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Techno-Economic Analysis of Biochemical Scenarios for Production of Cellulosic Ethanol

Technical Report
NREL/TP-6A2-46588
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Foreword

The purpose of this techno-economic analysis is to compare a set of biofuel conversion technologies selected for their promise and near-term technical viability. Every effort has been made to make this comparison on an equivalent basis using common assumptions. The process design and parameter value choices underlying this analysis are based on public domain literature only. For these reasons, the results are not indicative of potential performance. Rather they are meant to represent the most likely performance given the current state of public knowledge.

List of Acronyms

AFEX	ammonia fiber explosion (or expansion)
ASPEN	Advanced Simulator For Process Engineering (software)
BTU	British thermal unit
CAFI	Consortium for Applied Fundamentals and Innovation
COD	chemical oxygen demand
DB	declining balance (depreciation)
EtOH	ethanol
EVD	experimentally validated data
FCI	fixed capital investment
FPU	filter paper units
GGE	gallons of gasoline equivalent
IRS	Internal Revenue Service
MACRS	(IRS) modified accelerated cost recovery system
PV	product value
MM	million
MT	metric ton
NREL	National Renewable Energy Laboratory
SSCF	simultaneous saccharification and cofermentation
SSF	simultaneous saccharification and fermentation
ST	short ton
TD&IC	total direct and indirect costs
TCI	total capital investment

Executive Summary

A techno-economic analysis on the production of cellulosic ethanol by fermentation was conducted to understand the viability of liquid biofuel production processes within the next 5-8 years. Initially, 35 technologies were reviewed and a matrix was prepared considering economics, technological soundness and maturity, environmental aspects, process performance, and technical and economic risks. Then, a two-step down selection was performed to choose scenarios to be evaluated in a more detailed economic analysis. In the first screening, the lignocellulosic ethanol process was selected because it is well studied and portions of the process have been tested at pilot scales. In the second screening, seven scenarios of process variations were selected: four variations involved pretreatment (dilute acid, two-stage dilute acid, hot water, and ammonia fiber explosion) and three variations involved downstream processes (pervaporation, separate 5-carbon and 6-carbon sugar fermentation, and on-site enzyme production). Each of these scenarios was examined in detail. Given the time needed for design, construction, and startup of large process plants, plants operating in the 5-8 year timeframe would likely need to be based on recent experimental data. For this work, process designs were constrained to public data published in 2007 or earlier, without projecting for future process improvements. Economic analysis was performed for an "n^{th} plant" (mature technology) to obtain total investment and product value (PV) (defined as value of the product needed for a net present value of zero with a 10% internal rate of return). The final selection among the scenarios was performed primarily based on the PV. Sensitivity analysis was performed on PV to assess the impact of variations in process and economic parameters. Results show that the modeled dilute acid pretreatment process without any downstream process variation had the lowest PV of \$3.40/gal of ethanol (which is \$5.15/gallon of gasoline equivalent, GGE) in 2007 dollars. Sensitivity analysis shows that PV is most sensitive to feedstock and enzyme costs.

The cellulosic ethanol process is a new technology, for which a pioneer plant is expected to be significantly more expensive than the n^{th} plant. To assess the impact of technology maturity on pioneer plant cost, a cost growth analysis was performed following a method documented in a RAND Corporation report. This methodology attempts to incorporate added expenses and start-up time for a new process. There is some subjectivity in choosing the parameters for the pioneer plant analysis, so a range of parameters was used to estimate pioneer plant costs for three scenarios: optimistic, most probable, and pessimistic. The PV obtained from cost-growth analysis is substantially larger for a pioneer plant, increasing from \$3.40/gal (which is \$5.15/GGE), before including added expenses, to \$5.01/gal (\$7.59/GGE), \$5.76/gal (\$8.72/GGE), and \$7.08/gal (\$10.71/GGE) for the optimistic, most probable, and pessimistic scenarios, respectively.

The PV obtained from the 2007 published data is much higher than the market gasoline price. Also, published technological data may not be adequate to accurately project a competitive PV from a commercial plant of 2000 MT/day capacity. However, this analysis identifies some of the more cost intensive operations and areas. The current process can only reach approximately 4.5% ethanol in the fermentor beer, which is a third of what grain ethanol plants are achieving today. The analysis also assumed an enzyme price as \$0.69/gal of ethanol produced. Based on this analysis, we believe that high-performance enzymes at a cheaper price are required and that more research is needed to achieve higher ethanol concentration in the fermentor for this process to compete in the current energy market.

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Methodology

The objective of this study was to select a biochemical conversion process from published literature to produce a second generation liquid biofuel from lignocellulosic biomass. The studies involved several steps defining the objective and scope of the project. They started with a review of various biochemical processes published in journals and reports and then applied two steps of screening to make the final selection. The cost analysis is performed assuming n^{th} plant technology (mature technology) and then subjected to cost growth analysis for a pioneer plant. The following steps explain the basic methodology for performing the techno-economic analysis:

- Search literature for technologies under consideration
- Perform down selection with criteria to narrow to a few scenarios
- Design process models using AspenPlus process simulation using available experimental data
- Size and cost equipment using traditional methods such as literature references and vendor quotations
- Determine project investments and perform discounted cash flow analysis
- Adjust sensitivity parameters and document results
- Perform pioneer plant cost growth and performance analysis.

Summary of the Down-Selection Process

Thirty-five biochemical fuel production technologies were initially selected for consideration (Table I-1). Each technology option was evaluated based on economic and technical feasibility, environmental performance, and uncertainty criteria to determine, at a high level, the technologies with the greatest overall promise in the 5-8 year timeframe. For a plant to be operating at commercial levels in this timeframe, the plant would likely need to be designed based on current data because of the time needed for design, construction, and startup of a plant. A matrix of techno-economic studies was developed for "down selecting" the most promising fuels and processes. Major process and economic parameters such as feedstock, capital expenditure, operating cost, yield, capacity factor, complexity of the process, level of technology development, and internal rate of return were included in the matrix. These parameters cover the major process aspects that reflect the overall economics of the processes. This evaluation was conducted using past models, if they existed. It was also conducted through publicly available literature in 2007. If quantitative numbers were available for the matrix, they were used. But since the technologies are not mature, many matrix entries were qualitative.

Both butanol and ethanol processes were initially considered. Butanol has properties such as higher energy density and immiscibility with water that may make it a better transportation fuel. However, butanol production processes are currently at the lab scale or very early pilot stage of development, and published data on butanol-producing organisms indicate low yields relative to ethanol production. Therefore, only ethanol technologies were adopted for analysis.

Ethanol-producing processes were categorized by pretreatment method. Substantial research has been done on biomass pretreatment for biochemical conversion at the bench scale by the Consortium for Applied Fundamentals and Innovation (CAFI) projects [1, 2, 3]. Based on the results of these studies, concentrated acid, SO_2-steam, lime, and ammonia-recycle-percolation pretreatment were rejected as less attractive due to the high costs associated with the processes.

The pretreatment processes selected for further analysis were:

1. Acid pretreatment (single-stage dilute and two-stage dilute)
2. Ammonia Fiber Explosion (or Expansion) (AFEX)
3. Hot water pretreatment.

Dilute acid is a pretreatment technology that showed promise in the original CAFI study [1, 2] and serves as the base technology for the National Renewable Energy Laboratory's (NREL's) cellulosic ethanol design report [4]. AFEX is a pretreatment that does not require as much water as other pretreatments. Hot water pretreatment has low capital investment requirements. Two-stage dilute acid is a pretreatment that eliminates the need for enzymes for saccharification, one major source of uncertainty in the other pretreatments. Single-stage dilute acid, AFEX, and hot water pretreatment conversions are based on bench-scale experiments from CAFI [3]. Two-stage dilute acid pretreatments are also based on bench-scale experiments [5, 6].

In addition, process variations in combination with pretreatment that may offer attractive system performance were also considered. Among those process variations selected for further study was the production of enzymes on-site using hydrolyzate as a carbon source. On-site enzyme production may be cheaper by precluding the use of stabilizers and other additives that are needed when enzymes are purchased.

Another variation considered was parallel fermentation of 5-carbon (C5) and 6-carbon (C6) sugars. Current cofermentative organisms do not have high ethanol yields and are not highly robust to system variations. We therefore chose to include parallel fermentations of C5 and C6 sugars for comparison with the baseline fermentation.

Another technology that was considered worthy of further analysis was the use of pervaporation instead of the beer distillation column to separate ethanol and potentially allow higher titer and substrate utilization in the fermentor. Pervaporation also has the advantage of lower steam and utility requirements than a distillation column.

The three downstream technologies—parallel fermentation of C5 and C6 sugars, on-site enzyme production, and pervaporation—were compared to a base case using dilute acid pretreatment.

All together, seven scenarios were examined:

- Dilute Acid Pretreatment
- Two-Stage Dilute Acid Pretreatment
- Hot Water Pretreatment

- AFEX Pretreatment
- On-site Enzyme Production with Dilute Acid Pretreatment
- Ethanol Separation using Pervaporation with Dilute Acid Pretreatment
- Separate C5 and C6 Fermentation with Dilute Acid Pretreatment.

Project Assumptions

For each down-selected process a common list of assumptions on process operations and economic analysis was made. The scope of the work was to determine the product value of cellulosic ethanol for a plant operating in the 5-8 year timeframe. This timeframe was chosen because of the renewable fuel standard volumes mandated in the Energy Independence and Security Act of 2007 [7]. For a plant to be operating at commercial levels in this timeframe, the plant would likely need to be designed based on current data as large process plant projects typically take more than four years for design, construction, and startup [8]. The major assumptions are highlighted below, and the complete list is provided in Appendix A.

- The plant capacity is 2000 MT/day.
- The processes use corn stover as feedstock.
- The feedstock contains 25% moisture and the composition is assumed to be the same as that obtained in CAFI II feedstock analysis (Table A-1).
- Plant capacity factor is 96% (350 on-stream days/year).
- Feedstock cost is $83/dry MT ($75/dry ST).
- Purchased enzyme cost from off-site source is $507/MT ($460/ST) of broth of 10% protein used at a loading of 31.3 mg protein per gram cellulose in the feed. This cost was calculated based on the cost of producing enzyme on-site using hydrolyzate. This cost comes to $0.69/gallon of ethanol.
- The processes use 2007 lab-scale, experimentally validated data (2007 EVD) and equipment prices (indexed).
- Plant depreciation is calculated following the Internal Revenue Service (IRS) modified accelerated cost recovery system (MACRS) over 7 years for the main plant and 20 years for the cogeneration area.
- Plant life is 20 years.
- The plant is 100% equity financed.
- Contingency factor and working capital are 20% of total direct and indirect costs (TD&IC) and 15% of fixed capital investment (FCI), respectively, for n^{th} plant.
- The plant initiates operation in 5-8 years.
- Adopted units: cost of all purchased chemicals and feedstocks, plant capacity, and yields are reported in metric tons (MT). Operating conditions: temperature (in ^{o}C), pressure (in bar), and mass flow rates (in MT/day).

Process Description and Flow Diagram

The different scenarios detailed above were generated by modifying the NREL 2002 production process [4]. Appendix G contains a general description of the process steps that have not been altered from previous NREL studies. Appendix F contains details of operating conditions for the major process steps. The basic cellulosic ethanol production process, modified from NREL's 2002 design report [4], comprises nine sections as shown in Figure 1 and listed below.

- Feed Handling (Area 100)
- Pretreatment and Detoxification (Area 200)
- Enzymatic Hydrolysis and Fermentation (Area 300)
- On-site Enzyme Production (Area 400)
- Product Recovery (Area 500)
- Wastewater Pretreatment (Area 600)
- Storage (Area 700)
- Burner/Boiler Turbo-Generator (Area 800)
- Utilities (Area 900) sections.

The AspenPlus Process Simulator is used in process modeling. Current technological data are used in the simulation and described as the 2007 experimentally validated data (2007 EVD). In the pretreatment area, data for AFEX and hot water 2007 EVD are obtained from CAFI II research [1, 2, 3]. Two-stage dilute acid data are based on available literature [5, 6]. Dilute acid pretreatment data are taken from both CAFI II and NREL research. For all other areas, 2007 EVD conversion data are obtained from NREL research [9]. The overall process block diagram is shown in Figure 1. Operating conditions of the major unit operations of each area and mass flow rates are included in Figure 1.

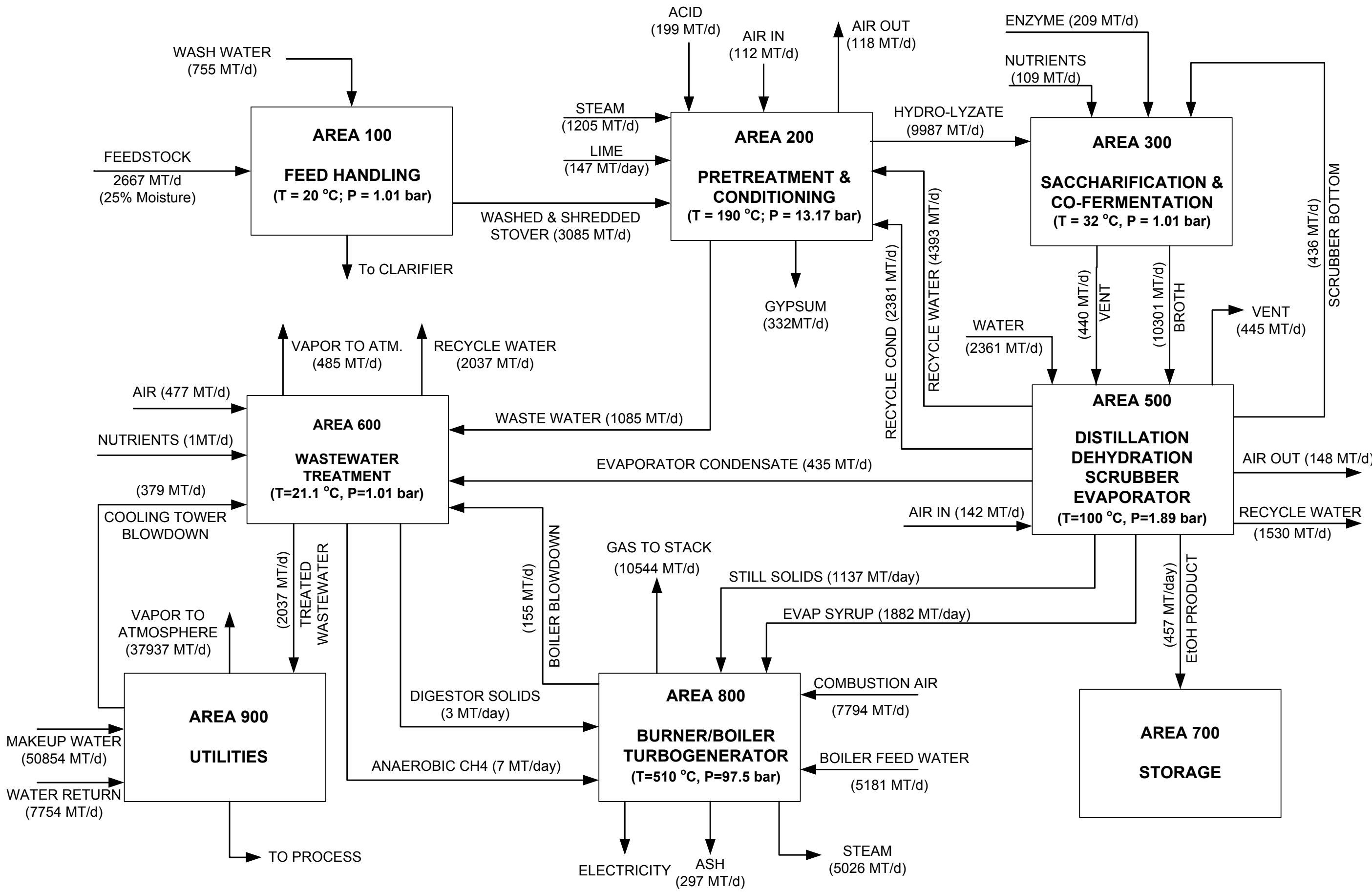


Figure 1. Overall process block diagram of a typical cellulosic ethanol process plant (based on NREL's 2002 design report and modified to 2007 EVD)

Process Variations

This section includes a description of each of the process scenarios. Results from the seven process variations are shown in Table 2 of the Results and Discussion section.

Dilute Acid Pretreatment

Table F-1 includes a summary of the dilute acid operating conditions and model parameters. The biomass from Area 100 is fed by a screw feeder to the presteamer, where low pressure steam (163°C, 4.46 bar) is added to maintain a temperature of around 100°C. The presteamer allows a portion of the pretreatment heat requirement to be met with low pressure steam. The biomass then enters the pretreatment reactor, where high pressure steam (268°C, 13.17 bar) is added as shown in Figure 2. Sulfuric acid, diluted with process water, is added to the reactor at a rate necessary to achieve 1.9 wt% of the liquid phase in the reactor. The reactor temperature, pressure, and residence time are maintained at 190°C, 11.6 bar, and 2 minutes, respectively. The biomass slurry is then flashed to 1.0 bar in the blow-down tank. The solid fraction is separated from the slurry in a Pneumapress pressure filter. In order to reduce toxicity to the fermentation organisms downstream, a liming step is added to neutralize excess H_2SO_4 in the hydrolyzate.

Solid lime is added to the overliming tank along with the aqueous fraction from the Pneumapress to raise the pH to 10. The reaction of lime and H_2SO_4 forms gypsum, which is separated from the hydrolyzate as solid cake. The residence time in the overliming tank is 1 hour. The slurry is pumped to a second tank where additional H_2SO_4 is added to reduce the pH to 4.5. The residence time in the second tank is 4 hours, which allows the gypsum crystals to grow to an adequate size for solid/liquid separation. The gypsum is separated from the slurry in a two-step process, with the first being a hydrocyclone and the second being a rotary drum filter. The conditioned hydrolyzate is mixed with the solid biomass fraction from which it was previously separated in the Pneumapress pressure filter, and the resulting slurry is ready for enzymatic hydrolysis. Mass balance and operating conditions of major unit operations are shown in the process flow diagram (Figure 2).

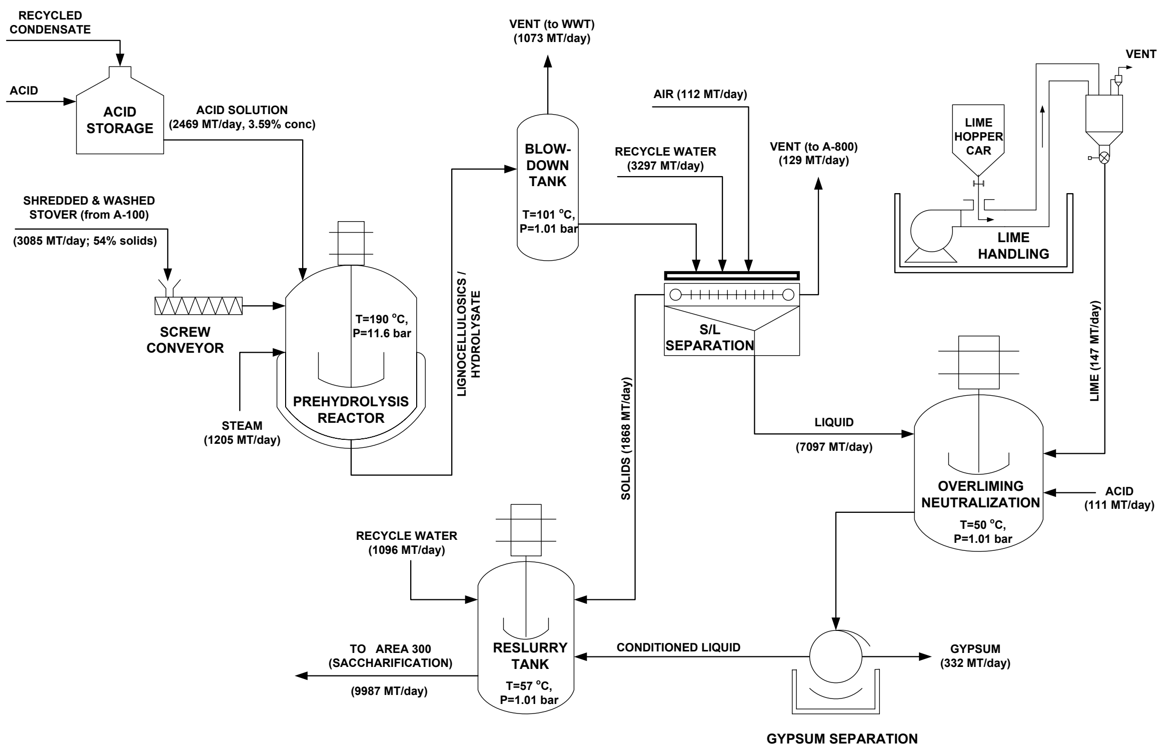


Figure 2. Process flow diagram for dilute acid pretreatment

Two-Stage Dilute Acid Pretreatment

An operating summary of the two-stage dilute acid pretreatment is provided in Table F-2. In the two-stage dilute acid pretreatment process, the first stage solubilizes most of the hemicellulose, just as in the dilute acid pretreatment process. In the second stage, a higher concentration of acid is added to hydrolyze the cellulose and remaining hemicellulose (Figure 3). This contrasts with the dilute acid pretreatment process, where enzymes are used to hydrolyze the cellulose (Figure 4). Mass balance and operating conditions of the major unit operations are shown in the flow diagrams (Figures 3 and 4). Two-stage dilute acid conversion data were taken from literature for softwood and assumed to be similar for corn stover [5, 6].

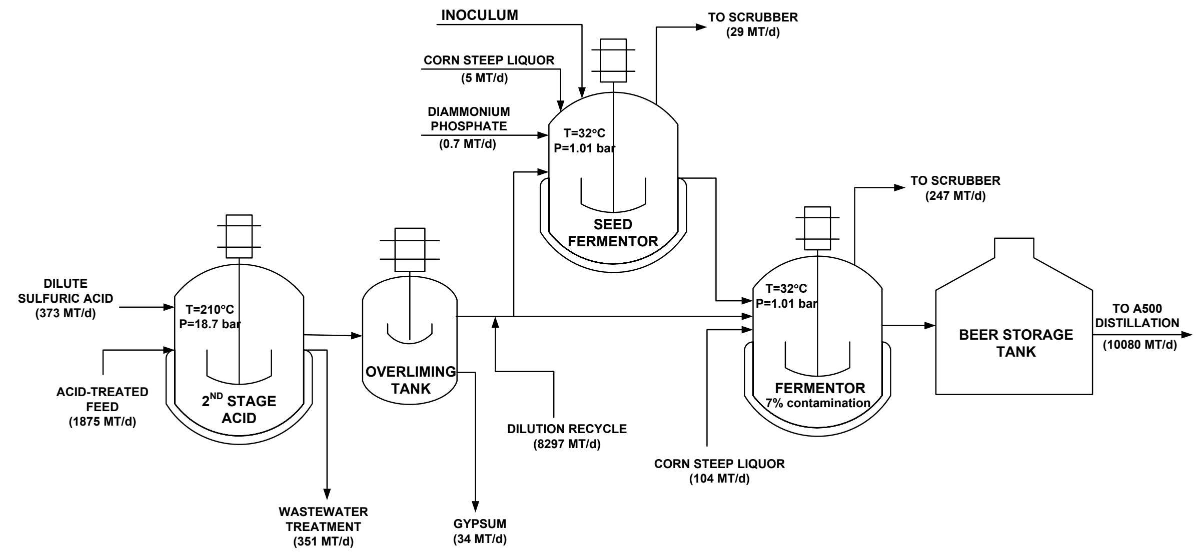


Figure 3. Process flow diagram for two-stage dilute acid pretreatment/hydrolysis and fermentation

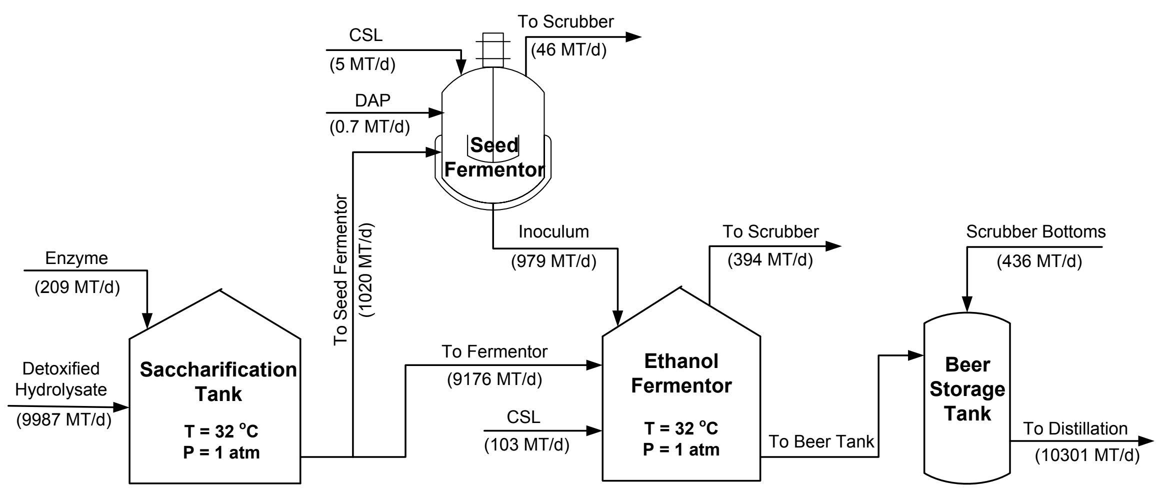


Figure 4. Process flow diagram for base case pretreatment, enzymatic hydrolysis, and fermentation

Hot Water Pretreatment
Table F-3 contains a summary of model parameters for hot water pretreatment. The chopped and washed biomass from Area 100 is mixed with recycled hot water from Area 500. The slurry is fed to a plug flow pretreatment reactor. The reactor pressure is maintained at 12.7 bar and the temperature is held constant at 190°C. The residence time in the pretreatment reactor is 5 minutes. The slurry is then cooled to 65°C and flashed to 1.0 bar in the flash tank. Ammonia is added to the reactor to neutralize acetic acid formed during the pretreatment process. The xylose and cellulose pretreatment yields are shown in Table F-3. Mass flow rates and operating conditions of reactor and major units are shown in the process flow diagram (Figure 5).

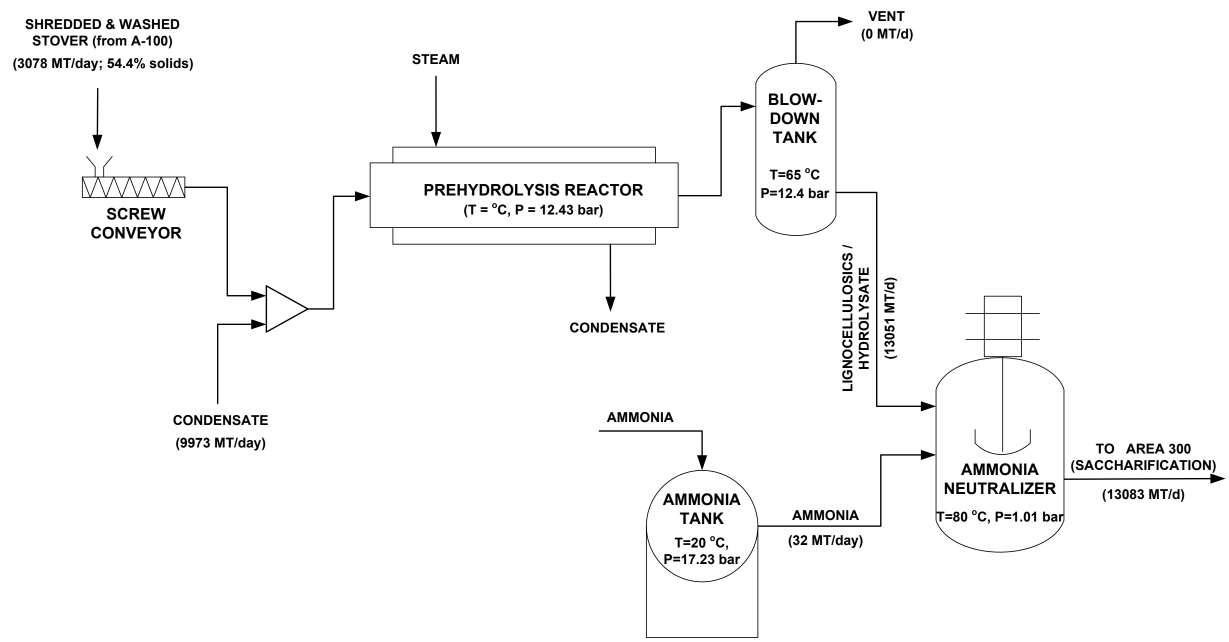


Figure 5. Hot water pretreatment process flow diagram

AFEX Pretreatment
A summary of operating parameters for AFEX pretreatment can be found in Table F-4. In the AFEX pretreatment process, the biomass is treated with liquid anhydrous ammonia under high pressure (17.2 bar) and 60°C for about 5 minutes [10]. The pressure is rapidly released, causing the fibers to explode and increasing the access of enzymes to cellulose. Most of the ammonia is recovered from the blow-down tank. Residual ammonia is recovered from the solids by a flash followed by fractionation from other volatiles. Recovered ammonia vapor is then compressed, condensed, and recycled back into the AFEX reactor [10]. The pretreated biomass is slurried into a holding tank to be sent to Area 300 for enzymatic hydrolysis (Figure 6).

