost in 2005$
D501	Beer Column	1	0.47	$636,976	1996	281.7	468.2	$636,976	$1,058,687	0.78	$587,495	2.1	$1,233,739
D502	Rectification column	1	0.50	$525,800	1996	281.7	468.2	$525,800	$873,907	0.78	$504,960	2.1	$1,060,417
E501	1st Effect Evaporation	2	0.62	$544,595	1996	281.7	468.2	$1,089,190	$1,810,290	0.68	$1,300,718	2.1	$2,731,509
E502	2nd Effect Evaporation	1	0.62	$435,650	1996	281.7	468.2	$435,650	$724,073	0.68	$520,256	2.1	$1,092,538
E503	3rd Effect Evaporation	2	0.62	$435,650	1996	281.7	468.2	$871,300	$1,448,146	0.68	$1,040,513	2.1	$2,185,077
H501	Beer Column Reboiler	1	0.50	$158,374	1996	281.7	468.2	$158,374	$263,226	0.68	$163,177	2.1	$342,673
H502	Rectification Column Reboiler	1	0.50	$29,600	1997	386.5	468.2	$29,600	$35,857	0.68	$22,228	2.1	$46,679
H504	Beer Column Condenser	1	0.45	$29,544	1996	281.7	468.2	$29,544	$49,104	0.68	$28,314	2.1	$59,459
H505	Rectification Column Condenser	1	0.50	$86,174	1996	281.7	468.2	$86,174	$143,226	0.68	$88,788	2.1	$186,454
H512	Beer Column Feed Interchange	2	0.50	$19,040	1996	281.7	468.2	$38,080	$63,291	0.68	$39,504	2.1	$82,958
H517	Evaporator Condenser	2	0.59	$121,576	1996	281.7	468.2	$243,152	$404,131	0.68	$282,294	2.1	$592,818
M503	Molecular Sieve (9 pieces)	1	0.46	$2,700,000	1998	389.5	468.2	$2,700,000	$3,245,546	0.7	$1,870,233	1	$1,870,233
P501	Beer Column Bottoms Pump	2	0.50	$42,300	1997	386.5	468.2	$84,600	$102,483	0.79	$59,270	2.8	$165,957
P503	Beer Column Reflux Pump	2	0.45	$1,357	1998	389.5	468.2	$2,714	$3,262	0.79	$1,721	2.8	$4,818
P504	Rectification Column Bottoms Pump	2	0.49	$4,916	1998	389.5	468.2	$9,832	$11,819	0.79	$6,727	2.8	$18,836
P505	Rectification Column Reflux Pump	2	0.50	$4,782	1998	389.5	468.2	$9,564	$11,496	0.79	$6,596	2.8	$18,470
P511	1st Effect Pump	3	0.49	$19,700	1997	386.5	468.2	$59,100	$71,593	0.79	$40,421	2.8	$113,178
P512	2nd Effect Pump	2	0.42	$13,900	1997	386.5	468.2	$27,800	$33,676	0.79	$16,811	2.8	$47,070
P513	3rd Effect Pump	3	0.26	$8,000	1997	386.5	468.2	$24,000	$29,073	0.79	$9,878	2.8	$27,658
P514	Evaporator Condensate Pump	2	0.59	$12,300	1997	386.5	468.2	$24,600	$29,800	0.79	$19,642	2.8	$54,998
P515	Scrubber Bottoms Pump	1	0.44	$2,793	1998	389.5	468.2	$2,793	$3,357	0.79	$1,755	2.8	$4,915
T503	Beer Column Reflux Drum	1	0.45	$11,900	1997	386.5	468.2	$11,900	$14,415	0.93	$6,789	2.1	$14,257
T505	Rectification Column Reflux Drum	1	0.50	$45,600	1997	386.5	468.2	$45,600	$55,239	0.72	$33,294	2.1	$69,917
T512	Vent Scrubber	1	0.50	$99,000	1998	389.5	468.2	$99,000	$119,003	0.78	$69,304	2.1	$145,538
TOTAL								$7,245,343	$10,604,701		$6,720,688		$12,170,163

Table C-25
Installed Equipment Cost Summary: Waste Water Treatment (A600)
Small Plant (1103 BD tons of biomass per day, 26 million gallons of ethanol produced per year)

A600	Waste Water Treatment (SOLIDS SEPARATION)												
Equip. No.		No. Reqd. + No Spares	Size Ratio Half Capacity	Original Cost (Per Unit)	Base Year	CEPCI	2005 CEPCI	Total Original Equip. Cost in Base Year	Equip. Cost 2005	Scaling Exponent	Scaled Cost 2005 Half Capacity	Inst. Factor (f)	Half Capacity Plant Installed Cost in 2005$
A602	Equalization Basin Agitator	1	0.48	$28,400	1997	386.5	468.2	$28,400	$34,403	0.51	$23,535	1.2	$28,242
A606	Anaerobic Agitator	4	0.51	$30,300	1997	386.5	468.2	$121,200	$146,820	0.51	$104,147	1.2	$124,976
A608	Aerobic Lagoon Agitators	16	0.51	$31,250	1998	389.5	468.2	$500,000	$601,027	0.51	$426,339	1.4	$596,874
A630	Recycled Water Tank Agitator	1	0.48	$5,963	1998	389.5	468.2	$5,963	$7,168	0.51	$4,903	1.3	$6,374
C601	Lignin Wet Cake Screw	1	0.50	$31,700	1997	386.5	468.2	$31,700	$38,401	0.78	$22,189	1.4	$31,064
C614	Aerobic Sludge Screw	1	0.47	$5,700	1997	386.5	468.2	$5,700	$6,905	0.78	$3,832	1.4	$5,364
H602	Anaerobic Digestor Feed Cooler	1	0.49	$128,600	1997	386.5	468.2	$128,600	$155,784	0.74	$91,890	2.1	$192,969
M606	Biogas Emergency Flare	1	0.51	$20,793	1998	389.5	468.2	$20,793	$24,994	0.6	$16,687	1.68	$28,034
P602	Anaerobic Reactor Feed Pump	2	0.48	$11,400	1997	386.5	468.2	$22,800	$27,620	0.79	$15,339	2.8	$42,950
P606	Aerobic Digestor Feed Pump	2	0.48	$10,700	1997	386.5	468.2	$21,400	$25,924	0.79	$14,397	2.8	$40,313
P608	Aerobic Sludge Recycle Pump	1	0.47	$11,100	1997	386.5	468.2	$11,100	$13,446	0.79	$7,406	1.4	$10,368
P610	Aerobic Sludge Pump	1	0.47	$11,100	1997	386.5	468.2	$11,100	$13,446	0.79	$7,406	1.4	$10,368
P611	Aerobic Digestion Outlet Pump	2	0.48	$10,700	1997	386.5	468.2	$21,400	$25,924	0.79	$14,397	2.8	$40,313
P614	Sludge Filtrate Recycle Pump	2	0.47	$6,100	1997	386.5	468.2	$12,200	$14,779	0.79	$8,140	2.8	$22,791
P616	Treated Water Pump	2	0.48	$10,600	1997	386.5	468.2	$21,200	$25,681	0.79	$14,263	2.8	$39,936
P630	Recycled Water Pump	2	0.48	$10,600	1997	386.5	468.2	$21,200	$25,681	0.79	$14,263	2.8	$39,936
S600	Bar Screen	1	0.48	$117,818	1991	361.3	468.2	$117,818	$152,678	0.3	$122,119	1.2	$146,543
S601	Beer Column Bottoms Centrifuge	3	0.48	$659,550	1998	389.5	468.2	$1,978,650	$2,378,444	0.6	$1,531,220	1.2	$1,837,464
S614	Belt Filter Press	1	0.51	$650,223	1998	389.5	468.2	$650,223	$781,603	0.72	$481,323	1.8	$866,382
T602	Equalization Basin Agitator	1	0.48	$350,800	1998	389.5	468.2	$350,800	$421,681	0.51	$288,468	1.42	$409,624
T606	Anaerobic Digestor	4	0.51	$881,081	1998	389.5	468.2	$3,524,324	$4,236,427	0.51	$3,005,111	1.04	$3,125,316
T608	Aerobic Digestor	1	0.48	$635,173	1998	389.5	468.2	$635,173	$763,512	1	$362,668	1	$362,668
T610	Clarifier	1	0.48	$174,385	1998	389.5	468.2	$174,385	$209,620	0.51	$143,399	1.96	$281,062
T630	Recycled Water Tank	1	0.48	$14,515	1998	389.5	468.2	$14,515	$17,448	0.745	$10,020	1.4	$14,028
M604	Nutrient Feed System	1	0.50	$31,400	1998	389.5	468.2	$31,400	$37,744	1	$18,872	2.58	$48,690
M612	Filter Precoat System	1	0.50	$3,000	1998	389.5	468.2	$3,000	$3,606	1	$1,803	1.4	$2,524
TOTAL								$8,465,044	$10,190,766		$6,754,137		$8,355,175

Table C-26

Installed Equipment Cost Summary: Storage (A700)

Small Plant (1103 BD tons of biomass per day, 26 million gallons of ethanol produced per year)

A700	Storage												
Equip. No.		No. Reqd. + No Spares	Size Ratio Half Capacity	Original Cost (Per Unit)	Base Year	CEPCI	2005 CEPCI	Total Original Equip. Cost in Base Year	Equip. Cost 2005	Scaling Exponent	Scaled Cost 2005 Half Capacity	Inst. Factor (f)	Half Capacity Plant Installed Cost in 2005$
A701	Denaturant In-Line Mixer	1	0.50	$1,900	1997	386.5	468.2	$1,900	$2,302	0.48	$1,650	1	$1,650
P701	Ethanol Product Pump	3	0.50	$7,500	1997	386.5	468.2	$22,500	$27,256	0.79	$15,763	2.8	$44,138
P703	Sulfuric Acid Pump	2	0.57	$8,000	1997	386.5	468.2	$16,000	$19,382	0.79	$12,346	2.8	$34,568
P704	Firewater Pump	2	0.50	$18,400	1997	386.5	468.2	$36,800	$44,579	0.79	$25,782	2.8	$72,190
P706	Ammonia Pump	2	0.60	$5,000	1997	386.5	468.2	$10,000	$12,114	0.79	$8,091	2.8	$22,656
P707	Antifoam Store Pump	2	0.50	$5,700	1997	386.5	468.2	$11,400	$13,810	0.79	$7,987	2.8	$22,363
P708	Diesel Pump	2	0.50	$6,100	1997	386.5	468.2	$12,200	$14,779	0.79	$8,547	2.8	$23,932
P710	Gasoline Pump	2	0.49	$4,500	1997	386.5	468.2	$9,000	$10,902	0.79	$6,206	2.8	$17,376
P720	CSL Pump	2	0.49	$8,800	1997	386.5	468.2	$17,600	$21,320	0.79	$12,037	2.8	$33,705
T701	Ethanol Product Storage Tank	2	0.50	$165,800	1997	386.5	468.2	$331,600	$401,695	0.51	$282,079	1.4	$394,911
T703	Sulfuric Acid Storage Tank	1	0.57	$42,500	1997	386.5	468.2	$42,500	$51,484	0.51	$38,478	1.2	$46,174
T704	Firewater Storage Tank	1	0.50	$166,100	1997	386.5	468.2	$166,100	$201,211	0.51	$141,295	1.4	$197,813
T706	Ammonia Storage Tank	1	0.60	$287,300	1997	386.5	468.2	$287,300	$348,031	0.72	$240,928	1.4	$337,299
T707	Antifoam Storage Tank	1	0.50	$14,400	1997	386.5	468.2	$14,400	$17,444	0.71	$10,664	1.4	$14,929
T708	Diesel Storage Tank	1	0.50	$14,400	1997	386.5	468.2	$14,400	$17,444	0.51	$12,250	1.4	$17,149
T710	Gasoline Storage Tank	1	0.49	$43,500	1997	386.5	468.2	$43,500	$52,695	0.51	$36,624	1.4	$51,274
T720	CSL Storage Pump	1	0.49	$88,100	1997	386.5	468.2	$88,100	$106,723	0.79	$60,255	1.4	$84,357
TOTAL								$1,125,300	$1,363,171		$920,983		$1,416,483

Table C-27
Installed Equipment Cost Summary: Burner, Boiler, and Turbogenerator (A800)
Small Plant (1103 BD tons of biomass per day, 26 million gallons of ethanol produced per year)

A800	Burner/Boiler Turbogenerator												
Equip. No.		No. Reqd. + No Spares	Size Ratio Half Capacity	Original Cost (Per Unit)	Base Year	CEPCI	2005 CEPCI	Total Original Equip. Cost in Base Year	Equip. Cost 2005	Scaling Exponent	Scaled Cost 2005 Half Capacity	Inst. Factor (f)	Half Capacity Plant Installed Cost in 2005$
H811	BFW Preheater	1	0.52	$58,400	1997	386.5	468.2	$58,400	$70,745	0.68	$45,053	2.1	$94,611
M801	Solid Feed Rotary Dryer	1	0.50	$1,620,000	1998	389.5	468.2	$1,620,000	$1,947,327	0.45	$1,425,527	1.6	$2,280,843
M803	Fluidized Bed Combustion Reactor	1	0.35	$24,900,000	1998	389.5	468.2	$24,900,000	$29,931,142	0.75	$13,473,723	1.3	$17,515,840
M804	Combustion Gas Baghouse	1	0.11	$2,536,300	1998	389.5	468.2	$2,536,300	$3,048,769	0.58	$847,486	1.5	$1,271,228
M811	Turbine/Generator	1	0.42	$10,000,000	1998	389.5	468.2	$10,000,000	$12,020,539	0.71	$6,492,784	1.5	$9,739,176
M820	Hot Process Water Softener System	1	0.48	$1,381,300	1999	390.6	468.2	$1,381,300	$1,655,721	0.82	$906,995	1.3	$1,179,094
M830	Hydrazine Addition Pkg.	1	0.53	$19,000	1994	368.1	468.2	$19,000	$24,167	0.6	$16,511	1	$16,511
M832	Ammonia Addition Pkg.	1	0.53	$19,000	1994	368.1	468.2	$19,000	$24,167	0.6	$16,511	1	$16,511
M834	Phosphate Addition Pkg.	1	0.53	$19,000	1994	368.1	468.2	$19,000	$24,167	0.6	$16,511	1	$16,511
P804	Condensate Pump	2	1.18	$7,100	1997	386.5	468.2	$14,200	$17,202	0.79	$19,605	2.8	$54,893
P811	Turbine Condensate Pump	2	0.70	$7,800	1997	386.5	468.2	$15,600	$18,898	0.79	$14,257	2.8	$39,920
P824	Deaerator Feed Pump	2	0.37	$9,500	1997	386.5	468.2	$19,000	$23,016	0.79	$10,493	2.8	$29,381
P826	BFW Pump	5	0.22	$52,501	1998	389.5	468.2	$262,505	$315,545	0.79	$93,689	2.8	$262,329
P828	Blowdown Pump	2	0.55	$5,100	1997	386.5	468.2	$10,200	$12,356	0.79	$7,705	2.8	$21,574
P830	Hydrazine Transfer Pump	1	0.53	$5,500	1997	386.5	468.2	$5,500	$6,663	0.79	$4,035	2.8	$11,297
T804	Condensate Collection Tank	1	0.32	$7,100	1997	386.5	468.2	$7,100	$8,601	0.71	$3,787	1.4	$5,302
T824	Condensate Surge Drum	1	0.49	$49,600	1997	386.5	468.2	$49,600	$60,085	0.72	$35,686	1.7	$60,666
T826	Deaerator	1	0.46	$165,000	1998	389.5	468.2	$165,000	$198,339	0.72	$112,506	2.8	$315,017
T828	Blowdown Flash Drum	1	0.56	$9,200	1997	386.5	468.2	$9,200	$11,145	0.72	$7,294	2.8	$20,423
T830	Hydrazine Drum	1	0.53	$12,400	1997	386.5	468.2	$12,400	$15,021	0.93	$8,323	1.7	$14,149
TOTAL								$41,123,305	$49,433,614		$23,558,482		$32,965,279

Table C-28
Installed Equipment Cost Summary: Utilities (A900)
Small Plant (1103 BD tons of biomass per day, 26 million gallons of ethanol produced per year)

