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Techno-economic comparison of hot water and dilute acid pretreatment for biochemical production of ethanol from corn stover and evaluation of alternative scenarios to purchasing cellulase enzymes

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Techno-economic comparison of hot water and dilute acid pretreatment for biochemical production of ethanol from corn stover and evaluation of alternative scenarios to purchasing cellulase enzymes

by

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A thesis submitted to the graduate faculty
in partial fulfillment of the requirements for the degree of
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Abstract

Techno-economic analysis was conducted to compare hot water pretreatment and dilute acid pretreatment for biochemical production of ethanol from corn stover, and to compare several enzyme production schemes as alternatives to on-site enzyme production. Each of these scenarios was modeled in detail and economic analysis was performed to estimate the total capital investment (TCI) and Minimum Ethanol Selling Price (MESP), and Equivalent Purchased Enzyme Price for the enzyme production scenarios.

In Chapter 2 hot water and dilute acid pretreatment technologies are compared for both an nth plant design as well as a pioneer plant. Plants are modeled assuming they receive 2000 MT/day (metric tonne per day) of biomass. The dilute acid pretreatment process has the lowest MESP, which is estimated to be \$3.40/gal EtOH, compared to \$4.29/gal EtOH for the hot water pretreatment scenario. Sensitivity analysis shows that the MESP for the dilute acid pretreatment scenario is most sensitive to feedstock costs, enzyme loading, enzyme cost, and installed equipment costs. The MESP ranges from \$3.37 to \$3.93 under the assumed ranges for sensitivity parameters.

Cellulosic ethanol production has yet to be commercialized and a pioneer plant is expected to be significantly more expensive than an nth plant. To assess the impact of technological maturity on pioneer plant capital cost and plant performance a cost growth analysis was performed using a method developed by the RAND Corporation. Pioneer plant costs are estimated for three scenarios: optimistic, most probable and pessimistic. The estimated range of MESP for the pioneer plant were substantially larger than for the nth plant. The MESP for the model with dilute acid pretreatment were \$4.19, \$5.22 and \$6.68/Gal EtOH for the optimistic, most probable and pessimistic scenarios, respectively. The Total Capital Investment (TCI) for the three respective scenarios increased by 53%, 104%, and 183% above the nth plant TCI.

Enzymes are one of the most significant costs of cellulosic ethanol production. Chapter 3 analyzes two enzyme production schemes as alternatives to purchasing enzymes. The first is the production of enzymes on-site for a stand-alone plant. The competitiveness of on-site enzyme production with purchasing enzymes is compared among plant scales varying from 500 MT/day to 3000 MT/day. The second scheme is the production of both ethanol and excess enzymes at a central plant for export to satellite plants producing only ethanol. Two cases were examined for this scheme—one in which the central plant supplies enzymes to two satellite plants and another with four satellite plants. Both the central plant and satellite plants in this scheme receive 2000 MT/day of corn stover.

For the on-site enzyme production scheme it was found that the competitiveness of on-site enzyme production with purchasing enzymes varies significantly with plant scale. At the 500 MT/day scale on-site enzyme production is economically advantageous at enzyme prices above \$1.07/gal EtOH (\$7.82/kg protein). The competitive price drops to \$0.66/gal EtOH (\$4.81/kg protein) for the 3000 MT/day plant scale. The MESPs for the 500 MT/day and 3000 MT/day are \$4.70 and \$3.24/gal EtOH, respectively.

The results from the central enzyme production scheme are similar to those of the 2500 MT/day on-site enzyme plant. This scheme is more economically advantageous than purchasing enzymes at enzyme prices higher than \$0.71 and \$0.69/gal EtOH (\$5.15 and \$5.02/kg protein) for the cases with two and four satellite plants, respectively. The central enzyme production scheme with both two and four satellite plants has lower MESPs than a stand-alone 2000 MT/day plant with on-site enzyme production. This is due primarily to the economy of scale that exists with the enzyme production equipment, resulting in a lower capital cost per gallon of ethanol production capacity.

Chapter 1 Importance of this Study

Due to concerns in a number of arenas, including national security, environmental, and economic, there has been a recent surge of interest in renewable liquid fuels. Additionally, biomass represents the only indefinitely available feedstock for producing organic chemicals and fuels (1). However, the production of fuels from food crops may place upward pressure on the price and availability of food today, and the likelihood of such a food vs. fuel conflict may increase as the world's population grows. Thus, it is imperative that technologies are developed for the production of transportation fuels and chemicals from lignocellulosic sources. The resource base of renewable feedstocks for fuel production could potentially provide a significant portion of US fuel consumption. A study conducted by the USDA and DOE indicates that with relatively small land use changes it may be feasible to derive 30% of the transportation fuel needs of the US, at current consumption levels, from biomass. In this scenario the vast majority of feedstock is lignocellulosic material (2).

The number of ligno-cellulosic biomass-to-fuel conversion process variations discussed in peer reviewed literature is extensive, and the degree of technological maturity among the conversion processes is wide ranging. The Energy Independence and Security Act of 2007 mandates that 36 billion gallons of renewable fuels be produced by 2022, while the 2008 production was only 9 billion gallons (3). With the timeframe of these biofuels mandates it is important to identify those processes which are likely to prove feasible in the relative near term. The scope of near term feasibility encompasses a number of areas including the degree of technological maturity, compatibility with current fuel infrastructure, and production cost.

Two techno-economic studies of biochemical ethanol production are discussed in this document. The objective of these studies is to compare several biochemical ethanol production process scenarios that are compatible with a 5-8 year timeframe for plant construction, on the basis of production cost and capital risk associated with a pioneer plant. Because of the short-term timeframe, the data used in modeling is based on the current state of technology.

In Chapter 2 dilute-acid pretreatment and hot water pretreatment are compared. Chapter 3 examines two enzyme production schemes as alternatives to purchasing enzymes from a supplier. The first scheme is the production of cellulase enzymes on-site, rather than purchasing enzymes. The relationship between plant scale and the economic competitiveness of on-site enzyme production with purchasing enzymes is examined. For the second scheme, the production of both ethanol and excess enzymes at a central plant for export to satellite plants is proposed and it is also compared with

purchasing enzymes. These processes were modeled using Aspen Plus, and discounted cash flow analysis was performed to estimate the cost of ethanol production.

Chapter 2 Techno-economic comparison of dilute acid pretreatment and hot water pretreatment

Introduction

Ligno-cellulosic biomass is primarily composed of cellulose (30-60 wt%), hemicellulose (20-40 wt%), and lignin (10-25 wt%) (4) (5). In order to produce fermentable sugars, the cellulose and hemicellulose must be hydrolyzed to monomers. Cellulose is a linear polymer of β -glucose that forms a rigid structure which is difficult to break because of the orientation of the linkages (5). Hemicellulose is a highly branched polymer of various five-carbon and six-carbon sugars of which xylose is typically the main component. Other hemicellulose sugars include mannose, galactose, and arabinose. Lignin is comprised of polyphenolic compounds which act as a binder between plant structures by surrounding plant cell walls (5). The complex matrix formed by lignin and hemicellulose is largely responsible for the recalcitrance of lignocellulosic biomass (6).

Because of the recalcitrant nature of ligno-cellulosic biomass, a pretreatment step is necessary to increase the exposure of cellulose during enzymatic hydrolysis. Experimental data indicates very low sugar yields during enzymatic hydrolysis from biomass that has not undergone pretreatment, thus, making such a process economically unviable (7) (6). Pretreatment methods are chemical or physical treatments, or a combination of both. Methods include acid catalyzed treatments such as dilute sulfuric acid pretreatment and hot water pretreatment (autohydrolysis), alkali treatments such as ammonia recycle percolation (ARP), and physico-chemical treatments such as ammonia fiber explosion (AFEX) and steam explosion.

This study is a techno-economic comparison of dilute acid pretreatment and hot water pretreatment. Dilute-acid pretreatment uses sulfuric acid at low concentrations (< 3 wt%) in aqueous solution with the biomass slurry at moderate temperatures (130-200°C) and relatively short residence times (1-30 min) (4) (8). During treatment the lignin-hemicellulose matrix is disrupted and a significant portion of the hemicellulose is hydrolyzed. A small portion of the cellulose may also be hydrolyzed to glucose and glucose oligomers (9). Because of the severity of the treatment, degradation products such as hydroxy-methylfurfural (HMF) and furfural may be produced from sugars.