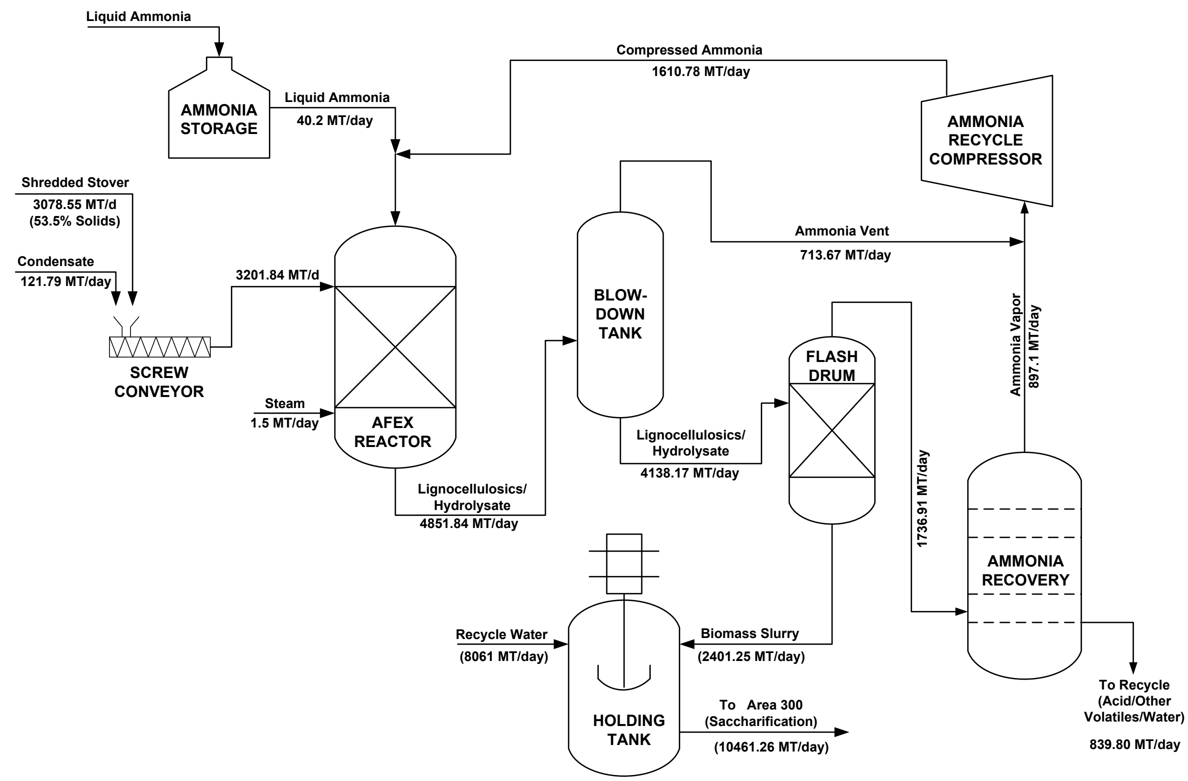


Figure 6. AFEX process flow diagram

Separate C5 and C6 Fermentation

Separate solid and liquid processing (C6 and C5 sugars respectively) takes advantage of enhanced yields where xylose is fermented separately using *Zymomonas mobilis* and glucose is fermented separately with yeast (*Saccharomyces cerevisiae* or *Saccharomyces pastorianus*). This avoids the issue of lower xylose to ethanol yields seen in the current cofermenting organisms. One disadvantage of separate processing is that more water is needed to dilute the solids stream because the best yields are achieved at low solids loading. The additional water increases ethanol recovery costs. The process shown in Figure 7 attempts to mitigate the water issue by using the product stream from the liquor (xylose) fermentation to dilute the stream prior to saccharification. Table F-5 contains a summary of operating parameters for the separate C5 and C6 fermentation scenario, based on work by Dutta et al. [11].

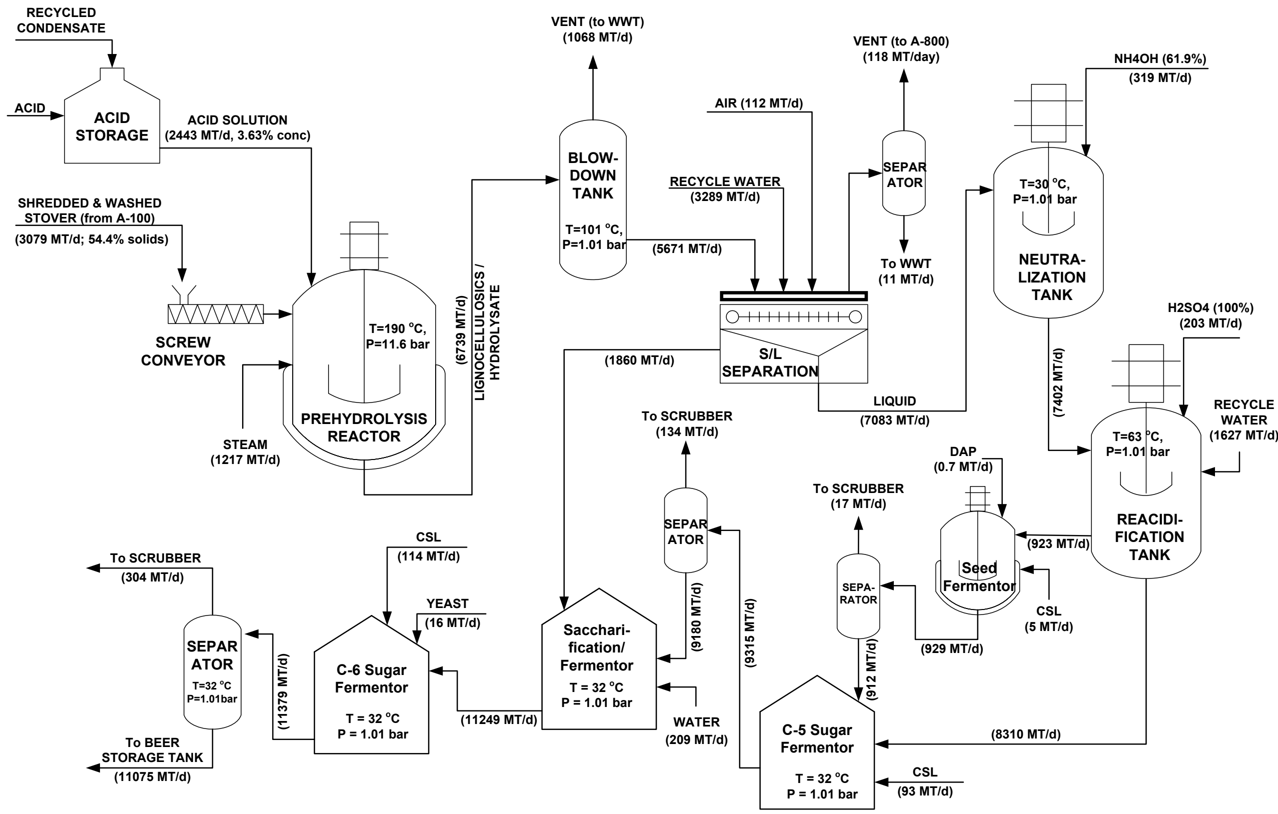


Figure 7. Separate C5 and C6 fermentation configuration

Ethanol Separation Using Pervaporation

Pervaporation refers to separation using a membrane with liquid feed on one side and a low-pressure, gaseous permeate output on the other side. Components in the liquid feed preferentially permeate through the membrane and then evaporate into the gaseous phase. Because pervaporation does not involve a large heat input, the process could save on costs associated with the heat and steam needed for the reboiler of a conventional distillation column. A summary of operating parameters for the pervaporation separation scenario is located in Table F-6.

In contrast to the base case distillation configuration (Figure 8), the pervaporation variation modifies the distillation section of the dilute acid model by inserting a pervaporation system in place of the beer column (Figure 9). The pervaporation system output is calculated from a separation factor and a total material flux needed to achieve the same separation as in the beer column [12]. The separation factors and material fluxes are based on literature [12]. In the base case dilute acid model, the beer column also serves to separate carbon dioxide from the ethanol stream. In the pervaporation model, a flash tank is added to separate out CO_2 at 110°C and a heat exchanger cools the stream to 41°C. The separation of CO_2 is assumed to be easier than what is modeled here, so neither the capital nor utility costs of the flash tank and heat exchanger are included in the economic analysis. The membrane system costs \$200/m^2 in 1999 dollars, with a replacement needed every five years at a cost of \$100/m^2 [13].

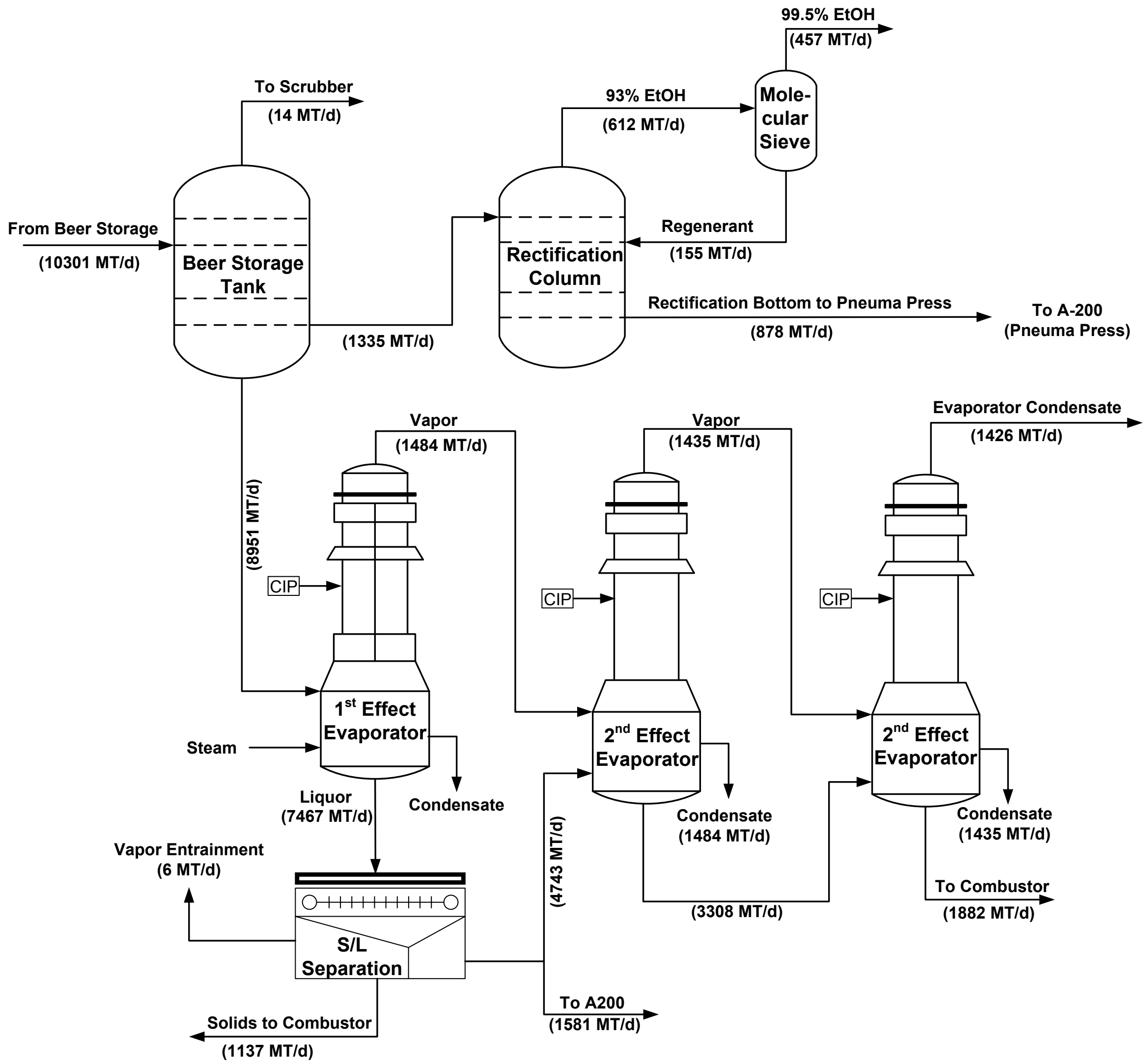


Figure 8. Base case distillation configuration

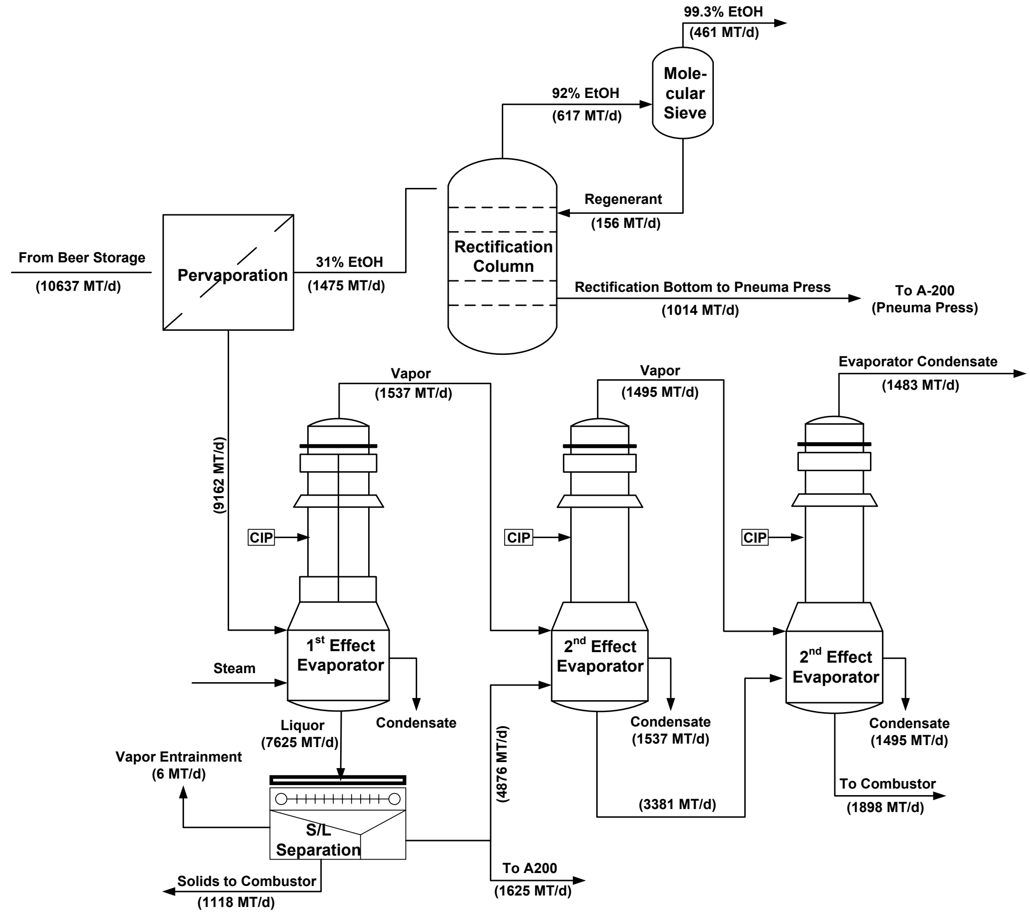


Figure 9. Pervaporation separation process

On-Site Enzyme Production

Table F-7 contains a summary of the on-site enzyme production operating parameters. On-site enzyme production (Area 400) is modeled as a variation to the baseline dilute acid process model (where enzyme is assumed to be purchased from suppliers). This work was based on a design report by NREL in 1999 [14]. Cellulase enzyme is a mixture of endoglucanases, exoglucanases, and β-glucosidase enzymes. In the present case, *Trichoderma reesei* (a fungal strain) is used for on-site cellulase enzyme production. A portion of the conditioned pretreated biomass from Area 200 is pumped to Area 400, where a fraction of the stream is used for the growth of *Trichoderma reesei* inoculums in seed bioreactors. The remainder is pumped into the jacketed aerobic bioreactors where inocula are added from the seed bioreactors to produce enzymes. Corn steep liquor and other trace nutrients are also added to the reactors. Ammonia is used to control pH and to provide additional fixed nitrogen to the fungus; oxygen is supplied by an air compressor. To control foam formation, corn oil is added into the bioreactors. Mass flow rates and process conditions of major unit operations are shown in Figure 10.

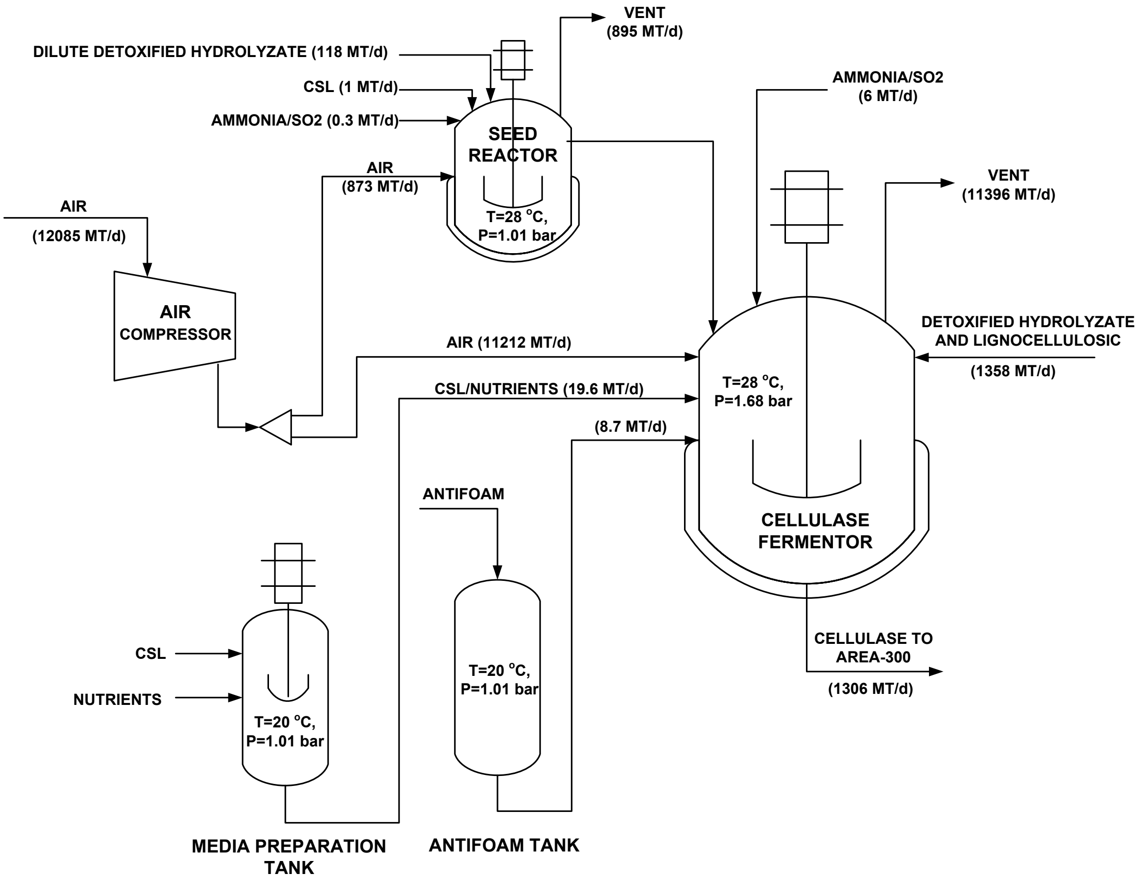


Figure 10. On-site enzyme production process flow diagram

Methodology for Economic Analysis

The base case cost estimation and analysis is performed assuming the plant is an n^{th} plant design. This means that a similar plant was previously constructed and operated without unexpected delays in startup and capacity loss. The process design is assumed to become the n^{th} plant, operating at the reported yields from experimental, bench-scale data. This is a major assumption for a process that is still early in its development, and when an n^{th} plant for cellulosic ethanol becomes commercial, it may not look like this design. However, the n^{th} plant analysis was chosen to provide an analysis parallel to analyses done by different groups, including NREL, the CAFI group, and Pacific Northwest National Laboratory.

However, because of the new technological elements of cellulosic ethanol production there are a number of engineering design and performance uncertainties. These uncertainties are accounted for in the cost growth analysis for a pioneer plant using the risk analysis methodology developed by RAND [15, 16]. The details of the cost estimation for the n^{th} plant and the cost growth analysis for the pioneer plant are discussed in the following subsections.

Methodology for Discounted Cash Flow Analysis for n^{th} Plant

Mass and energy balances for each of the seven selected process variations are performed by AspenPlus Process Simulator. Pinch analysis is performed to optimize energy balances. The stream flow rates, from AspenPlus simulations, are used to size unit operations. The costs for most of the equipment are obtained from previous vendor quotes obtained by NREL. Individual equipment is scaled and the scaled cost is estimated following exponential correlations as described in Appendix A. The scaled cost is then indexed to a year 2007 dollar value using the Chemical Engineering Plant Cost Index [17]. Separate installation factors are used for each of the unit operations to obtain individual installed equipment costs. The installation factors are obtained from the vendors who provided equipment quotes. The cost analysis is performed following NREL's approach [4] and that found in Peters et al. [18], with modified terminologies. Total installed cost (TIC) is defined as the sum of total installed equipment cost, warehouse, and site development costs.

Engineering and supervision costs, construction expenses, and legal and contractor's fees are assumed to be 32%, 34%, and 23% of purchased equipment cost, respectively. Total direct and indirect costs (TD&IC) is the sum of total installed cost and indirect costs. Contingency is assumed to be 20% of TD&IC, and working capital is 15% of fixed capital investment (FCI). FCI is the sum of TD&IC and contingency, and the total capital investment is the sum of FCI and working capital.

The manufacturing costs include raw materials costs (such as corn stover), variable costs (such as process chemicals, enzyme, nutrients) and fixed operating costs (employee salaries, overhead, insurance, and maintenance). The feedstock cost is assumed to be \$83/dry MT (\$75/dry ST) (Appendix A). The chemical and nutrients costs are obtained from NREL's previous quotes from suppliers and are indexed to 2007 dollar values following the Inorganic Chemical Index of the SRI International Economics Handbook, Economic Environment of the Chemical Industry [19]. NREL estimated the required number of employees and their salaries in their 2002 design report [4]. In the present study, the same number of employees is required for the same plant capacity, and the salaries are indexed to 2007 dollar values following the labor index from the Bureau of Labor Statistics [20].

In the present analysis, the process and steam generation plants are assumed to depreciate in 7 and 20 years, respectively, following IRS MACRS, and the plant life is 20 years. The project is assumed to be 100% equity financed and internal rate of return is 10%. For the present cost analysis, the capital investment is spread over 3 years at a rate of 8%, 60%, and 32% in the first, second, and third years, respectively. The product value (PV) of ethanol is calculated by iterating to reach a net present value of \$0 with a 10% internal rate of return.

Pioneer Plant Analysis

The pioneer plant analysis is performed using a method developed by Merrow et al. of RAND Corporation [15]. This methodology considers two sources of production cost growth in pioneer plants: less than expected plant performance and low capital cost estimation. These sources are regressed with two multi-factor ordinary least squares correlations to estimate the unexpected reduced plant performance (Equation 1) and capital cost growth (Equation 2) associated with pioneer plants.

Equation 1 estimates pioneer plant performance as a percentage of design capacity in the second 6 months after startup.

$$\text{Plant Performance} = 85.77 - 9.69\times\text{NEWSTEPS} + 0.33\times\text{BALEQS} - 4.12\times\text{WASTE} - 17.91\times\text{SOLIDS} \quad \text{(Eq. 1)}$$

Where,

NEWSTEPS ≡ The number of steps in the process that have not been proven commercially.

BALEQS ≡ The percentage of mass and energy balance equations used in plant design that are validated with commercial-scale data. The RAND report also mentions that some weight is given to rigorous theoretical models.

WASTE ≡ Potential problems that may be associated with waste handling. A 0-5 scale is used, with 0 meaning no waste handling issues and 5 meaning significant waste handling issues.

SOLIDS: The scale used is 0 or 1. If the process handles solids then the value is 1; otherwise it's 0.

Equation 2 estimates the capital cost growth, defined as the ratio of estimated to actual costs.

$$\text{Cost Growth} = 1.12196 - 0.00297\times\text{PCTNEW} - 0.02125\times\text{IMPURITIES} - 0.01137\times\text{COMPLEXITY} + 0.00111\times\text{INCLUSIVENESS} - C_1\times\text{PROJECT DEFINITION} \quad \text{(Eq. 2)}$$

Where,

PCTNEW ≡ The installed cost of all commercially undemonstrated equipment as percentage of total installed equipment cost.

IMPURITIES: Represents the potential process issues that may arise due to impurity buildup from recycle streams or problems due to equipment corrosion. The value ranges from 0 to 5, with 0 being given to processes with no impurity buildup or corrosion issues.

COMPLEXITY ≡ The number of continuously linked process steps.

INCLUSIVENESS ≡ The percentage of three factors: pre-startup personnel costs, pre-startup inventory cost, and land purchase. For example, if two of these factors have been rigorously considered, the variable would be given a value of 67%.

C1: C1 is 0.06361 if the design is at pre-development/exploratory or research and development stage and 0.04011 if the design is in commercial or pre-commercial stage. For the present studies C1 is assumed to be 0.06361.

PROJECT DEFINITION: Includes commitment of funds to define the plant scope, basic plant layout, and process flow conditions. Most major equipment is defined and examination of site begins at this point. The amount of work involved here depends on how much information is already available from previous project experience. Often some critical level of engineering (heat and mass balances, equipment need, and so forth) and site-specific information (on-site and off-site unit configurations, soils and hydrology data, health and safety requirements, and environmental requirements) is completed here. A numerical value is assigned to define the level of engineering completed at the time of estimation, following the level of completeness: (1) engineering completed, (2) moderate or extensive engineering, (3) limited engineering, and (4) screening design stage. Similarly, a value for site-specific information is assigned by the following: (1) definitive or completed work, (2) preliminary or limited work, (3) assumed or implicit analysis, and (4) not used in the cost estimation at all. The range of values given to project definition is 2 (for maximum definition) to 8 (for no definition).

Cost growth analysis for all seven process variations was performed. There was some subjectivity in choosing the parameters for the pioneer plant analysis, so a range of parameters was used to estimate pioneer plant costs for three scenarios: optimistic, most probable, and pessimistic. For the dilute acid pretreatment process the selected variable values of Equations 1-2 are shown in Table 1; the selection justification is discussed below. For all other process variations, the variable values are shown in Appendix J.

Table 1. Plant Performance and Cost Growth Variables for Dilute Acid Pretreatment Processes

Plant Performance (Equation 1)				**Cost Growth (Equation 2)**			
Variables	**Values**			**Variables**	**Values**		
	Opti-mistic	**Most Probable**	**Pessi-mistic**		**Opti-mistic**	**Most Probable**	**Pessi-mistic**
NEWSTEPS[a]	6	6	6	PCTNEW	61.76	61.76	61.76
BALEQS	50	40	30	IMPURITIES	0	3	5
WASTE	1	2	3	COMPLEXITY[b]	6	6	6
SOLIDS	1	1	1	INCLUSIVE-NESS	33	0	0
				PROJECT DEFINITION	6	6	7
Plant Performance (%)	**22.10**	**14.68**	**7.26**	**Cost Growth**	**0.53**	**0.42**	**0.32**

[a] New steps/units**:** Feedstock Handling, Pretreatment, Saccharification, Cofermentation, Beer Column, and Combustor. [b] Continuously linked steps: Feedstock Handling, Pretreatment, Saccharification, Cofermentation, Distillation, and Steam/Power Generation.

Justification of Correlation Variable Value Selection for Plant Performance (Equation 1)

For all three scenarios, the operations considered new steps/units are feedstock handling, pretreatment, saccharification, cofermentation, beer column, and the fluidized bed combustor (for converting lignin to heat and power), resulting in a value of 6 for NEWSTEPS.

Some of the steps and units are being used commercially, so the BALEQS variable is assigned a value of 50, 40, and 30 for optimistic, most probable, and pessimistic scenarios, respectively.

The wastewater contains a small amount of furfural that may not be degraded by the anaerobic or aerobic treatments used in the model, meaning that an additional chemical treatment may be necessary. No additional complications with waste are foreseen. The WASTE variable was assigned a value of 2 for the most probable case. Values of 1 and 3 were assigned for the optimistic and pessimistic cases, respectively.

These variable values are used in Equation 1 to calculate the percentage of Plant Performance for the three cases (Table 1).

Justification of Correlation Variable Value Selection for Cost Growth (Equation 2)
The feedstock handling area, pretreatment area, saccharification, cofermentation, beer column, and fluidized bed combustor are selected as new technologies/units to calculate the parameter PCTNEW for all three of the cases.

Some of the degradation products inhibit the saccharification and fermentation process, and buildup of those inhibitors in the process may result in yield loss. For the most probable case, the assigned value for the variable IMPURITIES is 3. For the optimistic and pessimistic cases the assigned values are 0 and 5, respectively.

The process design has six continuously linked process steps, which include feedstock handling, pretreatment, saccharification, cofermentation, distillation, and steam/power generation. Therefore, the value for the variable COMPLEXITY is assigned as 6 for all three cases.

Some of the initial plant inventory is included in the base case cost estimate, although it is not validated in a commercial plant. So, for the optimistic scenario a value of 33% is assigned for the variable INCLUSIVENESS. The value of 0% is assigned for both pessimistic and most probable cases.

A plant site has not been chosen, so none of the site-specific information has been procured. Some level of engineering has been completed. Therefore, a value of 6 is assigned for most probable and optimistic cases for the variable PROJECT DEFINITION and the pessimistic case is assigned a 5.

The assigned variable values are used in Equation 2 to calculate the percentage of cost growth as shown in Table 1. The base case n^{th} plant TCI is divided by the cost growth, obtained from Equation 2 (Table 1), to obtain an estimate for the pioneer plant TCI. The contingency factor is increased to 30%, compared with 20% for an n^{th} plant, to account for additional construction uncertainties. The plant performance, obtained from Equation 1 (Table 1), is multiplied by the first year ethanol sales to account for the reduced production of a pioneer plant. For the discounted cash flow analysis, plant performance is increased by 20% per year until design capacity is reached.