A900	Utilities												
Equip. No.		**No. Reqd. + No Spares**	**Size Ratio Half Capacity**	**Original Cost (Per Unit)**	**Base Year**	**CEPCI**	**2005 CEPCI**	**Total Original Equip. Cost in Base Year**	**Equip. Cost 2005**	**Scaling Exponent**	**Scaled Cost 2005 Half Capacity**	**Inst. Factor (f)**	**Half Capacity Plant Installed Cost in 2005$**
M902	Cooling Tower System	1	0.40	$1,659,000	1998	389.5	468.2	$1,659,000	$1,994,207	0.78	$966,309	1.2	$1,159,571
M904	Plant Air Compressor	3	0.50	$60,100	1997	386.5	468.2	$180,300	$218,413	0.34	$172,555	1.3	$224,321
M908	Chilled Water Package	3	0.48	$380,000	1997	386.5	468.2	$1,140,000	$1,380,978	0.8	$767,679	1.2	$921,215
M910	CIP System	1	0.50	$95,000	1995	381.1	468.2	$95,000	$116,712	0.6	$77,001	1.2	$92,402
P902	cooling Water Pumps	2	0.38	$332,300	1997	386.5	468.2	$664,600	$805,086	0.79	$374,862	2.8	$1,049,613
P912	Make-up Water Pump	2	0.38	$10,800	1997	386.5	468.2	$21,600	$26,166	0.79	$12,183	2.8	$34,113
P914	Process Water Circulating Pump	3	0.39	$11,100	1997	386.5	468.2	$33,300	$40,339	0.79	$19,172	2.8	$53,682
S904	Instrument Air Dryer	2	0.50	$15,498	1999	390.6	468.2	$30,996	$37,154	0.6	$24,512	1.3	$31,866
T904	Plant Air Receiver	1	0.50	$13,000	1997	386.5	468.2	$13,000	$15,748	0.72	$9,561	1.3	$12,429
T914	Process Water Tank	1	0.39	$195,500	1997	386.5	468.2	$195,500	$236,826	0.51	$146,511	1.4	$205,116
TOTAL								**$4,033,296**	**$4,871,629**		**$2,570,346**		**$3,784,328**

Appendix D

ESTIMATED CAPITAL COST

Table D-1
Installed Equipment Cost by Plant Section: Greenfield Plant

	Total Installed Equipment Cost ($Millions)		
Plant Section	**Base Case Plant 2005 (2205 Tons/Day)**	**Double Capacity Plant 2005 (4410 Tons/Day)**	**Half Capacity Plant 2005 (1102.5 Tons/Day)**
Feed Handling (A100)	6.0	9.1	4.0
Pretreatment/Detox (A200)	32.0	52.0	19.9
Fermentation (A300)	16.7	32.4	8.6
Cellulase Production (A400)	18.1	32.7	10.4
Distillation (A500)	19.9	32.5	12.2
Waste Water Treatment (A600)	12.6	19.2	8.4
Storage (A700)	2.2	3.5	1.4
Boiler/Turbogenerator (A800)	54.1	89.1	33.0
Utilities (A900)	6.3	10.7	3.8
Total Installed Equipment Cost	**167.9**	**281.2**	**101.6**

Table D-2
Installed Equipment Cost Equations by Plant Section: Greenfield Plant

Plant Section	**2005 Installed Equipment Cost (CT_j) Equation**
Feed Handling (A100)	$CT_{A100} = 62709*(HW)^{0.5929}$
Pretreatment/Detoxification (A200)	$CT_{A200} = 154923*(HW)^{0.6929}$
Fermentation (A300)	$CT_{A300} = 10559*(HW)^{0.9568}$
Cellulase Production (A400)	$CT_{A400} = 31641*(HW)^{0.8264}$
Distillation (A500)	$CT_{A500} = 86294*(HW)^{0.7068}$
Waste Water Treatment (A600)	$CT_{A600} = 128499*(HW)^{0.5963}$
Storage (A700)	$CT_{A700} = 13625*(HW)^{0.661}$
Boiler/Turbogenerator (A800)	$CT_{A800} = 217961*(HW)^{0.7165}$
Utilities (A900)	$CT_{A900} = 20234*(HW)^{0.7468}$
Total Installed Equipment Cost	$TC_{Installed} = 591236*(HW)^{0.7343}$

Table D-3
Total Project Investment Summary: Greenfield Plant

Total Project Investment Summary			
	Base Case Plant 2005 (2205 Tons/Day)	Double Capacity Plant 2005 (4410 Tons/Day)	Half Capacity Plant 2005 (1102.5 Tons/Day)
Direct Costs	**$Millions**	**$Millions**	**$Millions**
Total Installed Equipment Costs	167.9	281.2	101.6
Warehouse	2.5	4.2	1.5
Site Development	8.3	14.3	5.0
Total Installed Cost	178.8	299.7	108.1
Indirect Costs			
Field Expenses & Prorateable Costs	35.8	59.9	21.6
Home Office & Construction Fee	44.7	74.9	27.0
Project Contingency	5.4	9.0	3.2
Total Capital Investment	264.6	443.6	160.0
Other Costs	26.5	44.4	16.0
Total Project Investment	**291.0**	**487.9**	**176.0**

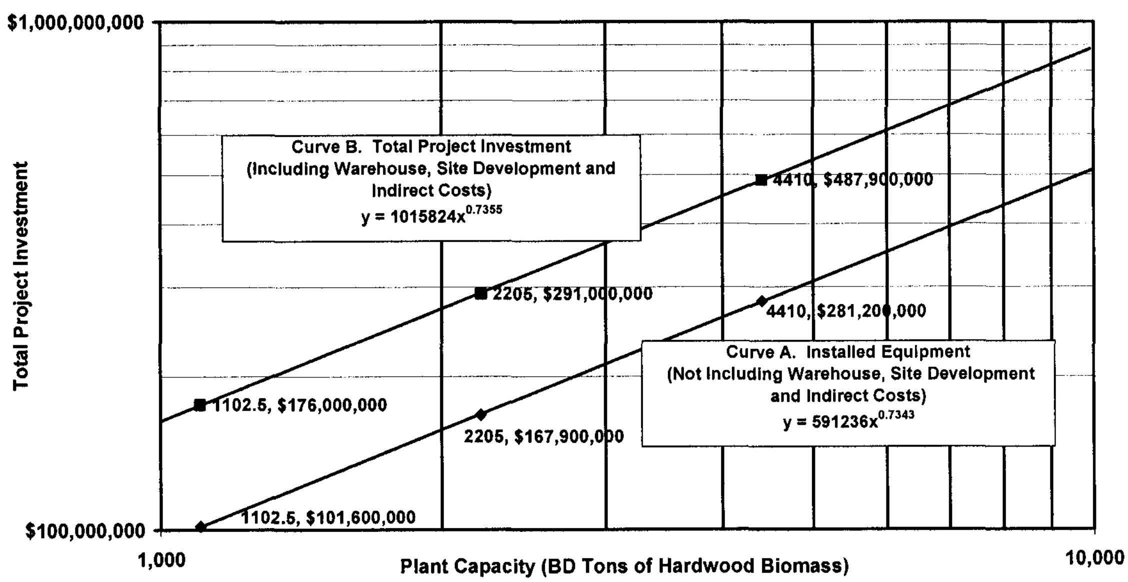

Figure D-1. Total Installed Equipment Cost and Total Project Investment (Greenfield)

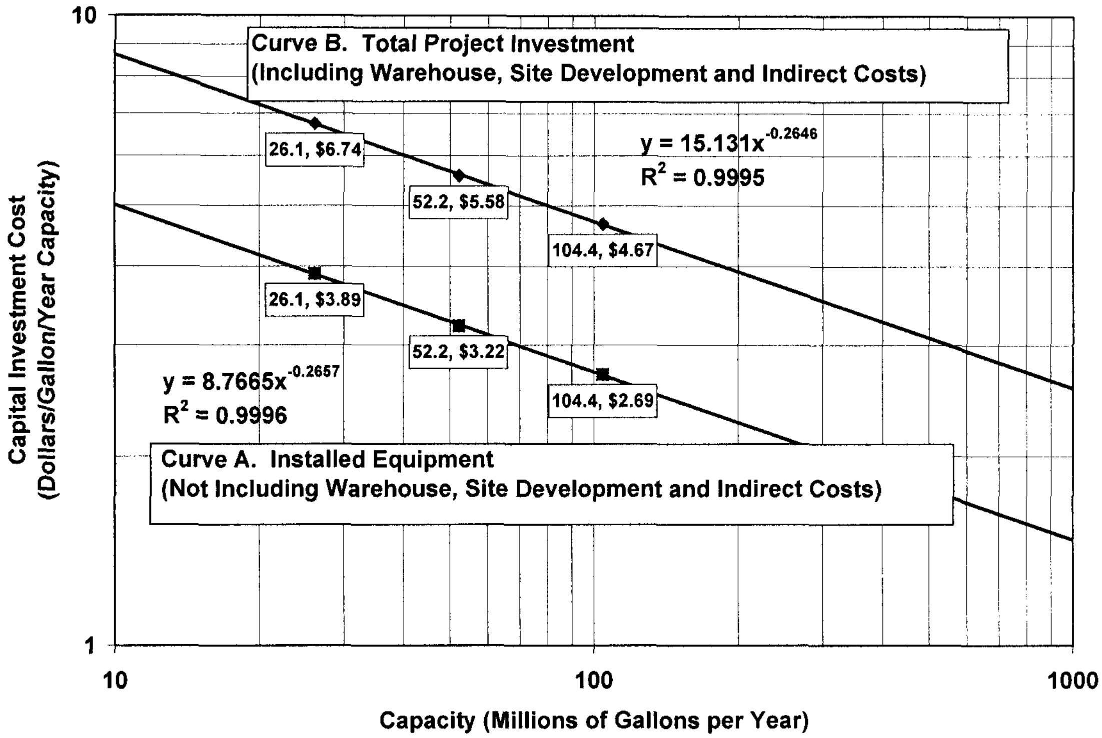

Figure D-2. Capital Investment per Gallon of Annual Production (Greenfield)

Table D-4
Installed Equipment Cost by Plant Section: Co-Location A

	Total Installed Equipment Cost ($Millions)		
Plant Section	**Base Case Plant 2005 (2205 Tons/Day)**	**Double Capacity Plant 2005 (4410 Tons/Day)**	**Half Capacity Plant 2005 (1102.5 Tons/Day)**
Feed Handling (A100)	6.0	9.1	4.0
Pretreatment/Detox (A200)	32.0	52.0	19.9
Fermentation (A300)	16.7	32.4	8.6
Cellulase Production (A400)	18.1	32.7	10.4
Distillation (A500)	19.9	32.5	12.2
Waste Water Treatment (A600)	12.6	19.2	8.4
Storage (A700)	2.2	3.5	1.4
Boiler/Turbogenerator (A800)	0.0	0.0	0.0
Utilities (A900)	6.3	10.7	3.8
Total Installed Equipment Cost	**113.8**	**192.1**	**68.7**

Table D-5
Total Project Investment Summary: Co-Location A

Total Project Investment Summary			
	Base Case Plant 2005 (2205 Tons/Day)	Double Capacity Plant 2005 (4410 Tons/Day)	Half Capacity Plant 2005 (1102.5 Tons/Day)
Direct Costs	**$Millions**	**$Millions**	**$Millions**
Total Installed Equipment Costs	113.8	192.1	68.7
Warehouse	1.7	2.9	1.0
Site Development	8.3	14.3	5.0
Total Installed Cost	123.8	209.3	74.7
Indirect Costs			
Field Expenses & Prorateable Costs	24.8	41.9	14.9
Home Office & Construction Fee	31.0	52.3	18.7
Project Contingency	3.7	6.3	2.2
Total Capital Investment	183.3	309.7	110.5
Other Costs	18.3	31.0	11.0
Total Project Investment	**201.6**	**340.7**	**121.5**

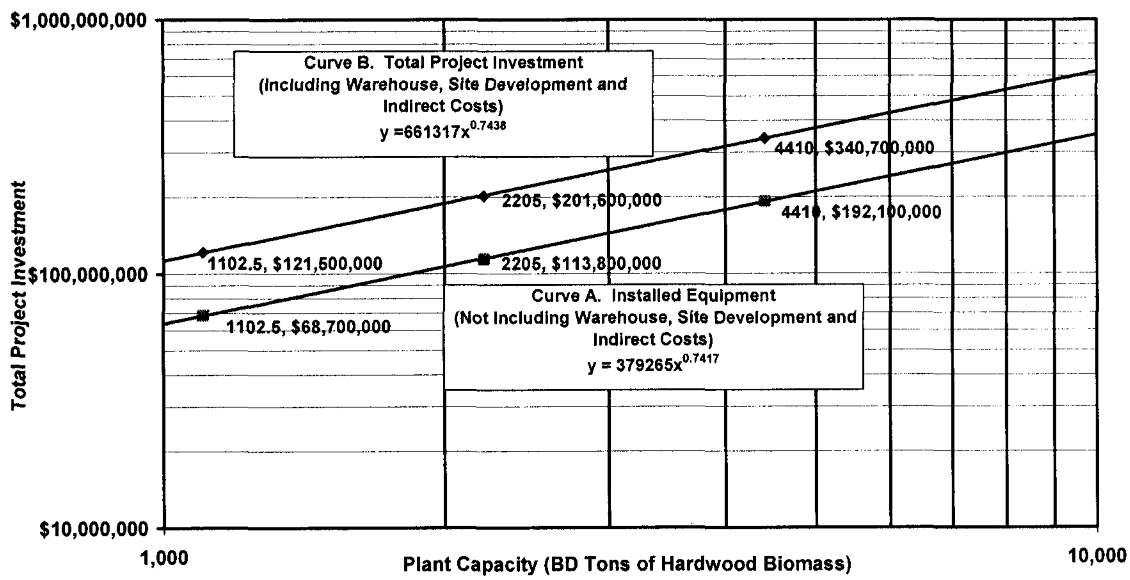

Figure D-3. Total Installed Equipment Cost and Total Project Investment: Co-Location A

Table D-6
Installed Equipment Cost by Plant Section: Co-Location B

	Total Installed Equipment Cost ($Millions)		
Plant Section	**Base Case Plant 2005 (2205 Tons/Day)**	**Double Capacity Plant 2005 (4410 Tons/Day)**	**Half Capacity Plant 2005 (1102.5 Tons/Day)**
Feed Handling (A100)	6.0	9.1	4.0
Pretreatment/Detox (A200)	32.0	52.0	19.9
Fermentation (A300)	16.7	32.4	8.6
Cellulase Production (A400)	0.0	0.0	0.0
Distillation (A500)	19.9	32.5	12.2
Waste Water Treatment (A600)	12.6	19.2	8.4
Storage (A700)	2.2	3.5	1.4
Boiler/Turbogenerator (A800)	0.0	0.0	0.0
Utilities (A900)	6.3	10.7	3.8
Total Equipment Cost (Installed)	**95.6**	**159.4**	**58.2**

Table D-7
Total Project Investment Summary: Co-Location B

Total Project Investment Summary			
	Base Case Plant 2005 (2205 Tons/Day)	**Double Capacity Plant 2005 (4410 Tons/Day)**	**Half Capacity Plant 2005 (1102.5 Tons/Day)**
Direct Costs	**$Millions**	**$Millions**	**$Millions**
Total Installed Equipment Costs	95.6	159.4	58.2
Warehouse	1.4	2.4	0.9
Site Development	6.7	11.3	4.0
Total Installed Cost	103.8	173.1	63.1
Indirect Costs			
Field Expenses & Prorateable Costs	20.8	34.6	12.6
Home Office & Construction Fee	25.9	43.3	15.8
Project Contingency	3.1	5.2	1.9
Total Capital Investment	153.6	256.3	93.4
Other Costs	15.4	25.6	9.3
Total Project Investment	**169.0**	**281.9**	**102.8**