Hot water pretreatment relies on the decreased pH of water at elevated temperatures and the increasing acidity due to acetic acid formation from acetate in the biomass to disrupt the structure of

the lignocellulosic material (10). The biomass slurry is typically treated at temperatures of 160-230°C with a residence time of around 10-60 minutes (4) (11).

While pretreatment is necessary to achieve high sugar yields, it has been estimated in previous techno-economic studies to represent a significant portion the total capital investment in cellulosic ethanol plants, and the capital costs vary considerably between different pretreatment methods (5) (12). A list of pretreatment capital cost estimates as a percentage of the total installed equipment cost from previous techno-economic studies is presented in Table 2.1.

Table 2.1 Pretreatment capital costs as a percentage of total installed cost from previous techno-economic studies

Reference	Pretreatment	Pretreatment Capital Cost, % of Total Installed Cost
Hamelink et al., 2005	Dilute Acid	15.5%
Aden et al., 2002	Dilute Acid	23.6%
Wooley et al., 1999	Dilute Acid	19.5%
Wingren et al., 2003	SO ₂ Steam	16-20%

In addition, each pretreatment technology has a range of impacts on the biomass, including the hemicellulose sugar yields, availability of cellulose to enzymatic saccharification, and yields of sugar degradation products, all of which have an impact on the ethanol yield, and therefore production cost. Because of the relatively high capital cost of pretreatment equipment and its significant effect on downstream processes, it is imperative that pretreatment technologies are compared on an economic basis in order to identify those technologies offering the greatest production-scale viability. Such analysis can provide valuable insight in two critical areas: (1) identifying the most economically promising pretreatment technologies, and (2) recognizing the process areas that offer the most potential for improving the economics, thus, effectively directing research focus. In this chapter dilute-acid pretreatment and hot water pretreatment are compared on the basis of ethanol production cost for an nth plant and the capital risk associated with a pioneer plant. The process parameters and conversions were chosen with the intent of modeling a plant constructed in a 5-8 year timeframe.

Methods

Two lignocellulose-to-ethanol process models were developed using Aspen Plus, one including dilute acid pretreatment and the other with hot water pretreatment. The process models were originally developed by the National Renewable Energy Laboratory (NREL) and were modified for this study. The pretreatment and enzymatic saccharification process parameters and reaction

conversions used are from lab scale experiments published by Wyman et al. (13) as part of the Biomass Refining Consortium for Applied Fundamentals and Innovation (CAFI). A third model was developed which uses pretreatment reaction conversions from experiments conducted at NREL in a 1 MT/day pilot scale dilute acid pretreatment reactor. This model is used for comparison to the dilute acid pretreatment model using lab scale conversions in order to understand how scaling up the process may affect the economics. However, comparing pretreatment technologies based on data from different scale experiments may lead to misleading results. Thus, the following discussion in the Nth Plant Analysis section comparing dilute acid and hot water pretreatment technologies focuses on the two models using lab scale data.

The properties for components not included in the Aspen Plus database were obtained from a custom property database developed by NREL. Heat, work, and material stream flows, as well as process conditions from Aspen Plus simulations are imported into Excel for use in the economic analysis. Discounted cash flow analysis is conducted to calculate the Minimum Ethanol Selling Price (MESp), which is the selling price (\$/gal) of ethanol that results in a project net present value of zero. Risk analysis is conducted to estimate the MESPs and capital cost for pioneer plants. A linear model for estimating capital cost increases and reduced performance for a plant incorporating new technology was developed by the RAND Corporation and is used in the risk analysis.

Process Model

The process models for both pretreatment scenarios consist of nine process areas. All process areas other than pretreatment are modeled identically for both scenarios. Both plants are scaled to receive 2000 MT/day of corn stover. The NRTL (Non-random, two liquids) property method is used for modeling. The NRTL property method uses binary interaction coefficients for chemical components to estimate vapor-liquid equilibria. Exceptions to the NRTL property method for specific unit operations are noted below. Each process area is described in detail in the following sections. Process flow diagrams for the dilute acid pretreatment scenario are located in Appendix 2.A and flow diagrams for the hot water pretreatment scenario are shown in Appendix 2.B.

Area 100—Feedstock Handling

The feedstock handling area (Area 100) receives the corn stover in bale form. The moisture content is 25%. Table 2.2 shows the feedstock composition.

Table 2.2 Corn stover composition (14)

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

Some on-site storage is necessary, however, storage area costs and inventory requirements are not considered in this analysis. The bales are unloaded by forklift from trucks and conveyed to the bale unwrapping system. In previous work at NREL it was estimated that eight forklifts are needed per shift for the 2000 MT/day plant scale (12). The unwrapped biomass is washed with water in the wash table to remove dirt before passing through a magnet to remove tramp metal, and finally conveyed to the shredders for size reduction. The dirty wash water is pumped to a clarifier to which a polymer is added to enhance solids separation. The polymer is included in operating costs. The solids from the clarifier are dewatered in a belt press and the water is recycled. After subtracting the water loss to the atmosphere during washing and the water remaining in the solids after separation, 45% of the wash water is recycled. An equipment list for Area 100 is located in Appendix 2.E and a process diagram can be seen in Appendix 2.A. The original cost quotes for the forklifts, wash tables, shredders, water clarifier, and belt press were obtained from Harris Group, Inc. and the bale unwrapping system was quoted by Cross Wrap (12). All equipment in Area 100 has been used in commercial operation except the bale unwrapping system.

The power requirements for Area 100 process equipment were estimated in previous NREL studies and are included in the utilities requirement for the process model.

Area 200—Dilute Acid Pretreatment

During dilute acid pretreatment most hemicellulose is hydrolyzed to sugar monomers and oligomers, and a small fraction of cellulose is hydrolyzed to glucon and glucose. The resulting hydrolyzate is much more accessible to subsequent enzymatic hydrolysis.

The biomass from Area 100 is pumped to a presteamer, where steam at 163°C and 4.4 atm is added to reach a temperature of around 100°C. The presteamer allows for 31% of the pretreatment heat load to be met by low pressure steam. The biomass residence time in the presteamer is 20 minutes, after which the biomass is fed via screw conveyor to the blow tank, which acts as a seal to prevent backflow of biomass from the pretreatment reactor to the presteamer. A screw conveyor

transfers the biomass from the blow tank to the reactor where additional steam at 268°C and 13 atm is added. Concentrated H₂SO₄ is added to the reactor along with evaporator condensate from Area 500 to dilute the H₂SO₄ concentration to 1.9 wt%. The enthalpy of reaction between water and H₂SO₄ is not included in the energy balance. Heat loss from the reactor and presteamer are also neglected. A summary of pretreatment reaction conditions are shown in Table 2.3.

Table 2.3 Dilute acid pretreatment parameters

Parameters	
Acid Concentration (wt%)	1.9
Acid Loading (g acid/g dry biomass)	0.0443
Total Solids (wt%)	0.2959
Temperature (°C)	190
Pressure (atm)	11.4
Retention Time (min)	2

Three continuous flow presteamer/blow-tank/reactor units operate in parallel. Cost estimates for the units were obtained from Anco-Eaglin, Inc. A list of reactions occurring in the pretreatment reactor is shown in Table 2.4. The conversions for mannan, galactan, and arabinan to monomers and oligomers are the same as those for xylan.

Table 2.4 Pretreatment reaction and conversions

Reaction	Fractional Conversions – Lab Scale	Fractional Conversions – Pilot Scale
H₂O + Cellulose → Glucose	6.26%	9.9%
Cellulose → Glucose Oligomers	--	0.3%
Cellulose → HMF + 2 H₂O	--	0.3%
Xylan → Xylose Oligomers	2.65%	21%
H₂O + Xylan → Xylose	82.49%	60%
Xylan → Furfural + 2 H₂O	--	11%
Acetate → Acetic Acid	100%	100%
Lignin → Soluble Lignin	10.0%	10%

The pretreated biomass slurry is flashed to 1 atm in a blowdown tank. Approximately 6.8% of the acetic acid formed during pretreatment is removed in the overhead vapor of the blowdown tank. The NRTL-HOC (Hayden and O’Connell equation of state) is the property method used for the blowdown tank. This method provides a more accurate model for acetic acid vaporization. The overhead vapor is used to preheat the feed stream to the beer column in Area 500.