Results and Discussion

nth Plant Cost Analysis

Ethanol yield, byproduct credit, total installed equipment cost, total project investment, and estimated PV for each of the process variations are shown in Table 2.

Table 2. Product Value for Various Pretreatment and Downstream Process Variations

Process Variations	Total Capital Investment ($MM)	Total Installed Equipment Cost ($MM)	Ethanol Yield (Gal/MT)[a]	Ethanol Production (MM Gal/Yr)[b]	Electricity Export ($MM/Yr)	Product Value ($/Gal)[c]
Dilute Acid Pretreatment (base case)	376	164	76.3 (288.8)	53.4 (202.2)	11.7	3.40 (0.90)
Dilute Acid Pretreatment (high solids)	389	169	72.5 (274.5)	50.8 (192.1)	12.6	3.60 (0.95)
Two-Stage Dilute Acid Pretreatment	391	173	46.8 (177.5)	32.8 (124.2)	16.8	4.38 (1.16)
Hot Water Pretreatment	361	156	55.8 (211.0)	39.0 (147.7)	11.3	4.44 (1.21)
AFEX Pretreatment	386	167	65.9 (249.7)	46.2 (174.8)	16.9	3.69 (0.97)
Pervaporation-Distillation	501	209	76.9 (291.3)	53.9 (203.9)	13.6	3.75 (0.99)
Separate C5 and C6 Fermentation	386	168	79.3 (300)	55.5 (210.1)	6.5	3.67 (0.97)
On-site Enzyme Production	434	188	67.7 (256.3)	47.4 (179.4)	-0.8	3.54 (0.94)

[a] Values in parentheses are in liter/MT. [b] Values in parentheses are in MM liter/year. [c] Values in parentheses are in $/liter.

Each pretreatment process has some variation in yield (47-76 gal/MT) with dilute acid pretreatment being the highest. The lowest PV among all process variations is from the dilute acid pretreatment scenario, which is $3.40/gal, and those for other pretreatment processes are in the range of $3.60/gal to $4.44/gal. The yield from two-stage dilute acid pretreatment is lowest (46.8 gal/MT) and the process requires higher TCI ($391 million), which drives the PV as high as $4.38/gal.

The installed equipment cost and TCI for all four pretreatment scenarios are in the range of $156-$173 million and $361-$391 million, respectively, with hot water pretreatment being the lowest and two-stage dilute acid pretreatment being the highest. The reason for lower installed equipment cost for hot water is because it uses a relatively simple horizontal tubular pretreatment reactor. The installed cost of the tubular reactor is $311,000 for hot water pretreatment, whereas pretreatment reactors for dilute acid and AFEX pretreatment processes are $22.99 million and $9.15 million, respectively. The dilute acid pretreatment process requires long retention time for overliming, which requires large expensive vessels. Although the AFEX reactor cost is lower than the dilute acid pretreatment reactor, the AFEX pretreatment process has additional expensive unit operations (such as the ammonia compressor) that increase the total installed equipment cost to slightly more than that of dilute acid pretreatment processes.

For the comparison of pretreatment technologies, lab-scale experimental data at low solids loading (25%) were used in the model [3]. In order to understand how process scale-up may impact ethanol production cost, a process model was developed using data from experiments conducted at NREL at higher solids loading (40%). Table 3 presents a summary comparison of the results from the dilute acid pretreatment models.

Table 3. Comparison of Dilute Acid Pretreatment Results from Lab- and Pilot-Scale Data

	Base Case (2007 EVD)	**High Solids**
Solids Loading	25%	40%
PV ($/gal)	3.40	3.60
Ethanol Yield (gal/MT)	76.3	72.5
Installed Equipment Cost (MM$)	164.1	169.4
Fixed Capital Investment (MM$)	326.8	337.8
Total Capital Investment (MM$)	375.9	388.5
Lang Factor	3.44	3.43

The PV of the model using high solids loading is $0.20/gal higher than that of the base case model. This is primarily due to the decreased ethanol production caused by lower yields of monosaccharides at higher solids loadings.

The costs of each process area for the dilute acid pretreatment scenario are presented in Table 4; costs for other scenarios are given in Appendix H. The expensive areas of the dilute acid pretreatment scenario are the pretreatment, saccharification and fermentation, distillation and solids recovery, and boiler and turbo-generator sections. Among these, the boiler and turbo-generator section is the most expensive area, accounting for 56% of the total installed equipment costs.

Table 4. Costs by Area of the Dilute Acid Pretreatment Process Scenario

Cost Areas/Factor	Installed Cost		Purchased Equipment Cost	
	(MM$)	(%)	(MM$)	(%)
Feedstock Handling (Area 100)	10.9	6.6	6.0	5.5
Pretreatment (Area 200)	36.2	22.1	19.7	18.0
Saccharification and Fermentation (Area 300)	21.8	13.3	17.3	15.8
Distillation and Solids Recovery (Area 500)	26.1	15.9	17.2	15.8
Wastewater Treatment (Area 600)	3.5	2.1	2.6	2.4
Storage (Area 700)	3.2	2.0	2.0	1.8
Boiler/Turbogenerator (Area 800)	56.1	34.2	40.3	36.9
Utilities (Area 900)	6.3	3.8	4.2	3.8
Total Installed Equipment Cost	164.1	100	109.3	100
Fixed Capital Investment (FCI)	326.8			
Working Capital (WC)	49.0			
Total Capital Investment (TCI)	375.9			
Lang Factor[a]	3.44			

[a] The Lang factor is calculated by dividing TCI by the total purchased equipment cost.

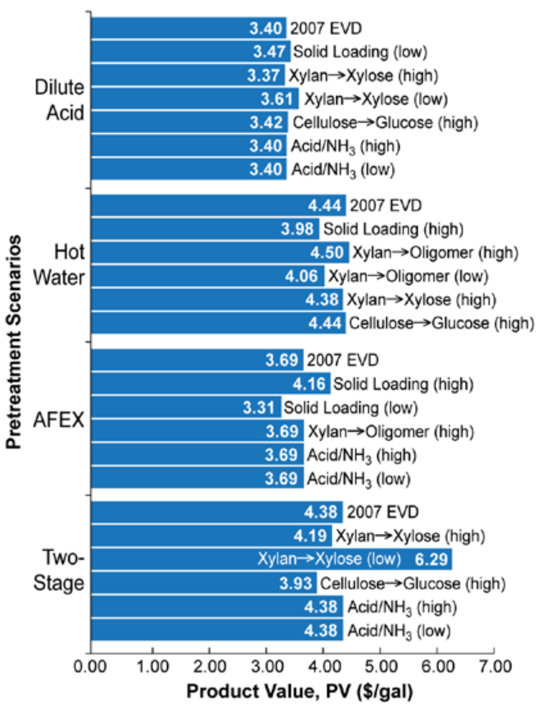
The PV from the pervaporation scenario is $3.75/gal, which is higher than the PV for the base case process (dilute acid pretreatment process, $3.40/gal). The cost differential comes from the high capital cost of the pervaporation membrane, which is not well developed commercially.

Ethanol yield in the dilute acid pretreatment process with cofermentation is 76.3 gal/MT. It was thought that separate C5 and C6 sugar fermentation using selective yeast might reduce the PV because of higher yields (79.3 gal/MT). However, the PV is 8% higher than the base case scenario because of increased capital costs primarily due to the additional fermentation vessels required.

It was also thought that on-site enzyme production might provide economic advantages over purchasing enzymes because it eliminates the need to use stabilizing chemicals and to concentrate the enzyme broth prior to transportation. The PV from the on-site enzyme production process is $3.54/gal, which is $0.14/gal higher than the base case dilute acid pretreatment process. The cost of the enzyme is affected by a lower electricity credit than the base case. This is because of high electricity consumption by the compressor supplying air to the enzyme production bioreactors that leads to lower net excess electricity and lower ethanol yield (67.7 gal/MT). In the on-site enzyme production process, part of the feedstock (hydrolyzate from Area 200) is diverted to enzyme production, which reduces the overall plant capacity by 6 MMGal of ethanol per year. It is assumed that *Trichoderma reesei* is the enzyme-producing fungal strain. The specific activity of enzyme, yield, and productivity are 600 FPU/g protein, 0.33 g protein/g cellulose and xylose, and 0.125 g protein/L-hr, respectively. The PV is higher for the on-site case than for the off-site case because the off-site case's enzyme cost was determined after increasing the feedstock in order to produce the same amount of ethanol. This increase in scale provided some economic benefit. This comparison does not include costs of stabilizing chemicals associated with purchased enzymes, so the enzyme cost continues to be uncertain. However, the comparison of an on-site case helps demonstrate what yields and electricity tradeoffs occur when enzyme is produced on-site.

Sensitivity Analysis

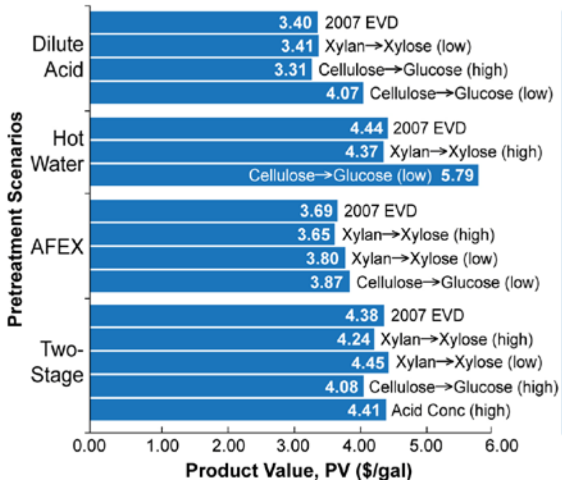
Process-specific sensitivity analysis of pretreatment and saccharification operations was performed on all pretreatment process scenarios. This analysis showed the impact of process operation parameters including operating temperature, retention time, acid concentrations, and yields on PV. The results are shown in Figures 11 and 12 (the detailed parameter values and results are tabulated in Table B-1 and Table C-1, respectively). PV is most sensitive to pretreatment solid consistency, retention time, and xylan and cellulose conversions. When the retention time of the dilute acid pretreatment reactor is increased from 2 to 10 minutes, an increase in PV of 16% is observed (Figure 11). When the conversion of xylan to xylose in the pretreatment reactor is reduced from 82.5% (2007 EVD) to 33%, the PV increases by 6%. And when the solid consistency in the hot water pretreatment reactor is increased from 13% to 20%, the PV is reduced by 10%. Xylose and cellulose conversions in the pretreatment reactor of the two-stage dilute acid scenario showed significant impact on PV. The PV increased by 44% and is reduced by 10% when xylan to xylose conversion is reduced from 82.5% to 33% and cellulose to glucose conversion is increased from 6.3% to 23%, respectively. The impact of other pretreatment parameters on PV is not so significant.



	Dilute Acid	Hot Water	AFEX	Two-Stage
Solid loading [%] (EVD)	29.6	12.9	48	30
Solid loading [%] (high)	–	20.0	70	–
Solid loading [%] (low)	18.0	–	20	–
Xylan→Oligomer [%] (EVD)	2.7	0	0.5	2.7
Xylan→Oligomer [%] (high)	–	60	20	–
Xylan→Oligomer [%] (low)	–	25	–	–
Xylan→Xylose [%] (EVD)	82.5	2.4	0	82.5
Xylan→Xylose [%] (high)	89.7	7.3	–	89.7
Xylan→Xylose [%] (low)	33	–	–	33
Cellulose→Glucose [%] (EVD)	6.3	0.32	0	6.3
Cellulose→Glucose [%] (high)	23	2	–	23
Cellulose→Glucose [%] (low)	–	–	–	–
Acid [%] or NH_3 loading [ratio to dry stover] (EVD)	1.9	0	1:1	1.64
Acid [%] or NH_3 loading [ratio to dry stover] (high)	2.4	–	1:1.3	2.66
Acid [%] or NH_3 loading [ratio to dry stover] (low)	0.71	–	1:0.8	0.5

Figure 11. Impact of pretreatment parameters on PV

Among the saccharification parameters, cellulose to glucose conversions showed significant impact on PV for all scenarios (Figure 12; the detailed parameter values and results are tabulated in Table B-1 and Table C-1, respectively). For the dilute acid pretreatment process, the PV increased by 20% when the cellulose to glucose conversion was reduced from 91% to 67%. The impact on PV for the hot water pretreatment scenario was much higher. For the hot water scenario, the PV increased by 31% when the cellulose to glucose conversion was reduced from 90% to 65%.



	Dilute Acid	Hot Water	AFEX	Two-Stage
Xylan→Xylose [%] (EVD)	57.13	56.61	77.7	14.3
Xylan→Xylose [%] (high)	–	63	90	41.5
Xylan→Xylose [%] (low)	52.4	–	55	0
Cellulose→Glucose [%] (EVD)	91.09	89.97	95.9	42.4
Cellulose→Glucose [%] (high)	97	–	–	48
Cellulose→Glucose [%] (low)	67	65	75	–
Acid [%] or NH_3 loading [ratio to dry stover] (EVD)	–	–	–	1.80
Acid [%] or NH_3 loading [ratio to dry stover] (high)	–	–	–	2.6
Acid [%] or NH_3 loading [ratio to dry stover] (low)	–	–	–	–

Figure 12. Impact of saccharification parameters on PV

A sensitivity analysis for the major economic assumptions was performed on the dilute acid pretreatment process to test the robustness of the estimated PV. The selected sensitivity parameters were feedstock cost, enzyme loading, enzyme cost (for purchased enzymes), contingency factor, installation factor (or corresponding installed equipment cost), and byproduct (electricity) credit. The feedstock cost and enzyme price have the dominant impact on PV. Enzyme loading, contingency factor, and the total installed equipment cost showed moderate impact on PV. Results are shown in Figure 13 (the detailed parameter values and results are tabulated in Table B-2 and Table C-2, respectively).

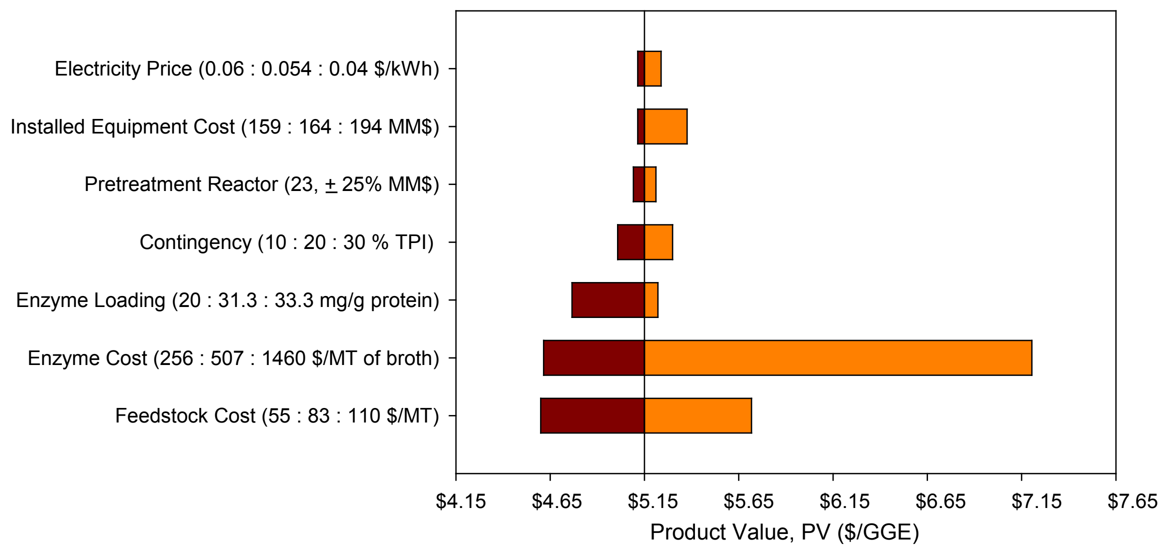


Figure 13. Impact of overall process/economic parameters on PV (dilute acid pretreatment)

When feedstock cost was increased from $83/dry MT (base case scenario) to $110/dry MT, PV increased by 11%. PV increased by 38% to $4.70/gal when enzyme cost was increased from $507/MT broth to $1460/MT (equivalent to $2.00/gal of ethanol produced). When enzyme cost was reduced to $256/MT (equivalent to $0.35/gal of ethanol produced), a 10% decrease in PV was observed. The exact cost of cellulase enzymes for large scale needs is not yet publicly available. A recent Novozymes presentation estimated enzyme costs in 2009 to be around $1–$2/gal of ethanol, and this range is included in the sensitivity analysis [21]. It may be important to further study the on-site enzyme production process with emphasis on microbial strain, protein yield, specific activity, residence time, oxygen requirement, and overall process area optimization. The current study is limited to publicly available data on those important parameters.

The enzyme loading in the saccharification reactor showed some impact on PV. When the enzyme loading was reduced from 31.3 mg protein/g cellulose (equivalent to 18.8 FPU/g cellulose) to 20 mg protein/g cellulose (equivalent to 12 FPU/g cellulose), the PV was reduced by 7%; the PV increased by 1% when the loading was increased from 31.3 to 33.3 mg protein/g cellulose (equivalent to 20 FPU/g cellulose).

Sensitivity analysis on total installed equipment cost was also performed. When the installed equipment cost was increased from $MM164.1 (corresponding weighted average installed factor of 2.58 for the base case scenario) to $MM194 (corresponding installed factor of 3.05 obtained from Peters and Timmerhaus [18]), the PV increased by 4%. Contingency factor showed a similar impact on PV. The impact of other parameters such as reactor cost and electricity price on PV was not significant.

Pioneer Plant Analysis Results

Table 5 shows the pioneer-plant PV results for all three cases. The PV for the most-probable case is \$5.76/gal, and that for the optimistic and pessimistic cases is \$5.01/gal and \$7.08/gal, respectively. The PV for the most probable, optimistic, and pessimistic cases is 69%, 47%, and 108% more, respectively, than the PV estimated for the n^{th} plant. The cost growth analysis shows that the TCI and Lang factor increased significantly from the base case n^{th} plant. For the most probable cost growth scenario, the TCI and Lang factor are MM\$886.4 and 8.11, respectively, which are an increase of 136% from the base case n^{th} plant values.

Table 5. Pioneer Plant Analysis Results for the Dilute Acid Pretreatment Process Scenario

Cost Item	Cost Growth (Pioneer Plant)		
	Most Probable	Optimistic	Pessimistic
PV (\$/gal)	5.76	5.01	7.08
Fixed Capital Investment (MM\$)	833	674	1,111
Total Capital Investment (MM\$)	886	727	1,164
Lang Factor	8.11	6.65	10.65

Comparison with Previous Studies

The results of this study deviate considerably from a number of previous techno-economic analyses of cellulosic ethanol production. There are many contributing factors to this deviation, and an explanation of the most significant of these factors is discussed here. Figure 14 presents a plot of estimated ethanol prices from seven previous studies as a function of feedstock price. The ethanol and feedstock prices were updated to 2007 dollars using the Consumer Price Index. The solid line on the plot represents the PV for the dilute acid pretreatment scenario using the model developed in this study as a function of feedstock price.

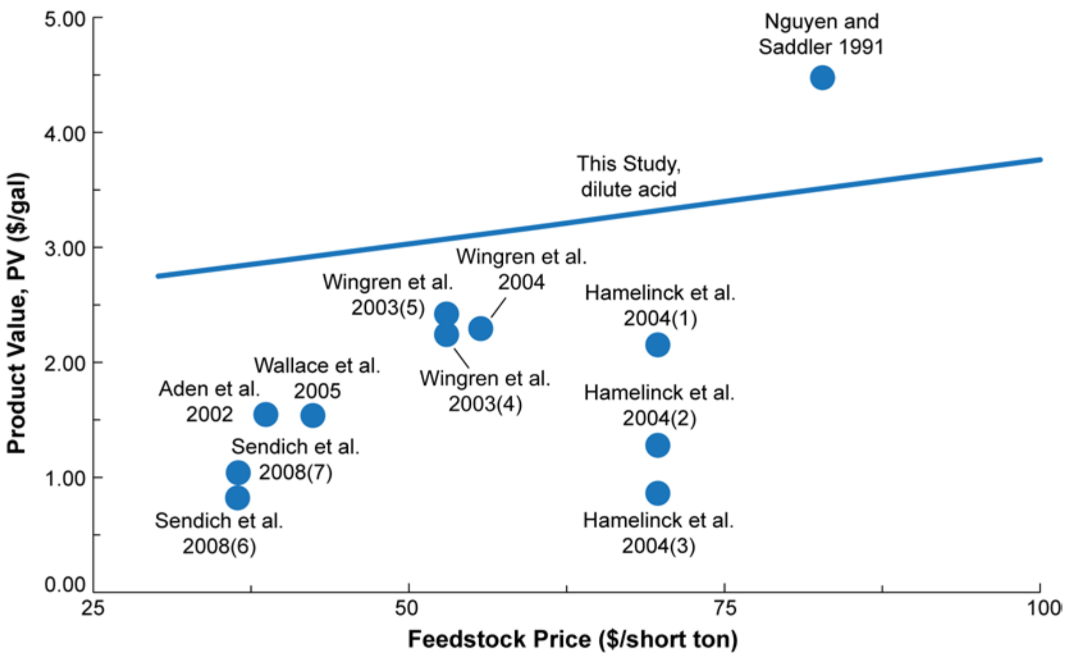


Figure 14. Ethanol cost estimations from previous techno-economic studies.
(1) Short term technology—Simultaneous saccharification and fermentation (SSF). (2) Middle term technology—Simultaneous saccharification and cofermentation (SSCF). (3) Long term technology—Consolidated bioprocessing (CBP). (4) Separate hydrolysis and fermentation (SHF). (5) SSF. (6) CBP. (7) SSCF.

After updating the feedstock and ethanol prices to 2007 dollars, much of the difference from previous studies can be explained by the clear correlation that exists between feedstock price and ethanol price. However, all of the studies except that of Nguyen and Saddler [22] remain lower than the line derived from this study. The study by Hamelinck et al. [23] represents a significant outlier from the apparent correlation between feedstock price and ethanol price. The three ethanol price estimates are for short-term (5 years from time of study), middle-term (10–15 years), and long-term (20+ years) technology implementation. The short-term estimate is closest to the time frame considered in this study. However, it also deviates from the trend of other studies. The assumptions for the short-term estimate—including feedstock input, rate of return, and reaction conversions—are quite similar to those used in this study, and the TCI (updated to 2007 dollars) is nearly equal as well. The most significant difference from this study is the non-feedstock operating cost, which is approximately $0.32/gal ethanol (EtOH) compared to

$1.68/gal. This is partly due to lower costs for corn steep liquor, cellulase, and other raw materials. This factor accounts for most of the discrepancy between ethanol price estimates.

The ethanol price from the study published by Sendich et al. [24] is also slightly lower than the apparent correlation of feedstock and ethanol price. The lowest estimate in that study assumes the use of consolidated bioprocessing, which is an advanced technology also modeled in the long-term estimate from Hamelinck et al. [23]. The higher ethanol price estimate of $1.03/gal is from a model using simultaneous saccharification and cofermentation (SSCF). SSCF is a more advanced technology than was considered in this study; this factor results in lower capital and operating costs by combining enzymatic saccharification and fermentation. A new AFEX pretreatment scheme was also employed, which may have contributed to lower capital and operating costs of pretreatment.

The enzyme cost used in this study is much higher than that used in other studies. Because enzyme cost is such a significant fraction of the PV, it contributes significantly to the discrepancy between the current study and previous studies. For example, the enzyme prices used in the prior studies of Wingren et al. (2003), Wingren et al. (2004), and Aden et al. [25, 26, 4] are approximately 30%, 30%, and 17% of the price used in this study, respectively.

Conclusions

The present study is based on published technological and economical data that in many ways lacks specifics and details. Thus, a list of assumptions was developed. Sensitivity analysis on those assumptions was performed, and their impact on TCI and PV was wide. The studies identify the strengths and weaknesses of the technology. The major strength of the technology is its pre-commercial maturity. Several pilot-scale cellulosic ethanol plants are in development today. The U.S. Department of Energy has funded 10 pioneer plant projects producing biofuels using a biochemical pathway [27]. However, technological and economic data from these pilot and demo plants are not publically available for use in this study.

The published data in 2007 shows that the technology can reach 4.5% of ethanol concentration in the fermented beer, which is approximately one-third of what commercial grain ethanol plants are achieving. The enzyme and feedstock costs are two major cost contributors. In our studies, cellulase price is assumed to be nearly $0.70/gal of ethanol produced and its loading is 18.8 FPU/g of cellulose. The high cost of enzymes represents a significant opportunity to reduce the PV through improved biotechnology. The current feedstock cost is assumed to be $83/MT. However, the cost is likely to be location sensitive. It is important to critically estimate the feedstock cost, which may limit the plant locations to certain areas. Published improvements in process technology since 2007 have not been considered in this work, and these improvements would likely decrease PV for a project being designed today.

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Appendices

Appendix A - Assumptions for Techno-economic Studies of Biochemical Conversion Processes

Plant Size, Location, and Construction

- Optimum plant size is regarded as economically feasible plant size, which would be a plant with capacity of 2000 MT/day (dry feedstock).
- The plant produces EtOH (as product) and electricity (as byproduct).
- The plant is considered to be located in the middle of corn farmland.
- 25% of the land will be tied up in infrastructure (roads and buildings).
- 75% of the farm land plants corn.
- The plant will be designed based on the state of the technology (as of 2007), and it would be the n^{th} plant of its kind. RAND/risk analysis will inflate price from mature n^{th} plant to immature pioneer plant.
- The online time would be 350 days per year (equivalent capacity factor of 96%).
- Construction time of 24 months is considered.
- Startup period would be 25% of the construction time (6 months).
- During this period, an average of 50% production will be achieved with expenditure of about 75% of variable expenses and 100% of fixed expenses.

Units

- Plant capacity, mass flow, and yields are based on metric tons (MT) per day.
- Feedstock cost and purchased chemicals costs are based on short tons (ST).
- Ethanol sale price is in $/gallon, and byproduct credit (electricity) is in $/kWh.
- Process operating conditions are:
    - Temperature in ^{o}C
    - Pressure in bar
    - Mass flow rates in MT/day.

Feedstock and Enzymes

- Corn stover (composed of stalks, leaves, cobs, and husks) is considered as feedstock.
    - The feedstock will be delivered to the feed handling area of the plant.
    - Moisture content in the feedstock is 25% (wet basis).
    - Feed composition is assumed to be the same as CAFI II analysis results. Table A-1 displays feedstock composition.
    - Feedstock is delivered in bales.
    - The feedstock transportation and management protocol are not considered.

- Feed cost is assumed to be $83/MT ($75/dry short ton) at the gate.

- 72 hours of on-site storage (corn stover bales) is considered, and long term storage costs are not included in the analysis.
- Stover is washed of dirt and metal is removed; no biomass is lost in washing.
- Enzyme will be purchased or produced on-site through purchase/licensing agreements with enzyme suppliers.
  - Enzyme loading is assumed as 31.3 mg protein/g cellulose in original feed.
  - Enzyme cost is $507/MT broth when purchased from off-site sources.

Table A-1. Corn Stover Feedstock Composition

Components	Composition (%)		Components	Composition (%)
Extractives	8.26		Lignin	10.69
Cellulose	33.43		Ash	5.93
Xylan	22.16		Acetate	5.44
Galactan	1.36		Protein	2.24
Arabinan	4.08		Soluble Solids	5.83
Mannan	0.58		Moisture	25

Material and Energy Balance

Material Balance

- Reactions and conversions of hemicellulose carbohydrates (arabinan, mannan, galactan) are assumed to have the same value as xylan in the pretreatment hydrolyzer (depends on pretreatment).
- A total of 7% of fermenting sugars is assumed to be lost to contamination, which may be regarded as bad batches (the bad batches are dumped and regarded as loss).
- The total amount of water from the pressure filter (bottom product of the first distillation column) that is directly recycled is set to be 25% (to minimize contaminant buildup in the stream).
- Boiler blow down is considered to be 3% of steam production.
- Carbon efficiency can be calculated based on carbohydrate carbon content, as follows:

$$Carbon\ efficiency\ (\%) = \frac{Carbon\ in\ Ethanol}{Carbon\ in\ Biomass\ Carbohydrate} \times 100$$

- Cooling tower windage is 0.1% of the total flow to the tower.
  - The tower blow down is 10% of the of the sum of the evaporative loss plus windage.
- Well water will be used as process makeup water (lost to evaporation, blow down, windage, in solid waste).

Energy Balance

- Heat loss from the reactor will not be accounted for in energy balance calculations.
- Heat loss from the combustor is accounted for, totaling about 2.2%.
- Electricity will be generated by burning lignin and waste (process waste and pressed solids from wastewater treatment), which will be used in the plant, and the net surplus will be sold to grid at a price of $0.054/kW (2007 dollar value).

Equipment Design, Material of Construction, and Costing

Equipment Design

- The reactors will be modeled using experimentally determined conversions of specific reactions (kinetic expressions will be not used because of the level of their development).
- If the size of any equipment is known to change linearly with the inlet flow, that information can be used for equipment scaling (a characteristic of the size might be the heat duty for a heat exchanger if the log-mean temperature difference is known not to change).
- For some equipment, nothing can be easily related to the size, in which case the unit will be resized with each process change (for example heat exchangers with varying temperature profiles; in this case, the heat exchanger area will be calculated each time the model will be run and the cost will be scaled using the ratio of the new and original areas).

Material of Construction

- The materials of construction for all equipment (except the pretreatment reactor, flash tank, and Pneumapress equipment) will be as follows (Delta-T/NREL experience):
    - SS316 for flash tank (for the solid-liquid separation equipment)
    - Incoloy 825-clad steel for continuous pretreatment reactor and its parts in contact with acid, and SS316L for most other parts including presteamer.