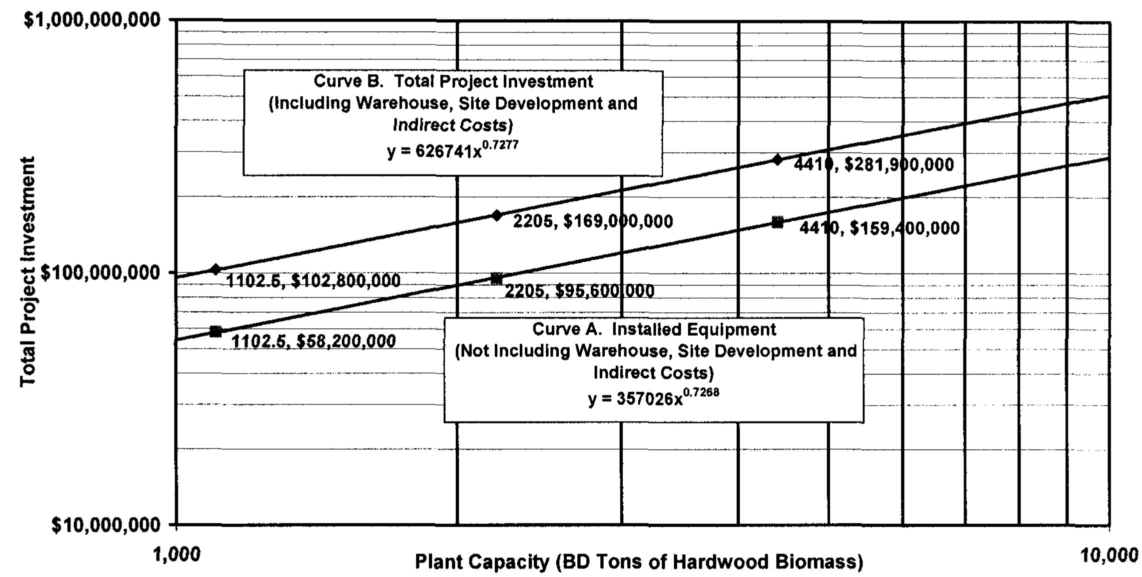

Figure D-4. Installed Equipment Cost and Total Project Investment: Co-Location B

Table D-8
Installed Equipment Cost by Plant Section: Co-Location C

Plant Section	Total Installed Equipment Cost ($Millions) Base Case Plant 2005 (2205 Tons/Day)	Double Capacity Plant 2005 (4410 Tons/Day)	Half Capacity Plant 2005 (1102.5 Tons/Day)
Feed Handling (A100)	0.0	0.0	0.0
Pretreatment/Detox (A200)	32.0	52.0	19.9
Fermentation (A300)	16.7	32.4	8.6
Cellulase Production (A400)	0.0	0.0	0.0
Distillation (A500)	19.9	32.5	12.2
Waste Water Treatment (A600)	0.0	0.0	0.0
Storage (A700)	2.2	3.5	1.4
Boiler/Turbogenerator (A800)	0.0	0.0	0.0
Utilities (A900)	6.3	10.7	3.8
Total Installed Equipment Cost	**77.1**	**131.1**	**45.9**

Table D-9
Total Project Investment Summary: Co-Location C

Total Project Investment Summary			
	Base Case Plant 2005 (2205 Tons/Day)	Double Capacity Plant 2005 (4410 Tons/Day)	Half Capacity Plant 2005 (1102.5 Tons/Day)
Direct Costs	**$Millions**	**$Millions**	**$Millions**
Total Installed Equipment Costs	77.1	131.1	45.9
Warehouse	1.2	2.0	0.7
Site Development	6.2	10.5	3.7
Total Installed Cost	84.4	143.6	50.3
Indirect Costs			
Field Expenses & Prorateable Costs	16.9	28.7	10.1
Home Office & Construction Fee	21.1	35.9	12.6
Project Contingency	2.5	4.3	1.5
Total Capital Investment	124.9	212.5	74.4
Other Costs	12.5	21.3	7.4
Total Project Investment	**137.4**	**233.8**	**81.8**

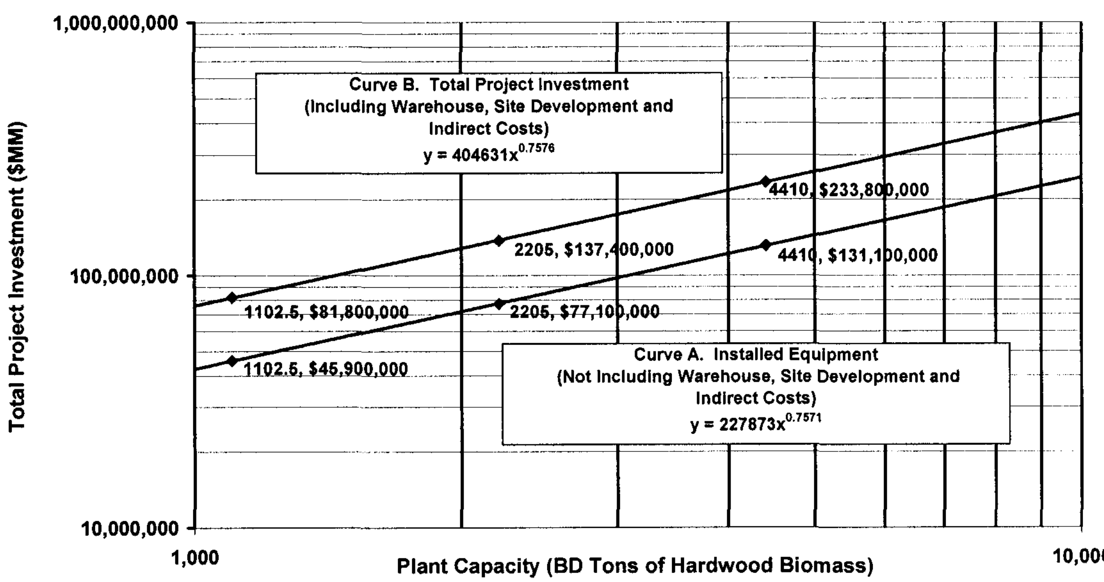

Figure D-5. Installed Equipment Cost and Total Project Investment: Co-Location C

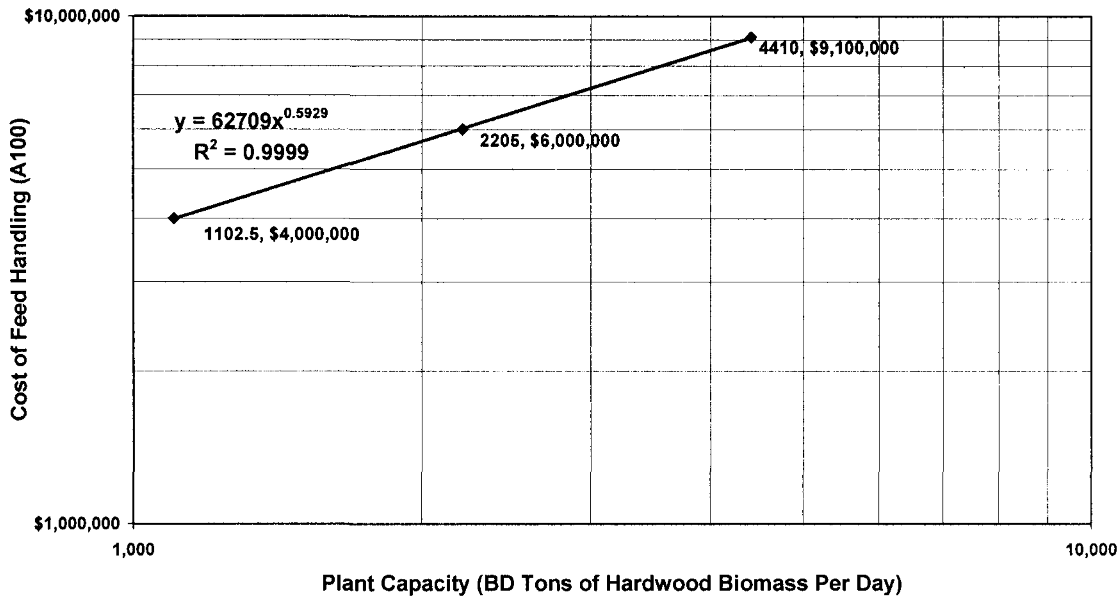

Figure D-6. Total Installed Equipment Cost vs. Hardwood Feedrate: Feed Handling (A100)

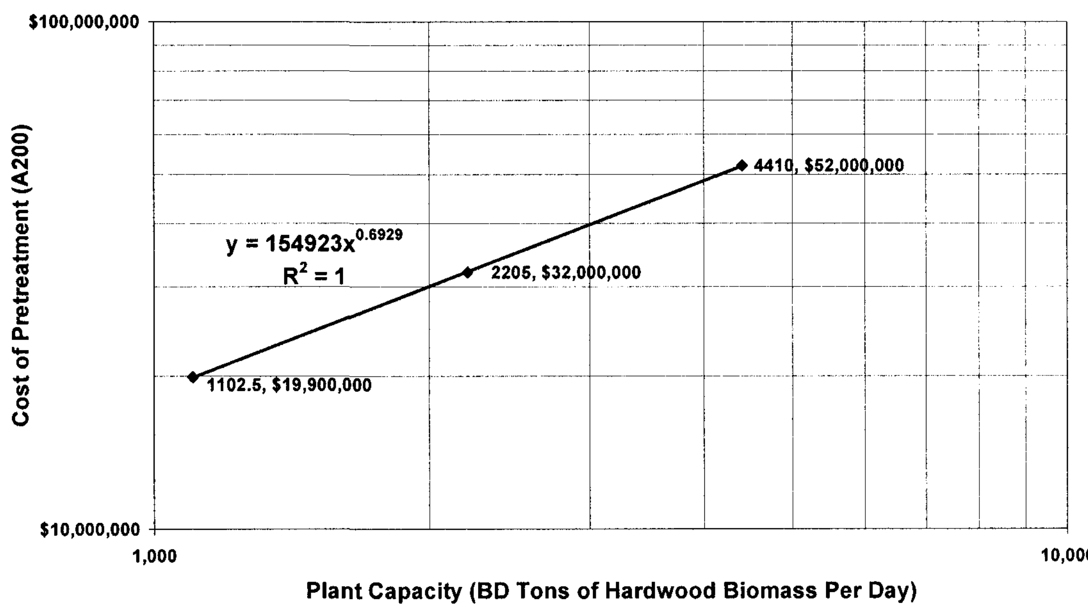

Figure D-7. Total Installed Equipment Cost vs. Hardwood Feedrate: Pretreatment (A200)

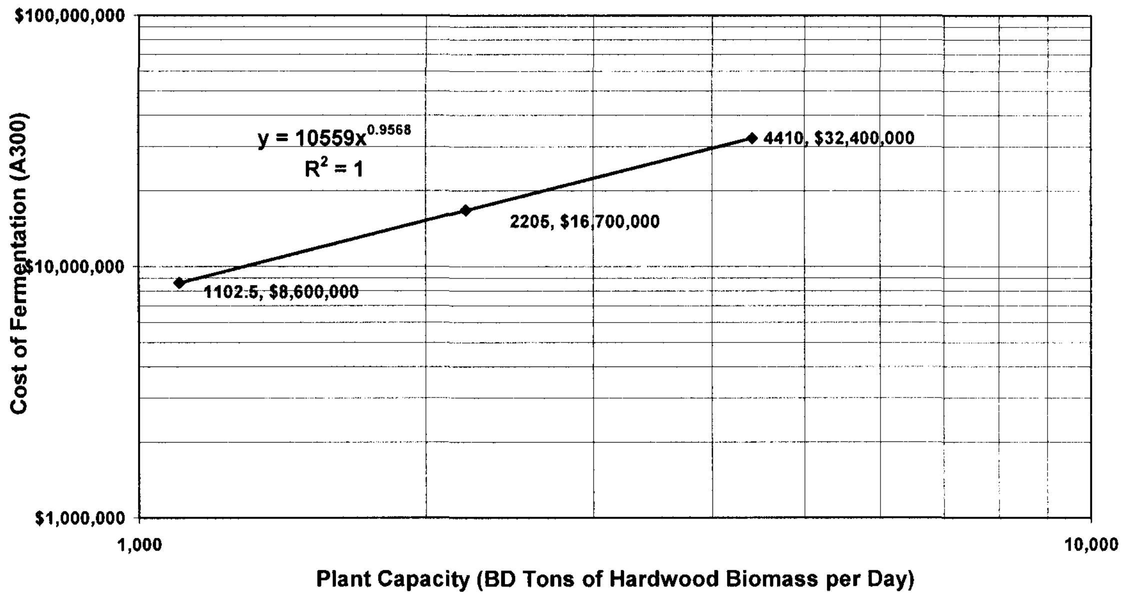

Figure D-8. Total Installed Equipment Cost vs. Hardwood Feedrate: Fermentation (A300)

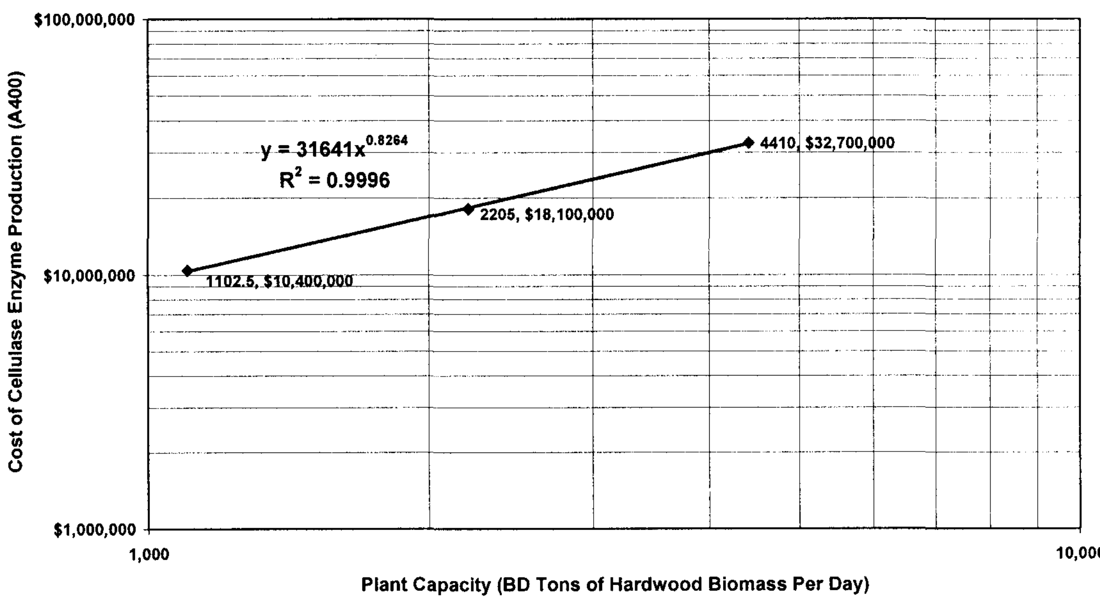

Figure D-9. Total Installed Equipment Cost vs. Hardwood Feedrate: Cellulase Production (A400)

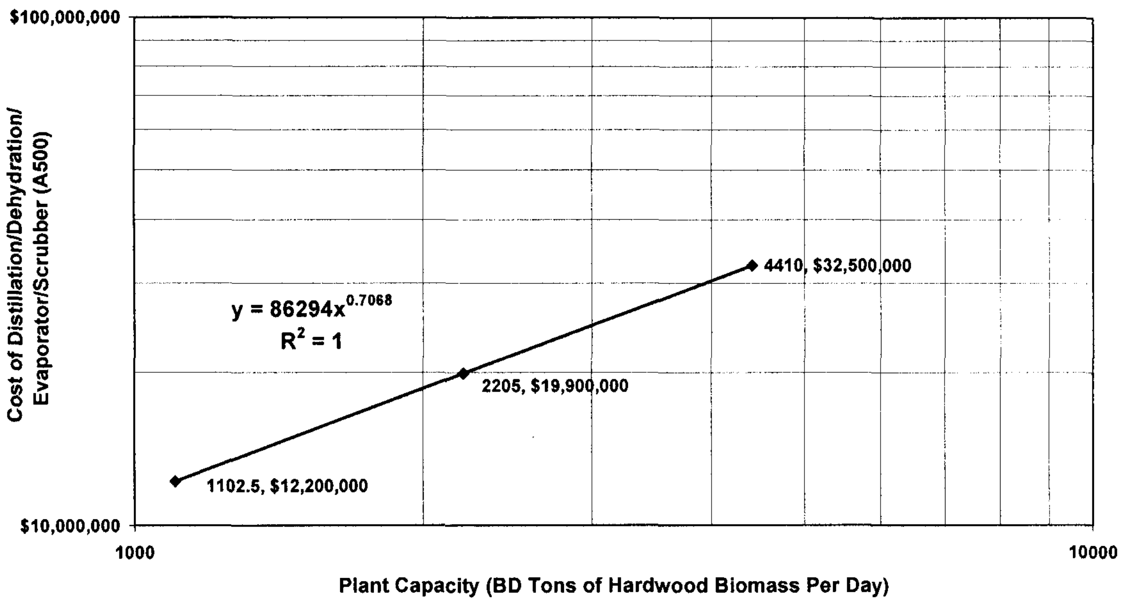

Figure D-10. Total Installed Equipment Cost vs. Hardwood Feedrate:
Distillation/Dehydration/Evaporator/Scrubber(A500)