The pretreated biomass slurry from the blowdown tank bottoms is separated into solid and liquid fractions in a Pnuemapress pressure filter which was sized for previous NREL studies by The Harris Group. Compressed air (9.5 atm) is forced through the cake along with water. Recycled water from the evaporators and distillation column in Area 500 is used to rinse the filter cake. The liquid fraction contains solubilized components as well as the H_2SO_4 which must be neutralized to increase the pH to levels that will not hinder downstream biological processes. The liquid fraction is pumped to the overliming tank where lime is added to raise the pH to 10. The residence time in the overliming tank is one hour, after which the hydrolyzate is pumped to the reacidification tank. Additional H_2SO_4 is added to the slurry to adjust the pH to 4.5, which is appropriate for fermentation. The residence time in the reacidification tank is four hours. Gypsum crystals are formed and subsequently filtered with a hydrocylcone followed by a rotary drum. The hydrolyzate liquids are combined with pretreated solids and the slurry is pumped to Area 300 for enzymatic saccharification and fermentation.

Area 200—Hot Water Pretreatment

Biomass from Area 100 is conveyed to the mix tank where recycled water from Area 500 is added to produce a slurry with a solids concentration of 17.2 wt%. The slurry is pumped to a pressure of 13 atm and passed through a heat exchanger with the hot side fluid being the slurry exiting the pretreatment reactor. The slurry is heated to 162.5°C in the heat exchanger before passing through a second trim heater which increases the slurry temperature to 190°C. The heat source for the trim heater is steam at 268°C and 13 atm. The slurry enters the plug flow pretreatment reactor where the residence time is 15 min. A list of reactions and conversions used in the process model is shown in Table 2.5.

Table 2.5 Hot water pretreatment reactions and conversions

Reaction	Fractional Conversion
Cellulose → Glucose Oligomers	5.3%
H₂O + Cellulose → Glucose	0.32%
Xylan → Xylose Oligomers	55.4%
H₂O + Xylan → Xylose	2.39%
Acetate → Acetic Acid	100%
Lignin → Soluble Lignin	5%

Other hemicelluloses including mannose, arabinose and galactose undergo the same reactions and have the same conversions as xylose. As previously noted, the slurry exiting the pretreatment reactor

passes through a heat exchanger to preheat the untreated slurry. The pretreated slurry also passes through a second heat exchanger to preheat the beer entering the beer column in Area 500, where the slurry temperature is reduced from 77.7°C to 65°C. The pretreated slurry is then flashed to 1 atm and ammonia is added to neutralize the acetic acid formed during pretreatment.

Area 300—Enzymatic Saccharification and Fermentation

The biomass hydrolyzate from Area 200 is pumped to a saccharification vessel where cellulase and hemicellulase enzymes are added. There are 20 saccharification vessels for the hot water pretreatment scenario and 15 for the dilute acid pretreatment scenario, each of which is 1.0 MMGal. The vessels are operated in batches with a residence time of five days and sequenced to provide continuous flow to the rest of the plant. The number of vessels is calculated from the volume of biomass from pretreatment. Less hemicellulose and cellulose are hydrolyzed during hot water pretreatment, and thus the volume of the slurry is larger than that of dilute acid pretreatment because of higher insoluble solids content. Before entering the saccharification vessels, the slurry passes through a heat exchanger where it is cooled to 32°C, which is the saccharification temperature. Cooling is provided to the vessels to maintain a constant temperature. Enzymes are added to the vessels at a rate of 18.8 FPU/g cellulose in the untreated biomass. This means that the enzyme use is equal for both pretreatment scenarios because the enzyme feed is based on cellulose content prior to pretreatment. Enzymatic saccharification conversions are different between the two pretreatment models because the extent of the structural disruption—and therefore accessibility by enzymes—is different for each pretreatment technology. Saccharification reactions and conversions are shown in Table 2.6.

Table 2.6 Saccharification reactions and conversions.

Reaction	Fractional Conversion- Dilute acid pretreatment scenario	Fractional Conversion- Hot water pretreatment scenario
H₂O + Cellulose → Glucose	91.09%	89.97%
H₂O + Xylan → Xylose	57.13%	56.61%

Approximately 10% of the hydrolyzate leaving the saccharification vessel is sent to the fermentation seed vessel train to be used as a carbon source for growth of *Z. mobilis*, along with corn steep liquor (CSL) and diammonium phosphate as nutrients. Two trains of fermentation seed vessels are used. Each train contains five consecutive seed vessels, with the first being 20 gallons, and each subsequent vessel being scaled up by a factor of 10. The initial inoculum is 10% of the total needed

for fermentation. The total seed production time of a single train from the first to fifth vessel is 180 hours. The seed vessels are cooled to maintain the temperature at 41°C. The reactions used to model cell production of *Z. mobilis* are shown in Table 2.7.

Table 2.7 *Z. mobilis* seed production vessel reactions.

Reaction	Fractional Conversion
Glucose → 2 EtOH + 2 CO₂	90%
Glucose + 0.04696 CSL + 0.018 (NH₄)₂HPO₄ → 6 <i>Z. mobilis</i> + 2.4 H₂O	4%
Glucose + 2 H₂O → 2 Glycerol + O₂	0.4%
Glucose + 2 CO₂ → 2 Succinic Acid + O₂	0.6%
Glucose → 3 Acetic Acid	1.50%
Glucose → 2 Lactic Acid	0.20%
3 Xylose → 5 EtOH + 5 CO₂	80%
Xylose + 0.03913 CSL + 0.015 (NH₄)₂HPO₄ → 5 <i>Z. mobilis</i> + 2 H₂O	4%
3 Xylose + 5 H₂O → 5 Glycerol + 2.5 O₂	0.30%
Xylose + H₂O → Xylitol + 0.5 O₂	4.60%
3 Xylose + 5 CO₂ → 5 Succinic Acid + 2.5 O₂	0.90%
2 Xylose → 5 Acetic Acid	1.40%
3 Xylose → 5 Lactic Acid	0.20%

The saccharified slurry is pumped to a fermentation vessel. There are 8 fermentation vessels in the hot water pretreatment scenario and 6 in the dilute acid pretreatment scenario. As with the saccharification vessels, the difference in number between the two scenarios is due to volume differences of the biomass slurries. The vessel volume is 1.0 MMGal. They are operated in batches and the residence time is 2 days. The temperature is maintained at 32°C during fermentation. The reactions used to model fermentation are shown in Table 2.8.

Table 2.8 Reactions used to model fermentation and reaction conversions.

Reaction	Fractional Conversion
Glucose \rightarrow 2 EtOH + 2 CO₂	95%
Glucose + 0.04696 CSL + 0.018 (NH₄)₂HPO₄ \rightarrow 6 <i>Z. mobilis</i> + 2.4 H₂O	2%
Glucose + 2 H₂O \rightarrow 2 Glycerol + O₂	0.4%
Glucose + 2 CO₂ \rightarrow 2 Succinic Acid + O₂	0.6%
Glucose \rightarrow 3 Acetic Acid	1.50%
Glucose \rightarrow 2 Lactic Acid	0.20%
3 Xylose \rightarrow 5 EtOH + 5 CO₂	75.6%
Xylose + 0.03913 CSL + 0.015 (NH₄)₂HPO₄ \rightarrow 5 <i>Z. mobilis</i> + 2 H₂O	1.9%
3 Xylose + 5 H₂O \rightarrow 5 Glycerol + 2.5 O₂	0.30%
Xylose + H₂O \rightarrow Xylitol + 0.5 O₂	4.60%
3 Xylose + 5 CO₂ \rightarrow 5 Succinic Acid + 2.5 O₂	0.90%
2 Xylose \rightarrow 5 Acetic Acid	1.40%
3 Xylose \rightarrow 5 Lactic Acid	0.20%

Due to biological contamination, it is assumed that 7% of monosaccharides are lost because of conversion to lactic acid. These losses are modeled in a dummy reactor in parallel with the fermentation vessels, where 7% of the stream is split and flows to the dummy reactor. The reactions are shown in Table 2.9.