Costing

- All pumps, tanks, screw conveyors, agitators, heat exchangers, and surge tanks will be estimated using ICARUS Process Evaluator and NREL database.
- Large vessels (saccharification tank, fermentors, seed hold tanks), boiler feed water softening equipment, anaerobic and aerobic digesters, and filter press will be estimated from quotation.
- Smaller vessels (seed fermentor) and coil coolers will be estimated by ICARUS and NREL database.
- If process changes are made and the equipment size changes, the equipment will be re-costed following the exponential scaling expression:

$$New\ Cost = Original\ \mathrm{Cost}\left(\frac{\mathrm{New\ size}*}{\mathrm{Original\ size}*}\right)^{\mathrm{exp}}$$

*or characteristic linearly related to the size

    - The scaling exponents are obtained from NREL's vendor quote database.
- The purchased equipment cost obtained in a particular year (before 2007) will be indexed to the year of 2007 using the Chemical Engineering Index.
- Installation factors (from Delta-T/NREL experience for aqueous-based process) will be applied to purchased equipment costs to determine the installation cost (not ICARUS).

Chemical Costing

- Cost of acids and other chemicals, including boiler feed water softening chemical cost, will be obtained from quotation.
    - The cost of the chemicals will also be indexed following Industrial Inorganic Chemical Index (from SRI) to estimate the cost of the chemicals in the year of 2007.

Operating Cost

- Working capital is assumed to be 15% of fixed capital investment.
    - It is assumed that the product will be made and shipped and payment received in 30 days.
- Annual maintenance materials will be 2% of the total installed equipment cost.
- Employee salaries will be indexed, if required, to the year 2007 following the data of the Bureau of Labor Statistics.
- Salaries of the yard employees will not include benefits and will be covered in the general overhead category.
    - General overhead will be a factor of 60% applied to the total salaries and covers items such as safety, general engineering, general plant maintenance, payroll overhead (including benefits), plant security, janitorial and similar services, phone, light, heat, and plant communications.

Wastewater Treatment Plant

- The process will be designed for zero discharge to a municipal treatment plant in a steady-state mode, and the treated water will be suitable for recycling to the process.
- Any process upset (sudden increase of solids in the wastewater) will not be considered in the model.
- Rain and snow run-off, equipment washing, and other non-process waters are assumed to flow to the municipal wastewater treatment system; other intermittent loads (process spills) will not be considered in the design.
- No insoluble components (cellulose, xylan) will be included in the chemical-oxygen-demand (COD) calculations because of uncertainty of their reactivity.
- Biological oxygen demand in the anaerobic digester is assumed to be 70% of the COD.
    - COD reduction in both digesters is considered as 99.4%.

Greenhouse Emissions and Control

- All of the sulfur entering into the combustor is converted to SO_2.
  - 1% of the generated SO_2 is converted to sulfuric acid.
  - Flue gas temp will be kept above the dew point of sulfuric acid.
- Carbon monoxide is assumed to be generated at a rate of 0.31 kg/MWhr.
  - Unburned carbon (char) in the ash is low at 1%.
- NO_x is generated at 0.31 kg/MWhr.
- Baghouse efficiency is taken as 98.8% (from Foster Wheeler Energy Limited experience) and will control the emission level below new source performance standard limit.
- The nitrogen level in the combined feed to the combustor is similar to coal when ammonia species are included, but more like untreated biomass when they are not.
  - Impact of global warming potential of NOx is small relative to global warming potential of CO_2 emission from the process.

Cost Analysis

- The total plant investment cost will be determined by applying overhead and contingency factors (NREL, ConocoPhillips Company experience and literature) to installed equipment costs.
  - 20% (of total direct and indirect cost) contingency is assumed for n^{th} plant, and 30% for pioneer plant.
- Warehouse cost is 1.5% of total installed equipment cost, and site development cost is 9% of the installed cost of process equipment areas (A100, A200, A300 and A500) (Delta-T/NREL/published data).
  - The estimates are location sensitive.
- Total installed cost (TIC) includes total installed equipment cost, warehouse cost, and site development cost.
- Indirect cost involves:
  - Engineering and supervision (32% of purchased equipment cost)
  - Construction expenses (34% of purchased equipment cost)
  - Legal and contractor's fees (23% of purchased equipment cost).
- Total direct and indirect cost (TD&IC) includes TIC and indirect costs.
- Working capital is assumed as 15% of FCI (FCI is the sum of TD&IC and contingency).
- Total capital investment is the sum of FCI and working capital.
- To determine the product value per gallon of ethanol (PV), a discounted cash flow analysis will be used (after knowing the major three costs areas: (i) total capital investment, (ii) variable operating costs, and (iii) fixed operating cost).

    - A 10% discounted cash flow rate of return will be used over a 20-year plant life.
    - The plant is considered 100% equity financed.
- For federal tax return purposes, depreciation will be determined as follows :
    - IRS modified accelerated cost recovery system, which includes general depreciation system, will be followed that allows both the 200% and 150% declining balance (DB) methods of depreciation.
        - This allows the shortest recovery period and the largest deductions.
    - According to the IRS, the steam production plant should use a 20-year recovery period (depreciated over 20 years).
    - Any other property not specifically described in the publication should be depreciated using a 7-year recovery period.
        - Property listed with a recovery period less than 10 years will use the 200% DB depreciation method and a 20-year recovery period property will use the 150% DB depreciation.
    - State tax will not be considered for the calculation (because the location of the plant is not specified).
- Return on investment will be calculated on a per gallon basis. Income tax will be averaged over the plant life and that average will be calculated on a per gallon basis.

Appendix B - Sensitivity Parameters and Values

Table B-1. Sensitivity Parameters for Pretreatment and Saccharification (AREA 200)

Parameters	**Dilute Acid (2007 EVD)**		**2-Stage Dilute acid**				**Hot Water**		**AFEX**	
	Base Case (50%)	Sensitivity (0-100%)	Base Case (50%)		Sensitivity (0-100%)		Base Case (50%)	Sensitivity (0-100%)	Base Case (50%)	Sensitivity (0-100%)
			1st Stage	2nd Stage	1st Stage	2nd Stage				
Pretreatment										
Temperature (^oC)	190	190-200	190	210	180		190	190-200	108	90-110
Pressure (atm)	11.6		12	18.5			12.6	12.6-15.4	18.7	
Retention time (min)	2	1-10	2				15	15-5	5	5
Catalyst (Acid/NH_3) conc (%)	1.9	0.71-2.4	1.64	1.80	0.5-2.66	2.6	0	N/A	1.1	0.5-1.3
Solid consistency (%)	29.6	18-29.6%	30	30			12.9	12.9-20	48	20-70
Conversion (%)										
Cellulose → Glucose	6.3	6.3-23	6.3	42.4	23	48	0.32	2-2.5	0	N/A
Xylan → Xylose oligomer	2.7		2.7	0			0	N/A	0.5	N/A
Xylan → Xylose	82.5	33-89.7	82.5	14.3	33-89.7	0-41.5	2.4	7.3-25	0	N/A
Saccharification										
Conversion (%)										
Cellulose → Glucose	91.09	67-97					89.97	65	95.9	75
Xylan → Xylose	57.13	52.4					56.61	63	77.7	55-90

Table B-2. Sensitivity Parameters for Overall Process

Parameters	Dilute Acid (2007 EVD)		2-Stage Dilute acid		Hot Water		AFEX	
	Base Case (50%)	Sensitivity (0-100%)	Base Case (50%)	Sensitivity (0-100%)	Base Case (50%)	Sensitivity (0-100%)	Base Case (50%)	Sensitivity (0-100%)
Feedstock Cost ($/MT)[a]	83	55-110	55	83-110	83	55-110	83	55-110
Pretreatment Reactor Installed Cost (MM$)[b]	23	14-32.7			0.32	0.22-0.62	25.7	20-40
Project Contingency (%)	20	10-30	20	20-30	20	10-30	20	10-30
Installation Factor or Cost (MM$)								
Factor	2.58	2.50-3.05			2.59	2.50-3.05	2.05	2-4
Total Installed Equipment Cost	164.1	159-194			156	151-184	211.5	200-275
Enzyme Cost ($/MT broth)	507	256-1460			507	256-1460	507	256-1460
Enzyme loading (mg protein/g cellulose)	31.3	20-33.3	31.3	20-33.3	31.3	20-33.3	31.3	20-33.3
Value of Excess Electricity/ Byproduct Credit ($/kWh)	0.054	0.03-0.06	0.054	0.03-0.05	0.04	0.03-0.05	0.04	0.03-0.05

[a] Feedstock contains 25% moisture. [b] 2007 indexed installed equipment cost.

Appendix C - Sensitivity Results

Table C-1. Impact of Pretreatment Parameters on PV

		Sensitivity		Production (MMGal/Yr)	PV	
		Parameter	**Values**		**($/Gal)**	**Change (%)[a]**
Dilute Acid Pretreatment — 2007 EVD Scenario						
Base Case				53.4	3.40	
Pretreatment	High[b] Scenarios	Reactor temperature (°C)	200.0	53.4	3.40	0
		Residence time (min)	10	53.4	3.93	15.6
		Acid concentration (%)	2.4	53.4	3.40	0
		Cellulose to Glucose (% conv)	23	53.4	3.42	0.6
		Xyl to Xylose (% conv)	89.7	53.9	3.37	-0.9
	Low[b] Scenarios	Residence time (min)	1	53.4	3.34	-1.8
		Solid consistency (%)	18.0	53.4	3.47	2.1
		Acid concentration (%)	0.71	53.4	3.40	0
		Xyl to Xylose (% conv)	33	50.1	3.61	6.2
Saccharification	High Scenarios	Cellulose to Glucose (% conv)	97	54.9	3.31	-2.6
	Low Scenarios	Cellulose to Glucose (% conv)	67	44.6	4.07	19.7
		Xylan to Xylose (% conv)	52.4	53.3	3.41	0.3
Two-Stage Dilute Acid Pretreatment — 2007 EVD Scenario						
Base Case				42.5	4.38	
1st Stage Acid Pretreatment (Pretreatment)	High Scenarios	Cellulose to Glucose (% conv)	23	47.3	3.93	-10.3
		Xylan to Xylose (% conv)	89.7	44.4	4.19	-4.3

		Sensitivity		Production (MMGal/Yr)	PV	
		Parameter	Values		($/Gal)	Change (%)[a]
		Other hemicellulose to monomers (% conv)	90	42.5	4.38	0
		Acid Concentration (%)	2.66	42.5	4.38	0
	Low Scenarios	Reactor temperature (°C)	180	42.5	4.37	-0.2
		Acid concentration (%)	0.5	42.5	4.38	0
		Xylan to Xylose (% conv)	33	29.5	6.29	43.6
2nd Stage Acid Pretreatment (Saccharification)	High Scenarios	Cellulose to glucose (% conv)	48	45.7	4.08	-6.8
		Xylan to Xylose (% conv)	41.5	43.9	4.24	-3.2
		Acid concentration	2.6	42.4	4.41	0.7
	Low Scenarios	Xylan to Xylose (% conv)	0	41.8	4.45	1.6
Hot Water Pretreatment — 2007 EVD Scenario						
Base Case				39.0	4.29	
Pretreatment	High Scenarios	Reactor temperature (°C)	200	39.0	4.29	0
		Residence time (min)	20	39.0	4.29	0
		Solid consistency (%)	20.0	39.0	3.84	-10.2
		Cellulose to Glucose (% conv)	2	39.1	4.29	0
		Xylan to Oligomer (% conv)	60	38.5	4.36	1.6
		Xylan to Xylose (% conv)	7.3	39.5	4.24	-1.1
	Low Scenarios	Residence time (min)	5	39.0	4.29	0
		Xylan to oligomer (% conv)	25	42.9	3.92	-8.4

		Sensitivity		Production (MMGal/Yr)	PV	
		Parameter	Values		($/Gal)	Change (%)[a]
		Xylan to Xylose (% conv)	63	39.6	4.23	-1.4
Saccharification	High Scenarios	Cell Cellulose to Glucose (% conv)	65	29.7	5.61	30.7
	Low Scenarios	Cellulose to Glucose (% conv)	65	29.7	5.61	30.7
AFEX — 2007 EVD Scenario						
Base Case				46.2	3.69	
Pretreatment	High Scenarios	Reactor temperature (^{o}C)	110	46.1	3.69	0
		Solids consistency (%)	70	46.2	4.16	12.7
		Xylan to Oligomer (%conv)	20	46.2	3.69	0
		Ammonia loading (ratio to dry stover)	1:0.8	46.2	3.69	0
	Low Scenarios	Reactor temperature (^{o}C)	80	46.0	3.68	-0.3
		Solid consistency (%)	20	46.2	3.31	-10.3
		Xylan to Xylose (%conv)	90	46.9	3.65	-1.1
		Ammonia loading (ratio to dry stover)	1:1.3	46.2	3.70	0.3
Saccharification	High Scenarios	Xylan to Xylose (%conv)	55	45.8	3.80	3.0
	Low Scenarios	Cellulose to Glucose (%conv)	75	43.2	3.87	4.9

[a] Percentage change of PV from 2007 EVD scenarios. [b] High and low refer to the value of each parameter with respect to the base case. Not all parameters were tested at both high and low scenarios.

Table C-2. Impact of Overall Process and Economic Parameters on PV

Sensitivity Parameters	**Base Case**		**Low**		**High**	
	Value	PV ($/gal)	Value	PV ($/gal)	Value	PV ($/gal)
Feedstock Cost ($/MT)	83	3.40	55	3.04	110	3.77
Pretreatment Reactor (MM$)	23	3.40	14	3.34	32.7	3.47
Installed Equipment Cost (MM$)	164.1	3.40	159	3.38	194.0	3.55
(Corresponding Installation factor)	(2.58)		(2.5)		(3.05)	
Enzyme Cost ($/MT of broth)	507	3.40	256	3.05	1460	4.73
Electricity Price ($/kWh)	0.054	3.40	0.06	3.38	0.03	3.50
Enzyme Loading (mg/g protein)	31.3	3.40	20	3.15	33.3	3.45
Contingency (%)	20	3.40	10	3.31	30	3.50

Appendix D - Cost Analysis Result Summary

Table D-1. Cost Analysis Result Summary for Dilute Acid Pretreatment Processes

Ethanol Production Process Engineering Analysis			
UCR Dilute Acid - Corn Stover, Current Case Dilute Acid Prehydrolysis with Saccharification and Cofermentation All Values in 2007$			
Product Value	**$3.40**		
Ethanol Production (MM Gal. / Year)	53.4	Ethanol at 68°F	
Ethanol Yield (Gal / Dry Metric Ton Feedstock)	76.3		
Feedstock Cost $/Dry MetricTon	$83		
Internal Rate of Return (After-Tax)	10%		
Equity Percent of Total Investment	100%		
Capital Costs		**Operating Costs (cents/gal ethanol)**	
Feed Handling	$10,900,000	Feedstock	108.4
Pretreatment	$36,200,000	Biomass to Boiler	0.0
Saccharification & Fermentation	$21,800,000	CSL	16.0
Distillation and Solids Recovery	$26,100,000	Cellulase	69.5
Wastewater Treatment	$3,500,000	Other Raw Materials	17.8
Storage	$3,200,000	Waste Disposal	12.7
Boiler/Turbogenerator	$56,100,000	Electricity	-21.9
Utilities	$6,300,000	Fixed Costs	18.5
Total Installed Equipment Cost	**$164,100,000**	Capital Depreciation	30.5
		Average Income Tax	26.7
Added Costs[a]	$211,800,000	Average Return on Investment	62.2
(% of TCI)	56%	**Operating Costs ($/yr)**	
Working Capital	49,030,000	Feedstock	$57,900,000
Total Capital Investment	**$375,900,000**	Biomass to Boiler	$0
		CSL	$8,500,000
Installed Equipment Cost/Annual Gallon	$3.07	Cellulase	$37,100,000
Total Project Investment/Annual Gallon	$7.04	Other Raw Matl. Costs	$9,500,000
		Waste Disposal	$6,800,000
Loan Rate	N/A	Electricity	-$11,700,000
Term (years)	N/A	Fixed Costs	$9,900,000
Capital Charge Factor	0.170	Capital Depreciation	$16,300,000
		Average Income Tax	$14,300,000
Denatured Fuel Prod. (MMgal / yr)	55.9	Average Return on Investment	$33,300,000
Denatured Fuel Min. Sales Price	$3.29	Excess Electricity (KWH/gal)	4.06
Denaturant Cost ($/gal denaturant)	$0.739	Plant Electricity Use (KWH/gal)	2.23
		Plant Steam Use (kg steam/gal)	17.5
Maximum Yields (100% of Theoretical)		Boiler Feed–LHV (Btu/lb)	2,209
Ethanol Production (MM Gal/yr)	82.5	Boiler Feed–Water Fraction	0.542
Theoretical Yield (Gal/ton)	106.9	**Specific Operating Conditions**	
Current Yield (Actual/Theoretical)	65%	Enzyme Loading (mg/g cellulose)	33.5
		Saccharification Time (days)	5.0
		Conversion Cellulose-->Glucose	0.9109
		Fermentation Time (days)	2.0

[a] Added costs include indirect costs (engineering and supervision, construction expenses, legal and contractor fees) and contingency.

Table D-2. Cost Analysis Result Summary for Dilute Acid Pretreatment (High Solids) Processes

Ethanol Production Process Engineering Analysis			
UCR Dilute Acid - Pretreatment yields based on NREL FY08 SOT Dilute Acid Prehydrolysis with Saccharification and Cofermentation All Values in 2007$			
Product Value	**$3.60**		
Ethanol Production (MM Gal. / Year)	50.8	Ethanol at 68°F	
Ethanol Yield (Gal / Dry Metric Ton Feedstock)	72.5		
Feedstock Cost $/Dry Metric Ton	$83		
Internal Rate of Return (After-Tax)	10%		
Equity Percent of Total Investment	100%		

Capital Costs		**Operating Costs (cents/gal ethanol)**	
Feed Handling	$10,900,000	Feedstock	114.0
Pretreatment	$38,000,000	Biomass to Boiler	0.0
Saccharification & Fermentation	$21,800,000	CSL	16.5
Distillation and Solids Recovery	$25,700,000	Cellulase	73.1
Wastewater Treatment	$5,800,000	Other Raw Materials	18.8
Storage	$3,100,000	Waste Disposal	13.0
Boiler/Turbogenerator	$57,600,000	Electricity	-24.8
Utilities	$6,600,000	Fixed Costs	19.8
Total Installed Equipment Cost	**$169,400,000**	Capital Depreciation	33.3
		Average Income Tax	29.0
Added Costs[a]	$219,100,000	Average Return on Investment	67.5
(% of TCI)	56%	**Operating Costs ($/yr)**	
Working Capital	50,670,000	Feedstock	$57,900,000
Total Capital Investment	**$388,500,000**	Biomass to Boiler	$0
		CSL	$8,400,000
Installed Equipment Cost/Annual Gallon	$3.34	Cellulase	$37,100,000
Total Project Investment/Annual Gallon	$7.65	Other Raw Matl. Costs	$9,600,000
		Waste Disposal	$6,600,000
Loan Rate	N/A	Electricity	-$12,600,000
Term (years)	N/A	Fixed Costs	$10,100,000
Capital Charge Factor	0.169	Capital Depreciation	$16,900,000
		Average Income Tax	$14,700,000
Denatured Fuel Prod. (MMgal / yr)	53.1	Average Return on Investment	$34,200,000
Denatured Fuel Min. Sales Price	$3.47		
Denaturant Cost ($/gal denaturant)	$0.739	Excess Electricity (KWH/gal)	4.58
		Plant Electricity Use (KWH/gal)	2.39
Maximum Yields (100% of Theoretical)			
Ethanol Production (MM Gal/yr)	82.5	Plant Steam Use (kg steam/gal)	18.2
Theoretical Yield (Gal/ton)	106.9	Boiler Feed -- LHV (Btu/lb)	2,286
Current Yield (Actual/Theoretical)	62%	Boiler Feed -- Water Fraction	0.539
		Specific Operating Conditions	
		Enzyme Loading (mg/g cellulose)	35.1
		Saccharification Time (days)	5.0
		Conversion Cellulose --> Glucose	0.9109
		Fermentation Time (days)	2.0

[a] Added costs include indirect costs (engineering and supervision, construction expenses, legal and contractor fees) and contingency.

Table D-3. Cost Analysis Result Summary for Hot Water Pretreatment Process

Ethanol Production Process Engineering Analysis

Hot Water - Corn Stover, Current Case
Hot Water Prehydrolysis with Saccharification and Cofermentation
All Values in 2007$

Product Value	**$4.44**	
Ethanol Production (MM Gal. / Year)	39.0	Ethanol at 68°F
Ethanol Yield (Gal / Dry Metric Ton Feedstock)	55.7	
Feedstock Cost $/Dry MetricTon	$83	
Internal Rate of Return (After-Tax)	10%	
Equity Percent of Total Investment	100%	

Capital Costs	
Feed Handling	$10,900,000
Pretreatment	$6,700,000
Saccharification & Fermentation	$30,200,000
Distillation and Solids Recovery	$30,900,000
Wastewater Treatment	$1,900,000
Storage	$3,300,000
Boiler/Turbogenerator	$65,800,000
Utilities	$6,700,000
Total Installed Equipment Cost	**$156,300,000**
Added Costs[a]	$204,800,000
(% of TCI)	57%
Working Capital	47,100,000
Total Capital Investment	**$361,100,000**
Installed Equipment Cost/Annual Gallon	$4.01
Total Project Investment/Annual Gallon	$9.26
Loan Rate	N/A
Term (years)	N/A
Capital Charge Factor	0.181
Denatured Fuel Prod. (MMgal / yr)	40.8
Denatured Fuel Min. Sales Price	$4.27
Denaturant Cost ($/gal denaturant)	$0.739
Maximum Yields (100% of Theoretical)	
Ethanol Production (MM Gal/yr)	82.5
Theoretical Yield (Gal/ton)	106.9
Current Yield (Actual/Theoretical)	47%

Operating Costs (cents/gal ethanol)	
Feedstock	148.4
Biomass to Boiler	0.0
CSL	28.5
Cellulase	95.1
Other Raw Materials	5.1
Waste Disposal	3.5
Electricity	-29.0
Fixed Costs	24.5
Capital Depreciation	40.2
Average Income Tax	36.3
Average Return on Investment	90.9

Operating Costs ($/yr)	
Feedstock	$57,900,000
Biomass to Boiler	$0
CSL	$11,100,000
Cellulase	$37,100,000
Other Raw Matl. Costs	$2,000,000
Waste Disposal	$1,400,000
Electricity	-$11,300,000
Fixed Costs	$9,600,000
Capital Depreciation	$15,700,000
Average Income Tax	$14,200,000
Average Return on Investment	$35,500,000
Excess Electricity (KWH/gal)	5.37
Plant Electricity Use (KWH/gal)	3.30
Plant Steam Use (kg steam/gal)	40.7
Boiler Feed -- LHV (Btu/lb)	2,354
Boiler Feed -- Water Fraction	0.529

Specific Operating Conditions	
Enzyme Loading (mg/g cellulose)	33.1
Saccharification Time (days)	5.0
Conversion Cellulose --> Glucose	0.8997
Fermentation Time (days)	2.0

[a] Added costs include indirect costs (engineering and supervision, construction expenses, legal and contractor fees) and contingency.

Table D-4. Cost Analysis Result Summary for Two-Stage Dilute Acid Pretreatment Processes

Ethanol Production Process Engineering Analysis

Corn Stover Design Report Case: 2020 Market Target Case (2002 Design Report Target Case)
Dilute Acid Prehydrolysis and Hydrolysis
All Values in 2007$

Product Value	**$4.38**	
Ethanol Production (MM Gal. / Year)	32.8	Ethanol at 68°F
Ethanol Yield (Gal / Dry Metric Ton Feedstock)	46.9	
Feedstock Cost $/Dry Metric Ton	$83	
Internal Rate of Return (After-Tax)	10%	
Equity Percent of Total Investment	100%	

Capital Costs		**Operating Costs (cents/gal ethanol)**	
Feed Handling	$10,900,000	Feedstock	176.4
Pretreatment	$44,900,000	Biomass to Boiler	0.0
Saccharification & Fermentation	$9,700,000	CSL	26.2
Distillation and Solids Recovery	$26,700,000	Cellulase	0.0
Wastewater Treatment	$4,600,000	Other Raw Materials	32.3
Storage	$2,800,000	Waste Disposal	22.0
Boiler/Turbogenerator	$66,200,000	Electricity	-51.3
Utilities	$6,900,000	Fixed Costs	31.0
Total Installed Equipment Cost	**$172,700,000**	Capital Depreciation	51.8
		Average Income Tax	45.8
Added Costs[a]	$218,300,000	Average Return on Investment	103.5
(% of TCI)	56%	**Operating Costs ($/yr)**	
Working Capital (WC)	51,000,000	Feedstock	$57,900,000
Total Capital Investment	**$391,000,000**	Biomass to Boiler	$0
		CSL	$8,600,000
Installed Equipment Cost/Annual Gallon	$5.26	Cellulase	$0
Total Project Investment/Annual Gallon	$11.92	Other Raw Matl. Costs	$10,600,000
		Waste Disposal	$7,200,000
Loan Rate	N/A	Electricity	-$16,800,000
Term (years)	N/A	Fixed Costs	$10,200,000
Capital Charge Factor	0.169	Capital Depreciation	$17,000,000
		Average Income Tax	$15,000,000
Denatured Fuel Prod. (MMgal / yr)	34.3	Average Return on Investment	$34,000,000
Denatured Fuel Min. Sales Price	$4.21		
Denaturant Cost ($/gal denaturant)	$0.739	Excess Electricity (KWH/gal)	9.50
		Plant Electricity Use (KWH/gal)	3.46
		Plant Steam Use (kg steam/gal)	35.4
Maximum Yields (100% of Theoretical)		Boiler Feed -- LHV (Btu/lb)	2,379
Ethanol Production (MM Gal/yr)	82.5	Boiler Feed -- Water Fraction	0.521
Theoretical Yield (Gal/ton)	106.9	**Specific Operating Conditions**	
Current Yield (Actual/Theoretical)	40%	Feed rate (dry tonnes/day)	2,000
		Feed rate (dry tons/day)	2,205
		Lignin Residue (dry tonnes/day)	979
		Lignin Residue (dry tons/day)	1,079
		Pretreatment Total Solids (wt%)	30
		Saccharification Total Solids (wt%)	34

[a] Added costs include indirect costs (engineering and supervision, construction expenses, legal and contractor fees) and contingency.

Table D-5. Cost Analysis Result Summary for AFEX Pretreatment Processes

Ethanol Production Process Engineering Analysis		
Corn Stover Design Report Case: 2020 Market Target Case (2002 Design Report Target Case) AFEX with Saccharification and Cofermentation All Values in 2007$		
Product Value	**$3.69**	
Ethanol Production (MM Gal. / Year)	46.2	Ethanol at 68°F
Ethanol Yield (Gal / Dry Metric Ton Feedstock)	66.0	
Feedstock Cost $/Dry Metric Ton	$83	
Internal Rate of Return (After-Tax)	10%	
Equity Percent of Total Investment	100%	

Capital Costs		**Operating Costs (cents/gal ethanol)**	
Feed Handling	$10,900,000	Feedstock	125.3
Pretreatment	$30,800,000	Biomass to Boiler	0.0
		CSL	19.4
Saccharification & Fermentation	$23,500,000	Cellulase	80.4
Distillation and Solids Recovery	$27,500,000	Other Raw Materials	14.1
Wastewater Treatment	$1,600,000	Waste Disposal	2.9
Storage	$2,800,000	Electricity	-36.7
Boiler/Turbogenerator	$62,000,000	Fixed Costs	21.6
Utilities	$8,400,000	Capital Depreciation	36.4
Total Installed Equipment Cost	**$167,400,000**	Average Income Tax	32.0
		Average Return on Investment	73.4
Added Costs[a]	$218,600,000	**Operating Costs ($/yr)**	
(% of TCI)	57%	Feedstock	$57,900,000
Working Capital	50,350,000	Biomass to Boiler	$0
Total Capital Investment	**$386,000,000**	CSL	$8,900,000
		Cellulase	$37,100,000
Installed Equipment Cost/Annual Gallon	$3.63	Other Raw Matl. Costs	$6,500,000
Total Project Investment/Annual Gallon	$8.36	Waste Disposal	$1,400,000
		Electricity	-$16,900,000
Loan Rate	N/A	Fixed Costs	$10,000,000
Term (years)	N/A	Capital Depreciation	$16,800,000
Capital Charge Factor	0.170	Average Income Tax	$14,800,000
		Average Return on Investment	$33,900,000
Denatured Fuel Prod. (MMgal / yr)	48.3	Excess Electricity (KWH/gal)	5.23
Denatured Fuel Min. Sales Price	$3.56	Plant Electricity Use (KWH/gal)	2.57
Denaturant Cost ($/gal denaturant)	$0.739	Plant Steam Use (kg steam/gal)	35.9
		Boiler Feed -- LHV (Btu/lb)	2,295
Maximum Yields (100% of Theoretical)		Boiler Feed -- Water Fraction	0.538
Ethanol Production (MM Gal/yr)	82.5	**Specific Operating Conditions**	
Theoretical Yield (Gal/ton)	106.9	Feed rate (dry tonnes/day)	2,000
Current Yield (Actual/Theoretical)	56%	Feed rate (dry tons/day)	2,205
		Lignin Residue (dry tonnes/day)	730
		Lignin Residue (dry tons/day)	804
		Pretreatment Total Solids (wt%)	48
		Saccharification Total Solids (wt%)	20

[a] Added costs include indirect costs (engineering and supervision, construction expenses, legal and contractor fees) and contingency.