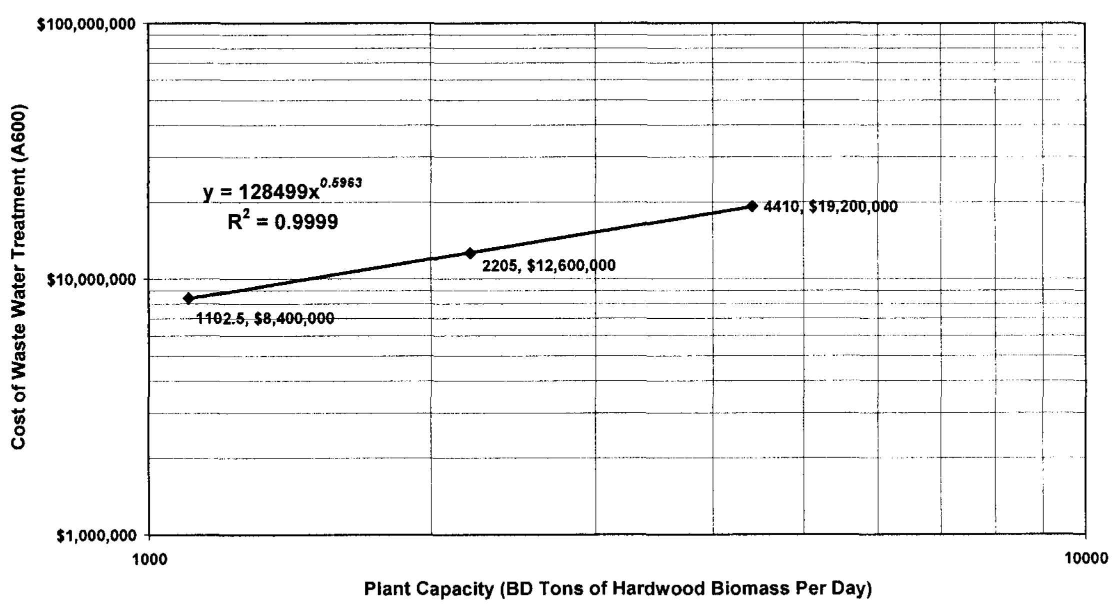

Figure D-11. Total Installed Equipment Cost vs. Hardwood Feedrate:
Waste Water Treatment (A600)

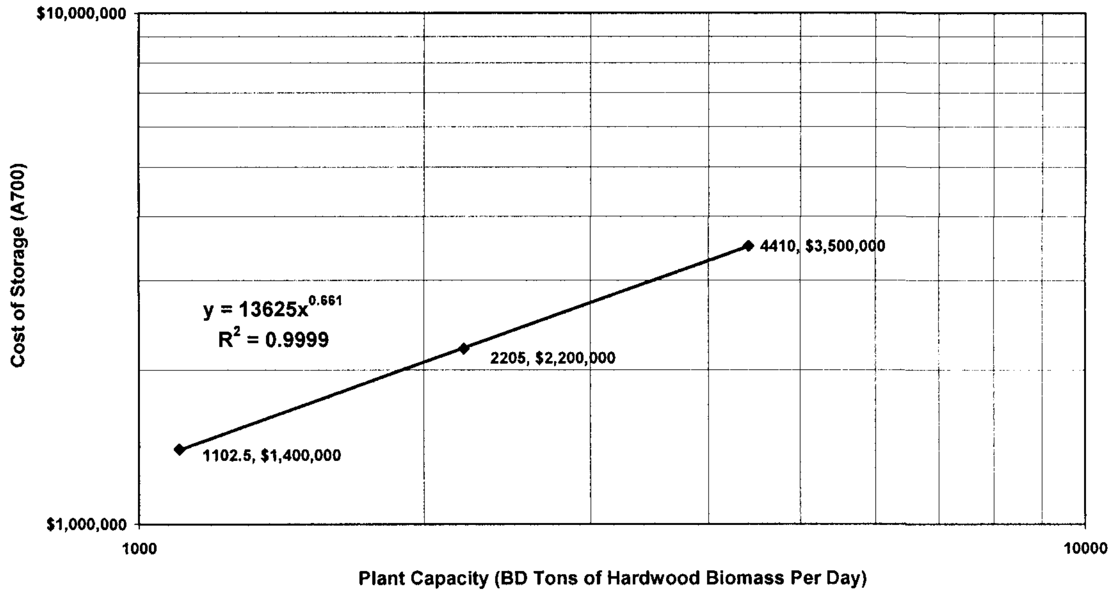

Figure D-12. Total Installed Equipment Cost vs. Hardwood Feedrate: Storage (A700)

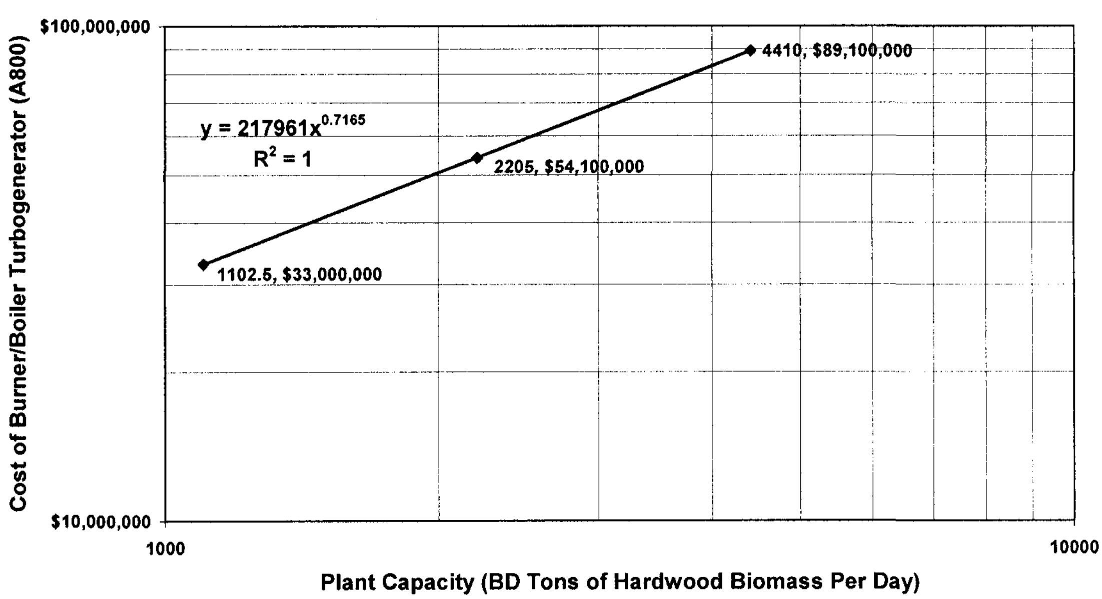

Figure D-13. Total Installed Equipment Cost vs. Hardwood Feedrate: Burner, Boiler, and Turbogenerator (A800)

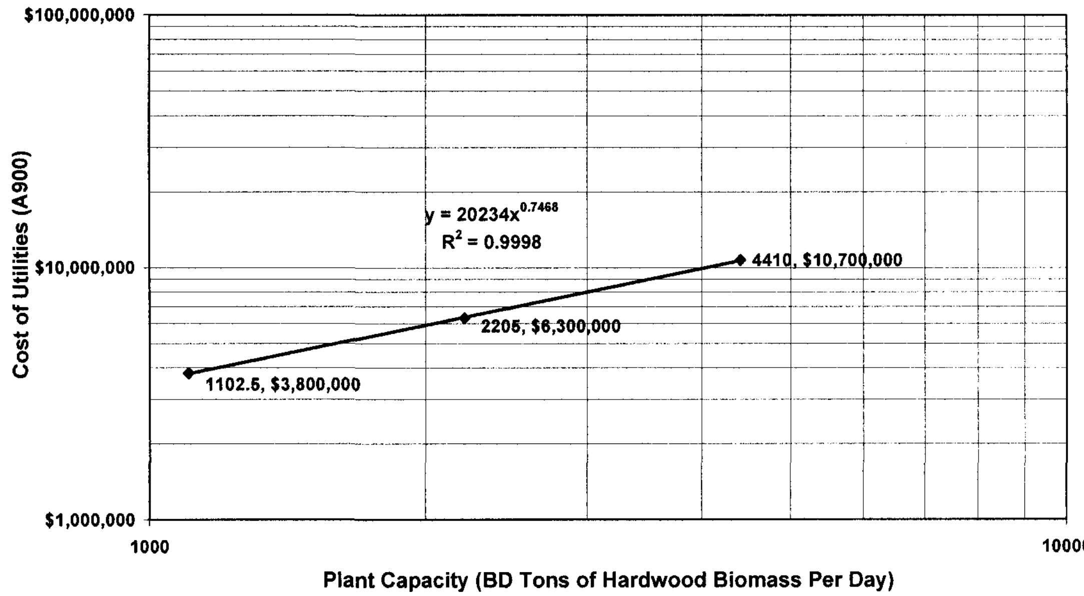

Figure D-14. Total Installed Equipment Cost vs. Hardwood Feedrate: Utilities (A900)

Appendix E

YEARLY OPERATING COSTS

Table E-1
Raw Materials Summary for the Greenfield Plant and Co-Location Case A

Raw Material	1996 Base Case MM$/yr	Inorganic Chemical Index 1996	Inorganic Chemical Index 2005 (Projected)	2005 Base Case MM$/yr	2005 Double Capacity MM$/yr	2005 Half Capacity MM$/yr
Biomass Feedstock	19.31	1996 $25/ton	2005 $60/ton	46.34	92.69	23.17
Sulfuric Acid	0.41	119.5	131.2	0.45	0.90	0.23
Lime	0.44	119.5	131.2	0.48	0.97	0.24
Ammonia	2.2	119.5	131.2	2.42	4.83	1.21
Corn Steep Liquor	2.63	119.5	131.2	2.89	5.77	1.44
Nutrients	0.43	119.5	131.2	0.47	0.94	0.24
Ammonium Sulfate	0.16	119.5	131.2	0.18	0.35	0.09
Antifoam (Corn Oil)	1.01	119.5	131.2	1.11	2.22	0.55
WWT Nutrients	0.45	119.5	131.2	0.49	0.99	0.25
BFW Chemicals	0.01	119.5	131.2	0.01	0.02	0.01
CW Chemicals	0.1	119.5	131.2	0.11	0.22	0.05
WWT Chemicals	0.03	119.5	131.2	0.03	0.07	0.02
Make-up Water	0.45			0.45	0.90	0.23
Diesel	0.48	$0.407/gallon	$1.778/gallon	2.10	4.19	1.05
Ash Disposal	0.19	1996($20/Mt)	2005($40/Mt)	0.38	0.76	0.19
Gypsum Disposal	0.42	1996($20/Mt)	2005($40/Mt)	0.84	1.68	0.42
TOTAL	28.72			58.75	117.50	29.38

Table E-2
Raw Materials Summary for Co-Location Case B and Co-Location Case C

Raw Material	1996 Base Case MM$/yr	Inorganic Chemical Index 1996	Inorganic Chemical Index 2005 (Projected)	2005 Base Case MM$/yr	2005 Double Capacity MM$/yr	2005 Half Capacity MM$/yr
Biomass Feedstock	19.31	1996 $25/ton	2005 $60/ton	46.34	92.69	23.17
Cellulase	0			4.60	9.20	2.30
Sulfuric Acid	0.41	119.5	131.2	0.45	0.90	0.23
Lime	0.44	119.5	131.2	0.48	0.97	0.24
Ammonia	2.2	119.5	131.2	2.17	4.34	1.08
Corn Steep Liquor	2.63	119.5	131.2	2.89	5.77	1.44
Nutrients	0.43	119.5	131.2	0.00	0.00	0.00
Ammonium Sulfate	0.16	119.5	131.2	0.18	0.35	0.09
Antifoam (Corn Oil)	1.01	119.5	131.2	0.00	0.00	0.00
WWT Nutrients	0.45	119.5	131.2	0.49	0.99	0.25
BFW Chemicals	0.01	119.5	131.2	0.01	0.02	0.01
CW Chemicals	0.1	119.5	131.2	0.11	0.22	0.05
WWT Chemicals	0.03	119.5	131.2	0.03	0.07	0.02
Make-up Water	0.45			0.45	0.90	0.23
Diesel	0.48	$0.407/gallon	$1.778/gallon	2.10	4.19	1.05
Ash Disposal	0.19	1996($20/Mt)	2005($40/Mt)	0.38	0.76	0.19
Gypsum Disposal	0.42	1996($20/Mt)	2005($40/Mt)	0.84	1.68	0.42
TOTAL	28.72			61.52	123.05	30.76

Table E-3
Cost of Labor Summary for All Plant Cases

Job Description	Salary	Number	Total Cost 1998	Labor Index 1998	Labor Index 2005 Projected	Total Cost 2005 All Plant Sizes
Plant Manager	$80,000	1	$80,000	17.17	19.90	$92,720
Plant Engineer	$65,000	1	$65,000	17.17	19.90	$75,335
Maintenance Supervisor	$60,000	1	$60,000	17.17	19.90	$69,540
Lab Manager	$50,000	1	$50,000	17.17	19.90	$57,950
Shift Supervisor	$37,000	5	$185,000	17.17	19.90	$214,415
Lab Technician	$25,000	2	$50,000	17.17	19.90	$57,950
Maintenance Technician	$28,000	8	$224,000	17.17	19.90	$259,616
Shift Operators	$25,000	20	$500,000	17.17	19.90	$579,499
Yard Employees	$20,000	8	$160,000	17.17	19.90	$185,440
General Manager	$100,000	1	$100,000	17.17	19.90	$115,900
Clerks & Secretaries	$20,000	5	$100,000	17.17	19.90	$115,900
Total Salaries			**$1,574,000**			**$1,824,263**

Table E-5
Yearly Operating Costs (Co-Location A)

	Yearly Operating Costs ($Millions)		
Yearly Costs	Base Case Plant 2005 (2205 Tons/Day)	Double Capacity Plant 2005 (4410 Tons/Day)	Half Capacity Plant 2005 (1102.5 Tons/Day)
Raw Materials	58.8	117.5	29.4
Total Salaries	1.8	1.8	1.8
Overhead/Maintenance	1.1	1.1	1.1
Maintenance	2.3	3.8	1.4
Insurance & Taxes	1.9	3.1	1.1
Recovery of Capital	36.2	61.2	21.8
TOTAL	102.0	188.6	56.6
Cost Per Gallon of Ethanol Produced	$1.95	$1.81	$2.17

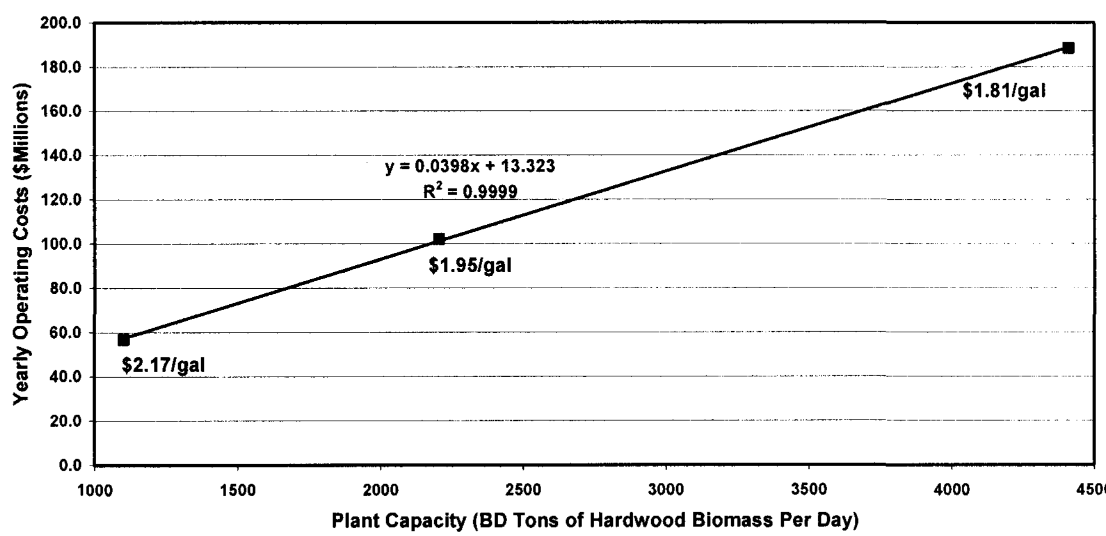

Figure E-2. Yearly Operating Cost vs. Plant Capacity (Co-Location A)

Table E-6
Yearly Operating Costs (Co-Location B)

	Yearly Operating Costs ($Millions)		
Yearly Costs	Base Case Plant 2005 (2205 Tons/Day)	Double Capacity Plant 2005 (4410 Tons/Day)	Half Capacity Plant 2005 (1102.5 Tons/Day)
Raw Materials	61.5	123.0	30.8
Total Salaries	1.8	1.8	1.8
Overhead/Maintenance	1.1	1.1	1.1
Maintenance	1.9	3.2	1.2
Insurance & Taxes	1.6	2.6	0.9
Recovery of Capital	30.3	50.6	18.5
TOTAL	98.2	182.4	54.2
Cost Per Gallon of Ethanol Produced	$1.88	$1.75	$2.08