Table 2.9 Reactions due to biological contamination.

Reaction	Fractional Conversion
Glucose \rightarrow 2 Lactic Acid	100%
3 Xylose \rightarrow 5 Lactic Acid	100%
3 Arabinose \rightarrow 5 Lactic Acid	100%
Galactose \rightarrow 2 Lactic Acid	100%
Mannose \rightarrow 2 Lactic Acid	100%

Vent gas from the fermentation vessels is piped to the beer column for ethanol recovery. The fermented beer is pumped to Area 500 for ethanol distillation and solids separation.

Area 500—Ethanol Distillation, Solids Separation, and Water Recovery

As previously noted, the beer is preheated to 95°C by the biomass slurry exiting the pretreatment reactor in Area 200. It is heated further to 100°C by the bottoms stream of the beer column before entering the beer column. The beer column removes most of the CO₂ in the overhead stream. Ethanol is removed in a side stream as vapor which is 26 wt% ethanol. The bottoms contain over 90% of the water entering the beer column, as well as unfermented sugars and insoluble solids. Steam at 4.4 atm and 163°C (medium pressure steam) provides heat to the reboiler, which operates at 124°C.

The overhead stream, which contains approximately 11% ethanol, is sent to a water scrubber along with the vent gas from the ethanol fermentors and fermentation seed vessels. The scrubber recovers 99% of the ethanol and the bottoms stream is recycled to the beer column feed.

The side stream is sent to a second distillation column where the ethanol/water separation proceeds so that the overhead is near the azeotropic ratio. The bottoms stream is almost entirely water and is recycled for use as process water. Medium pressure steam provides heat to the second distillation column.

The near-azeotropic mixture is further purified in a molecular sieve adsorption column to produce neat ethanol. The overhead is superheated with medium pressure steam before entering the adsorption column. Water is adsorbed resulting in 99.5% pure ethanol. There are two adsorption columns, with one always in operation while the other regenerates. Neat ethanol vapor is passed through the column under vacuum conditions during regeneration, removing adsorbed water.

Stillage from the beer column is pumped to the 1st effect evaporator where the water concentration is reduced from 85wt% to 77wt%. Heat is provided by condensing the reflux vapor from the second distillation column, as well as from steam at 1.7 atm and 115°C. The evaporator operates at 0.57 atm and 87°C. The overhead water vapor from the evaporator is condensed and used to provide heat to the 2nd effect evaporator which operates at a lower temperature.

The bottoms stream from the evaporator is separated into liquid and solid fractions in a Pneumapress filter. The solid fraction is comprised of mostly lignin and unhydrolyzed cellulose and hemicellulose. Compressed air is fed to the pressure filter to aid in liquid removal. After leaving the Pneumapress the solids fraction, containing 45% water, is used as combustor fuel. The liquid fraction is pumped to the 2nd effect evaporator where additional water is removed. The 2nd effect evaporator operates at 0.32 atm and 74°C. The overhead vapor is condensed and provides heat for the 3rd effect evaporator which operates at 67°C and 0.21 atm. The syrup leaving the 3rd effect evaporator is

about 60 wt% water and contains unfermented sugars and other soluble compounds. It is used as combustor fuel.

The overhead vapor from the three evaporators contains a small amount of acetic acid. Therefore, it cannot be entirely recycled or acetic acid buildup would occur in the process, which can inhibit fermentation if present in large enough quantity. Ten percent of the condensed vapor is pumped to the waste water treatment area and the remainder is recycle as process water.

Area 600—Wastewater Treatment

Four streams are received for treatment in Area 600. These include the boiler blowdown, evaporator overhead, pretreatment blow-tank overhead, Pneumapress vent from the pretreatment hydrolyzate solid/liquid separation, cooling tower blowdown, and waste from seed production/fermentation vessel cleaning. Waste water flow to Area 600 is 85,600 kg/hr for the dilute acid pretreatment scenario and 52,400 kg/hr for the hot water pretreatment scenario. A two step treatment is used for waste water treatment (Area 600). The first is an anaerobic treatment with a small amount of urea, phosphoric acid, and micronutrients being added as nutrients for the anaerobic organisms. The anaerobic model converts 90% of organic components to methane and CO₂ at a ratio of 3 moles of methane per mole of CO₂. Biogas from the anaerobic digester is used in the combustor. The waste water is secondly 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 combusted in the fluidized bed combustor. The water from the clarifying tank is recycled as process water.

Merrick Engineering designed the wastewater treatment system under contract with NREL.

Area 700—Ethanol and Chemical Storage

Ethanol product and gasoline used as a denaturant are stored in Area 700. Process chemicals are also stored which include CSL, sulfuric acid (dilute acid pretreatment scenario only), diammonium phosphate, enzymes, fire suppression water, and propane.

Both ethanol and gasoline storage capacities are enough for seven days of production. Ethanol is stored in two tanks and a single tank is used to gasoline. Sulfuric acid is storage capacity is five days. The storage tank is stainless steel because of the corrosiveness of H₂SO₄. CSL has five days of storage capacity. CSL and diammonium phosphate are mixed in Area 700 and are pumped to Area 300 for seed production and fermentation nutrients. Diammonium phosphate is received as a solid and storage volume is sufficient for seven days. Enzyme broth is stored in two tanks with a capacity of 4 days. Use of two tanks allows lots to remain separate.

The ethanol storage, sulfuric acid storage, and fire suppression system were sized by Delta-T Corp. for previous NREL studies.

Area 800—Combined Heat and Power Generation

Fuel streams fed to the boiler include evaporator syrup, insoluble solids from the Pnuemapress, waste water treatment sludge, and biogas. The moisture content in the combined fuels is 54% for the dilute acid pretreatment scenario and 53% for the hot water pretreatment scenario. The fuel streams are combusted in a circulating fluidized bed. Combustion air is preheated by the flue gas before entering the fluidized bed.

Boiler water is superheated to 510°C and 86 atm. The superheated steam enters the multi-stage turbogenerator and exits at the three conditions (268°C, 13 atm; 164°C, 4.4 atm; 115°C, 1.7 atm) needed in the plant. The remaining steam exits the turbine at 46°C and 0.1 atm and is condensed with cooling water, pressurized and returned to the boiler. The generator produces more electricity than is needed in the plant, and it is assumed that excess electricity is sold through the grid. The circulating fluidized bed combustor was designed by Radian Corporation and the turbogenerator design and specifications were supplied by ABB Power Generation Systems.

The boiler feed water system includes a deaerator system. Hydrazine is added to the deaerator to aid in oxygen removal. Ammonia is added to the water for pH control and phosphate is added to reduce scaling. Boiler chemical costs are included in operating costs in the discounted cash flow analysis. The boiler is modeled so that 3% of the boiler water is sent to waste water treatment as boiler blowdown. The boiler feed water system was designed by Badger Engineering Inc.

A baghouse system to remove flue gas particulate was included in the cost estimation, which was provided by Hamon Research-Cottrell. Combustion was modeled so that all of the sulfur in the boiler fuel is converted to SO₂ and 1% of SO₂ is converted to H₂SO₄. The flue gas is used to preheat the incoming combustion air. However, to prevent corrosion it is important that the flue gas temperature remain higher than the H₂SO₄ dew point. The model includes a 35°C safety factor for the flue gas temperature.

Area 900—Utilities

The utilities area includes the cooling water system, plant air compressors, seed production and fermentation vessel cleaning system, and the process water system.

The cooling water system provides cooling water at 28°C. Make-up well water at 13°C is used to cool the seed production and fermentation vessels before being combined with the process water.

Windage loss in the cooling tower is assumed to be 0.1% of the total flow to the tower. Blowdown is 10% of the windage plus evaporative losses.

Compressed air at 9.5 atm is supplied to the Pneumapress pressure filter used to separate the solid and liquid fractions of the pretreatment hydrolyzate in the dilute acid pretreatment scenario, and is supplied to the Pneumapress used to separate the post-fermentation solid and liquid fractions in both pretreatment scenarios. Plant instrument air and miscellaneous requirements are assumed to be 1530 m³/hr, which was estimated by Delta-T Corp. for a previous NREL study. The compressor system cost was obtained from ICARUS Process Evaluator for screw compressors.