Table D-6. Cost Analysis Result Summary for On-site Enzyme Production Processes

Ethanol Production Process Engineering Analysis			
UCR Dilute Acid - Corn Stover, Current Case Dilute Acid Prehydrolysis with Saccharification and Cofermentation (On-site Enzyme Production) All Values in 2007$			
Product Value	**$3.54**		
Ethanol Production (MM Gal. / Year)	47.4	Ethanol at 68°F	
Ethanol Yield (Gal / Dry Metric Ton Feedstock)	67.7		
Feedstock Cost $/Dry Metric Ton	$83		
Internal Rate of Return (After-Tax)	10%		
Equity Percent of Total Investment	100%		
Capital Costs		**Operating Costs (cents/gal ethanol)**	
Feed Handling	$10,900,000	Feedstock	122.1
Pretreatment	$36,200,000	Biomass to Boiler	0.0
On-site Enzyme Production	$23,700,000	CSL	19.8
Saccharification & Fermentation	$21,800,000	Cellulase	0.0
Distillation and Solids Recovery	$25,400,000	Other Raw Materials	20.4
Wastewater Treatment	$3,500,000	Waste Disposal	14.3
Storage	$2,500,000	Electricity	1.6
Boiler/Turbogenerator	$57,200,000	Fixed Costs	22.7
Utilities	$6,800,000	Capital Depreciation	39.9
Total Installed Equipment Cost	**$187,800,000**	Average Income Tax	34.0
		Average Return on Investment	79.2
Added Costs[a]	$246,400,000	**Operating Costs ($/yr)**	
(% of TCI)	57%	Feedstock	$57,900,000
Working Capital	56,630,000	Biomass to Boiler	$0
Total Capital Investment	**$434,200,000**	CSL	$9,400,000
		Cellulase	$0
Installed Equipment Cost/Annual Gallon	$3.96	Other Raw Matl. Costs	$9,700,000
Total Project Investment/Annual Gallon	$9.16	Waste Disposal	$6,800,000
		Electricity	$800,000
Loan Rate	N/A	Fixed Costs	$10,700,000
Term (years)	N/A	Capital Depreciation	$18,900,000
Capital Charge Factor	0.167	Average Income Tax	$16,100,000
		Average Return on Investment	$37,500,000
Denatured Fuel Prod. (MMgal / yr)	49.6		
Denatured Fuel Min. Sales Price	$3.42	Excess Electricity (KWH/gal)	-0.30
Denaturant Cost ($/gal denaturant)	$0.739	Plant Electricity Use (KWH/gal)	7.63
		Plant Steam Use (kg steam/gal)	19.7
		Boiler Feed -- LHV (Btu/lb)	2,239
Maximum Yields (100% of Theoretical)		Boiler Feed -- Water Fraction	0.538
Ethanol Production (MM Gal/yr)		**Specific Operating Conditions**	
Theoretical Yield (Gal/ton)	82.5	Enzyme Loading (mg/g cellulose)	0.0
Current Yield (Actual/Theoretical)	106.9	Saccharification Time (days)	5.0
	57%	Conversion Cellulose --> Glucose	0.9109
		Fermentation Time (days)	2.0

[a] Added costs include indirect costs (engineering and supervision, construction expenses, legal and contractor fees) and contingency.

Table D-7. Cost Analysis Result Summary for Pervaporation Purification Processes

Ethanol Production Process Engineering Analysis		
Corn Stover Design Report Case: 2020 Market Target Case (2002 Design Report Target Case) Dilute Acid Prehydrolysis with Saccharification and Cofermentation (Pervaporation Separation) All Values in 2007$		
Product Value	**$3.75**	
Ethanol Production (MM Gal. / Year)	53.9	Ethanol at 68°F
Ethanol Yield (Gal / Dry Metric Ton Feedstock)	77.0	
Feedstock Cost $/Dry MetricTon	$83	
Internal Rate of Return (After-Tax)	10%	
Equity Percent of Total Investment	100%	

Capital Costs		**Operating Costs (cents/gal ethanol)**	
Feed Handling	$10,900,000	Feedstock	107.5
Pretreatment	$36,200,000	Biomass to Boiler	0.0
		CSL	15.8
Saccharification & Fermentation	$21,900,000	Cellulase	68.9
Distillation and Solids Recovery	$70,800,000	Other Raw Materials	18.4
Wastewater Treatment	$3,300,000	Waste Disposal	12.6
Storage	$3,300,000	Electricity	-25.3
Boiler/Turbogenerator	$55,900,000	Fixed Costs	21.4
Utilities	$7,000,000	Capital Depreciation	40.5
Total Installed Equipment Cost	**$209,200,000**	Average Income Tax	34.1
		Average Return on Investment	81.0
Added Costs[a]	$226,500,000	**Operating Costs ($/yr)**	
(% of TCI)	52%	Feedstock	$57,900,000
Working Capital	$65,350,000	Biomass to Boiler	$0
Total Capital Investment	**$501,000,000**	CSL	$8,500,000
		Cellulase	$37,100,000
Installed Equipment Cost/Annual Gallon	$3.88	Other Raw Matl. Costs	$9,900,000
Total Project Investment/Annual Gallon	$8.09	Waste Disposal	$6,800,000
		Electricity	-$13,600,000
Loan Rate	N/A	Fixed Costs	$11,500,000
Term (years)	N/A	Capital Depreciation	$21,800,000
Capital Charge Factor	0.192	Average Income Tax	$18,400,000
		Average Return on Investment	$43,600,000
Denatured Fuel Prod. (MMgal / yr)	56.4	Excess Electricity (KWH/gal)	4.68
Denatured Fuel Min. Sales Price	$3.61	Plant Electricity Use (KWH/gal)	2.25
Denaturant Cost ($/gal denaturant)	$0.739	Plant Steam Use (kg steam/gal)	11.9
Maximum Yields (100% of Theoretical)		Boiler Feed -- LHV (Btu/lb)	2,193
Ethanol Production (MM Gal/yr)	82.5	Boiler Feed -- Water Fraction	0.543
Theoretical Yield (Gal/ton)	106.9	**Specific Operating Conditions**	
Current Yield (Actual/Theoretical)	65%	Feed rate (dry tonnes/day)	2,000
		Feed rate (dry tons/day)	2,205
		Lignin Residue (dry tonnes/day)	626
		Lignin Residue (dry tons/day)	690
		Pretreatment Total Solids (wt%)	30
		Saccharification Total Solids (wt%)	20

[a] Added costs include indirect costs (engineering and supervision, construction expenses, legal and contractor fees) and contingency.

Table D-8. Cost Analysis Result Summary for Separate C5 and C6 Fermentation Processes

Ethanol Production Process Engineering Analysis		
Corn Stover Design Case: 2005 Post Enzyme-Subcontract Case - 90% Cellulose in 7 days @ 1.9X loading Dilute Acid Prehydrolysis with Saccharification and Cofermentation (Separate C5 & C6 Sugar fermentation) All Values in 2007$		
Product Value	**$3.67**	
Ethanol Production (MM Gal. / Year)	55.5	Ethanol at 68°F
Ethanol Yield (Gal / Dry Metric Ton Feedstock)	79.3	
Feedstock Cost $/Dry MetricTon	$83	
Internal Rate of Return (After-Tax)	10%	
Equity Percent of Total Investment	100%	

Capital Costs		**Operating Costs (cents/gal ethanol)**	
Feed Handling	$11,400,000	Feedstock	104.3
Pretreatment	$34,200,000	Biomass to Boiler	0.0
		CSL	33.1
Saccharification & Fermentation	$29,600,000	Cellulase	66.9
Distillation and Solids Recovery	$27,300,000	Other Raw Materials	35.9
Wastewater Treatment	$3,100,000	Waste Disposal	2.6
Storage	$5,100,000	Electricity	-11.6
Boiler/Turbogenerator	$50,900,000	Fixed Costs	18.0
Utilities	$6,200,000	Capital Depreciation	30.3
Total Installed Equipment Cost	**$167,800,000**	Average Income Tax	26.0
		Average Return on Investment	61.6
Added Costs[a]	$218,000,000	**Operating Costs ($/yr)**	
(% of TCI)	57%	Feedstock	$57,900,000
Working Capital	$50,320,000	Biomass to Boiler	$0
Total Capital Investment	**$385,800,000**	CSL	$18,400,000
		Cellulase	$37,100,000
Installed Equipment Cost/Annual Gallon	$3.02	Other Raw Matl. Costs	$19,900,000
Total Project Investment/Annual Gallon	$6.95	Waste Disposal	$1,400,000
		Electricity	-$6,500,000
Loan Rate	N/A	Fixed Costs	$10,000,000
Term (years)	N/A	Capital Depreciation	$16,800,000
Capital Charge Factor	0.170	Average Income Tax	$14,400,000
		Average Return on Investment	$34,200,000
Denatured Fuel Prod. (MMgal / yr)	58.1		
Denatured Fuel Min. Sales Price	$3.54	Excess Electricity (KWH/gal)	2.16
Denaturant Cost ($/gal denaturant)	$0.739	Plant Electricity Use (KWH/gal)	2.33
		Plant Steam Use (kg steam/gal)	19.9
		Boiler Feed -- LHV (Btu/lb)	1,489
Maximum Yields (100% of Theoretical)		Boiler Feed -- Water Fraction	0.547
Ethanol Production (MM Gal/yr)	82.5	**Specific Operating Conditions**	
Theoretical Yield (Gal/ton)	106.9	Enzyme Loading (mg/g cellulose)	33.4
Current Yield (Actual/Theoretical)	67%	Saccharification Time (days)	5.0
		Conversion Cellulose --> Glucose	0.9109
		Fermentation Time (days)	2.0

[a] Added costs include indirect costs (engineering and supervision, construction expenses, legal and contractor fees) and contingency.

Appendix E - Equipment List and Costs, Installation Factors, and Installed Equipment Costs for Dilute Acid Pretreatment Processes

Table E-1. Equipment Lists and Costs for Dilute Acid Pretreatment Processes

Equip-ment ID	Number Required	Spares Nos	Equipment Name	Scaling Stream Flow (Kg/hr)	New Stream Flow	Size Ratio	Original Equip Cost (per unit)	Base Year	Total Original Equip Cost (Req'd & Spare) in Base Year	Scaling Expo-nent	Scaled Cost in Base Year	Installa-tion Factor	Installed Cost in Base Year	Installed Cost in 2007$
C-101	2		Bale Transport Conveyor	98,040	111,111	1.13	$400,000	2000	$800,000	0.6	$862,388	1.62	$1,397,068	$1,862,521
C-102	2		Bale Unwrapping Conveyor	98,040	111,111	1.13	$150,000	2000	$300,000	0.6	$323,395	1.19	$384,840	$513,056
C-103	1		Belt Press Discharge Conveyor	98,040	111,111	1.13	$50,000	2000	$50,000	0.6	$53,899	1.89	$101,870	$135,809
C-104	4		Shredder Feed Conveyor	98,040	111,111	1.13	$60,000	2000	$240,000	0.6	$258,716	1.38	$357,028	$475,978
M-101	2		Truck Scales	98,040	111,111	1.13	$34,000	2000	$68,000	0.6	$73,303	2.47	$181,058	$241,380
M-102	4	1	Truck Unloading Forklift	98,040	111,111	1.13	$18,000	2000	$90,000	1	$101,999	1	$101,999	$135,982
M-103	4		Bale Moving Forklift	98,040	111,111	1.13	$18,000	2000	$72,000	1	$81,599	1	$81,599	$108,785
M-104	2		Corn Stover Wash Table	98,040	111,111	1.13	$104,000	2000	$208,000	0.6	$224,221	2.39	$535,888	$714,426
M-105	4		Shredder	98,040	111,111	1.13	$302,000	2000	$1,208,000	0.6	$1,302,205	1.38	$1,797,043	$2,395,754
M-106	1		Concrete Feedstock-Storage Slab	98,040	111,111	1.13	$450,655	2000	$450,655	1	$510,738	2.2	$1,123,624	$1,497,976
M-107	1		Polymer Feed System	98,040	111,111	1.13	$30,000	2000	$30,000	0.6	$32,340	2.28	$73,734	$98,300
P-101	2	1	Wash Table Pump	98,040	111,111	1.13	$20,000	2000	$60,000	0.79	$66,236	3.87	$256,332	$341,732
P-102	2	1	Wash Water Pump	98,040	111,111	1.13	$15,000	2000	$45,000	0.79	$49,677	5.19	$257,822	$343,719
P-103	1	1	Clarifier Underflow Pump	98,040	111,111	1.13	$6,000	2000	$12,000	0.79	$13,247	13.41	$177,644	$236,828
P-104	1	1	Clarified Water Pump	98,040	111,111	1.13	$15,000	2000	$30,000	0.79	$33,118	7.07	$234,143	$312,151
P-105	1	1	Belt Press Sump Pump	98,040	111,111	1.13	$19,000	2000	$38,000	0.79	$41,949	2.92	$122,492	$163,301
S-101	1		Clarifier Thickener	98,040	111,111	1.13	$135,000	2000	$135,000	0.6	$145,528	1.51	$219,747	$292,959

Equip-ment ID	Number Required	Spares Nos	Equipment Name	Scaling Stream Flow (Kg/hr)	New Stream Flow	Size Ratio	Original Equip Cost (per unit)	Base Year	Total Original Equip Cost (Req'd & Spare) in Base Year	Scaling Expo-nent	Scaled Cost in Base Year	Installa-tion Factor	Installed Cost in Base Year	Installed Cost in 2007$
S-102	1		Belt Press	98,040	111,111	1.13	$100,000	2000	$100,000	0.6	$107,798	1.25	$134,748	$179,641
S-103	1		Magnetic Separator	159,948	111,111	0.69	$13,863	1998	$13,863	0.6	$11,141	1.3	$14,483	$19,537
T-101	1		Wash Water Tank	98,040	111,111	1.13	$50,000	2000	$50,000	0.51	$53,296	2.8	$149,227	$198,945
T-102	1		Clarifier Thickener Tank	98,040	111,111	1.13	$135,000	2000	$135,000	0.51	$143,898	3.04	$437,450	$583,192
A100							**Subtotal**		**$4,135,518**		**$4,490,691**	**1.81**	**$8,139,839**	**$10,851,970**
A-201	1		In-line Sulfuric Acid Mixer	55,308	102,884	1.86	$1,900	1997	$1,900	0.48	$2,559	1	$2,559	$3,479
A-205	1		Hydrolyzate Mix Tank Agitator	358,810	236,880	0.66	$36,000	1997	$36,000	0.51	$29,129	1.2	$34,955	$47,517
A-209	1		Overliming Tank Agitator	167,050	301,831	1.81	$19,800	1997	$19,800	0.51	$26,773	1.3	$34,805	$47,313
A-224	1		Reacidification Tank Agitator	167,280	306,440	1.83	$65,200	1997	$65,200	0.51	$88,782	1.2	$106,539	$144,827
A-232	1		Reslurrying Tank Agitator	358,810	416,118	1.16	$36,000	1997	$36,000	0.51	$38,826	1.2	$46,591	$63,335
C-201	1		Hydrolyzate Screw Conveyor	225,140	236,880	1.05	$59,400	1997	$59,400	0.78	$61,802	1.3	$80,343	$109,217
C-202	1		Hydrolyzate Washed Solids Belt Conveyor	91,633	77,862	0.85	$80,000	2000	$80,000	0.76	$70,686	1.45	$102,495	$136,643
C-225	1		Lime Solids Feeder				$3,900	1997	$3,900		$3,900	1.3	$5,070	$6,892
H-200	1		Hydrolyzate Cooler	1,988	2,288	1.15	$45,000	1997	$45,000	0.51	$48,340	2.1	$101,515	$137,997
H-201	2	1	Beer Column Feed Economizer	12,532	12,217	0.97	$132,800	1997	$398,400	0.68	$391,574	2.1	$822,304	$1,117,823
H-205	1		Pneumapress Vent Condensor	120	176	1.46	$15,385	2000	$15,385	0.68	$19,940	2.1	$41,874	$55,825
H-244	2	1	Waste Vapor Condensor	12,532	1,710	0.14	$132,800	1997	$398,400	0.68	$102,824	2.1	$215,931	$293,532
M-202	3		Prehydrolysis/Screw Feeder/Reactor	271,313	281,602	1.04	$2,454,982	2000	$7,364,947	0.6	$7,531,285	2.29	$17,246,643	$22,992,607
P-201	1	1	Sulfuric Acid	1,647	8,300	5.04	$4,800	1997	$9,600	0.79	$34,446	2.8	$96,450	$131,112

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			Pump											
P-205	2	1	Pneumapress Feed Pump	50,299	42,588	0.85	$15,416	2000	$46,248	0.79	$40,551	3.34	$135,440	$180,563
P-209	1	1	Overlimed Hydrolyzate Pump	167,050	301,831	1.81	$10,700	1997	$21,400	0.79	$34,149	2.8	$95,617	$129,980
P-211	1	1	Primary Filtrate Pump	136,350	159,432	1.17	$32,549	2000	$65,098	0.79	$73,659	3.56	$262,225	$349,590
P-213	1	1	Wash Filtrate Pump	131,530	136,282	1.04	$49,803	2000	$99,606	0.79	$102,438	2.71	$277,607	$370,096
P-222	1	1	Filtered Hydrolyzate Pump	162,090	292,599	1.81	$10,800	1997	$21,600	0.79	$34,443	2.8	$96,441	$131,099
P-223	1		Lime Unloading Blower	547	6,117	11.18	$47,600	1998	$47,600	0.5	$159,176	1.4	$222,847	$300,600
P-224	2	1	Saccharification Feed Pump	358,810	416,118	1.16	$61,368	1998	$184,104	0.7	$204,225	2.8	$571,831	$771,348
P-239	1	1	Reacidified Liquor Pump	167,280	306,440	1.83	$10,800	1997	$21,600	0.79	$34,845	2.8	$97,567	$132,631
S-205	3		Pneumapress Filter	50,299	42,588	0.85	$1,575,000	2000	$4,725,000	0.6	$4,276,015	1.05	$4,489,816	$5,985,662
S-222	1		Hydroclone & Rotary Drum Filter	5,195	13,841	2.66	$165,000	1998	$165,000	0.39	$241,801	1.4	$338,521	$456,634
S-227	1		LimeDust Vent Baghouse	548	6,117	11.16	$32,200	1997	$32,200	1	$359,422	1.5	$539,133	$732,886
T-201	1		Sulfuric Acid Tank	1,647	8,300	5.04	$5,760	1996	$5,760	0.71	$18,160	1.4	$25,423	$34,995
T-203	1		Blowdown Tank	270,300	281,602	1.04	$64,100	1997	$64,100	0.93	$66,589	1.2	$79,907	$108,624
T-205	1		Hydrolyzate Mixing Tank	358,810	236,880	0.66	$44,800	1997	$44,800	0.71	$33,361	1.2	$40,033	$54,420
T-209	1		Overliming Tank	167,050	301,831	1.81	$71,000	1997	$71,000	0.71	$108,061	1.4	$151,286	$205,654
T-211	1		Primary Filtrate Tank	136,350	159,432	1.17	$36,000	2000	$36,000	0.71	$40,228	2.45	$98,558	$131,394
T-213	1		Wash Filtrate Tank	131,530	136,282	1.04	$18,000	2000	$18,000	0.71	$18,459	3.68	$67,930	$90,562
T-220	1		Lime Storage Bin	548	6,117	11.16	$69,200	1997	$69,200	0.46	$209,928	1.3	$272,906	$370,983
T-224	1		Reacidification Tank	167,280	306,440	1.83	$147,800	1997	$147,800	0.51	$201,258	1.2	$241,510	$328,304
T-232	1		Slurrying Tank	358,810	416,118	1.16	$44,800	1997	$44,800	0.71	$49,770	1.2	$59,724	$81,188

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A200							**Subtotal**		**$14,464,848**		**$14,757,407**	**1.84**	**$27,102,397**	**$36,235,330**
A-300	12		Ethanol Fermentor Agitator				$19,676	1996	$236,112		$236,112	1.2	$283,334	$390,002
A-301	1		Seed Hold Tank Agitator	41,777	40,806	0.98	$12,551	1996	$12,551	0.51	$12,401	1.2	$14,882	$20,484
A-304	2		4th Seed Vessel Agitator	41,777	40,806	0.98	$11,700	1997	$23,400	0.51	$23,121	1.2	$27,745	$37,716
A-305	2		5th Seed Vessel Agitator	41,777	40,806	0.98	$10,340	1996	$20,680	0.51	$20,433	1.2	$24,520	$33,751
A-306	2		Beer Surge Tank Agitator	381,700	429,208	1.12	$48,700	1998	$97,400	0.51	$103,405	1.2	$124,086	$167,381
A-310	30		Saccharification Tank Agitator				$19,676	1996	$590,280		$590,280	1.2	$708,336	$975,006
F-300	6		Ethanol Fermentor				$493,391	1998	$2,960,346		$2,960,346	1.2	$3,552,415	$4,791,884
F-301	2		1st Seed Fermentor				$14,700	1997	$29,400		$29,400	2.8	$82,320	$111,904
F-302	2		2nd Seed Fermentor				$32,600	1997	$65,200		$65,200	2.8	$182,560	$248,168
F-303	2		3rd Seed Fermentor				$81,100	1997	$162,200		$162,200	2.8	$454,160	$617,376
F-304	2		4th Seed Fermentor	41,777	40,806	0.98	$39,500	1997	$79,000	0.93	$77,291	1.2	$92,749	$126,081
F-305	2		5th Seed Fermentor	41,777	40,806	0.98	$147,245	1998	$294,490	0.51	$290,978	1.2	$349,174	$471,004
H-300	6	1	Fermentation Cooler	67,820	129,602	1.91	$4,000	1997	$28,000	0.78	$46,402	2.1	$97,444	$132,463
H-301	1	1	Hydrolyzate Heater	256	376	1.47	$22,400	2001	$44,800	0.68	$58,229	2.1	$122,282	$162,939
H-302	3		Saccharified Slurry Cooler	3,765	0	0.00	$25,409	1998	$76,227	0.78	$0	2.1	$0	$0
H-304	1		4th Seed Fermentor Coil	0.1380	0.2846	2.06	$3,300	1997	$3,300	0.83	$6,017	1.2	$7,220	$9,815
H-305	1		5th Seed Fermentor Coil	0.1380	0.2846	2.06	$18,800	1997	$18,800	0.98	$38,208	1.2	$45,849	$62,326
H-310	15	1	Saccharification Cooler	67,820	1,696	0.03	$4,000	1997	$64,000	0.78	$3,603	2.1	$7,566	$10,286

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P-300	6	1	Fermentation Recirc/Transfer Pump	67,737	129,602	1.91	$8,000	1997	$56,000	0.79	$93,497	2.8	$261,792	$355,875
P-301	1	1	Seed Hold Transfer Pump	41,777	40,806	0.98	$22,194	1998	$44,388	0.7	$43,663	1.4	$61,128	$82,457
P-302	2		Seed Transfer Pump	41,777	40,806	0.98	$54,088	1998	$108,176	0.7	$106,409	1.4	$148,973	$200,951
P-306	1	1	Beer Transfer Pump	381,701	429,208	1.12	$17,300	1997	$34,600	0.79	$37,960	2.8	$106,287	$144,484
P-310	15	1	Saccharification Recirc/Transfer Pump	67,737	1,696	0.03	$8,000	1997	$128,000	0.79	$6,952	2.8	$19,465	$26,461
T-301	1		Seed Hold Tank	41,777	40,806	0.98	$161,593	1998	$161,593	0.51	$159,666	1.2	$191,599	$258,450
T-306	1		Beer Storage Tank	381,700	429,208	1.12	$237,700	1998	$237,700	0.71	$258,345	1.2	$310,014	$418,181
T-310	15		Saccharification Tank				$493,391	1998	$7,400,865		$7,400,865	1.2	$8,881,038	$11,979,711
A300							**Subtotal**		**$12,977,508**		**$12,830,984**	**1.3**	**$16,156,940**	**$21,835,156**
A400							**Subtotal**		**$0**		**$0**		**$0**	**$0**
A-530	1		Recycled Water Tank Agitator	179,446	263,502	1.47	$5,963	1998	$5,963	0.51	$7,254	1.3	$9,430	$12,720
C-501	1		Lignin Wet Cake Screw	99,199	47,373	0.48	$31,700	1997	$31,700	0.78	$17,811	1.4	$24,936	$33,897
D-501	1		Beer Column	17	20	1.17	$478,100	1998	$478,100	0.68	$531,355	2.1	$1,115,846	$1,505,174
D-502	1		Rectification Column	56,477	62,095	1.10	$525,800	1996	$525,800	0.68	$560,821	2.1	$1,177,724	$1,621,105
E-501	2		1st Effect Evaporation	22,278	32,408	1.45	$544,595	1996	$1,089,190	0.68	$1,405,364	2.1	$2,951,265	$4,062,339
E-502	1		2nd Effect Evaporation	22,278	32,408	1.45	$435,650	1996	$435,650	0.68	$562,112	2.1	$1,180,436	$1,624,839
E-503	2		3rd Effect Evaporation	22,278	32,408	1.45	$435,650	1996	$871,300	0.68	$1,124,224	2.1	$2,360,871	$3,249,677
H-501	1	1	Beer Column Reboiler	-28.3092	-32	1.14	$158,374	1996	$316,748	0.68	$346,704	2.1	$728,079	$1,002,181

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H-502	1		Rectification Column Reboiler	-3.5547	-4	0.99	$29,600	1997	$29,600	0.68	$29,382	2.1	$61,702	$83,876
H-504	1		Beer Column Condenser	1.0002	0	0.45	$29,544	1996	$29,544	0.68	$17,260	2.1	$36,246	$49,892
H-505	1		Start-up Rect. Column Condenser	17.6595	19	1.06	$86,174	1996	$86,174	0.68	$89,788	2.1	$188,555	$259,542
H-512	1	1	Beer Column Feed Interchanger	909	914	1.01	$19,040	1996	$38,080	0.68	$38,214	2.1	$80,250	$110,462
H-517	1	1	Evaporator Condenser	24	33	1.35	$121,576	1996	$243,152	0.68	$297,600	2.1	$624,959	$860,240
M-503	1		Molecular Sieve (9 pieces)	20,491	19,052	0.93	$2,700,000	1998	$2,700,000	0.7	$2,565,867	1	$2,565,867	$3,461,120
P-501	1	1	Beer Column Bottoms Pump	5,053	6,003	1.19	$42,300	1997	$84,600	0.79	$96,938	2.8	$271,425	$368,970
P-503	1	1	Beer Column Reflux Pump	1.0002	0.4537	0.45	$1,357	1998	$2,714	0.79	$1,454	2.8	$4,070	$5,490
P-504	1	1	Rectification Column Bottoms Pump	31,507	36,584	1.16	$4,916	1998	$9,832	0.79	$11,064	2.8	$30,978	$41,787
P-505	1	1	Rectification Column Reflux Pump	18	19	1.06	$4,782	1998	$9,564	0.79	$10,030	2.8	$28,084	$37,883
P-511	2	1	1st Effect Pump	278,645	311,140	1.12	$19,700	1997	$59,100	0.79	$64,481	2.8	$180,547	$245,432
P-512	1	1	2nd Effect Pump	91,111	137,819	1.51	$13,900	1997	$27,800	0.79	$38,551	2.8	$107,944	$146,736
P-513	2	1	3rd Effect Pump	48,001	78,409	1.63	$8,000	1997	$24,000	0.79	$35,365	2.8	$99,022	$134,608
P-514	1	1	Evaporator Condensate Pump	140,220	181,051	1.29	$12,300	1997	$24,600	0.79	$30,104	2.8	$84,290	$114,582
P-515	1		Scrubber Bottoms Pump	15,377	18,179	1.18	$2,793	1998	$2,793	0.79	$3,188	2.8	$8,926	$12,040
P-530	1	1	Recycled Water Pump	179,446	263,502	1.47	$10,600	1997	$21,200	0.79	$28,718	2.8	$80,409	$109,306
S-505	4		Pneumapress Filter	26,601	18,737	0.70	$1,418,750	2000	$5,675,000	0.6	$4,598,800	1.04	$4,782,752	$6,376,193
T-503	1		Beer Column Relfux Drum	1.0002	0	0.45	$11,900	1997	$11,900	0.93	$5,706	2.1	$11,982	$16,288
T-505	1		Rectification Column Reflux	17.6627	19	1.06	$45,600	1997	$45,600	0.72	$47,621	2.1	$100,005	$135,944

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			Drum											
T-512	1		Vent Scrubber	18,523	18,916	1.02	$99,000	1998	$99,000	0.78	$100,635	2.1	$211,335	$285,071
T-514	1		Evaporator Condensate Drum	164,760	181,051	1.10	$37,200	1998	$37,200	0.93	$40,609	2.1	$85,279	$115,034
T-530	1		Recycled Water Tank	179,446	263,502	1.47	$14,515	1998	$14,515	0.745	$19,325	1.4	$27,055	$36,495
A500							**Subtotal**		**$13,030,419**		**$12,726,345**	**1.51**	**$19,220,269**	**$26,118,926**
A-602	1		Equalization Basin Agitator	188,129	85,599	0.46	$28,400	1997	$28,400	0.51	$19,007	1.2	$22,808	$31,005
A-606	1		Anaerobic Agitator	810,250	401,778	0.50	$30,300	1997	$30,300	0.51	$21,188	1.2	$25,425	$34,562
A-608	16		Aerobic Lagoon Agitator	812	108	0.13	$31,250	1998	$500,000	0.51	$178,402	1.4	$249,763	$336,908
C-614	1		Aerobic Sludge Screw	978	112	0.11	$5,700	1997	$5,700	0.78	$1,051	1.4	$1,471	$2,000
H-602	1		Anaerobic Digestor Feed Cooler	7,627	3,155	0.41	$128,600	1997	$128,600	0.74	$66,913	2.1	$140,518	$191,017
M-604	1		Nutrient Feed System				$31,400	1998	$31,400		$31,400	2.58	$81,012	$109,278
M-606	1		Biogas Emergency Flare	2,572	292	0.11	$20,739	1998	$20,739	0.6	$5,620	1.68	$9,442	$12,737
M-612	1		Filter Precoat System				$3,000	1998	$3,000		$3,000	1.4	$4,200	$5,665
P-602	1	1	Anaerobic Reactor Feed Pump	188,129	85,599	0.46	$11,400	1997	$22,800	0.79	$12,240	2.8	$34,271	$46,587
P-606	1	1	Aerobic Digestor Feed Pump	185,782	85,335	0.46	$10,700	1997	$21,400	0.79	$11,574	2.8	$32,408	$44,055
P-608	1		Aerobic Sludge Recycle Pump	5,862	679	0.12	$11,100	1997	$11,100	0.79	$2,022	1.4	$2,831	$3,848
P-610	1		Aerobic Sludge Pump	5,862	679	0.12	$11,100	1997	$11,100	0.79	$2,022	1.4	$2,831	$3,848
P-611	1	1	Aerobic Digestion Outlet Pump	187,827	85,560	0.46	$10,700	1997	$21,400	0.79	$11,498	2.8	$32,196	$43,766
P-614	1	1	Sludge Filtrate Recycle Pump	4,885	567	0.12	$6,100	1997	$12,200	0.79	$2,226	2.8	$6,234	$8,474