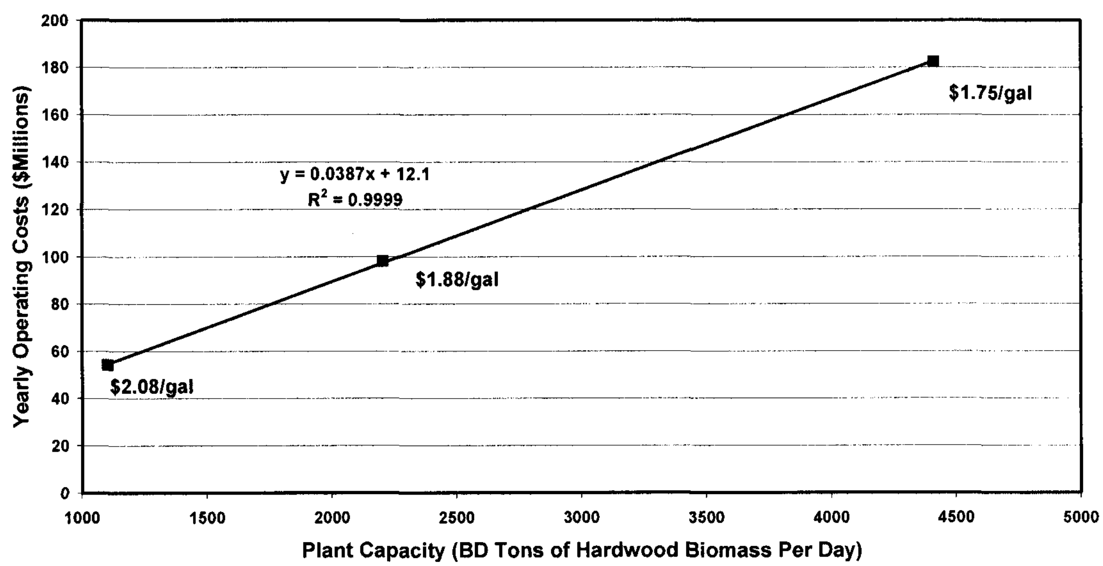

Figure E-3. Yearly Operating Cost vs. Plant Capacity (Co-Location B)

Table E-7
Yearly Operating Costs (Co-Location C)

	Yearly Operating Costs ($Millions)		
Yearly Costs	Base Case Plant 2005 (2205 Tons/Day)	Double Capacity Plant 2005 (4410 Tons/Day)	Half Capacity Plant 2005 (1102.5 Tons/Day)
Raw Materials	61.5	123.0	30.8
Total Salaries	1.8	1.8	1.8
Overhead/Maintenance	1.1	1.1	1.1
Maintenance	1.5	2.6	0.9
Insurance & Taxes	1.3	2.2	0.8
Recovery of Capital	24.7	42.0	14.7
TOTAL	91.9	172.7	50.0
Cost Per Gallon of Ethanol Produced	$1.76	$1.65	$1.92

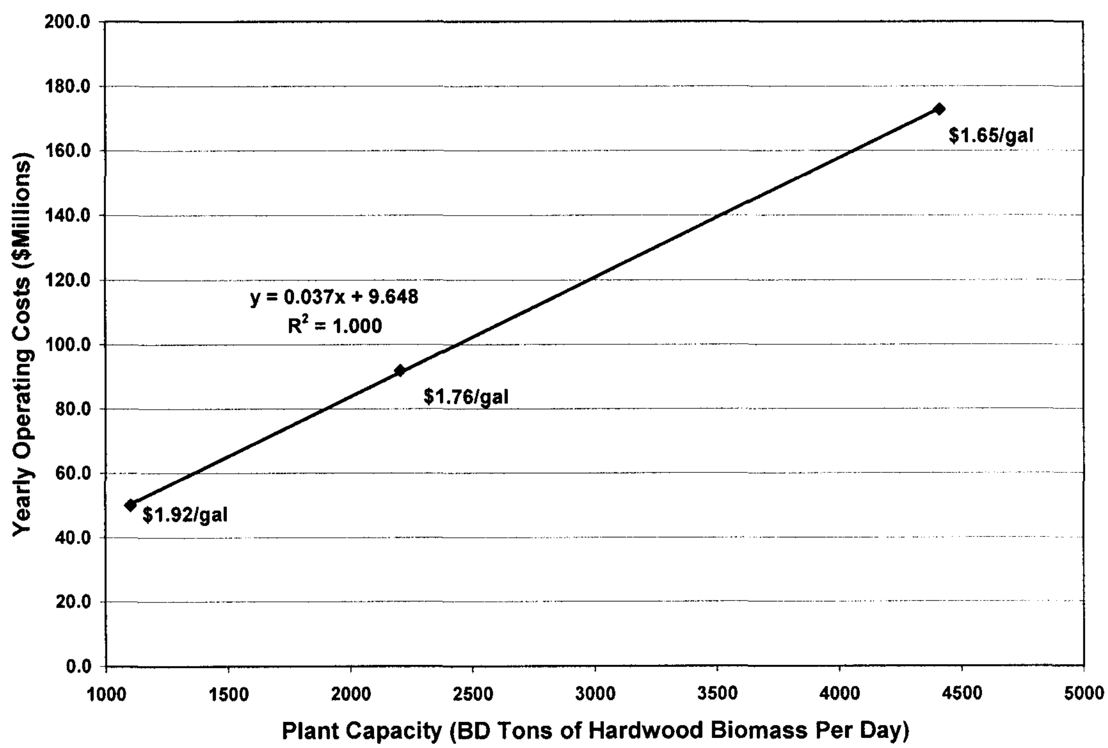

Figure E-4. Yearly Operating Cost vs. Plant Capacity (Co-Location Case C)

Appendix F

SUMMARY OF PROFITABILITY ANALYSIS

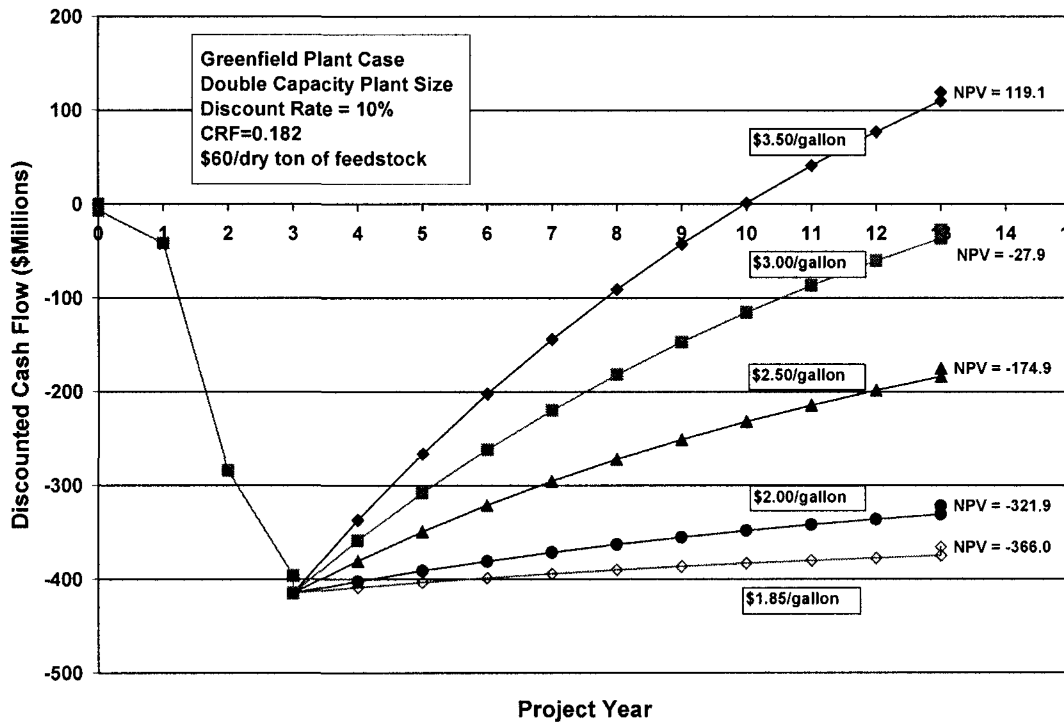

Figure F-1. Net Present Value at Varying Ethanol Selling Prices for the Double Capacity Plant Size (4,410 BD Tons/Day) (Greenfield Plant)

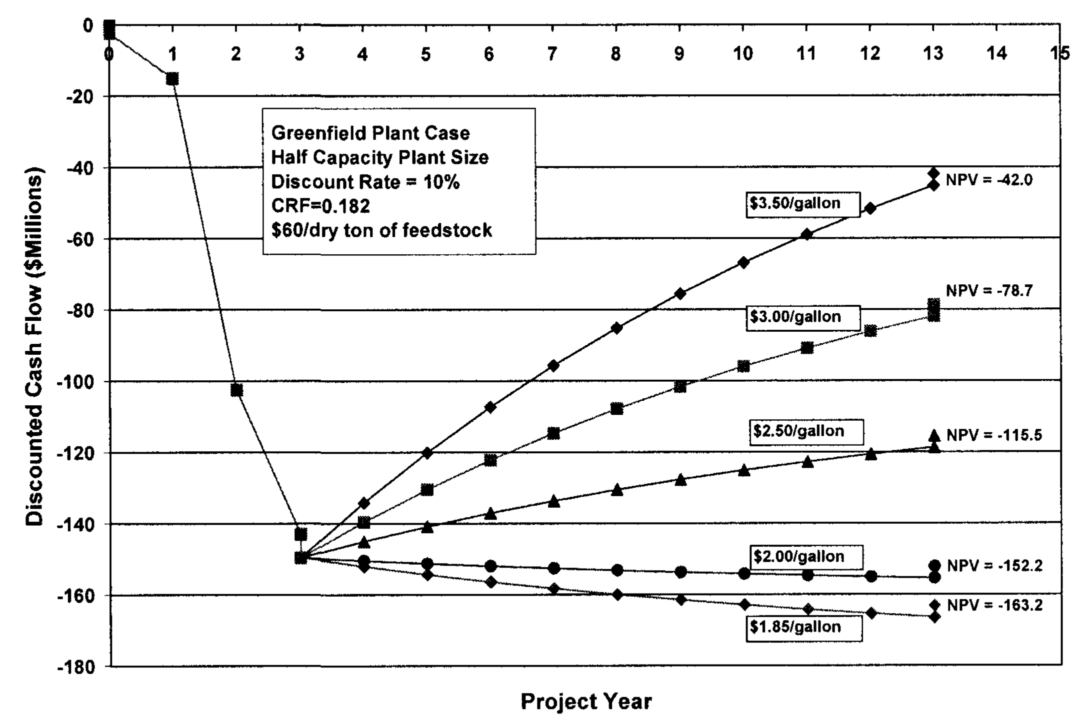

Figure F-2. Net Present Value at Varying Ethanol Selling Prices for the Half Capacity Plant Size (1,103 BD Tons/Day) (Greenfield Plant)

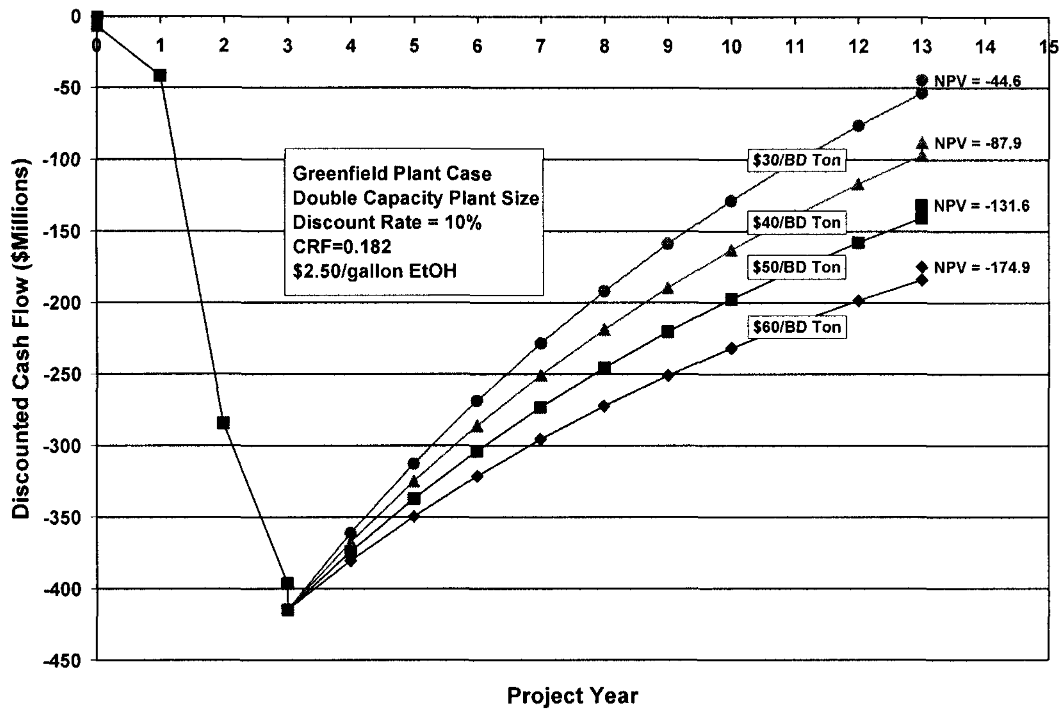

Figure F-3. Net Present Value at Varying Cost of Hardwood for the Double Capacity Plant Size (4,410 BD Tons/Day) (Greenfield Plant)

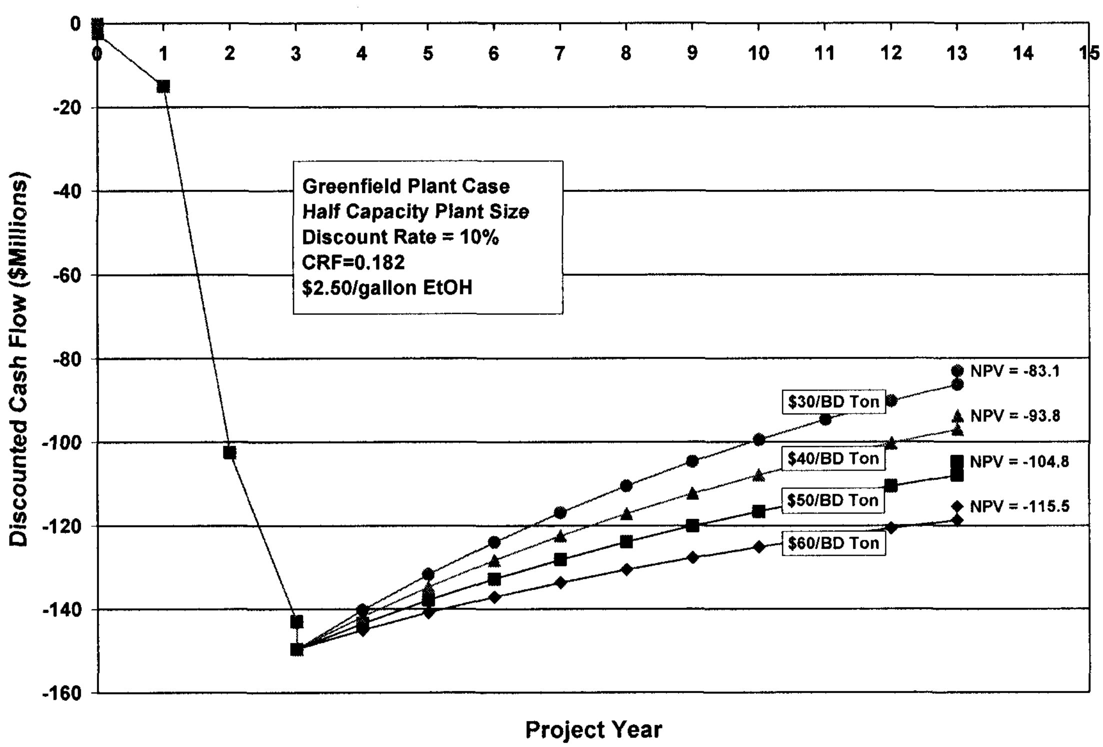

Figure F-4. Net Present Value at Varying Cost of Hardwood for the Half Capacity Plant Size (1,103 BD Tons/Day) (Greenfield Plant)