Treated wastewater is recycled as process water and is held in the process water storage tank. Well water is used to make up for losses. As previously noted, well water is first used to cool the fermentation and seed production vessels before entering the process water tank. Make-up water is needed at a rate of 228,800 kg/hr for the hot water pretreatment scenario and 211,900 kg/hr for the dilute acid pretreatment scenario.

The seed production and fermentation vessel cleaning system uses process water which is heated to 121°C by with process steam. The heat requirement is negligible and therefore the steam demand is not included in the process model. Chemicals used in sterilization and cleaning are not included in operating costs. The cleaning water supply system was designed and quoted for a flow rate of 454,000 kg/hr. However, cleaning water is used intermittently and the steady state flow rate used in the process model is only 63 kg/hr. The cleaning system was designed by Delta-T Corp. who also provided the cost quote for previous NREL studies.

Discounted Cash Flow Analysis

Stream flow results from the Aspen Plus process model are imported to Excel, which is used for the discounted cash flow analysis. Key economic assumptions include:

- Equipment, chemical and labor costs are indexed to 2007 dollars
- The process equipment and steam generation plant depreciate in 7 and 20 years, respectively, following the MACRS method
- The project is 100% equity financed
- Capital investment is spread over 3 years at a rate of 8, 60 and 32% in the first, second and third years, respectively
- Working capital is 15% of Fixed Capital Investment
- 20 year project life
- 10% return on investment

Individual equipment costs are estimated using original quotes from previous NREL studies as a basis. Equipment costs are scaled according to the scaling ratio, calculated by dividing the new equipment scaling attribute by the attribute of the original quoted equipment. The scaling ratio attributes include the major heat, work, or material stream associated with each piece of equipment. The appropriate scaling exponent is applied to individual equipment to account for economy of scale. Equation 1.1 is used to cost equipment based on the original NREL quote.

$$\text{Equation 1.1} \quad C = C_o \left(\frac{S}{S_o} \right)^n$$

Where C_o is the original price quote, S is the new value of the scaling attribute, S_o is the value of the original scaling attribute, and n is the scaling exponent. Appropriate installation factors are then applied, which are also obtained from previous NREL studies. The Chemical Engineering Plant Cost Index (15) is used to inflate the price in the quoted year to the 2007 price. Equipment lists and prices for the two scenarios are shown in Appendix 2.F.

The warehouse cost is estimated as 1.5% of total installed equipment cost and site development as 9% of ISBL cost as estimated by NREL. Indirect capital costs include Engineering and Supervision, Construction Expenses, and Legal and Contractor's Fees, and are estimated as 32%, 34%, and 23% of total FOB equipment cost, respectively. These cost factors follow those outlined by Peters and Timmerhaus (15) for a plant handling both solids and liquids. Contingency was estimated as 20% of total direct and indirect capital costs. Total capital investment (TCI) is the sum of contingency and direct and indirect costs.

Raw material prices are inflated to 2007 dollars from the previous NREL price list using the Industrial Inorganic Chemical Index (17), and annual costs are estimated from material stream flows. Raw material prices are shown in Table 2.10. Fixed operating costs included labor, overhead, insurance, and maintenance. The Labor Index was used to adjust the labor cost from previous NREL labor estimates to 2007 dollars.

Table 2.10 Raw material costs.

Feedstock/Chemical	Price	Feedstock/Chemical	Price
Corn Stover (US\$/MT)	82.5	Propane (US\$/MT)	339
Enzyme broth (US\$/MT)*	506	Boiler Chemicals (US\$/MT)	4986
Sulfuric Acid (US\$/MT)	35	Cooling Tower Chemicals (US\$/MT)	2988
Hydrated Lime (US\$/MT)	99	Wastewater Chemicals (US\$/MT)	462
Corn Steep Liquor (CSL) (US\$/MT)	226	Wastewater Polymer (US\$/MT)	7470
Diammonium Phosphate (US\$/MT)	200	Clarifier Polymer (US\$/ST)	3567
Electricity Price (¢/kWh)	5.4		

*Broth contains 100 g enzyme protein/L. It is assumed that specific activity is 600 FPU/g protein.

The Minimum Ethanol Selling Price (MESP) is calculated by iterating the selling price of ethanol to achieve a Net Present Value of \$0.

Cost Growth Analysis

A higher level of risk is associated with a first-of-a-kind plant, both in terms of capital cost overruns and reduced plant performance because of unexpected shutdowns. A method for estimating increased capital cost and reduced plant performance was developed by the RAND Corp. for plants which employ new technology. RAND Corp. developed linear regression models based on data collected from 44 chemical and mineral processing plants. Those regression models were used in the cost growth analysis in this study to estimate the possible impact of both the unexpected reduced plant performance of the pioneer plant, and the capital cost growth associated with the pioneer plant.

The Plant Performance factor according to the method developed by the RAND Corp. is defined as the online factor of a plant during the second six month period after startup. Equation 1.2 estimates pioneer Plant Performance as a percent of design capacity in the second six months after startup:

$$\text{Equation 1.2} \quad \text{Plant Performance} = 85.77 - 9.69 \times \text{NEWSTEPS} + 0.33 \times \text{BALEQS} - 4.12 \times \text{WASTE} - 17.91 \times \text{SOLIDS}$$

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. However, some weight is given to rigorous theoretical models.

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

SOLIDS: If the process handles solids a value of 1 is given, otherwise it is 0.

The Cost Growth is defined as the percentage cost increase over the original capital cost estimate. The Cost Growth correlation is shown in Equation 1.3.

Equation 1.3 ***Cost Growth = 1.1219 – 0.00297 × PCTNEW – 0.02125 × IMPURITIES – 0.01137 × COMPLEXITY + 0.00111 × INCLUSIVENESS – C₁ × PROJECT DEFINITION***

Where,

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

IMPURITIES: 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. Zero for no impurity buildup or corrosion issues.

COMPLEXITY: Number of continuously linked process steps.

INCLUSIVENESS: Percentage of three factors— pre-startup personnel costs, pre-startup inventory cost, and land purchase—that were included in the analysis. For example, if 2 of these factors have been rigorously considered, the variable would be given a value of 67%.

C₁: 0.06351 if the design is at pre-development/exploratory or R&D stage, and 0.04011 if the design is in commercial or pre-commercial stage.

PROJECT DEFINITION: Takes into account both the level of site-specific information included in the analysis and the level of engineering completed at the time of the original estimate. The value for the level of engineering completed at the time of the estimate is given as follows: (1) engineering completed, (2) moderate or extensive engineering, (3) limited engineering, or (4) screening design stage. The level of site-specific information in four areas, including on-site and off-site unit configurations, soils and hydrology data, health and safety requirements, and environmental requirements, is the second part of PROJECT DEFINITION. The work completed for each of those four areas

is rated based on the following scale: (1) definitive or completed work, (2) preliminary or limited work, (3) assumed or implicit analysis, or (4) not used in the cost estimate. The average of the values given to those four areas is then added to the value given for the level of engineering completed at the time of the original cost estimate, so that the value for PROJECT DEFINITION may vary from 2 to 8. Note that a higher value represents less definition.

In the cash flow spreadsheet the Total Capital Investment (TCI) of the base case n^{th} plant is divided by the percentage Cost Growth (Equation 1.3) to estimate the TCI of the pioneer plant. The first year ethanol sales, variable operating costs, and electricity export of the n^{th} plant is multiplied by the percentage of Plant Performance (Equation 1.2) to account for the reduced production of a pioneer plant. The contingency factor is increased to 30% for the cost growth analysis, up from 20% for the n^{th} plant. For the discounted cash flow analysis the plant performance is increased by 20% per year until nameplate capacity is reached.

Three cost growth scenarios are considered in the analysis—most probable, optimistic and pessimistic—representing the range of estimates for variables used in Equations 1.2 and 1.3. The variable values are shown in Table 2.11 and the selection justification is discussed below.

Table 2.11 Plant Performance and Cost Growth variables for the dilute acid pretreatment scenario.