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P-616	1	1	Treated Water Pump	181,965	84,881	0.47	$10,600	1997	$21,200	0.79	$11,607	2.8	$32,498	$44,178
S-600	1		Bar Screen	188,129	85,599	0.46	$117,818	1991	$117,818	0.3	$93,028	1.2	$111,634	$162,338
S-614	1		Belt Filter Press	438	58	0.13	$650,223	1998	$650,223	0.72	$151,646	1.8	$272,963	$368,203
T-602	1		Equalization Basin	188,129	85,599	0.46	$350,800	1998	$350,800	0.51	$234,772	1.42	$333,376	$449,694
T-606	1		Anaerobic Digestor	810,250	401,778	0.50	$881,081	1998	$881,081	0.51	$616,103	1.04	$640,747	$864,309
T-608	1		Aerobic Digestor	19,506,756	8,960,675	0.46	$635,173	1998	$635,173	1	$291,775	1	$291,775	$393,578
T-610	1		Clarifier	185,782	85,335	0.46	$174,385	1998	$174,385	0.51	$117,272	1.96	$229,852	$310,050
A600							**Subtotal**		**$3,678,819**		**$1,884,366**	**1.36**	**$2,558,255**	**$3,466,097**
A-701	1		Denaturant In-line Mixer	19,436	19,838	1.02	$1,900	1997	$1,900	0.48	$1,919	1	$1,919	$2,608
A-720	1		CSL Storage Tank Agitator	41,777	4,488	0.11	$12,551	1996	$12,551	0.51	$4,023	1.2	$4,828	$6,645
A-760	1		CSL/DAP Day Tank Agitator	1,400	4,488	3.21	$12,795	2001	$12,795	0.51	$23,177	1.2	$27,813	$37,060
C-755	1		DAP Solids Feeder				$3,900	1997	$3,900		$3,900	1.3	$5,070	$6,892
P-701	2	1	Ethanol Product Pump	18,549	19,052	1.03	$7,500	1997	$22,500	0.79	$22,981	2.8	$64,347	$87,472
P-703	1	1	Sulfuric Acid Pump	1,647	8,300	5.04	$8,000	1997	$16,000	0.79	$57,411	2.8	$160,750	$218,520
P-704	1	1	Firewater Pump	6,823	7,008	1.03	$18,400	1997	$36,800	0.79	$37,585	2.8	$105,238	$143,059
P-710	1	1	Gasoline Pump	887	786	0.89	$4,500	1997	$9,000	0.79	$8,181	2.8	$22,907	$31,139
P-720	1	1	CSL Pump	2,039	4,488	2.20	$8,800	1997	$17,600	0.79	$32,824	2.8	$91,908	$124,937
P-750	1	1	Cellulase Pump	6,823	8,712	1.28	$18,400	1997	$36,800	0.79	$44,637	2.8	$124,983	$169,900
P-755	1		DAP Unloading Blower	154	30	0.19	$47,600	1998	$47,600	0.5	$20,922	1.4	$29,291	$39,511
P-760	1	1	CSL/DAP Pump	2,039	4,488	2.20	$8,800	1997	$17,600	0.79	$32,824	2.8	$91,908	$124,937
S-755	1		DAP Vent Baghouse	548	30	0.05	$32,200	1997	$32,200	1	$1,748	1.5	$2,622	$3,565

Equip-ment ID	Number Required	Spares Nos	Equipment Name	Scaling Stream Flow (Kg/hr)	New Stream Flow	Size Ratio	Original Equip Cost (per unit)	Base Year	Total Original Equip Cost (Req'd & Spare) in Base Year	Scaling Expo-nent	Scaled Cost in Base Year	Installa-tion Factor	Installed Cost in Base Year	Installed Cost in 2007$
T-701	2		Ethanol Product Storage Tank	18,549	19,052	1.03	$165,800	1997	$331,600	0.51	$336,159	1.4	$470,623	$639,755
T-703	1		Sulfuric Acid Storage Tank	1,647	8,300	5.04	$42,500	1997	$42,500	0.51	$96,961	1.2	$116,353	$158,168
T-704	1		Firewater Storage Tank	6,822	7,008	1.03	$166,100	1997	$166,100	0.51	$168,392	1.4	$235,748	$320,472
T-709	1		Propane Storage Tank	15	16	1.04	$24,834	2001	$24,834	0.72	$25,580	1.4	$35,812	$47,719
T-710	1		Gasoline Storage Tank	887	786	0.89	$43,500	1997	$43,500	0.51	$40,902	1.4	$57,262	$77,841
T-720	1		CSL Storage Tank	2,039	4,488	2.20	$88,100	1997	$88,100	0.79	$164,307	1.4	$230,030	$312,699
T-750	2		Cellulase Storage Tank	9,234	8,712	0.94	$125,900	2001	$251,800	0.79	$240,483	1.4	$336,676	$448,617
T-755	1		DAP Storage Bin	154	30	0.19	$33,384	2001	$33,384	0.44	$16,195	1.3	$21,054	$28,054
T-760	1		CSL/DAP Day Tank	1,400	4,488	3.21	$30,084	2001	$30,084	0.79	$75,512	1.4	$105,717	$140,866
A700							**Subtotal**		**$1,279,148**		**$1,456,624**	**1.6**	**$2,342,860**	**$3,170,437**
H-801	1		Burner Combustion Air Preheater	24.0337	13	0.56	$1,049,900	1997	$1,049,900	0.6	$739,305	1.5	$1,108,957	$1,507,493
H-811	1		BFW Preheater	415	226	0.54	$58,400	1997	$58,400	0.68	$38,647	2.1	$81,159	$110,325
M-803	1		Fluidized Bed Combustion Reactor	341,270	209,403	0.61	24,900,000	1998	$24,900,00 0	0.75	$17,262,891	1.3	$22,441,758	$30,271,886
M-804	1		Combustion Gas Baghouse	652,517	450,947	0.69	$2,536,300	1998	$2,536,300	0.58	$2,047,058	1.5	$3,070,587	$4,141,941
M-811	1		Turbine/Generator	281,179	209,403	0.74	10000000	1998	$10,000,00 0	0.71	$8,111,862	1.5	$12,167,793	$16,413,244
M-820	1		Hot Process Water Softener System	225,889	192,701	0.85	$1,381,300	1999	$1,381,300	0.82	$1,212,546	1.3	$1,576,310	$2,120,311
M-830	1		Hydrazine Addition Pkg.	229,386	215,880	0.94	$19,000	1994	$19,000	0.6	$18,321	1	$18,321	$26,150
M-832	1		Ammonia Addition Pkg	229,386	215,880	0.94	$19,000	1994	$19,000	0.6	$18,321	1	$18,321	$26,150

Equip-ment ID	Number Required	Spares Nos	Equipment Name	Scaling Stream Flow (Kg/hr)	New Stream Flow	Size Ratio	Original Equip Cost (per unit)	Base Year	Total Original Equip Cost (Req'd & Spare) in Base Year	Scaling Expo-nent	Scaled Cost in Base Year	Installa-tion Factor	Installed Cost in Base Year	Installed Cost in 2007$
M-834	1		Phosphate Addition Pkg.	229,386	215,880	0.94	$19,000	1994	$19,000	0.6	$18,321	1	$18,321	$26,150
P-804	2		Condensate Pump	61,471	136,035	2.21	$7,100	1997	$14,200	0.79	$26,596	2.8	$74,470	$101,233
P-811	2		Turbine Condensate Pump	39,524	59,037	1.49	$7,800	1997	$15,600	0.79	$21,419	2.8	$59,972	$81,525
P-824	2		Deaerator Feed Pump	293,605	192,701	0.66	$9,500	1997	$19,000	0.79	$13,623	2.8	$38,145	$51,853
P-826	5		BFW Pump	564,626	215,880	0.38	$52,501	1998	$262,505	0.79	$122,821	2.8	$343,900	$463,890
P-828	2		Blowdown Pump	6,613	6,476	0.98	$5,100	1997	$10,200	0.79	$10,033	2.8	$28,093	$38,189
P-830	1		Hydrazine Transfer Pump	229,386	215,880	0.94	$5,500	1997	$5,500	0.79	$5,243	2.8	$14,679	$19,955
T-804	1		Condensate Collection Tank	228,862	136,035	0.59	$7,100	1997	$7,100	0.71	$4,907	1.4	$6,870	$9,339
T-824	1		Condensate Surge Drum	222,360	192,701	0.87	$49,600	1997	$49,600	0.72	$44,742	1.7	$76,062	$103,397
T-826	1		Deaerator	266,213	215,880	0.81	$165,000	1998	$165,000	0.72	$141,890	2.8	$397,292	$535,911
T-828	1		Blowdown Flash Drum	6,563	6,476	0.99	$9,200	1997	$9,200	0.72	$9,112	2.8	$25,515	$34,684
T-830	1		Hydrazine Drum	229,386	215,880	0.94	$12,400	1997	$12,400	0.93	$11,720	1.7	$19,923	$27,083
A800							**Subtotal**		**$40,553,205**		**$29,879,378**	**1.4**	**$41,586,447**	**$56,110,709**
M-902	1		Cooling Tower System	147.960 0	105	0.71	$1,659,000	1998	$1,659,000	0.78	$1,265,253	1.2	$1,518,304	$2,048,054
M-904	2	1	Plant Air Compressor	98,040	111,111	1.13	$278,200	2000	$834,600	0.34	$870,881	1.3	$1,132,145	$1,509,336
M-910	1		CIP System	63	63	1.00	$95,000	1995	$95,000	0.6	$95,036	1.2	$114,043	$157,225
P-902	1	1	Cooling Water Pump	182900 00	12,447, 929	0.68	$332,300	1997	$664,600	0.79	$490,386	2.8	$1,373,081	$1,866,538
P-912	1	1	Make-up Water Pump	244,160	211,894	0.87	$10,800	1997	$21,600	0.79	$19,312	2.8	$54,073	$73,506
P-914	2	1	Process Water Circulating Pump	352,710	254,604	0.72	$11,100	1997	$33,300	0.79	$25,741	2.8	$72,073	$97,975
S-904	1	1	Instrument Air Dryer	159,950	111,111	0.69	$15,498	1999	$30,996	0.6	$24,910	1.3	$32,383	$43,558

Equip-ment ID	Number Required	Spares Nos	Equipment Name	Scaling Stream Flow (Kg/hr)	New Stream Flow	Size Ratio	Original Equip Cost (per unit)	Base Year	Total Original Equip Cost (Req'd & Spare) in Base Year	Scaling Expo-nent	Scaled Cost in Base Year	Installa-tion Factor	Installed Cost in Base Year	Installed Cost in 2007$
T-902	3		Prehydrolysis Filter Air Receiver	5,259	4,687	0.89	$17,000	2000	$51,000	0.72	$46,942	1.2	$56,331	$75,098
T-904	1		Plant Air Receiver	159,950	111,111	0.69	$13,000	1997	$13,000	0.72	$10,000	1.3	$13,001	$17,673
T-905	4		Product Recovery Filter Air Receiver	5,700	5,921	1.04	$17,000	2000	$68,000	0.72	$69,887	1.2	$83,865	$111,806
T-914	1		Process Water Tank	352,710	254,604	0.72	$195,500	1997	$195,500	0.51	$165,560	1.4	$231,784	$315,082
A900							**Subtotal**		**$3,666,596**		**$3,083,909**	**1.5**	**$4,681,083**	**$6,315,850**
							Equipment Cost		**$93,786,061**		**$81,109,704**	**1.50**	**$121,788,091**	**$164,104,477**

Appendix F - Process Operating Summaries

Table F-1. Operating Summary for Dilute Acid Pretreatment Processes

Pretreatment		**Saccharification**		**Fermentation**	
Acid Conc (wt%)	0.0190	Enzyme Loading (mg/g cell)	35.1	Total Solids (wt%)	20.1%
Acid Loading (g acid/g dry biomass)	0.0445	Total Solids (wt%)	20.0%	Insoluble Solids (wt%)	4.3%
Total Solids (wt%)	0.2956	Insoluble Solids (wt%)	10.0%	Temperature (°C)	32
Temperature (°C)	190	Temperature (°C)	32	Pressure (atm)	1.0
Pressure (atm)	11.5	Pressure (atm)	1.0	Residence Time (days)	2
Residence Time (min.)		Residence Time (days)	5	Conversions:	
Conversions:		Conversions:		Glucose to Ethanol	0.95
Cellulose to Glucolig	0.003	Cellulose to Glucolig	0	Glucose to Zymo	0.02
Cellulose to Cellobiose	0	Cellulose to Cellobiose	0	Glucose to Glycerol	0.004
Cellulose to Glucose	0.099	Cellulose to Glucose	0.9109	Glucose to Succinic Acid	0.006
Cellulose to HMF	0.003	Glucolig to Cellobiose	0	Glucose to Acetic Acid	0.015
Xylan to Oligomer	0.21	Glucolig to Glucose	0	Glucose to Lactic Acid	0.002
Xylan to Xylose	0.6	Cellobiose to Glucose	0	Xylose to Ethanol	0.756
Xylan to Furfural	0.11	Xylan to Oligomer	0	Xylose to Zymo	0.019
Xylan to Tar	0	Xylan to Xylose	0.5713	Xylose to Glycerol	0.003
Mannan to Oligomer	0.21	Xylose Oligomer to Xylose	0	Xylose to Xylitol	0.046
Mannan to Mannose	0.6	Xylan to Tar	0	Xylose to Succinic Acid	0.009
Mannan to HMF	0.08	Arabinan to Oligomer	0.5713	Xylose to Acetic Acid	0.014
Galactan to Oligomer	0.21	Arabinan to Arabinose	0	Xylose to Lactic Acid	0.002
Galactan to Galactose	0.6	Galactan to Oligomer	0	Arabinose to Ethanol	0
Galactan to HMF	0.08	Galactan to Galactose	0.5713	Arabinose to Zymo	0
Arabinan to Oligomer	0.21	Galactose Oligomer to Galactose	0	Arabinose to Glycerol	0
Arabinan to Arabinose	0.6	Mannan to Oligomer	0	Arabinose to Succinic Acid	0
Arabinan to Furfural	0.08	Mannan to Mannose	0.5713	Arabinose to Acetic Acid	0
Arabinan to Tar	0	Mannose Oligomer to Mannose	0	Arabinose to Lactic Acid	0
Acetate to Oligomer	0	**Sugar & Solids Flow Rates (kg/hr)**		Galactose to Ethanol	0
Acetate to Acetic Acid	1	Soluble Sugars From PT	22,989	Galactose to Zymo	0
Furfural to Tar	1	Other Soluble Solids From PT	18,543	Galactose to Glycerol	0
HMF to Tar	1	Soluble Sugars in Purchased Cellulase	0	Galactose to Succinic Acid	0
Lignin to Soluble Lignin	0.1	Other Sol Solids in Purchased Cellulase	0	Galactose to Acetic Acid	0

Pretreatment, cont.			Saccharification, cont.			Fermentation, cont.	
S/L Separation			Soluble Sugars in Produced Cellulase	0		Galactose to Lactic Acid	0
Water/Hydrolyzate Ratio (kg/kg)	0.58		Other Sol Solids in Produced Cellulase	0		Mannose to Ethanol	0
Water to S/L Separator (kg/hr)	131,829					Mannose to Zymo	0
Dilution Water (kg/hr)	52,436		Soluble Sugars From Sacc	44,457		Mannose to Glycerol	0
Conditioning			Other Soluble Solids From Sacc	16,688		Mannose to Succinic Acid	0
Ca(OH)2 to Gypsum	0.145		Soluble Sugars From Seed Train	880		Mannose to Acetic Acid	0
Sugar Conversion to TAR:			Other Soluble Solids From Seed Train	1,955		Mannose to Lactic Acid	0
Xylose	0.13		Soluble Sugars From DAP	0			
Arabinose	0.2		Other Soluble Solids From DAP	0		Contamination Loss	7.0%
Glucose	0.12		Soluble Sugars From CSL	0			
Galactose	0.28		Other Soluble Solids From CSL	2,095		Ethanol Out of Fermenters (wt%)	4.7%
Mannose	0						
Cellobiose	0.36						

Table F-2. Operating Summary for Dilute Acid Pretreatment (Pilot) Processes

Pretreatment		**Saccharification**		**Fermentation**	
Acid Conc (wt%)	0.0190	Enzyme Loading (mg/g cell)	35.1	Total Solids (wt%)	20.1%
Acid Loading (g acid/g dry biomass)	0.0445	Total Solids (wt%)	20.0%	Insoluble Solids (wt%)	4.3%
Total Solids (wt%)	0.2956	Insoluble Solids (wt%)	10.0%	Temperature (°C)	32
Temperature (°C)	190	Temperature (°C)	32	Pressure (atm)	1.0
Pressure (atm)	11.5	Pressure (atm)	1.0	Residence Time (days)	2
Residence Time (min.)		Residence Time (days)	5	Conversions:	
Conversions:		Conversions:		Glucose to Ethanol	0.95
Cellulose to Glucolig	0.003	Cellulose to Glucolig	0	Glucose to Zymo	0.02
Cellulose to Cellobiose	0	Cellulose to Cellobiose	0	Glucose to Glycerol	0.004
Cellulose to Glucose	0.099	Cellulose to Glucose	0.9109	Glucose to Succinic Acid	0.006
Cellulose to HMF	0.003	Glucolig to Cellobiose	0	Glucose to Acetic Acid	0.015
Xylan to Oligomer	0.21	Glucolig to Glucose	0	Glucose to Lactic Acid	0.002
Xylan to Xylose	0.6	Cellobiose to Glucose	0	Xylose to Ethanol	0.756
Xylan to Furfural	0.11	Xylan to Oligomer	0	Xylose to Zymo	0.019
Xylan to Tar	0	Xylan to Xylose	0.5713	Xylose to Glycerol	0.003
Mannan to Oligomer	0.21	Xylose Oligomer to Xylose	0	Xylose to Xylitol	0.046
Mannan to Mannose	0.6	Xylan to Tar	0	Xylose to Succinic Acid	0.009
Mannan to HMF	0.08	Arabinan to Oligomer	0.5713	Xylose to Acetic Acid	0.014
Galactan to Oligomer	0.21	Arabinan to Arabinose	0	Xylose to Lactic Acid	0.002
Galactan to Galactose	0.6	Galactan to Oligomer	0	Arabinose to Ethanol	0
Galactan to HMF	0.08	Galactan to Galactose	0.5713	Arabinose to Zymo	0
Arabinan to Oligomer	0.21	Galactose Oligomer to Galactose	0	Arabinose to Glycerol	0
Arabinan to Arabinose	0.6	Mannan to Oligomer	0	Arabinose to Succinic Acid	0
Arabinan to Furfural	0.08	Mannan to Mannose	0.5713	Arabinose to Acetic Acid	0
Arabinan to Tar	0	Mannose Oligomer to Mannose	0	Arabinose to Lactic Acid	0
Acetate to Oligomer	0			Galactose to Ethanol	0
Acetate to Acetic Acid	1	**Sugar & Solids Flow Rates (kg/hr)**		Galactose to Zymo	0
Furfural to Tar	1	Soluble Sugars From PT	22,989	Galactose to Glycerol	0
HMF to Tar	1	Other Soluble Solids From PT	18,543	Galactose to Succinic Acid	0
Lignin to Soluble Lignin	0.1	Soluble Sugars in Purchased Cellulase	0	Galactose to Acetic Acid	0
		Other Sol Solids in Purchased Cellulase	0	Galactose to Lactic Acid	0
		Soluble Sugars in Produced Cellulase	0	Mannose to Ethanol	0
		Other Sol Solids in Produced Cellulase	0	Mannose to Zymo	0

Pretreatment, cont.			Saccharification, cont.			Fermentation, cont.	
S/L Separation			Soluble Sugars From Sacc	44,457		Mannose to Glycerol	0
Water/Hydrolyzate Ratio (kg/kg)	0.58		Other Soluble Solids From Sacc	16,688		Mannose to Succinic Acid	0
Water to S/L Separator (kg/hr)	131,829		Soluble Sugars From Seed Train	880		Mannose to Acetic Acid	0
Dilution Water (kg/hr)	52,436		Other Soluble Solids From Seed Train	1,955		Mannose to Lactic Acid	0
Conditioning			Soluble Sugars From DAP	0			
Ca(OH)2 to Gypsum	0.145		Other Soluble Solids From DAP	0		Contamination Loss	7.0%
Sugar Conversion to TAR:			Soluble Sugars From CSL	0			
Xylose	0.13		Other Soluble Solids From CSL	2,095		Ethanol Out of Fermenters (wt%)	4.7%
Arabinose	0.2						
Glucose	0.12						
Galactose	0.28						
Mannose	0						
Cellobiose	0.36						

Table F-3. Operating Summary for Two-Stage Dilute Acid Pretreatment Processes

Pretreatment		**Saccharification**		**Fermentation**	
Acid Conc (wt%)	0.0190	Enzyme Loading (mg/g cell)	N/A	Total Solids (wt%)	19.7%
Acid Loading (g acid/g dry biomass)	0.0443	Total Solids (wt%)	34.2%	Insoluble Solids (wt%)	7.1%
Total Solids (wt%)	0.2956	Insoluble Solids (wt%)	28.6%	Temperature (°C)	32
Temperature (°C)	190	Temperature (°C)	144.0873	Pressure (atm)	1.0
Pressure (atm)	11.5	Pressure (atm)	4.0	Residence Time (days)	2
Residence Time (min.)		Residence Time (days)	0	Conversions:	
Conversions:		Conversions:		Glucose to Ethanol	0.95
Cellulose to Glucolig	0	Cellulose to Glucolig	N/A	Glucose to Zymo	0.02
Cellulose to Cellobiose	0	Cellulose to Cellobiose	N/A	Glucose to Glycerol	0.004
Cellulose to Glucose	0.0626	Cellulose to Glucose	N/A	Glucose to Succinic Acid	0.006
Cellulose to HMF	0	Glucolig to Cellobiose	N/A	Glucose to Acetic Acid	0.015
Xylan to Oligomer	0.0265	Glucolig to Glucose	N/A	Glucose to Lactic Acid	0.002
Xylan to Xylose	0.8249	Cellobiose to Glucose	N/A	Xylose to Ethanol	0.756
Xylan to Furfural	0	Xylan to Oligomer	N/A	Xylose to Zymo	0.019
Xylan to Tar	0	Xylan to Xylose	N/A	Xylose to Glycerol	0.003
Mannan to Oligomer	0.0265	Xylose Oligomer to Xylose	N/A	Xylose to Xylitol	0.046
Mannan to Mannose	0.8249	Xylan to Tar	N/A	Xylose to Succinic Acid	0.009
Mannan to HMF	0	Arabinan to Oligomer	N/A	Xylose to Acetic Acid	0.014
Galactan to Oligomer	0.0265	Arabinan to Arabinose	N/A	Xylose to Lactic Acid	0.002
Galactan to Galactose	0.8249	Galactan to Oligomer	N/A	Arabinose to Ethanol	0
Galactan to HMF	0	Galactan to Galactose	N/A	Arabinose to Zymo	0
Arabinan to Oligomer	0.0265	Galactose Oligomer to Galactose	N/A	Arabinose to Glycerol	0
Arabinan to Arabinose	0.8249	Mannan to Oligomer	N/A	Arabinose to Succinic Acid	0
Arabinan to Furfural	0	Mannan to Mannose	N/A	Arabinose to Acetic Acid	0
Arabinan to Tar	0	Mannose Oligomer to Mannose	N/A	Arabinose to Lactic Acid	0
Acetate to Oligomer	0	**Sugar & Solids Flow Rates (kg/hr)**		Galactose to Ethanol	0
Acetate to Acetic Acid	1	Soluble Sugars From PT	0	Galactose to Zymo	0
Furfural to Tar	1	Other Soluble Solids From PT	13,600	Galactose to Glycerol	0
HMF to Tar	1	Soluble Sugars in Purchased Cellulase	0	Galactose to Succinic Acid	0
Lignin to Soluble Lignin	0.1	Other Sol Solids in Purchased Cellulase	0	Galactose to Acetic Acid	0
S/L Separation		Soluble Sugars in Produced Cellulase	0	Galactose to Lactic Acid	0
Water/Hydrolyzate Ratio (kg/kg)	0.58	Other Sol Solids in Produced Cellulase	0	Mannose to Ethanol	0
Water to S/L Separator (kg/hr)	137,458			Mannose to Zymo	0

Pretreatment, cont.			**Saccharification, cont.**			**Fermentation, cont.**	
Conditioning			Soluble Sugars From Sacc	33,169		Mannose to Glycerol	0
Ca(OH)2 to Gypsum	0.183		Other Soluble Solids From Sacc	16,462		Mannose to Succinic Acid	0
Sugar Conversion to TAR:			Soluble Sugars From Seed Train	881		Mannose to Acetic Acid	0
Xylose	0.13		Other Soluble Solids From Seed Train	1,934		Mannose to Lactic Acid	0
Arabinose	0.2		Soluble Sugars From DAP	0			
Glucose	0.12		Other Soluble Solids From DAP	0		Contamination Loss	7.0%
Galactose	0.28		Soluble Sugars From CSL	0			
Mannose	0		Other Soluble Solids From CSL	2,159		Ethanol Out of Fermenters (wt%)	3.1%
Cellobiose	0.36						

Table F-4. Operating Summary for Hot Water Pretreatment Processes

Pretreatment		Saccharification		Fermentation	
Total Solids (wt%)	17.2%	Enzyme Loading (mg/g cell)	33.1	Total Solids (wt%)	16.5%
Total Insoluble Solids (wt%)	12.9%	Total Solids (wt%)	16.0%	Insoluble Solids (wt%)	3.9%
Temperature (°C)	190.0	Insoluble Solids (wt%)	9.2%	Temperature (°C)	32
Pressure (atm)	12.3	Temperature (°C)	32	Pressure (atm)	1.0
Conversions:		Pressure (atm)	1.0	Residence Time (days)	2
Cellulose to Glucolig	0.053	Residence Time (days)	5	Conversions:	
Cellulose to Cellobiose	0	Conversions:		Glucose to Ethanol	0.95
Cellulose to Glucose	0.0032	Cellulose to Glucolig	0	Glucose to Zymo	0.02
Cellulose to HMF	0	Cellulose to Cellobiose	0	Glucose to Glycerol	0.004
Xylan to Oligomer	0.554	Cellulose to Glucose	0.8997	Glucose to Succinic Acid	0.006
Xylan to Xylose	0.0239	Glucolig to Cellobiose	0	Glucose to Acetic Acid	0.015
Xylan to Furfural	0	Glucolig to Glucose	0	Glucose to Lactic Acid	0.002
Xylan to Tar	0	Cellobiose to Glucose	0	Xylose to Ethanol	0.756
Mannan to Oligomer	0.554	Xylan to Oligomer	0	Xylose to Zymo	0.019
Mannan to Mannose	0.0239	Xylan to Xylose	0.5661	Xylose to Glycerol	0.003
Mannan to HMF	0	Xylose Oligomer to Xylose	0	Xylose to Xylitol	0.046
Galactan to Oligomer	0.554	Xylan to Tar	0	Xylose to Succinic Acid	0.009
Galactan to Galactose	0.0239	Arabinan to Oligomer	0.5661	Xylose to Acetic Acid	0.014
Galactan to HMF	0	Arabinan to Arabinose	0	Xylose to Lactic Acid	0.002
Arabinan to Oligomer	0.554	Galactan to Oligomer	0	Arabinose to Ethanol	0
Arabinan to Arabinose	0.0239	Galactan to Galactose	0.5661	Arabinose to Zymo	0
Arabinan to Furfural	0	Galactose Oligomer to Galactose	0	Arabinose to Glycerol	0
Arabinan to Tar	0	Mannan to Oligomer	0	Arabinose to Succinic Acid	0
Acetate to Oligomer	0	Mannan to Mannose	0.5661	Arabinose to Acetic Acid	0
Acetate to Acetic Acid	1	Mannose Oligomer to Mannose	0	Arabinose to Lactic Acid	0
Furfural to Tar	1	Sugar & Solids Flow Rates (kg/hr)		Galactose to Ethanol	0
HMF to Tar	1	Soluble Sugars From PT	19,745	Galactose to Zymo	0
Lignin to Soluble Lignin	0.05	Other Soluble Solids From PT	18,174	Galactose to Glycerol	0
Glucose to HMF	0.5	Soluble Sugars in Purchased Cellulase	0	Galactose to Succinic Acid	0
Mannose to HMF	0.5	Other Sol Solids in Purchased Cellulase	0	Galactose to Acetic Acid	0
Galactose to HMF	0.5	Soluble Sugars in Produced Cellulase	0	Galactose to Lactic Acid	0
Xylose to Furfural	0.5	Other Sol Solids in Produced Cellulase	0	Mannose to Ethanol	0
Arabinose to Furfural	0.5			Mannose to Zymo	0

Pretreatment, cont.	
S/L Separation	
Water/Hydrolyzate Ratio (kg/kg)	#N/A
Water to S/L Separator (kg/hr)	#N/A
Dilution Water (kg/hr)	#N/A
Conditioning	
Ca(OH)2 to Gypsum	0.000
Sugar Conversion to TAR:	
Xylose	0
Arabinose	0
Glucose	0
Galactose	0
Mannose	0
Cellobiose	0