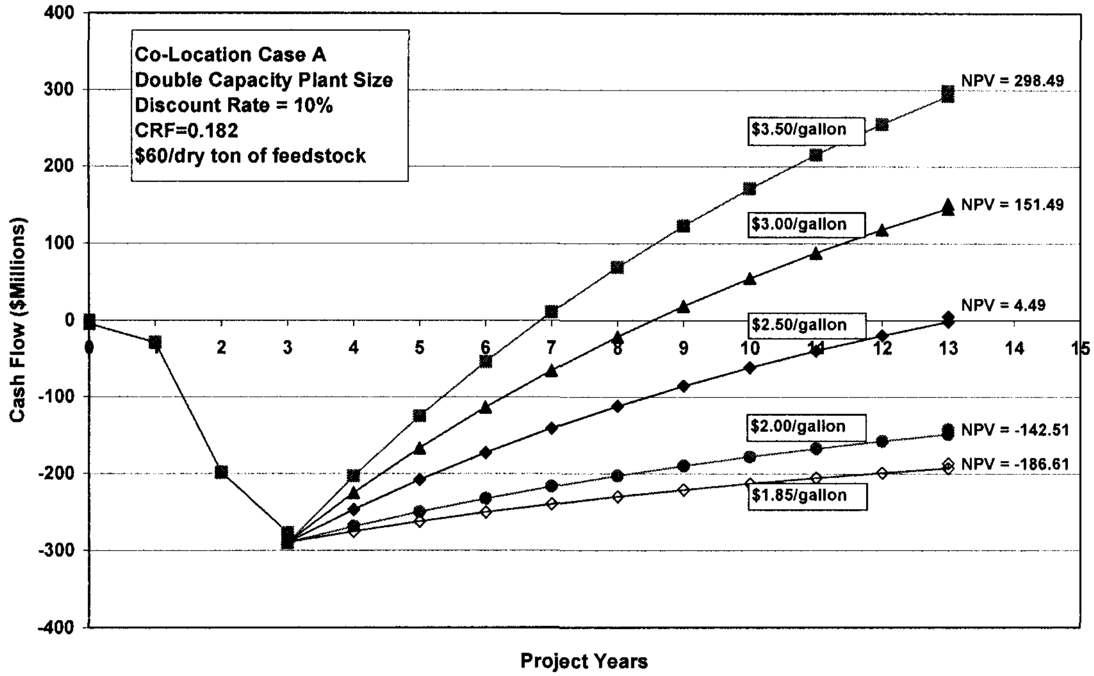

Figure F-5. Net Present Value at Varying Ethanol Selling Prices for Co-Location Case A at the Double Capacity Plant Size (4,410 BD Tons/Day)

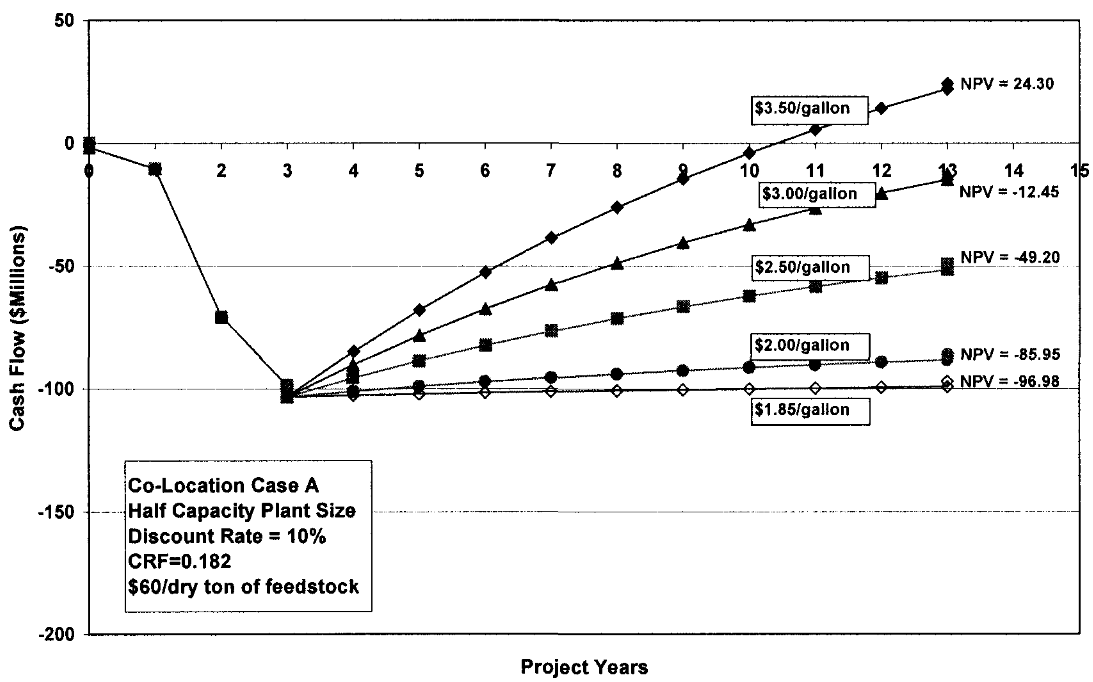

Figure F-6. Net Present Value at Varying Ethanol Selling Prices for Co-Location Case A at the Half Capacity Plant Size (1,103 BD Tons/Day)

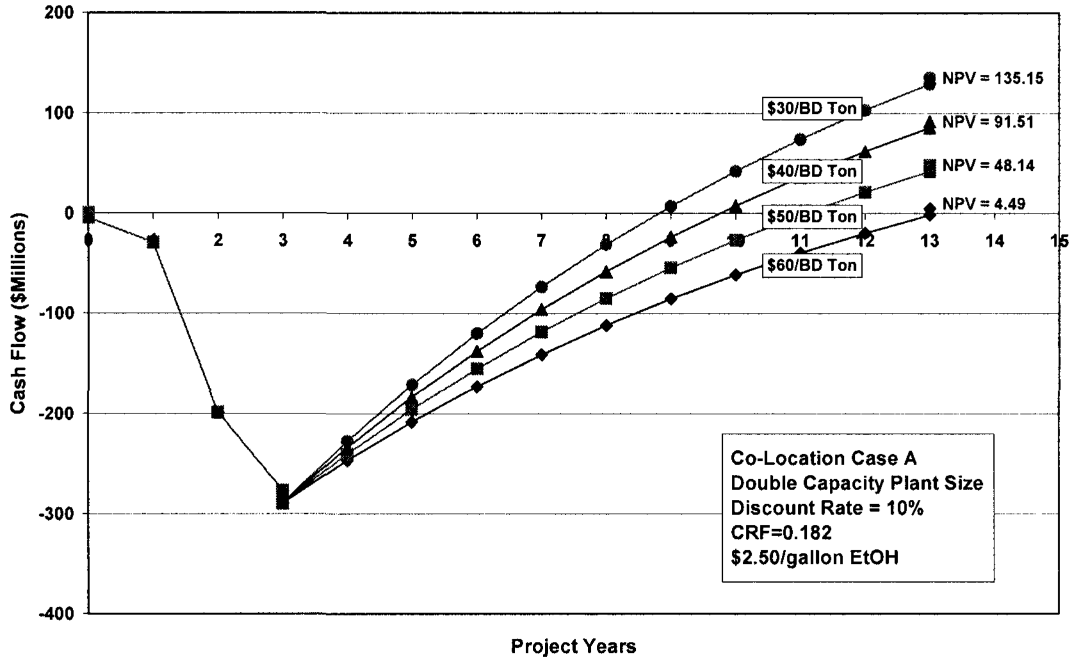

Figure F-7. Net Present Value at Varying Hardwood Cost for Co-Location Case A at the Double Capacity Plant Size (4,410 BD Tons/Day)

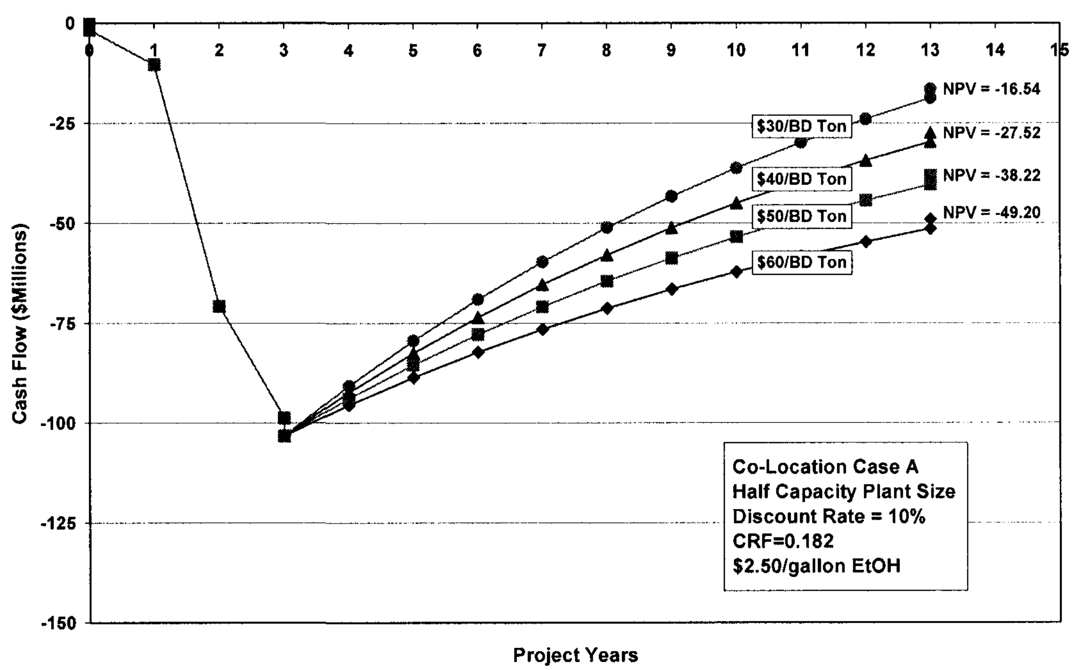

Figure F-8. Net Present Value at Varying Hardwood Cost for Co-Location Case A at the Half Capacity Plant Size (1,103 BD Tons/Day)

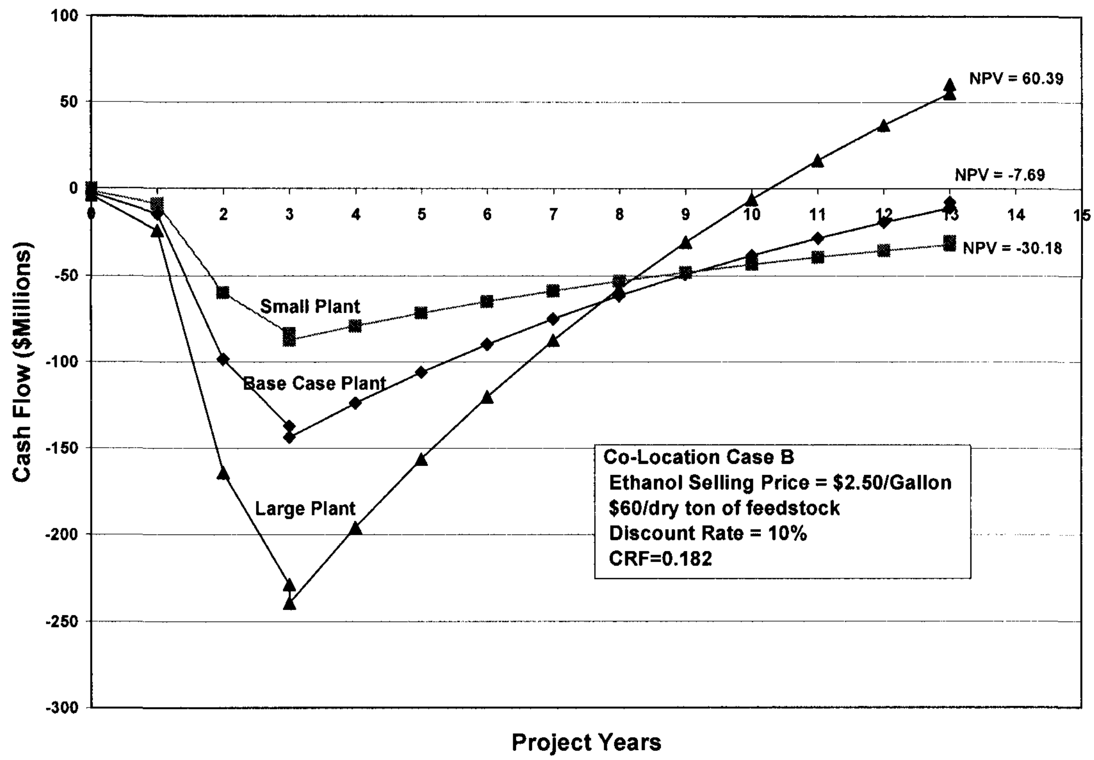

Figure F-9. Net Present Value for Co-Location Case B
(Discount Rate of 10%)

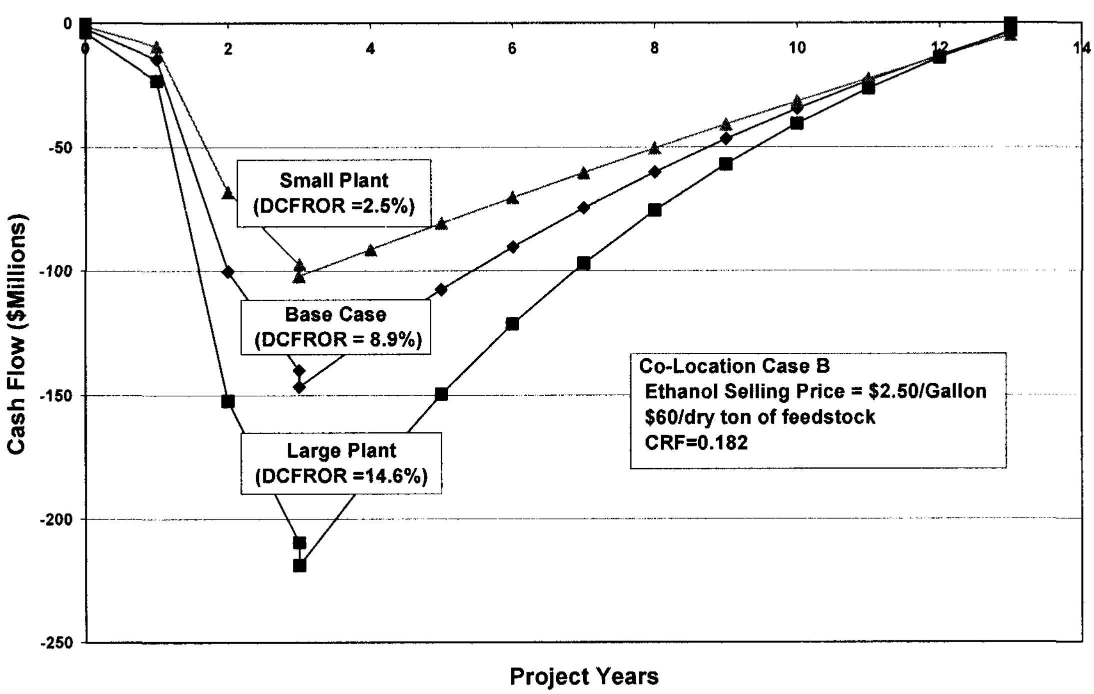

Figure F-10. Discounted Cash Flow Rate of Return for Co-Location Case B

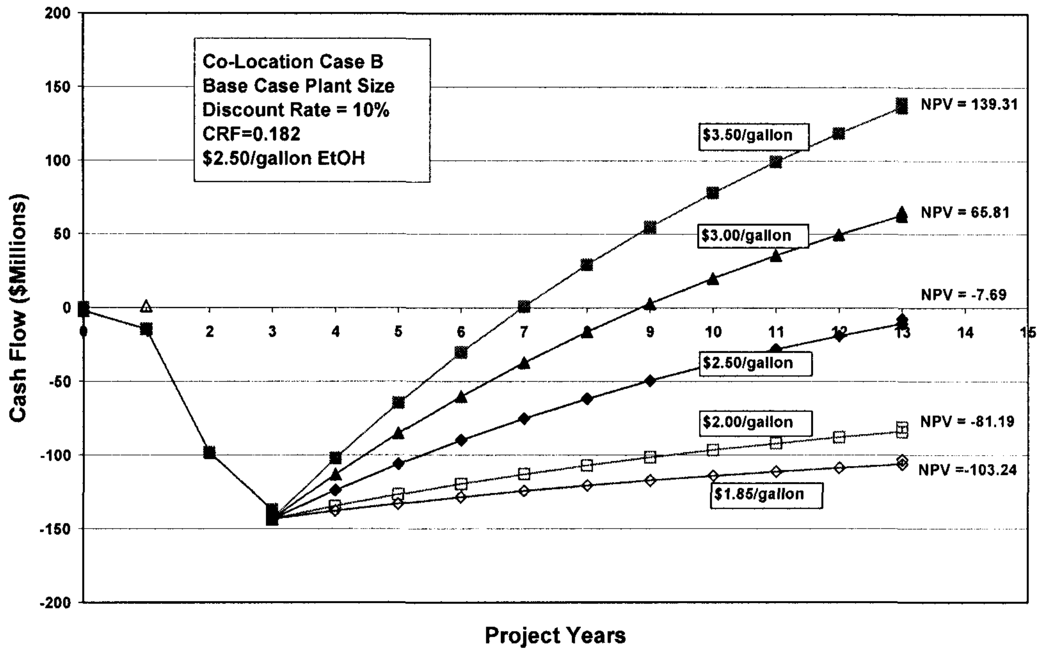

Figure F-11. Effect of Ethanol Selling Prices for Co-Location Case B (Base Case Plant Size of 2205 BD Tons Wood/Day and 52 Million Gallons per Year)