Plant Performance (Equation-1)				Cost Growth (Equation-2)			
Variables	Values			Variables	Values		
	Optimistic	Most Probable	Pessimistic		Optimistic	Most Probable	Pessimistic
NEWSTEPS	2	4	5	PCTNEW	2	4	5
BALEQS	20	0	0	IMPURITIES	1	3	5
WASTE	1	2.5	3	COMPLEXITY	5	5	5
SOLIDS	1	1	1	INCLUSIVENESS	50	33	0
				PROJECT DEFINITION	5	6	7
<i>Plant Performance (%)</i>	<i>51.00</i>	<i>19.00</i>	<i>7.00</i>	<i>Cost Growth (%)</i>	<i>75.00</i>	<i>53.00</i>	<i>33.00</i>

For the most probable scenario, the new process steps are considered to be feedstock handling, pretreatment, saccharification, and the fluidized bed combustor. For the optimistic scenario the assumed new steps are pretreatment and saccharification, and for the pessimistic case they are feedstock handling, pretreatment, saccharification, the beer column, and the fluidized bed combustor. None of the mass and energy balances can be verified with commercial production data, so the most probable and pessimistic cases are given a zero for BALEQS. However, some level of rigor is provided by the Aspen Plus process simulator. Therefore, BALEQS is assigned a value of 20% for the optimistic case. The process treats wastewater in Area 600, all of which is recycled to the process. The wastewater contains a small amount of furfural which will not be degraded through biological treatment in Area 600, thus additional treatment might be required. Therefore, WASTE is assigned values of 1 and 3 for the optimistic and pessimistic scenarios, respectively, and for the most probable scenario a mid-range value of 2.5 is assumed. These variable values are used in Equation 1.2 to calculate the percentage of Plant Performance for the three cases.

PCTNEW accounts for the new technologies of a pioneer plant. Feedstock handling, pretreatment, saccharification vessels, and the fluidized bed combustor are considered new technologies for the most probable case. For the optimistic case the new technologies include pretreatment and for the pessimistic scenario include feedstock handling, pretreatment, saccharification and fermentation, distillation columns and evaporators, and fluidized bed combustor. During pretreatment degradation products, such as furfural and hydroxy-methylfurfural, form and may build up in the process loop. Some of those products show inhibitory effects on fermentative organisms that may result in ethanol yield loss. For the present studies a value of 3 is assigned for the variable IMPURITIES. For the optimistic and pessimistic scenarios the values are 1 and 5, respectively, representing the full range of values because the extent of impact that may be caused by degradation products and the likelihood of buildup is not yet known. The plant has five continuously linked process steps that include feedstock handling, pretreatment, saccharification and fermentation, distillation and by-product combustion. Therefore, the variable COMPLEXITY is 5 for all three scenarios. Plant startup cost and permits are assumed to be 10% of the total capital investment. However, the startup costs have not been studied in detail. Therefore, INCLUSIVENESS is given a value of 33% for the most probable scenario. For the optimistic and pessimistic scenarios the values are assigned 0% and 50%, respectively. Specific plant site information has not been taken into account, so none of the site-specific information has been evaluated. Thus, the PROJECT DEFINITION is 6 for the most probable scenario, and for the optimistic and pessimistic scenarios are 5 and 7, respectively.

Results and Discussion

Nth Plant Analysis

Ethanol yield, byproduct credit, total installed equipment cost, total project investment, and estimated MESP for each of the process variations is shown in Table 2.12.

Table 2.12 MESP from the dilute acid and hot water pretreatment scenarios.

Process Variations	TCI (\$MM)	Total Installed Equipment Cost (\$MM)	Ethanol Yield (Gal/MT)	Ethanol Production (MM Gal/Yr)	Electricity Export (\$MM/Yr)	MESP (\$/Gal)
Dilute Acid Pretreatment (base case)	376	164	76.3	53.4	11.7	3.40
Dilute Acid Pretreatment (Pilot)*	389	169	72.5	50.8	12.6	3.60
Hot Water Pretreatment	361	156	55.8	39.0	11.3	4.29

* Conversions are from NREL's 1 MT/d pilot plant

The difference in ethanol yield between the two process scenarios is significant, with the yields being 76.3 Gal/MT for the dilute acid scenario and 55.8 Gal/MT for the hot water pretreatment scenario. This is the most significant factor in the difference in the MESP between the two scenarios. The MESP for the dilute acid pretreatment scenario is \$3.40/gal compared to \$4.29/gal for the hot water pretreatment scenario.

There is a relatively small difference in the TCI between the scenarios. However, the TCI of the hot water pretreatment scenario is slightly lower, due primarily to the low cost of the pretreatment area. The cost estimate for the pretreatment reactor for the hot water pretreatment scenario is a single continuous flow tubular reactor, while there are three reactors for the dilute acid pretreatment scenario, each with its own feed system and blowdown tank. The installed cost of the tubular reactor is \$MM 0.31 compared to \$MM 23.0 for the dilute acid pretreatment reactors. Further analysis is necessary to determine if a single reactor can meet the flow rate requirements for dilute acid pretreatment at the scale of this study. The slurry, after dilute acid pretreatment, also must be conditioned to remove sulfuric acid, which adds an additional \$10.8M to the pretreatment cost.

The costs of each process area for both scenarios are presented in Table 2.13. The most expensive costs for both scenarios are the cogeneration area, representing 34% and 42% of the total installed cost for the dilute acid and hot water pretreatment scenarios, respectively. The boiler/turbogenerator

is more costly in the hot water pretreatment scenario because the fuel feed is larger, and the fuel feed is the attribute used to size the boiler/turbogenerator for the cost estimate. The larger fuel feed is due to the decreased conversion of xylan and cellulose to monomers, resulting in more xylan and glucan oligomers as by-products to be combusted.

Table 2.13 Capital costs by process area.

Cost Areas / Factor	Dilute Acid Pretreatment Scenario-Installed Equipment Cost		Hot Water Pretreatment Scenario-Installed Equipment Cost	
	(MMS)	(%)	(MMS)	(%)
Feedstock Handling (Area 100)	10.9	6.6	10.9	7
Pretreatment (Area 200)	36.2	22.1	6.7	4.3
Saccharification & Fermentation (Area 300)	21.8	13.3	30.2	19.3
Distillation and Solids Recovery (Area 500)	26.1	15.9	30.9	19.8
Wastewater Treatment (Area 600)	3.5	2.1	1.9	1.2
Storage (Area 700)	3.2	2.0	3.3	2.1
Boiler/Turbogenerator (Area 800)	56.1	34.2	65.8	42
Utilities (Area 900)	6.3	3.8	6.7	4.3
Total Installed Equipment Cost	164.1		156.3	
Fixed Capital Investment (FCI)	326.8		284.3	
Working Capital (WC)	49.0		42.6	
Total Capital Investment (TCI)	375.9		327.0	
Lang Factor*	2.9		2.6	

* The Lang factor is calculated by dividing TCI by the total equipment cost

While less capital cost is incurred for the pretreatment area in the hot water pretreatment scenario, other areas require larger capital investment than the dilute acid pretreatment scenario. As previously noted, the boiler/turbogenerator is more costly because of the larger by-product stream. The saccharification and fermentation area is more costly because more vessels are needed due to the lower slurry density. The density of the hot water pretreatment slurry is lower because of lower solubilization than occurs due to dilute acid pretreatment. The solids recovery equipment is also more costly because of the increased solids stream—again due to lower solubilization—and the evaporators are more costly because of a larger flow of oligosaccharides that are not fermented by *Z. mobilis*.

Process Model Results

Process flow diagrams, including stream flows and conditions, for each of the process areas may be viewed in Appendices 2.A and 2.B. Energy flow diagrams are presented in Appendix 2.C. The energy in the ethanol product, excess electricity, and internal heat and electricity as percentages of the higher heating value of the feedstock is shown in Figure 2.1 for both scenarios.