Saccharification, cont.	
Soluble Sugars From Sacc	47,147
Other Soluble Solids From Sacc	16,357
Soluble Sugars From Seed Train	2,176
Other Soluble Solids From Seed Train	1,954
Soluble Sugars From DAP	0
Other Soluble Solids From DAP	0
Soluble Sugars From CSL	0
Other Soluble Solids From CSL	2,792

Fermentation, cont.	
Mannose to Glycerol	0
Mannose to Succinic Acid	0
Mannose to Acetic Acid	0
Mannose to Lactic Acid	0
Contamination Loss	7.0%
Ethanol Out of Fermenters (wt%)	2.7%

Table F-5. Operating Summary for AFEX Pretreatment Processes

Pretreatment		**Saccharification**		**Fermentation**	
Acid Conc (wt%)	0.0000	Enzyme Loading (mg/g cell)	31.3	Total Solids (wt%)	20.5%
Acid Loading (g acid/g dry biomass)	0.0000	Total Solids (wt%)	20.0%	Insoluble Solids (wt%)	3.6%
Total Solids (wt%)	0.4808	Insoluble Solids (wt%)	11.6%	Temperature (°C)	32
Temperature (°C)	108.1302	Temperature (°C)	32	Pressure (atm)	1.0
Pressure (atm)	18.7	Pressure (atm)	1.0	Residence Time (days)	2
Residence Time (min.)		Residence Time (days)	5	Conversions:	
Conversions:		Conversions:		Glucose to Ethanol	0.95
Cellulose to Glucolig		Cellulose to Glucolig	0	Glucose to Zymo	0.02
Cellulose to Cellobiose		Cellulose to Cellobiose	0	Glucose to Glycerol	0.004
Cellulose to Glucose		Cellulose to Glucose	0.959	Glucose to Succinic Acid	0.006
Cellulose to HMF		Glucolig to Cellobiose	0	Glucose to Acetic Acid	0.015
Xylan to Oligomer	0.5	Glucolig to Glucose	0	Glucose to Lactic Acid	0.002
Xylan to Xylose		Cellobiose to Glucose	1	Xylose to Ethanol	0.756
Xylan to Furfural		Xylan to Oligomer	0	Xylose to Zymo	0.019
Xylan to Tar		Xylan to Xylose	0.777	Xylose to Glycerol	0.003
Mannan to Oligomer	0.5	Xylose Oligomer to Xylose	0	Xylose to Xylitol	0.046
Mannan to Mannose		Xylan to Tar	0	Xylose to Succinic Acid	0.009
Mannan to HMF		Arabinan to Oligomer	0.777	Xylose to Acetic Acid	0.014
Galactan to Oligomer	0.5	Arabinan to Arabinose	0	Xylose to Lactic Acid	0.002
Galactan to Galactose		Galactan to Oligomer	0	Arabinose to Ethanol	0
Galactan to HMF		Galactan to Galactose	0.777	Arabinose to Zymo	0
Arabinan to Oligomer	0.5	Galactose Oligomer to Galactose	0	Arabinose to Glycerol	0
Arabinan to Arabinose		Mannan to Oligomer	0	Arabinose to Succinic Acid	0
Arabinan to Furfural		Mannan to Mannose	0.777	Arabinose to Acetic Acid	0
Arabinan to Tar		Mannose Oligomer to Mannose	0	Arabinose to Lactic Acid	0
Acetate to Oligomer		**Sugar & Solids Flow Rates (kg/hr)**		Galactose to Ethanol	0
Acetate to Acetic Acid		Soluble Sugars From PT	15,980	Galactose to Zymo	0
Furfural to Tar		Other Soluble Solids From PT	21,338	Galactose to Glycerol	0
HMF to Tar		Soluble Sugars in Purchased Cellulase	0	Galactose to Succinic Acid	0
Lignin to Soluble Lignin	0.33	Other Sol Solids in Purchased Cellulase	0	Galactose to Acetic Acid	0
S/L Separation		Soluble Sugars in Produced Cellulase	0	Galactose to Lactic Acid	0
Water/Hydrolyzate Ratio (kg/kg)	N/A	Other Sol Solids in Produced Cellulase	0	Mannose to Ethanol	0

Pretreatment, cont.	
Water to S/L Separator (kg/hr)	N/A
Dilution Water (kg/hr)	N/A
Conditioning	
Ca(OH)2 to Gypsum	#N/A
Sugar Conversion to TAR:	
Xylose	#N/A
Arabinose	#N/A
Glucose	#N/A
Galactose	#N/A
Mannose	#N/A
Cellobiose	#N/A

Saccharification, cont.	
Soluble Sugars From Sacc	50,400
Other Soluble Solids From Sacc	19,205
Soluble Sugars From Seed Train	1,962
Other Soluble Solids From Seed Train	2,242
Soluble Sugars From DAP	0
Other Soluble Solids From DAP	0
Soluble Sugars From CSL	0
Other Soluble Solids From CSL	2,242

Fermentation, cont.	
Mannose to Zymo	0
Mannose to Glycerol	0
Mannose to Succinic Acid	0
Mannose to Acetic Acid	0
Mannose to Lactic Acid	0
Contamination Loss	7.0%
Ethanol Out of Fermenters (wt%)	3.9%

Table F-6. Operating Summary for Separate C5 & C6 Fermentation Processes

Pretreatment		**Saccharification**		**Fermentation**	
Acid Conc (wt%)	0.0190	Enzyme Loading (mg/g cell)	33.4	Total Solids (wt%)	17.1%
Acid Loading (g acid/g dry biomass)	0.0443	Total Solids (wt%)	16.1%	Insoluble Solids (wt%)	3.9%
Total Solids (wt%)	0.2968	Insoluble Solids (wt%)	9.4%	Temperature (°C)	32
Temperature (°C)	190	Temperature (°C)	32	Pressure (atm)	1.0
Pressure (atm)	11.4	Pressure (atm)	1.0	Residence Time (days)	2
Residence Time (min.)		Residence Time (days)	5	Conversions:	
Conversions:		Conversions:		Glucose to Ethanol	0.97
Cellulose to Glucolig	0	Cellulose to Glucolig	0	Glucose to Zymo	0.01
Cellulose to Cellobiose	0	Cellulose to Cellobiose	0	Glucose to Glycerol	0.002
Cellulose to Glucose	0.0626	Cellulose to Glucose	0.9109	Glucose to Succinic Acid	0.002
Cellulose to HMF	0	Glucolig to Cellobiose	0	Glucose to Acetic Acid	0.005
Xylan to Oligomer	0.0265	Glucolig to Glucose	0	Glucose to Lactic Acid	0.001
Xylan to Xylose	0.8249	Cellobiose to Glucose	0	Xylose to Ethanol	0
Xylan to Furfural	0	Xylan to Oligomer	0	Xylose to Zymo	0
Xylan to Tar	0	Xylan to Xylose	0.5713	Xylose to Glycerol	0
Mannan to Oligomer	0.0265	Xylose Oligomer to Xylose	0	Xylose to Xylitol	0
Mannan to Mannose	0.8249	Xylan to Tar	0	Xylose to Succinic Acid	0
Mannan to HMF	0	Arabinan to Oligomer	0.5713	Xylose to Acetic Acid	0
Galactan to Oligomer	0.0265	Arabinan to Arabinose	0	Xylose to Lactic Acid	0
Galactan to Galactose	0.8249	Galactan to Oligomer	0	Arabinose to Ethanol	0
Galactan to HMF	0	Galactan to Galactose	0.5713	Arabinose to Zymo	0
Arabinan to Oligomer	0.0265	Galactose Oligomer to Galactose	0	Arabinose to Glycerol	0
Arabinan to Arabinose	0.8249	Mannan to Oligomer	0	Arabinose to Succinic Acid	0
Arabinan to Furfural	0	Mannan to Mannose	0.5713	Arabinose to Acetic Acid	0
Arabinan to Tar	0	Mannose Oligomer to Mannose	0	Arabinose to Lactic Acid	0
Acetate to Oligomer	0	**Sugar & Solids Flow Rates (kg/hr)**		Galactose to Ethanol	0
Acetate to Acetic Acid	1	Soluble Sugars From PT	10,676	Galactose to Zymo	0
Furfural to Tar	1	Other Soluble Solids From PT	20,699	Galactose to Glycerol	0
HMF to Tar	1	Soluble Sugars in Purchased Cellulase	0	Galactose to Succinic Acid	0
Lignin to Soluble Lignin	0.1	Other Sol Solids in Purchased Cellulase	0	Galactose to Acetic Acid	0
S/L Separation		Soluble Sugars in Produced Cellulase	0	Galactose to Lactic Acid	0
Water/Hydrolyzate Ratio (kg/kg)	0.58	Other Sol Solids in Produced Cellulase	0	Mannose to Ethanol	0
Water to S/L Separator (kg/hr)	137,034			Mannose to Zymo	0

Pretreatment, cont.			Saccharification, cont.			Fermentation, cont.	
Dilution Water (kg/hr)	67,796		Soluble Sugars From Sacc	39,360		Mannose to Glycerol	0
Conditioning			Other Soluble Solids From Sacc	20,699		Mannose to Succinic Acid	0
Ca(OH)2 to Gypsum	#N/A		Soluble Sugars From Seed Train	0		Mannose to Acetic Acid	0
Sugar Conversion to TAR:			Other Soluble Solids From Seed Train	0		Mannose to Lactic Acid	0
Xylose	0.018		Soluble Sugars From DAP	0			
Arabinose	0.018		Other Soluble Solids From DAP	0		Contamination Loss	7.0%
Glucose	0.006		Soluble Sugars From CSL	0			
Galactose	0.006		Other Soluble Solids From CSL	2,371		Ethanol Out of Fermenters (wt%)	4.3%
Mannose	0.006						
Cellobiose	0.006						

Table F-7. Operating Summary for Pervaporation Two-Stage Dilute Acid Pretreatment Processes

Pretreatment	
Acid Conc (wt%)	0.0190
Acid Loading (g acid/g dry biomass)	0.0443
Total Solids (wt%)	0.2960
Temperature (°C)	190
Pressure (atm)	11.4
Residence Time (min.)	
Conversions:	
Cellulose to Glucolig	0
Cellulose to Cellobiose	0
Cellulose to Glucose	0.0626
Cellulose to HMF	0
Xylan to Oligomer	0.0265
Xylan to Xylose	0.8249
Xylan to Furfural	0
Xylan to Tar	0
Mannan to Oligomer	0.0265
Mannan to Mannose	0.8249
Mannan to HMF	0
Galactan to Oligomer	0.0265
Galactan to Galactose	0.8249
Galactan to HMF	0
Arabinan to Oligomer	0.0265
Arabinan to Arabinose	0.8249
Arabinan to Furfural	0
Arabinan to Tar	0
Acetate to Oligomer	0
Acetate to Acetic Acid	1
Furfural to Tar	1
HMF to Tar	1
Lignin to Soluble Lignin	0.1
S/L Separation	
Water/Hydrolyzate Ratio (kg/kg)	0.58
Water to S/L Separator (kg/hr)	137,371

Saccharification	
Enzyme Loading (mg/g cell)	33.5
Total Solids (wt%)	20.0%
Insoluble Solids (wt%)	10.3%
Temperature (°C)	32
Pressure (atm)	1.0
Residence Time (days)	5
Conversions:	
Cellulose to Glucolig	0
Cellulose to Cellobiose	0
Cellulose to Glucose	0.9109
Glucolig to Cellobiose	0
Glucolig to Glucose	0
Cellobiose to Glucose	0
Xylan to Oligomer	0
Xylan to Xylose	0.5713
Xylose Oligomer to Xylose	0
Xylan to Tar	0
Arabinan to Oligomer	0.5713
Arabinan to Arabinose	0
Galactan to Oligomer	0
Galactan to Galactose	0.5713
Galactose Oligomer to Galactose	0
Mannan to Oligomer	0
Mannan to Mannose	0.5713
Mannose Oligomer to Mannose	0
Sugar & Solids Flow Rates (kg/hr)	
Soluble Sugars From PT	22,739
Other Soluble Solids From PT	18,576
Soluble Sugars in Purchased Cellulase	0
Other Sol Solids in Purchased Cellulase	0
Soluble Sugars in Produced Cellulase	0
Other Sol Solids in Produced Cellulase	0

Fermentation	
Total Solids (wt%)	20.1%
Insoluble Solids (wt%)	4.2%
Temperature (°C)	32
Pressure (atm)	1.0
Residence Time (days)	2
Conversions:	
Glucose to Ethanol	0.95
Glucose to Zymo	0.02
Glucose to Glycerol	0.004
Glucose to Succinic Acid	0.006
Glucose to Acetic Acid	0.015
Glucose to Lactic Acid	0.002
Xylose to Ethanol	0.756
Xylose to Zymo	0.019
Xylose to Glycerol	0.003
Xylose to Xylitol	0.046
Xylose to Succinic Acid	0.009
Xylose to Acetic Acid	0.014
Xylose to Lactic Acid	0.002
Arabinose to Ethanol	0
Arabinose to Zymo	0
Arabinose to Glycerol	0
Arabinose to Succinic Acid	0
Arabinose to Acetic Acid	0
Arabinose to Lactic Acid	0
Galactose to Ethanol	0
Galactose to Zymo	0
Galactose to Glycerol	0
Galactose to Succinic Acid	0
Galactose to Acetic Acid	0
Galactose to Lactic Acid	0
Mannose to Ethanol	0
Mannose to Zymo	0

Pretreatment, cont.		**Saccharification, cont.**		**Fermentation, cont.**	
Dilution Water (kg/hr)	45,921	Soluble Sugars From Sacc	46,152	Mannose to Glycerol	0
Conditioning		Other Soluble Solids From Sacc	16,718	Mannose to Succinic Acid	0
Ca(OH)2 to Gypsum	0.174	Soluble Sugars From Seed Train	839	Mannose to Acetic Acid	0
Sugar Conversion to TAR:		Other Soluble Solids From Seed Train	1,960	Mannose to Lactic Acid	0
Xylose	0.13	Soluble Sugars From DAP	0		
Arabinose	0.2	Other Soluble Solids From DAP	0	Contamination Loss	7.0%
Glucose	0.12	Soluble Sugars From CSL	0		
Galactose	0.28	Other Soluble Solids From CSL	2,138	Ethanol Out of Fermenters (wt%)	4.8%
Mannose	0				
Cellobiose	0.36				

Table F-8. Operating Summary for On-site Enzyme Production Processes

Pretreatment		**Saccharification**		**Fermentation**	
Acid Conc (wt%)	0.0190	Enzyme Loading (mg/g cell)	0.0	Total Solids (wt%)	20.1%
Acid Loading (g acid/g dry biomass)	0.0442	Total Solids (wt%)	20.0%	Insoluble Solids (wt%)	4.7%
Total Solids (wt%)	0.2959	Insoluble Solids (wt%)	10.2%	Temperature (°C)	32
Temperature (°C)	190	Temperature (°C)	32	Pressure (atm)	1.0
Pressure (atm)	11.4	Pressure (atm)	1.0	Residence Time (days)	2
Residence Time (min.)		Residence Time (days)	5	Conversions:	
Conversions:		Conversions:		Glucose to Ethanol	0.95
Cellulose to Glucolig	0	Cellulose to Glucolig	0	Glucose to Zymo	0.02
Cellulose to Cellobiose	0	Cellulose to Cellobiose	0	Glucose to Glycerol	0.004
Cellulose to Glucose	0.0626	Cellulose to Glucose	0.9109	Glucose to Succinic Acid	0.006
Cellulose to HMF	0	Glucolig to Cellobiose	0	Glucose to Acetic Acid	0.015
Xylan to Oligomer	0.0265	Glucolig to Glucose	0	Glucose to Lactic Acid	0.002
Xylan to Xylose	0.8249	Cellobiose to Glucose	0	Xylose to Ethanol	0.756
Xylan to Furfural	0	Xylan to Oligomer	0	Xylose to Zymo	0.019
Xylan to Tar	0.085	Xylan to Xylose	0.5713	Xylose to Glycerol	0.003
Mannan to Oligomer	0.0265	Xylose Oligomer to Xylose	0	Xylose to Xylitol	0.046
Mannan to Mannose	0.8249	Xylan to Tar	0	Xylose to Succinic Acid	0.009
Mannan to HMF	0	Arabinan to Oligomer	0.5713	Xylose to Acetic Acid	0.014
Galactan to Oligomer	0.0265	Arabinan to Arabinose	0	Xylose to Lactic Acid	0.002
Galactan to Galactose	0.8249	Galactan to Oligomer	0	Arabinose to Ethanol	0
Galactan to HMF	0	Galactan to Galactose	0.5713	Arabinose to Zymo	0
Arabinan to Oligomer	0.0265	Galactose Oligomer to Galactose	0	Arabinose to Glycerol	0
Arabinan to Arabinose	0.8249	Mannan to Oligomer	0	Arabinose to Succinic Acid	0
Arabinan to Furfural	0	Mannan to Mannose	0.5713	Arabinose to Acetic Acid	0
Arabinan to Tar	0.085	Mannose Oligomer to Mannose	0	Arabinose to Lactic Acid	0
Acetate to Oligomer	0	**Sugar & Solids Flow Rates (kg/hr)**		Galactose to Ethanol	0
Acetate to Acetic Acid	1	Soluble Sugars From PT	20,317	Galactose to Zymo	0
Furfural to Tar	1	Other Soluble Solids From PT	16,259	Galactose to Glycerol	0
HMF to Tar	1	Soluble Sugars in Purchased Cellulase	0	Galactose to Succinic Acid	0
Lignin to Soluble Lignin	0.1	Other Sol Solids in Purchased Cellulase	0	Galactose to Acetic Acid	0
S/L Separation		Soluble Sugars in Produced Cellulase	545	Galactose to Lactic Acid	0
Water/Hydrolyzate Ratio (kg/kg)	0.58	Other Sol Solids in Produced Cellulase	2,781	Mannose to Ethanol	0
Water to S/L Separator (kg/hr)	137,416			Mannose to Zymo	0

Pretreatment, cont.			Saccharification, cont.			Fermentation, cont.	
Dilution Water (kg/hr)	19,186		Soluble Sugars From Sacc	41,246		Mannose to Glycerol	0
Conditioning			Other Soluble Solids From Sacc	17,136		Mannose to Succinic Acid	0
Ca(OH)2 to Gypsum	0.175		Soluble Sugars From Seed Train	783		Mannose to Acetic Acid	0
Sugar Conversion to TAR:			Other Soluble Solids From Seed Train	2,003		Mannose to Lactic Acid	0
Xylose	0.13		Soluble Sugars From DAP	0			
Arabinose	0.2		Other Soluble Solids From DAP	0		Contamination Loss	7.0%
Glucose	0.12		Soluble Sugars From CSL	0			
Galactose	0.28		Other Soluble Solids From CSL	2,056		Ethanol Out of Fermenters (wt%)	4.5%
Mannose	0						
Cellobiose	0.36						

Appendix G - General Process Description

Area 100

The feedstock handling area receives the corn stover in bales. After being unwrapped, stover is washed to remove dirt before being passed through a magnetic separator to remove tramp metal and then conveyed to the shredders for size reduction.

Area 200

In the pretreatment area, biomass undergoes a physical and/or chemical treatment, which allows for improved exposure of the cellulose during enzymatic hydrolysis. The details of the four pretreatment processes are discussed under the Pretreatment Variation section of process variations in the results section.

Area 300

Enzymatic saccharification followed by fermentation of sugars occurs in this area. The pretreatment hydrolyzate from Area 200 is pumped into 18 parallel saccharification vessels (each has a capacity of 1 MM gal) where enzyme is added (enzyme loading of 31.3 mg protein/g cellulose) either from purchased stock preparation or from the on-site enzyme production section (Area 400). The residence time for saccharification is 5 days. Note that for two-stage dilute acid treatment, acid is used in place of enzyme to hydrolyze the cellulose and xylan.

A small fraction of the hydrolyzate leaving the saccharification vessel is sent to one of two sequenced batch fermentation seed vessel trains to be used as a carbon source for the growth of the fermentative organism, *Zymomonas mobilis.* Nutrients such as CSL and diammonium phosphate are also added into the bio-reactors.

The bulk of the hydrolyzate from the saccharification vessels is pumped to one of eight parallel sequenced batch fermentation reactors, where *Zymomonas mobilis* is introduced from the seed reactors. A 2 day residence time is used for fermentation. Vent gas from the fermentors is scrubbed to collect escaped ethanol in the scrubber and sent to the beer column along with fermented beer from the fermentors.

Area 500

Ethanol is separated and recovered from water and residual solids in this section. Fermented beer is distilled in two distillation columns: beer column and rectification column. The beer column removes dissolved CO_2 and vapor generating a concentrated ethanol stream. The rectification column purifies the ethanol to about 95 wt% concentration. Nearly all of the remaining water is removed in a molecular sieve adsorption column. The stillage from the beer column is pumped to the 1st-effect evaporator where water content is reduced, followed by dewatering in a Pneumapress filter and screw press to separate most of the insoluble solids, comprising mainly lignin, from the aqueous stream. The liquid fraction from the screw press is concentrated in the 2nd- and the 3rd-effect evaporators to high concentration syrup of soluble solids. The evaporator syrup and the dewatered solid fraction from the screw press are used as boiler fuel in the fluidized bed combustor.

Area 600

Wastewater is treated in anaerobic and aerobic digesters. In the anaerobic digester, a small amount of urea, phosphoric acid, and micronutrients are added as nutrients for the anaerobic organisms. Biogas from the anaerobic digester is used as boiler fuel. The wastewater is further treated aerobically. It is then held in a clarifying tank where the settled solids are separated from the water. The solids are dewatered in a belt filter press, with a polymer being added to aid in dewatering, followed by a screw press. The resulting sludge is used as boiler fuel. The water from the clarifying tank is recycled as process water.

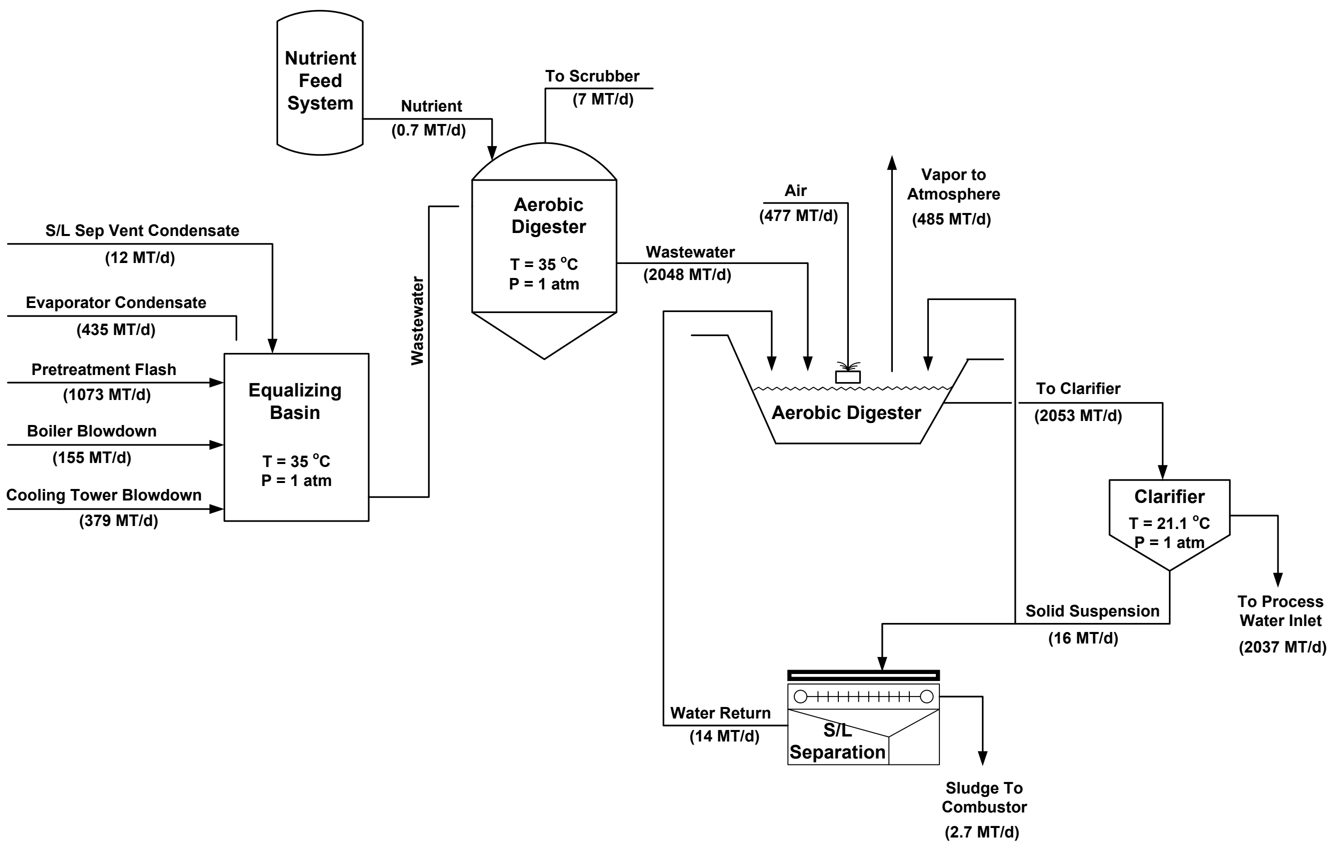

Figure G-1. Wastewater treatment section (Area 600)

Area 800

Cogeneration of steam for process heat and electricity occurs in Area 800. Evaporator syrup, insoluble solids from the Pneumapress, wastewater treatment sludge, and biogas are used as fuel in the fluidized bed combustor. Superheated steam is generated in the boiler and exits the multi-stage turbine at three different conditions needed in the process.