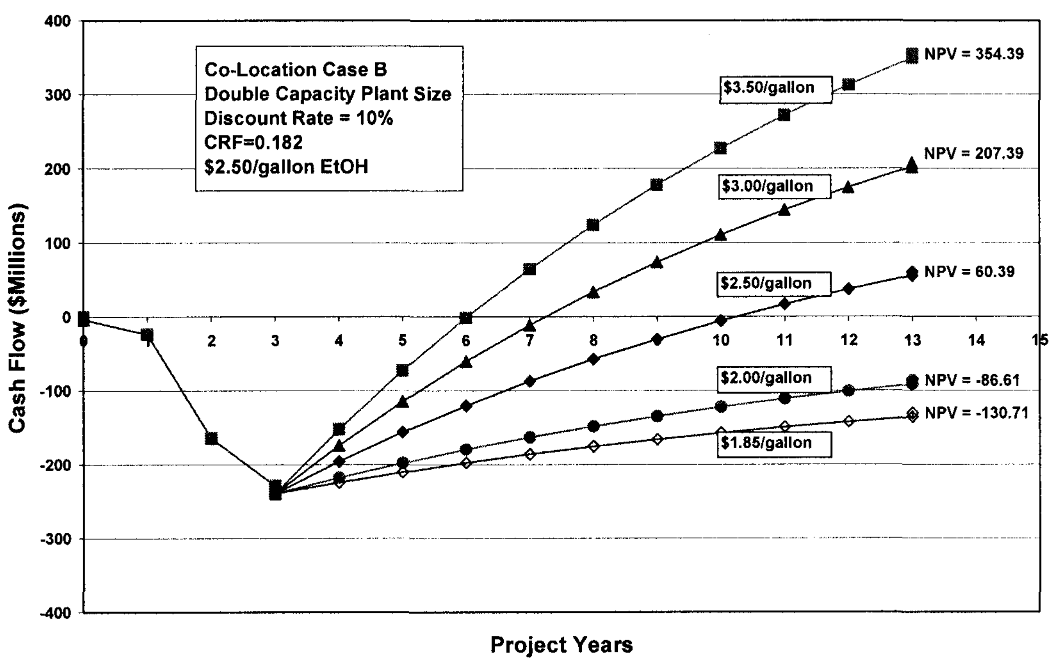

Figure F-12. Net Present Value at Varying Ethanol Selling Prices for Co-Location Case B at the Double Capacity Plant Size (4,410 BD Tons/Day)

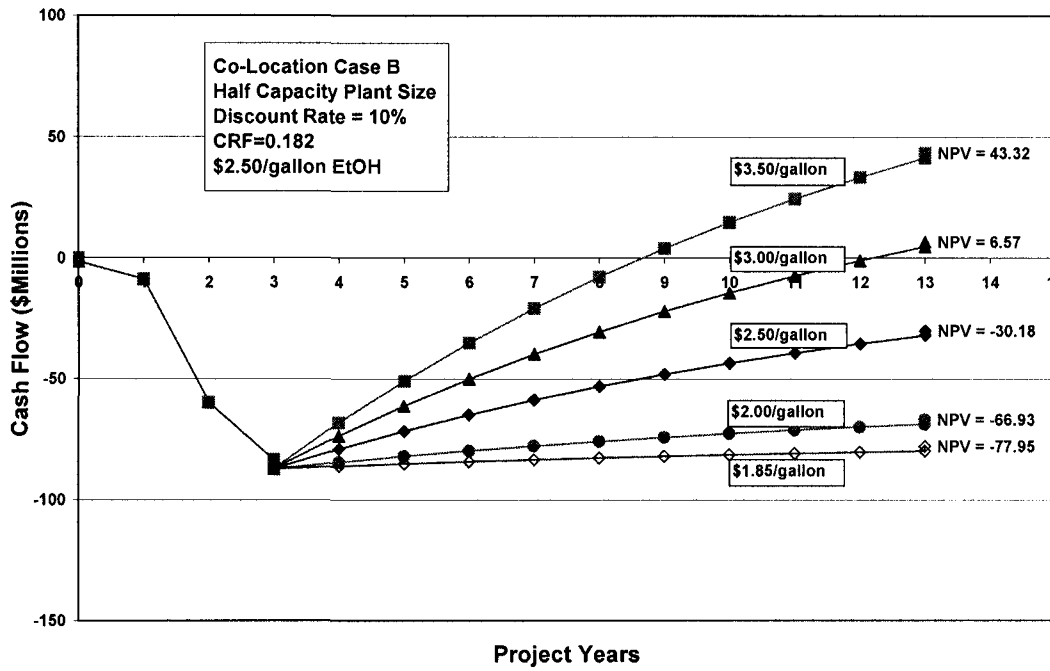

Figure F-13. Net Present Value at Varying Ethanol Selling Prices for Co-Location Case B at the Half Capacity Plant Size (1,103 BD Tons/Day)

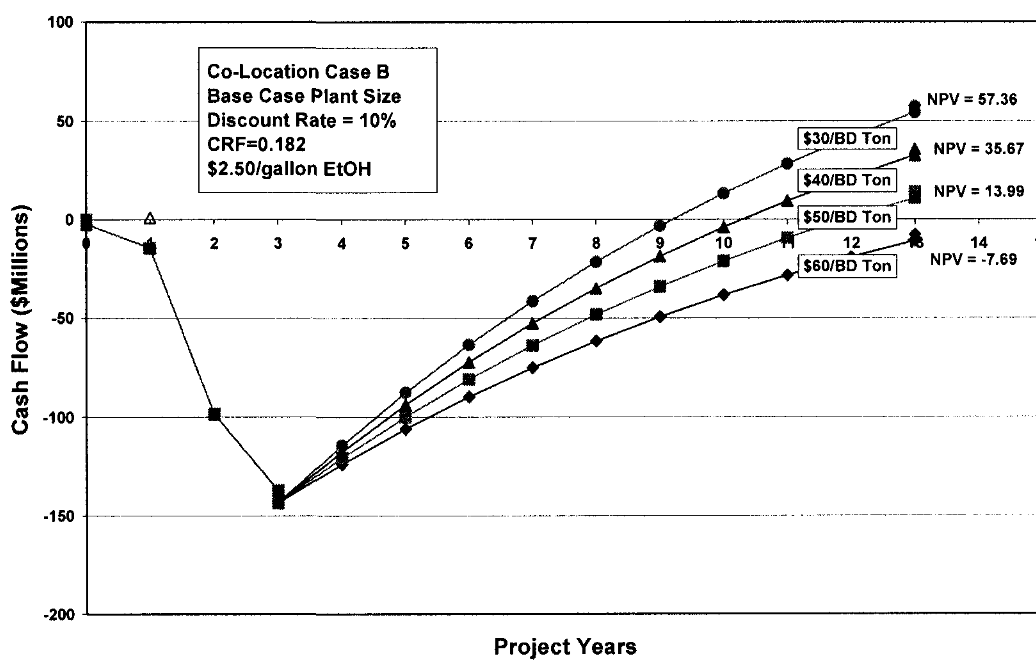

Figure F-14. Net Present Value at Varying Hardwood Cost for Co-Location Case B at the Base Case Plant Size (2,205 BD Tons/Day)

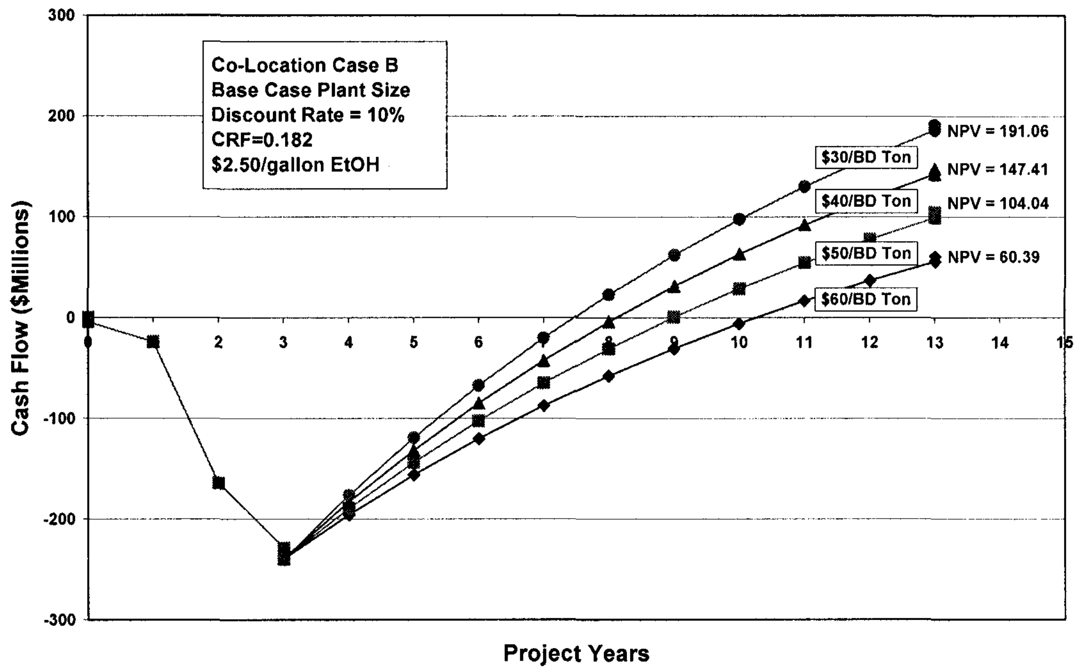

Figure F-15. Net Present Value at Varying Hardwood Cost for Co-Location Case B at the Double Capacity Plant Size (4,410 BD Tons/Day)

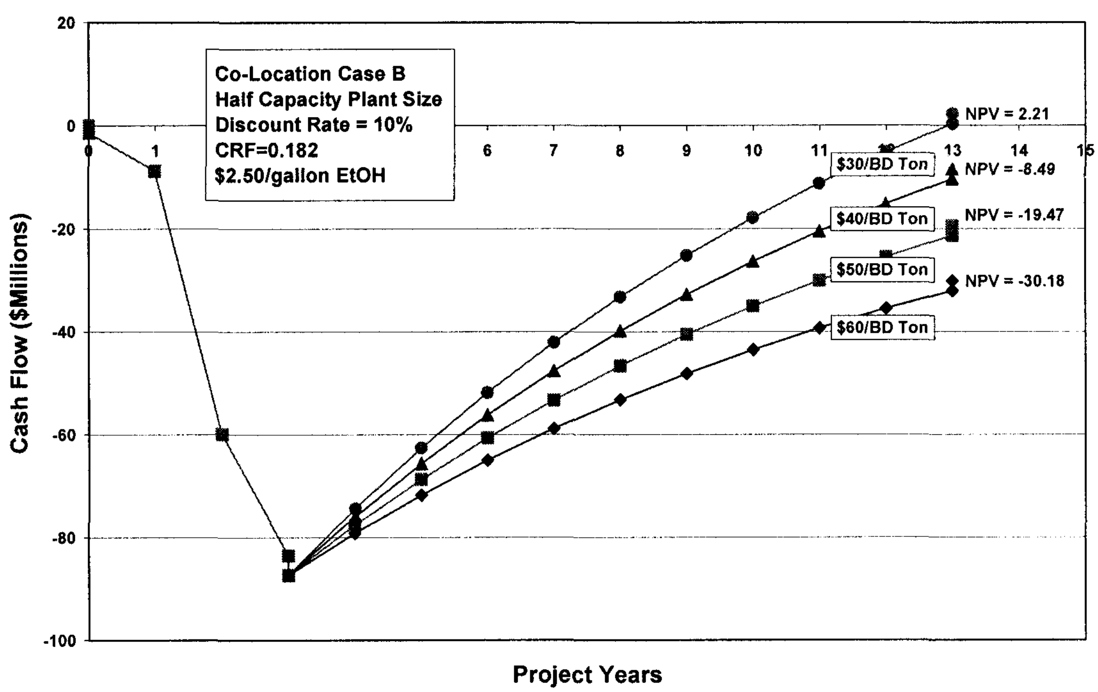

Figure F-16. Net Present Value at Varying Hardwood Cost for Co-Location Case B at the Base Case Plant Size (1,103 BD Tons/Day)

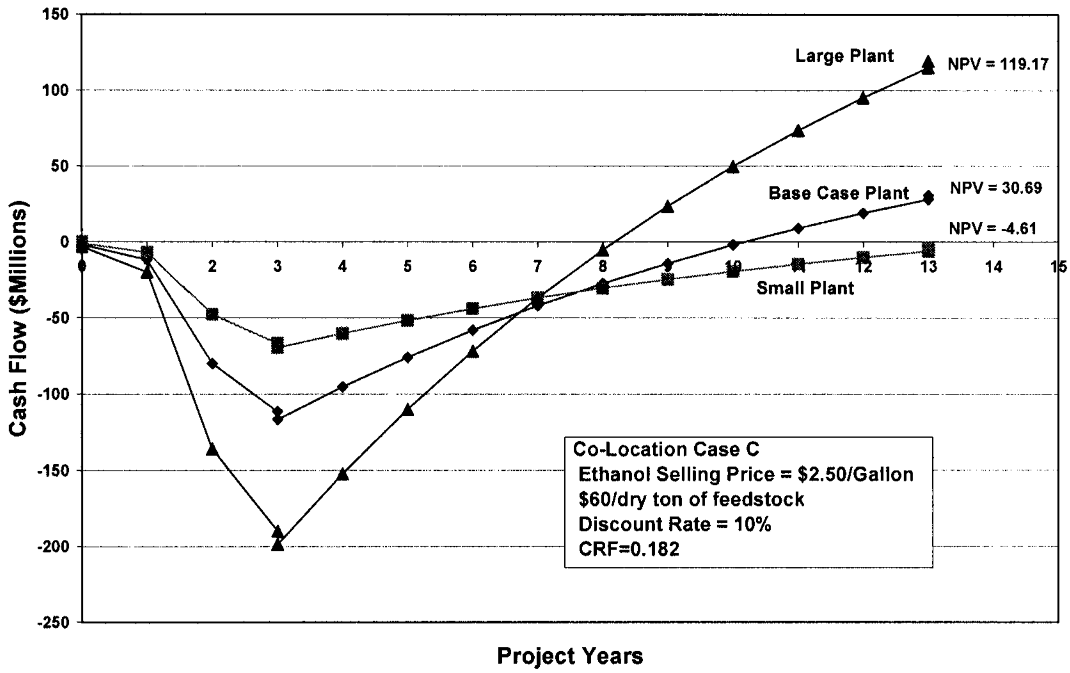

Figure F-17. Net Present Value at a Discount Rate of 10% (Co-Location Case C)

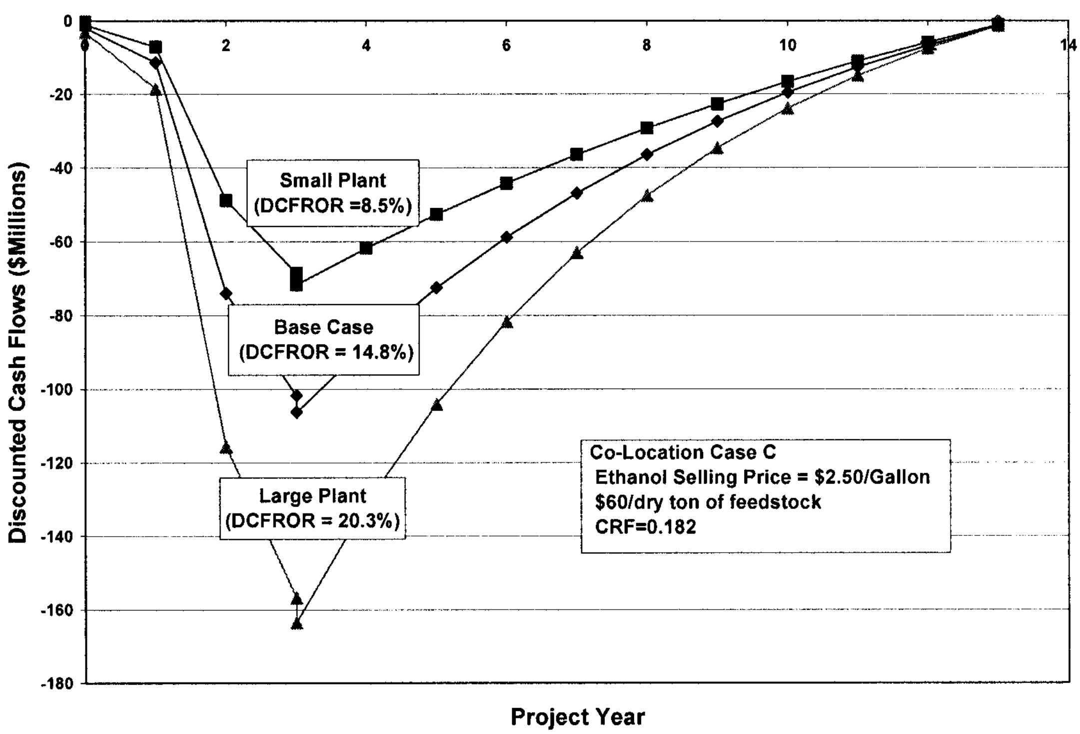

Figure F-18. Cash Flow at Discounted Cash Flow Rate of Return (Co-Location Case C)