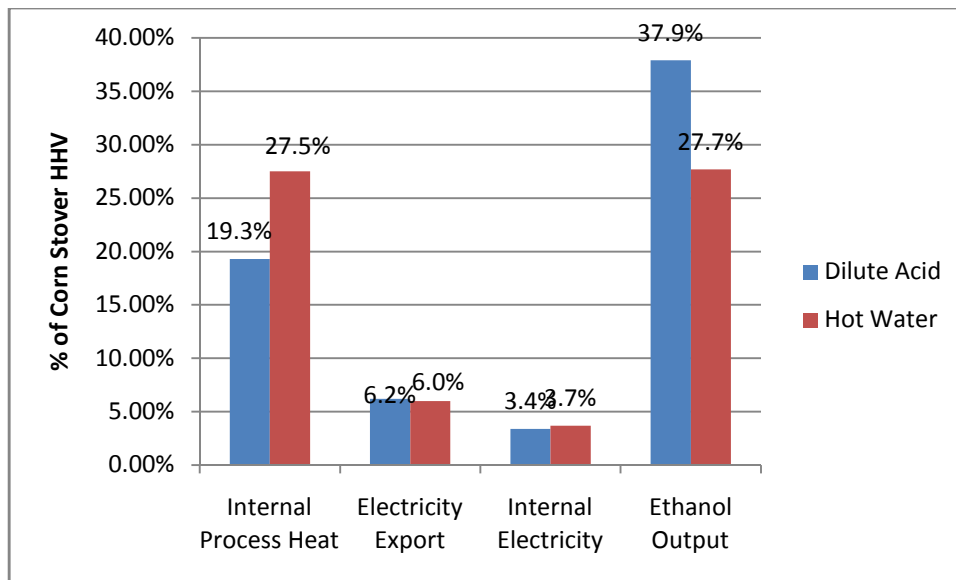


Figure 2.1 Energy in products and by-products and internal use.

Considering the total energy export in the ethanol and excess electricity, the overall efficiency of the dilute acid pretreatment scenario is 44.1% and that of the hot water pretreatment scenario is 33.7%. The theoretical maximum for the ethanol output alone is 45.7%.

The hot water pretreatment scenario uses considerably more process heat than the dilute acid scenario due to the indirect heating of biomass slurry prior to pretreatment. For dilute acid pretreatment the slurry is heated directly by adding steam in the presteamer and reactor. For hot water pretreatment the pretreated slurry exchanges heat with the incoming slurry to provide preheating. However, since steam is used to directly heat the biomass for dilute acid pretreatment, the pretreated slurry is used to completely preheat the fermented beer before distillation. Therefore, the heat load of the beer column is much greater in the hot water pretreatment scenario because the fermented beer is only slightly preheated.

Sensitivity Analysis

Process specific sensitivity analysis involving pretreatment and saccharification operations has been performed to study the impact of operating temperature, retention time, acid concentration, and reaction conversions on the MESP. The results of the pretreatment sensitivities are shown in Figures

2.2 and the results of the enzymatic saccharification sensitivities are shown in Figure 2.3. In the figures the base case MESP for both scenarios are represented by the dots. The parameter values and results used in Figures 2.2 and 2.3 are tabulated in Tables 2.D.1 and 2.D.2 of Appendix 2.D. The range of sensitivity parameters are chosen based on probable ranges from literature and are not equal positive and negative perturbations. This is meant to show a range of plausible MESP.

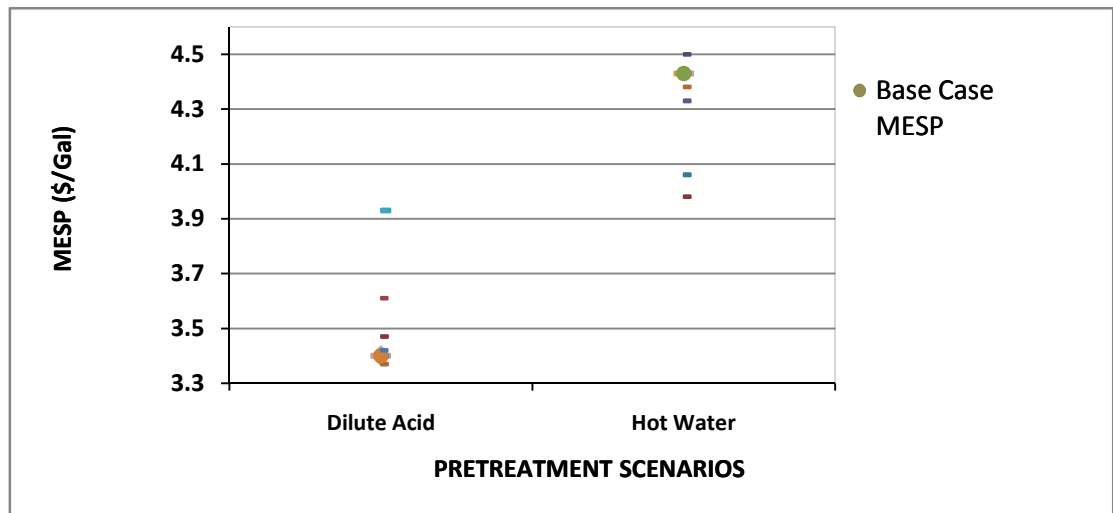


Figure 2.2 Impact of pretreatment parameters on MESP.

The MESP for the dilute acid pretreatment scenario sensitivity analysis ranges from \$3.37-\$3.93/gal EtOH and the hot water pretreatment scenario ranges from \$3.98-\$4.50/gal EtOH, meaning there is no overlap in MESP between the two scenarios. The lowest MESP for the dilute acid pretreatment scenario are results of the low parameter value for the conversion of xylan to xylose, for which 0.33 was used compared to a base case value of 0.825. The highest MESP for the dilute acid pretreatment scenario occurs with a retention time of 10 min, up from one minute for the base case. The large increase in MESP for longer retention time is due to the significant size increase of the pretreatment reactor, which comprises a very significant capital cost. The lowest MESP for the hot water pretreatment scenario occurs with a high solids loading of 20 wt% compared to 12.9% for the base case. With higher solids loading the stream heat requirement is reduced, resulting in increased electricity generation and export. The highest MESP occurs when the conversion of xylan to xylose oligomers is increased to 0.6 from the base case of 0.554. It is assumed that oligomers produced during pretreatment are not hydrolyzed further to monomers during enzymatic saccharification. Therefore, increasing the conversion to oligomers results in fewer monosaccharides after saccharification.

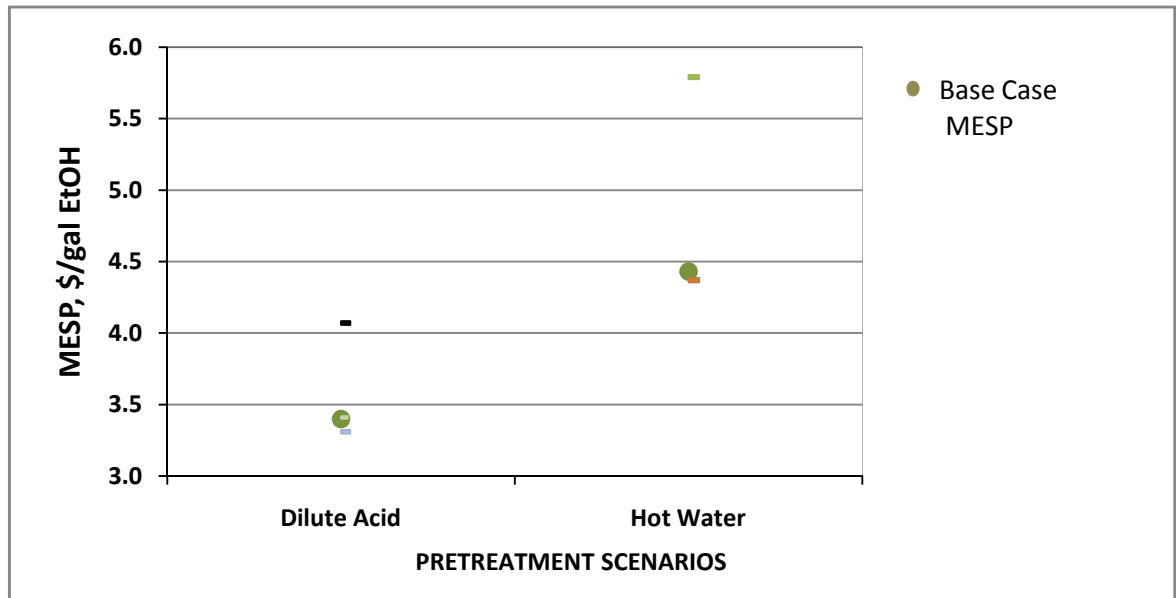


Figure 2.3 Impact of saccharification parameters on MESP.