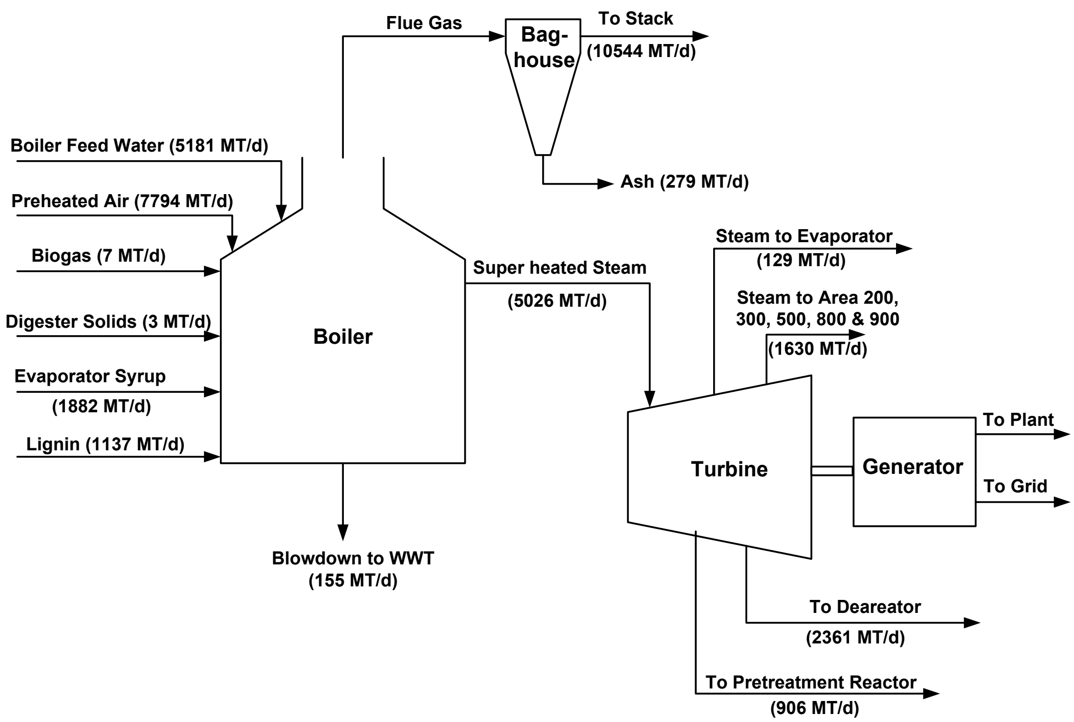

Figure G-2. Steam and power generation section (Area 800)

Appendix H - Cost By Area Of Process Scenarios

Table H-1. Costs by Area of the Dilute Acid Pretreatment (High Solids) Scenario

Cost Areas / Factor	Installed Cost		Purchased Equipment Cost	
	(MM$)	(%)	(MM$)	(%)
Feedstock Handling (Area 100)	10.9	6.4	6.0	5.3
Pretreatment (Area 200)	38.0	22.4	21.0	18.5
Saccharification & Fermentation (Area 300)	21.8	12.9	17.3	15.3
Distillation and Solids Recovery (Area 500)	25.7	15.1	16.9	14.9
Wastewater Treatment (Area 600)	5.8	3.4	4.5	4.0
Storage (Area 700)	3.1	1.8	1.9	1.7
Boiler/Turbogenerator (Area 800)	57.6	24.4	41.4	36.5
Utilities (Area 900)	6.6	3.9	4.3	3.8
Purchased Equipment Cost			**113.3**	**100**
Total Installed Equipment Cost	**169.4**	**100**		
Fixed Capital Investment (FCI)	**337.8**			
Working Capital (WC)	50.7			
Total Capital Investment (TCI)	**388.5**			
Lang Factor	3.43			

Table H-2. Costs by Area of the 2-Stage Dilute Acid Pretreatment Scenario

Cost Areas / Factor	Installed Cost		Purchased Equipment Cost	
	(MM$)	(%)	(MM$)	(%)
Feedstock Handling (Area 100)	10.9	6.3	6.0	5.4
Pretreatment (Area 200)	44.9	26.0	23.5	21.0
Saccharification & Fermentation (Area 300)	9.7	5.6	7.2	6.4
Distillation and Solids Recovery (Area 500)	26.7	15.5	18.1	16.1
Wastewater Treatment (Area 600)	4.6	2.7	3.5	3.1
Storage (Area 700)	2.8	1.6	1.7	1.5
Boiler/Turbogenerator (Area 800)	66.2	38.3	47.6	42.5
Utilities (Area 900)	6.9	4.0	4.5	4.0
Purchased Equipment Cost			**112.1**	**100**
Total Installed Equipment Cost	**172.7**	**100**		
Fixed Capital Investment (FCI)	**340.0**			
Working Capital (WC)	51.0			
Total Capital Investment (TCI)	**391.0**			
Lang Factor	3.49			

Table H-3. Costs by Area of the Hot Water Pretreatment Scenario

Cost Areas / Factor	Installed Cost		Purchased Equipment Cost	
	(MM$)	(%)	(MM$)	(%)
Feedstock Handling (Area 100)	10.9	7.0	6.0	5.6
Pretreatment (Area 200)	6.7	4.3	3.9	3.6
Saccharification & Fermentation (Area 300)	30.2	19.3	23.4	21.7
Distillation and Solids Recovery (Area 500)	30.9	19.8	19.4	18.0
Wastewater Treatment (Area 600)	1.9	1.2	1.3	1.2
Storage (Area 700)	3.3	2.1	2.1	2.0
Boiler/Turbogenerator (Area 800)	65.8	42.1	47.3	43.9
Utilities (Area 900)	6.7	4.3	4.3	4.0
Purchased Equipment Cost			**107.8**	**100**
Total Installed Equipment Cost	**156.3**	**100**		
Fixed Capital Investment (FCI)	**314.0**			
Working Capital (WC)	47.1			
Total Capital Investment (TCI)	**361.1**			
Lang Factor	3.35			

Table H-4. Costs by Area of the AFEX Pretreatment Scenario

Cost Areas / Factor	Installed Cost		Purchased Equipment Cost	
	(MM$)	(%)	(MM$)	(%)
Feedstock Handling (Area 100)	10.9	6.5	6.0	5.3
Pretreatment (Area 200)	30.8	18.4	18.4	16.1
Saccharification & Fermentation (Area 300)	23.5	14.0	18.8	16.5
Distillation and Solids Recovery (Area 500)	27.5	16.4	17.5	15.4
Wastewater Treatment (Area 600)	1.6	1.0	1.1	1.0
Storage (Area 700)	2.8	1.7	1.8	1.6
Boiler/Turbogenerator (Area 800)	62.0	37.0	44.6	39.1
Utilities (Area 900)	8.4	5.0	5.8	5.1
Purchased Equipment Cost			**113.9**	**100**
Total Installed Equipment Cost	**167.4**	**100**		
Fixed Capital Investment (FCI)	**335.6**			
Working Capital (WC)	50.4			
Total Capital Investment (TCI)	**386.0**			
Lang Factor	3.39			

Table H-5. Costs by Area of the Separate C-5 and C-6 Fermentation Process Scenario

Cost Areas / Factor	Installed Cost		Purchased Equipment Cost	
	(MM$)	(%)	(MM$)	(%)
Feedstock Handling (Area 100)	11.4	6.8	6.2	5.5
Pretreatment (Area 200)	34.2	20.4	18.4	16.4
Saccharification & Fermentation (Area 300)	29.6	17.6	23.7	21.1
Distillation and Solids Recovery (Area 500)	27.3	16.3	17.8	15.8
Wastewater Treatment (Area 600)	3.1	1.8	2.3	2.0
Storage (Area 700)	5.1	3.0	3.4	3.0
Boiler/Turbogenerator (Area 800)	50.9	30.3	36.5	32.5
Utilities (Area 900)	6.2	3.7	4.1	3.6
Purchased Equipment Cost			**112.4**	**100**
Total Installed Equipment Cost	**167.8**	**100**		
Fixed Capital Investment (FCI)	**335.4**			
Working Capital (WC)	50.3			
Total Capital Investment (TCI)	**385.8**			
Lang Factor	3.43			

Table H-6. Costs by Area of the Pervaporation Process Scenario

Cost Areas / Factor	Installed Cost		Purchased Equipment Cost	
	(MM$)	(%)	(MM$)	(%)
Feedstock Handling (Area 100)	10.9	5.2	6.0	3.9
Pretreatment (Area 200)	36.2	17.3	19.7	12.7
Saccharification & Fermentation (Area 300)	21.9	10.5	17.4	11.2
Distillation and Solids Recovery (Area 500)	70.8	33.8	62.9	40.5
Wastewater Treatment (Area 600)	3.3	1.6	2.4	1.5
Storage (Area 700)	3.3	1.6	2.1	1.4
Boiler/Turbogenerator (Area 800)	55.9	26.7	40.1	25.8
Utilities (Area 900)	7.0	3.3	4.6	3.0
Purchased Equipment Cost			**155.2**	**100**
Total Installed Equipment Cost	**209.2**	**100**		
Fixed Capital Investment (FCI)	**435.7**			
Working Capital (WC)	65.4			
Total Capital Investment (TCI)	**501.0**			
Lang Factor	3.23			

Table H-7. Costs by Area of the On-site Enzyme Production Processes Scenario

Cost Areas / Factor	Installed Cost		Purchased Equipment Cost	
	(MM$)	(%)	(MM$)	(%)
Feedstock Handling (Area 100)	10.9	5.8	6.0	4.7
Pretreatment (Area 200)	36.2	19.3	19.7	15.5
Saccharification & Fermentation (Area 300)	21.8	11.6	17.3	13.6
On-site Enzyme Production (Area 400)	23.7	12.6	17.9	14.0
Distillation and Solids Recovery (Area 500)	25.3	13.5	16.8	13.2
Wastewater Treatment (Area 600)	3.4	1.9	2.6	2.0
Storage (Area 700)	2.5	1.3	1.6	1.2
Boiler/Turbogenerator (Area 800)	57.2	30.4	41.1	32.3
Utilities (Area 900)	6.8	3.6	4.4	3.5
Purchased Equipment Cost			**127.4**	**100**
Total Installed Equipment Cost	**187.8**	**100**		
Fixed Capital Investment (FCI)	**377.5**			
Working Capital (WC)	56.6			
Total Capital Investment (TCI)	**434.2**			
Lang Factor	3.41			

Appendix I - Down Selection Matrix

Table I-1. Process Down-Selection Matrix

	Capital Expenditure ($x10^6$\$)	Operating Costs ($x10^6$\$)	Fuel Cost (\$/gal)	Internal Rate of Return (%)	Risk	Plant Efficiency (%)	Carbon Efficiency	Process Yield (fuel) [gal Ethanol or Butanol/ ton feed stock]	Capacity Factor	Optimal Plant Size (MTon/day)	Complexity of Process	Technology Development	Water Use/Discharge (Mton/day)	Greenhouse Gas Emissions	Particulates/Solid Wastes	Air Emissions	Waste Treatment	Toxics	Energy Content (Gasoline Equivalent)	Engine Compatibility	Use Emissions	Infrastructure Compatibility	Fuel Toxicity	Ability to Meet Fuel Standards	Bibliography[a]	Notes	Process Blocks
Ethanol Concentrated Acid/Fermentation																											
Ethanol Dilute Acid Pretreatment/Enzymatic / Fermentation	3.24	0.16	1.07	10			48	89.7	93	2000		Pilot	5440						0.658						1		Process Blocks: (1) Pretreatment (dilute H_2SO_4); (2) Continuous Saccharification; and (3) Cofermentation (hydrolysis & fermentation units are in series); (4) Separation/Purification (distillation and molecular sieve); (5) Water treatment (treated water recycle to process)
Ethanol Dilute Acid Pretreatment/Enzymatic / Fermentation			2.63	10					91	2000		Commercial/Pilot							0.628						2	Scenario-I (Short Term: 5 years, following pretreatment and SSF)	Process Blocks: (1) Pretreatment (dilute acid); (2) Hydrolysis (Enzymatic); (3) Fermentation; (4) Separation/Purification (distillation & desiccants)
Ethanol Dilute Acid Pretreatment/Enzymatic / Fermentation	4.53		1.71	12			48	76.2	95.9	5000		Commercial/Pilot													3	Current Process (pretreatment & SSF)	Process Blocks: (1) Pretreatment (dilute acid); (2a) SSF (simultaneous sccharification & fermentation); (2b) with separate fermentation for C5; (3) Separation/Purification (2-column distillation); (4) Water treatment (suspended sludge digester)
Ethanol Dilute Acid Pretreatment/Enzymatic / Fermentation	1.27	1.02	0.88				50	119.5	90.4	634									0.658						4	Feedstock: Corn	Process Blocks: (1) Pretreatment (dilute H_2SO_4); (2) Simultaneous Saccharification and Cofermentation (glucose & Xylose); and (3) Separation/Purification (distillation and molecular sieve); (5) Water treatment (treated water recycle to process)
Ethanol Dilute Acid Pretreatment/Enzymatic / Fermentation	4.83	1.68	1.45				42	79.2	90.4	957									0.658						5	Feedstock: Cornstover	Process Blocks: (1) Pretreatment (dilute H_2SO_4); (2) Simultaneous Saccharification and Cofermentation (glucose & Xylose); and (3) Separation/Purification (distillation and molecular sieve); (5) Water treatment (treated water recycle to process)
Ethanol Dilute Acid Pretreatment/Enzymatic / Fermentation	4.91	0.25	1.14	10			36	68	95.9	2000		Pilot	6307						0.661						6	SSCF Base Case; Feedstock: Yellow poplar	Process Blocks: (1) Pretreatment (dilute H_2SO_4); (2) Simultaneous sccharification and cofermentation (of glucose & xylan); (3) Separation/Purification; (5) Water treatment (treated water recycle to process)

	Capital Expenditure (x10^6$)	Operating Costs (x10^6$)	Fuel Cost ($/gal)	Internal Rate of Return (%)	Risk	Plant Efficiency (%)	Carbon Efficiency	Process Yield (fuel) [gal Ethanol or Butanol/ ton feed stock]	Capacity Factor	Optimal Plant Size (MTon/day)	Complexity of Process	Technology Development	Water Use/Discharge (Mton/day)	Greenhouse Gas Emissions	Particulates/Solid Wastes	Air Emissions	Waste Treatment	Toxics	Energy Content (Gasoline Equivalent)	Engine Compatibility	Use Emissions	Infrastructure Compatibility	Fuel Toxicity	Ability to Meet Fuel Standards	Bibliography[a]	Notes	Process Blocks
Ethanol Dilute Acid Pretreatment/Enzymatic / Fermentation	3.85		1.16	10			40	76	95.9	2000		Pilot	6307						0.661						7	SSCF Near Term; Feedstock: Yellow poplar	
Ethanol Dilute Acid Pretreatment/Enzymatic / Fermentation	2.98		0.94	10			42	81	95.9	2000		Pilot	6307						0.661						8	SSCF 2005; Feedstock: Yellow poplar	
Ethanol Dilute Acid Pretreatment/Enzymatic / Fermentation	2.37		0.82	10			49	94	95.9	2000		Pilot	6307						0.661						9	SSCF 2010; Feedstock: Yellow poplar	
Ethanol Dilute Acid Pretreatment/Enzymatic / Fermentation	2.29		0.76	10			52	99	95.9	2000		Pilot	6307						0.661						10	SSCF 2015; Feedstock: Yellow poplar	
Ethanol Dilute Acid Pretreatment/Enzymatic / Fermentation	8.67	1.76	1.25	15			78	33	100	4360		Pilot													11	SHF (TVA process); High profit from co-products: furfural	Process Blocks: (1) Pretreatment (dilute H_2SO_4); (2) Hydrolysis (Enzymatic); (3) Fermentation (Hexose & Pentose in separate vessels); (4) Separation/Purification (distillation); (5) Waste treatment
Ethanol SO_2-Steam Pretreatment/Enzymatic / Fermentation	5.58	0.45	0.98	25			42	78.5	100	500															12	SO_2 & NaOH in 2-step preteratment; Rate of return on working capital	Process Blocks: (1) Pretreatment (SO_2, Steam); (2) Hydrolysis (Enzymatic); (3a) Fermentation, and (3b) fermentation of C5 sugars in separate vessels; (4) Seperation; (5) water treatment
Ethanol SO_2-Steam Pretreatment/Enzymatic / Fermentation	13.61		0.75	5			36	69.6	91.2	586.8		lab	552						0.624						13	SHF process; SO_2 is used for pretreatment; Total Investment is 1184 MSKR	Process Blocks: (1) Pretreatment (Stake Tech: SO_2 & Steam); (2) Hydrolysis (Enzymatic); (3) Fermentation; (4) Separation/Purification (distillation & drying)
Ethanol SO_2-Steam Pretreatment/Enzymatic / Fermentation	9.18		0.83	5			41	79.2	91.2	586.8		lab	586						0.618						14	SSF process; SO_2 is used for pretreatment; Total Investment is 910 MSKR	Process Blocks: (1) Pretreatment (Stake Tech: SO_2 & Steam); (2) SSF (Simultaneous Saccharification & Fermentation); (3) Separation/Purification (distillation & drying)
Ethanol SO_2-Steam Pretreatment/Enzymatic / Fermentation	8.63		2.24	6			40	76.8	91.2	600		lab							0.621						15	SSF process; SO_2 is used for pretreatment; One-Step Pretreatment; Total Investment is 847 MSKR	Process Blocks: (1) Pretreatment (Stake Tech: SO_2 & Steam); (2) SSF (Simultaneous Saccharification & Fermentation; feed batch); (3) Separation/Purification (2-separate stripper & rectifier operations)

	Capital Expenditure ($x10^6$$)	Operating Costs ($x10^6$$)	Fuel Cost ($/gal)	Internal Rate of Return (%)	Risk	Plant Efficiency (%)	Carbon Efficiency	Process Yield (fuel) [gal Ethanol or Butanol/ ton feed stock]	Capacity Factor	Optimal Plant Size (MTon/day)	Complexity of Process	Technology Development	Water Use/Discharge (Mton/day)	Greenhouse Gas Emissions	Particulates/Solid Wastes	Air Emissions	Waste Treatment	Toxics	Energy Content (Gasoline Equivalent)	Engine Compatibility	Use Emissions	Infrastructure Compatibility	Fuel Toxicity	Ability to Meet Fuel Standards	Bibliography[a]	Notes	Process Blocks
Ethanol SO_2-Steam Double Pretreatment/Enzymatic / Fermentation	8.72		2.24	6			42	80.1	91.2	600		lab							0.621						16	SSF process; SO_2 is used for pretreatment; Two-Step Pretreatment; Total Investment is 847 MSKR	Process Blocks: (1) Pretreatment (2-step SO_2-steam pretreatment: double pretreatment); (2) SSF (Simultaneous Saccharification & Fermentation; feed batch); (3) Separation/Purification (2-separate stripper & rectifier operations)
Ethanol AFEX Pretreatment/Enzymatic / Fermentation	1.95		0.73	12			61	105.3	95.9	5000		Lab													17	Advanced EtOH Rankine process	Process Blocks: (1) Pretreatment (Ammonia fiber expansion, AFEX, pretreatment - no detoxification); (2) SSF (simultaneous sccharification & fermentation); (3) Separation/Purification (1-column distillation with heat integration); (4) Water treatment (attached film digester)
Ethanol AFEX Pretreatment/Enzymatic / Fermentation	2.89		0.77	12			61	105.3	95.9	5000		conceptual													18	Advanced EtOH-GTCC Process	Process Blocks: (1) Pretreatment (Ammonia fiber expansion, AFEX, pretreatment- no detoxification); (2) CBP (consolidated bio-processing); (3) Separation/Purification (1-column distillation with heat integration); (4) Water treatment (attached film digester)
E thanol Liquid Hotwater Pretreatment/Enzymatic / Fermentation			0.99	10					91	2000		conceptual							0.628						19	Scenario-III (Long Term: >20 years, following Consolidated BioProcessing, CBP)	Process Blocks: (1) Pretreatment (Liquid Hot Water); (2) Consolidated Bio-Processing; (3) Separation/Purification (distillation & desiccants)
Ethanol Steam Pretreatment/Enzymatic / Fermentation	681-907		1.48	10					91	2000		lab/pilot							0.628						20	Scenario-II (Mid Term: 5-10 years, following pretreatment and SSCF)	Process Blocks: (1) Pretreatment (steam explosion); (2) SSF & SSCF(Simultaneous saccharification & fermentation, and simultaneous saccharification and cofermentation); (3) Separation/Purification (distillation & desiccants)
Ethanol Enzymatic / Fermentation	9.06	1.33	1.1	15			47	79.4	100	15		Lab							0.658						21	Scenario-I (minerals sent to land fill; No ethanol from xylan)	Process Blocks: (1) SSF (Simultaneous saccharification and fermentation); (2) Separation/Purification (distillation and molecular sieve); (3) Water treatment
Ethanol Enzymatic / Fermentation	7.70	0.85	1.1	15			55	93.5	100	15		no-lab demo							0.658						22	Scenario-II (90% of minerals are recovered and sent to paper mill)	Process Blocks: (1) SSF (Simultaneous saccharification and fermentation); (2) Separation/Purification (distillation and molecular sieve); (3) Water treatment
Butanol Concentrated Acid / Fermentation																											
Butanol Dilute Acid Pretreatment / Fermentation							44	28.4				lab							0.957						23	Feedstock: corn fiber; dilute H_2SO_4 hydrolysis	Process Blocks: (1) Hydrolysis (dilute H_2SO_4); (2) Fermentation; (3) Separation/Purification (distillation)

	Capital Expenditure (x10[6]$)	Operating Costs (x10[6]$)	Fuel Cost ($/gal)	Internal Rate of Return (%)	Risk	Plant Efficiency (%)	Carbon Efficiency	Process Yield (fuel) [gal Ethanol or Butanol/ ton feed stock]	Capacity Factor	Optimal Plant Size (MTon/day)	Complexity of Process	Technology Development	Water Use/Discharge (Mton/day)	Greenhouse Gas Emissions	Particulates/Solid Wastes	Air Emissions	Waste Treatment	Toxics	Energy Content (Gasoline Equivalent)	Engine Compatibility	Use Emissions	Infrastructure Compatibility	Fuel Toxicity	Ability to Meet Fuel Standards	Bibliography[a]	Notes	Process Blocks
Butanol AFEX Pretreatment / Enzymatic / Fermentation																											
Butanol Hotwater Pretreatment / Enzymatic / Fermentation							65	57.6				Lab							0.957						24	Feedstock: Saccharified Liquefied Cornstarch (SLCS) with moisture content of approximately 60%	Process Blocks: (1) Hydrolysis (enzymatic hydrolysis of LCS to SLCS); (2) Fermentation; (3) Separation/Purification
Butanol Hotwater Pretreatment / Fermentation							60	54				Lab							0.957						25	Feedstock: Liquefied Cornstarch (LCS) with moisture content of approximately 60%	Process Blocks: (1) Simultaneous saccharification and Fermentation; (2) Separation/Purification
Butanol Enzyme / Fermentation	0.81	0.07	0.74	10-30			66	89.84	95.9	1469		lab							0.949						26	Feedstock: corn (14% moisture); Hydrolysis + Immobilized Cell Continuous Fermentation & pervaporative recovery, ICCFPR process	Process Blocks: (1) Hydrolysis (enzymatic); (2) Fermentation (contineous); (3) Separation/Purification (pervaporation & distillation)
Butanol Fermentation	1.31	0.15	0.81	15			58	5.52	100	2268		lab							0.957						27	Feedstock: Whey permeat (4.5% lactose); Fibrous Bed Reactor, FBR; Process: 2 step fermentation	Process Blocks: (1) 2-step Fermentation (i. fermentation of lactose to butyric acid, ii. Fermentation of butyric acis to butanil); (3) Separation/Purification (adsorption, desorption, distillation);
Butanol Fermentation							57	114				lab							0.957							Feedstock: Glucose; 2-step fermentation, FBR	Process Blocks: (1) hydrolysis of lignocellulosics/starch to fermentable sugars; (2) 2-step Fermentation of glucose (i. fermentation of lactose to butyric acid, ii. Fermentation of butyric acis to butanil); (3) Separation/Purification (adsorption, desorption, distillation);
Butanol Fermentation	2.37	0.22	1.68	20			65	89.79	95.9	1469		Pilot							0.947						28	Feedstock: corn (14% moisture)	Process Blocks: (1) Simultaneous saccharification and Fermentation; (2) Separation/Purification (distillation); (3) Wastewater treatment (recycling to process)
Butanol Fermentation	1.00	0.09	0.95	10-30			66	89.84	95.9	1469		Pilot							0.949						29	Feedstock: corn (14% moisture); Batch Fermentation & pervaporative recovery, BFPR process	Process Blocks: (1) Simultaneous saccharification and Fermentation; (2) Separation/Purification (pervaporation & distillation)

	Capital Expenditure (x$10^{6}\$$)	Operating Costs (x$10^{6}\$$)	Fuel Cost ($/gal)	Internal Rate of Return (%)	Risk	Plant Efficiency (%)	Carbon Efficiency	Process Yield (fuel) [gal Ethanol or Butanol/ ton feed stock]	Capacity Factor	Optimal Plant Size (MTon/day)	Complexity of Process	Technology Development	Water Use/Discharge (Mton/day)	Greenhouse Gas Emissions	Particulates/Solid Wastes	Air Emissions	Waste Treatment	Toxics	Energy Content (Gasoline Equivalent)	Engine Compatibility	Use Emissions	Infrastructure Compatibility	Fuel Toxicity	Ability to Meet Fuel Standards	Bibliography[a]	Notes	Process Blocks
Butanol Fermentation	0.93	0.08	0.81	10-30			66	89.84	95.9	1469		Pilot							0.949						30	Feedstock: corn (14% moisture); Fed-batch Fermentation & pervaporative recovery, FBFPR process	Process Blocks: (1) Simultaneous saccharification and Fermentation; (2) Separation/Purification (pervaporation & distillation)
Zeachem Process																											

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Appendix J - Analysis: Cost Growth Variables And Results

Table J-1. Plant Performance and Cost Growth Variables for Dilute Acid Pretreatment (High Solids Loading) Processes

Plant Performance (Equation 1)				Cost Growth (Equation 2)			
Variables	Values			Variables	Values		
	Opti-mistic	Most Prob-able	Pessi-mistic		Opti-mistic	Most Probable	Pessi-mistic
NEWSTEPS[a]	6	6	6	PCTNEW	61.4	61.4	61.4
BALEQS	50	40	30	IMPURITIES	0	3	5
WASTE	1	2	3	COMPLEXITY[b]	6	6	6
SOLIDS	1	1	1	INCLUSIVENESS	33	0	0
				PROJECT DEFINITION	6	6	7
Plant Performance (%)	**22.1**	**14.7**	**7.3**	**Cost Growth (%)**	**52.6**	**42.6**	**32.0**

[a] New steps/units: Feedstock handling, Pretreatment, Saccharification, Cofermentation, Beer Column, and Combustor. [b] Continuously linked steps: Feedstock Handling, Pretreatment, Saccharification, Cofermentation, Distillation, Steam/Power Generation.

Table J-2. Plant Performance and Cost Growth Variables for Hot water Pretreatment Processes

Plant Performance (Equation 1)				Cost Growth (Equation 2)			
Variables	Values			Variables	Values		
	Opti-mistic	Most Prob-able	Pessi-mistic		Opti-mistic	Most Probable	Pessi-mistic
NEWSTEPS[a]	6	6	6	PCTNEW	53.6	53.6	53.6
BALEQS	50	40	30	IMPURITIES	0	3	5
WASTE	1	2	3	COMPLEXITY[b]	6	6	6
SOLIDS	1	1	1	INCLUSIVENESS	33	0	0
				PROJECT DEFINITION	6	6	7
Plant Performance (%)	**22.1**	**14.7**	**7.3**	**Cost Growth (%)**	**55.0**	**44.9**	**34.3**

[a] New steps/units: Feedstock handling, Pretreatment, Saccharification, Cofermentation, Beer Column, and Combustor. [b] Continuously linked steps: Feedstock Handling, Pretreatment, Saccharification, Cofermentation, Distillation, Steam/Power Generation.

Table J-3. Plant Performance and Cost Growth Variables for Two-Stage Dilute Acid Pretreatment Processes

Plant Performance (Equation 1)				Cost Growth (Equation 2)			
Variables	Values			Variables	Values		
	Opti-mistic	Most Prob-able	Pessi-mistic		Opti-mistic	Most Probable	Pessi-mistic
NEWSTEPS[a]	6	6	6	PCTNEW	60.8	60.8	60.8
BALEQS	50	40	30	IMPURITIES	0	3	5
WASTE	1	2	3	COMPLEXITY[b]	6	6	6
SOLIDS	1	1	1	INCLUSIVENESS	33	0	0
				PROJECT DEFINITION	6	6	7
Plant Performance (%)	**22.1**	**14.7**	**7.3**	**Cost Growth (%)**	**50.8**	**42.8**	**32.2**

[a] New steps/units: Feedstock Handling, Pretreatment, Saccharification, Cofermentation, Beer Column, and Combustor. [b] Continuously linked steps: Feedstock Handling, Pretreatment, Saccharification, Cofermentation, Distillation, Steam/Power Generation.

Table J-4. Plant Performance and Cost Growth Variables for AFEX Pretreatment Processes

Plant Performance (Equation 1)				Cost Growth (Equation 2)			
Variables	Values			Variables	Values		
	Opti-mistic	Most Prob-able	Pessi-mistic		Opti-mistic	Most Probable	Pessi-mistic
NEWSTEPS[a]	7	7	7	PCTNEW	60.38	60.38	60.38
BALEQS	50	40	30	IMPURITIES	0	3	5
WASTE	1	2	2	COMPLEXITY[b]	7	7	7
SOLIDS	1	1	1	INCLUSIVENESS	33	0	0
				PROJECT DEFINITION	6	6	7
Plant Performance (%)	**12.4**	**5.0**	**1.7**	**Cost Growth (%)**	**51.8**	**41.8**	**31.2**

[a] New steps/units**:** Feedstock Handling, Pretreatment, Saccharification, Cofermentation, Ammonia Separation, Beer Column, and Combustor. [b] Continuously linked steps: Feedstock Handling, Pretreatment, Saccharification, Cofermentation, Ammonia recovery, Distillation, Steam/Power Generation.

Table J-5. Plant Performance and Cost Growth Variables for On-Site Enzyme Production Processes

Plant Performance (Equation 1)				Cost Growth (Equation 2)			
Variables	Values			Variables	Values		
	Opti-mistic	Most Prob-able	Pessi-mistic		Opti-mistic	Most Probable	Pessi-mistic
NEWSTEPS[a]	7	7	7	PCTNEW	66.9	66.9	66.9
BALEQS	50	40	30	IMPURITIES	0	3	5
WASTE	1	2	3	COMPLEXITY[b]	7	7	7
SOLIDS	1	1	1	INCLUSIVENESS	33	0	0
				PROJECT DEFINITION	6	6	7
Plant Performance (%)	**12.4**	**5.0**	**0**	**Cost Growth (%)**	**49.9**	**39.8**	**29.2**

[a] New steps/units: Feedstock handling, Pretreatment, Saccharification, Cofermentation, Beer Column, Combustor, and Enzyme Production. [b] Continuously linked steps: Feedstock Handling, Pretreatment, Saccharification, Cofermentation, Distillation, Steam/Power Generation, and Enzyme Production.

Table J-6. PV from Cost Growth Analysis for Various Process Scenarios

Cost Item	Cost Growth (Pioneer Plant)[a]		
	Most Probable	Optimistic	Pessimistic
Dilute Acid Pretreatment Processes (Pilot)			
PV ($/Gal)	6.11	5.27	7.59
Total Capital Investment (MM$)	913.9	750.1	1,198.8
Lang Factor	8.06	6.62	10.58
Hot Water Pretreatment Processes			
PV ($/Gal)	7.32	6.41	8.87
Total Capital Investment (MM$)	808.3	619.0	1,042.5
Lang Factor	7.50	6.21	9.67
2-Stage Dilute Acid Pretreatment Processes			
PV ($/Gal)	8.36	7.10	10.56
Total Capital Investment (MM$)	916.2	752.6	1,200.2
Lang Factor	8.17	6.71	10.70
AFEX Pretreatment Processes			
PV ($/Gal)	6.84	5.85	8.41
Total Capital Investment (MM$)	925.2	756.5	1,221.7
Lang Factor	8.12	6.64	10.72
On-site Enzyme Production Processes			
PV ($/Gal)	7.21	6.04	9.21
Total Capital Investment (MM$)	1,088.1	881.4	1460.9
Lang Factor	8.54	6.92	11.47

[a] 30% contingency is used for pioneer plant cost analysis.

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13. SUPPLEMENTARY NOTES

14. ABSTRACT *(Maximum 200 Words)*

A techno-economic analysis on the production of cellulosic ethanol by fermentation was conducted to understand the viability of liquid biofuel production processes within the next 5-8 years. Initially, 35 technologies were reviewed, then a two-step down selection was performed to choose scenarios to be evaluated in a more detailed economic analysis. The lignocellulosic ethanol process was selected because it is well studied and portions of the process have been tested at pilot scales. Seven process variations were selected and examined in detail. Process designs were constrained to public data published in 2007 or earlier, without projecting for future process improvements. Economic analysis was performed for an 'nth plant' (mature technology) to obtain total investment and product value (PV). Sensitivity analysis was performed on PV to assess the impact of variations in process and economic parameters. Results show that the modeled dilute acid pretreatment process without any downstream process variation had the lowest PV of $3.40/gal of ethanol ($5.15/gallon of gasoline equivalent) in 2007 dollars. Sensitivity analysis shows that PV is most sensitive to feedstock and enzyme costs.

15. SUBJECT TERMS

cellulosic ethanol; biofuels; product value; fermentation; biochemical; feedstock; enzyme; process design; economic analysis; dilute acid pretreatment

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