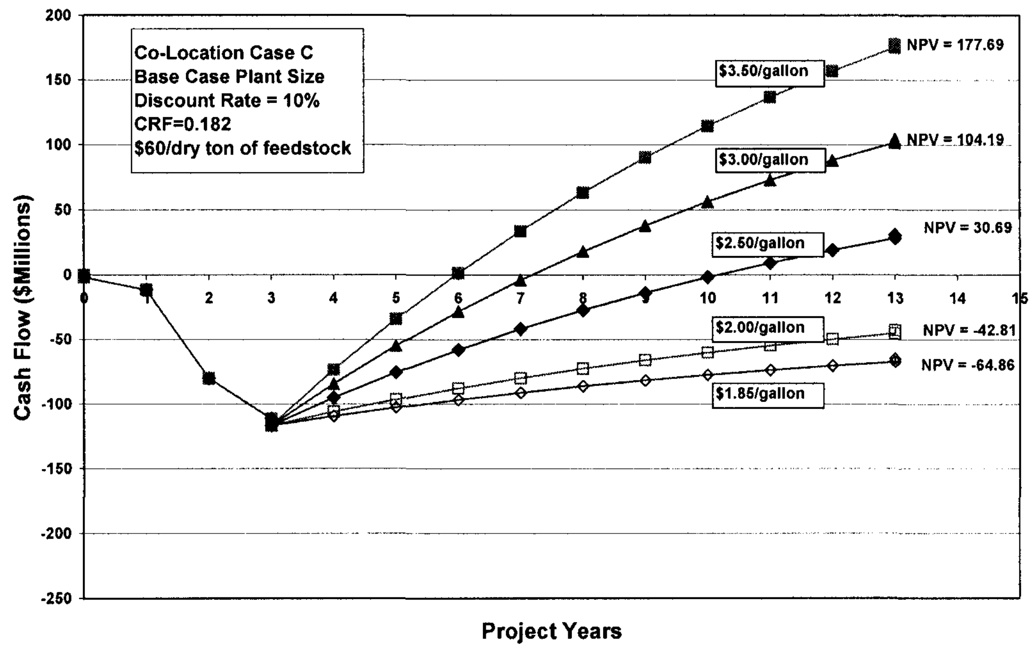

Figure F-19. Net Present Value at Varying Ethanol Selling Prices for Co-Location Case C at the Base Case Plant Size (2,205 BD Tons/Day)

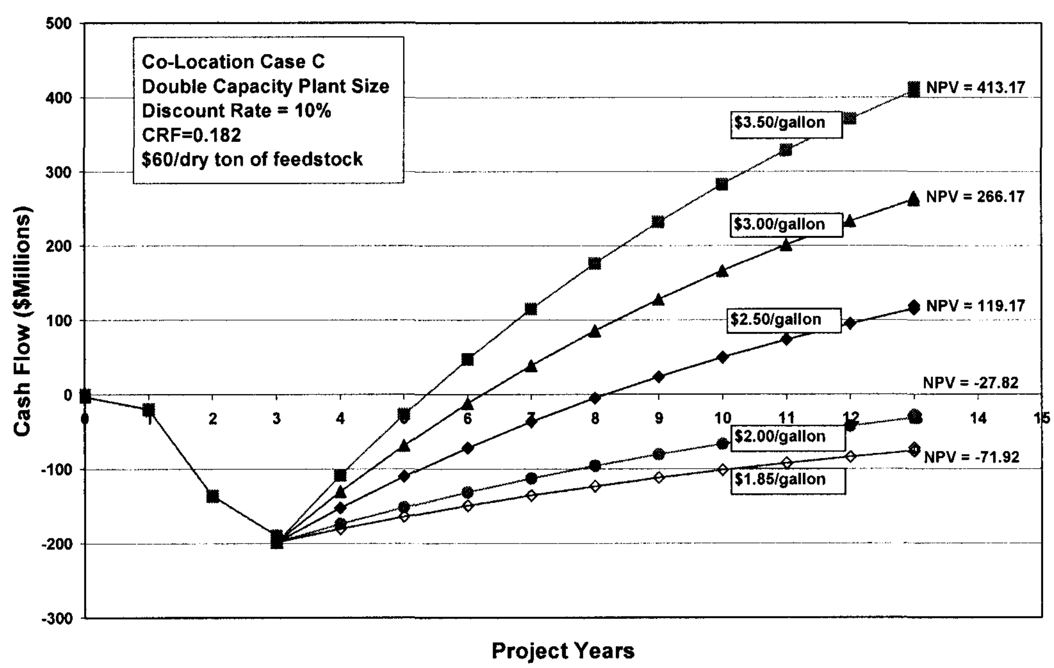

Figure F-20. Net Present at Varying Ethanol Selling Prices for Co-Location Case C at the Double Capacity Plant Size (4,410 BD Tons/Day)

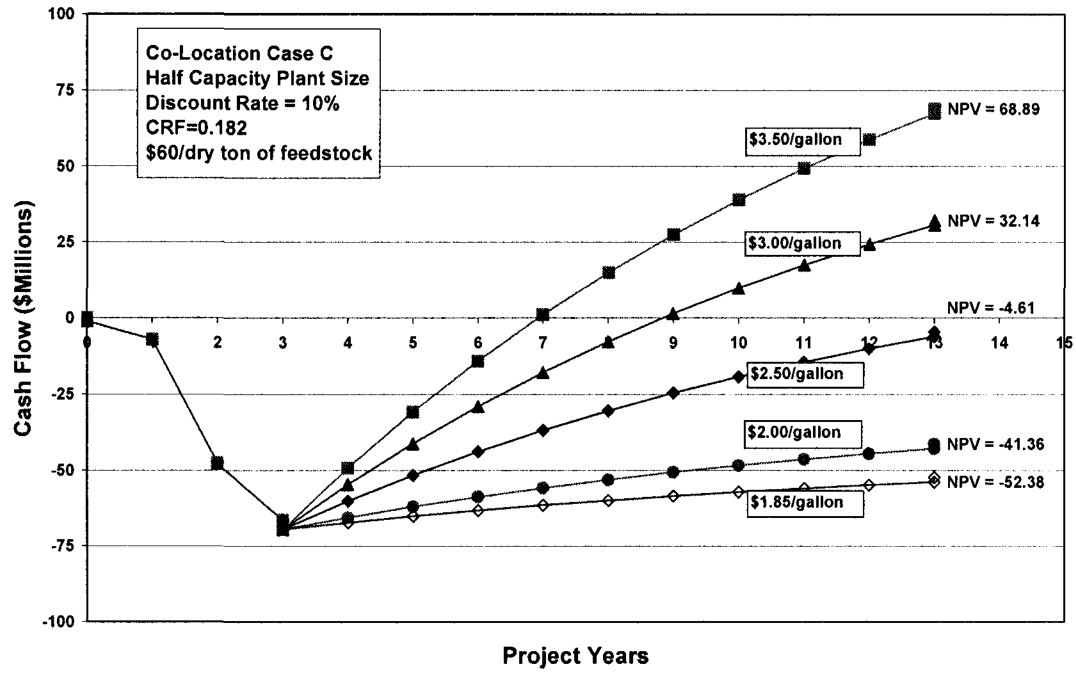

Figure F-21. Net Present at Varying Ethanol Selling Prices for Co-Location Case C at the Half Capacity Plant Size (1,103 BD Tons/Day)

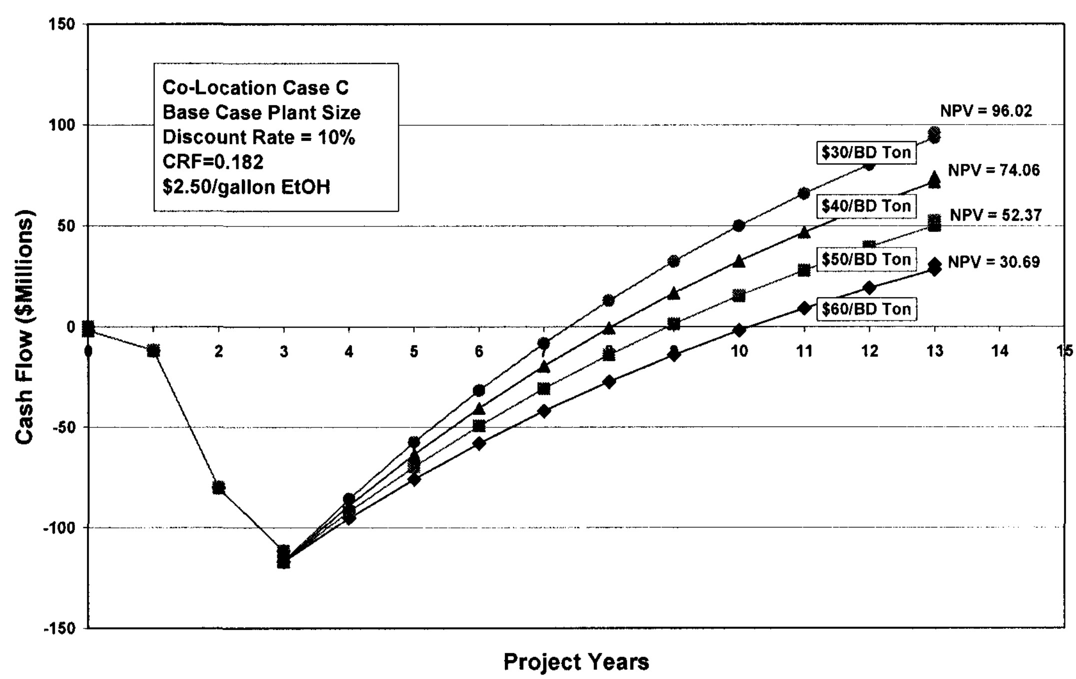

Figure F-22. Net Present Value at Varying Hardwood Cost for Co-Location Case C at the Base Case Plant Size (2,205 BD Tons/Day)

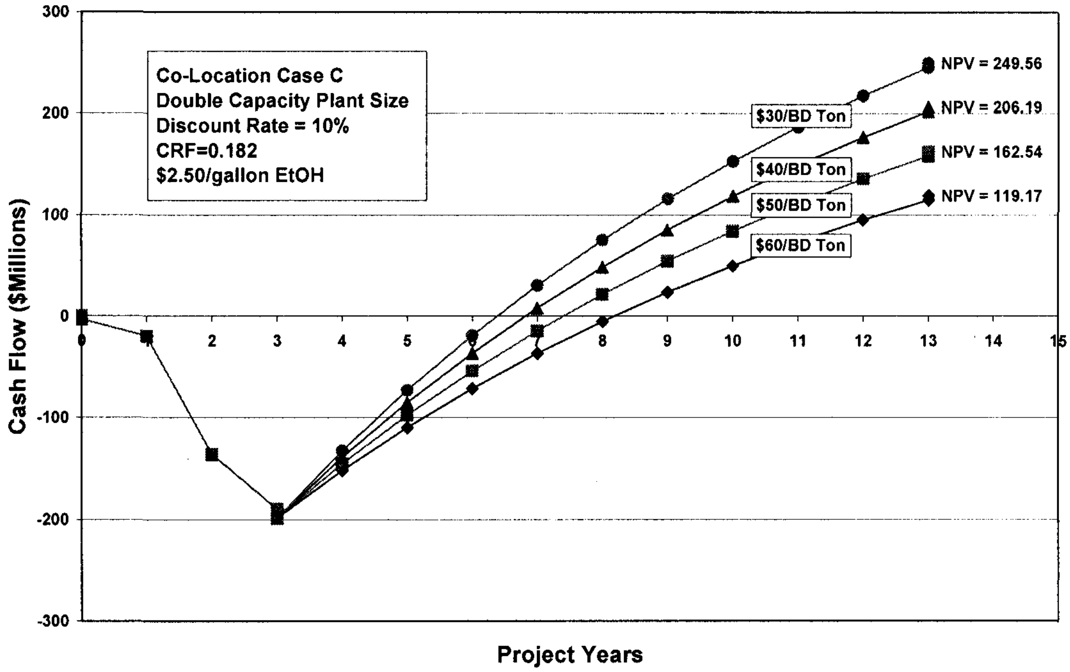

Figure F-23. Net Present at Varying Hardwood Cost for Co-Location Case C at the Double Capacity Plant Size (4,410 BD Tons/Day)

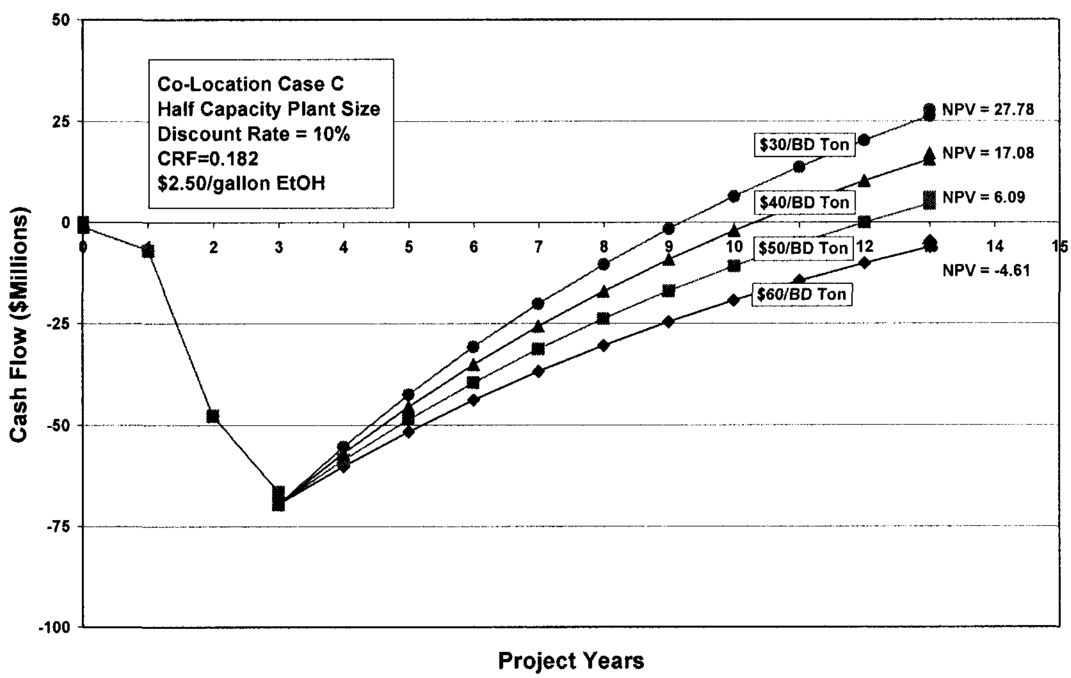

Figure F-24. Net Present at Varying Hardwood Cost for Co-Location Case C at the Base Case Plant Size (1,103 BD Tons/Day)

BIOGRAPHY OF THE AUTHOR

Jay Mitchell was born in Lewiston, Maine on September 22, 1981. He was raised in Jay, ME and graduated as valedictorian from Jay High School in 2000. He received the Pulp and Paper Foundation Scholarship from the University of Maine and entered the Chemical Engineering program in 2000. He graduated from the University of Maine in August, 2004 with a Bachelor's degree in Chemical Engineering and entered the graduate program in the fall of 2004.

After receiving his degree, Jay will be joining Hollingsworth and Vose Co. to begin his career as a process engineer. Jay is a candidate for the Master of Science degree in Chemical Engineering from The University of Maine in December, 2006.