The saccharification parameter sensitivities for the dilute acid pretreatment scenario result in an MESP ranging from \$3.31-\$4.07/gal EtOH. The highest and lowest MESP occur with the high and low sensitivity parameter value for the conversion of cellulose to glucose. The high conversion was 0.97 while the low conversion was 0.67 and the base case was 0.911.

The MESP ranges from \$4.37-\$5.79/gal EtOH as a result of the saccharification sensitivity parameter values chosen for the hot water pretreatment scenario. Only two points were used for this sensitivity study—a high value for the conversion of xylan to xylose and a low value for the conversion of cellulose to glucose. This is because the base case conversion of xylan to xylose is similar to the lowest values found in the literature, therefore, only a higher value was used. Similarly, the base case conversion of cellulose to glucose is similar to the highest values found in the literature, so only a low value was used as a point sensitivity.

A sensitivity study was also conducted for several major economic parameters. The selected sensitivity parameters are feedstock cost, enzyme cost, contingency factor, installation factor (or corresponding installed equipment cost), and export electricity value. The range of values for each parameter was chosen based on estimates for the most probable values. Figures 2.4 and 2.5 show the results of the economic sensitivity parameters for the dilute acid and hot water pretreatment scenarios, respectively.

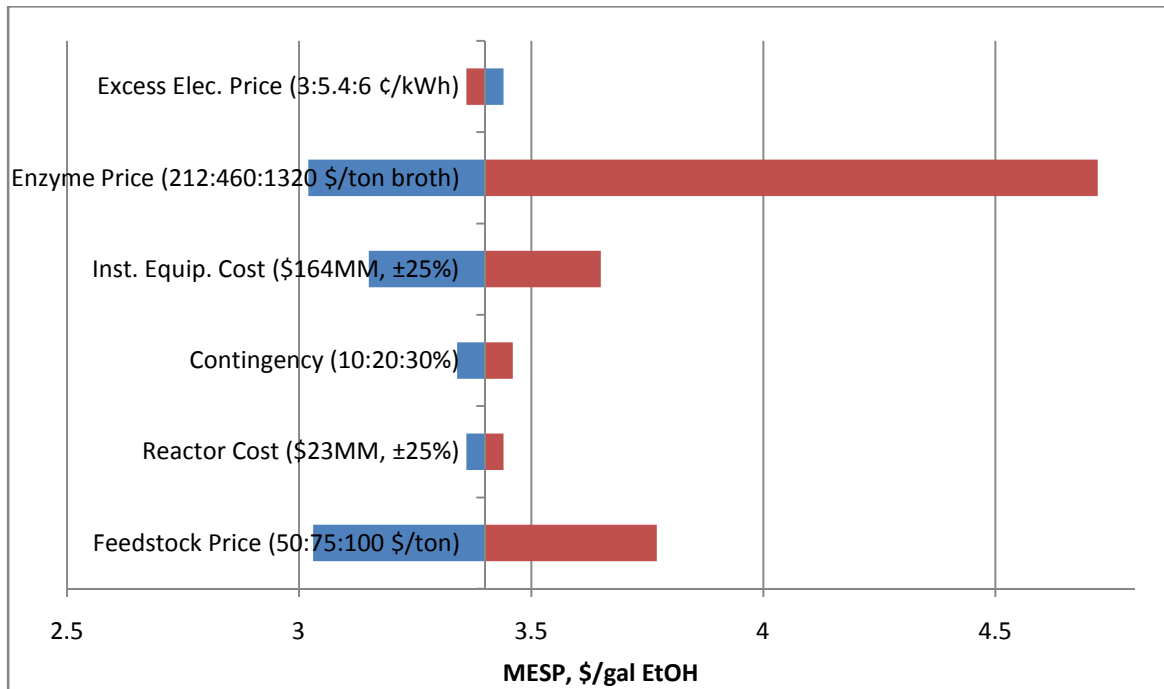


Figure 2.4 Impact of economic parameters on MESP for the dilute acid pretreatment scenario

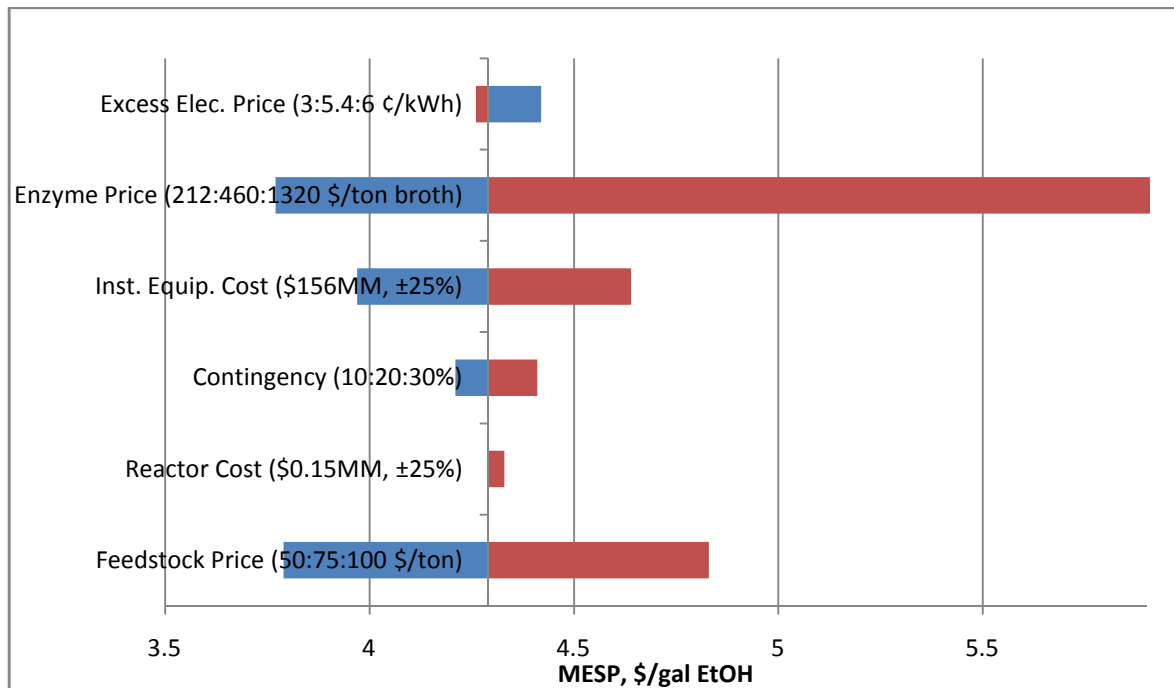


Figure 2.5 Impact of economic parameters on MESP for the hot water pretreatment scenario.

The feedstock cost and enzyme price have the most significant impact on MESP for both scenarios. Contingency factor and the total installed equipment cost showed a moderate impact on MESP. When feedstock cost is increased from \$75/dry ST (base case scenario) to \$100/dry ST, MESP increased by 11% and 13% for the dilute acid and hot water pretreatment scenarios, respectively. For the high enzyme cost the MESP increased by \$1.32 and \$1.84/gal EtOH for the two scenarios. The enzyme cost range was chosen based on a range of estimated cost per gallon of ethanol because most public estimates from enzyme companies mention cost in terms of dollars per gallon of ethanol produced. The sensitivity range of enzyme broth cost on a dollar per ton basis equates to a range of \$0.32-\$2.00/gal EtOH. This range was chosen based on publicly stated enzyme cost estimates and the cost per ton of broth was back-calculated using the dilute acid pretreatment discounted cash flow spreadsheet.

Using the same parameters, another sensitivity study was performed using equal perturbations for each parameter. The previous sensitivity analysis was based on probable ranges for each parameter. However, for this analysis each parameter was increased and decreased by 10% in order to understand which parameters are weighted the heaviest. Figure 2.6 shows the results of this study for the dilute acid pretreatment scenario.

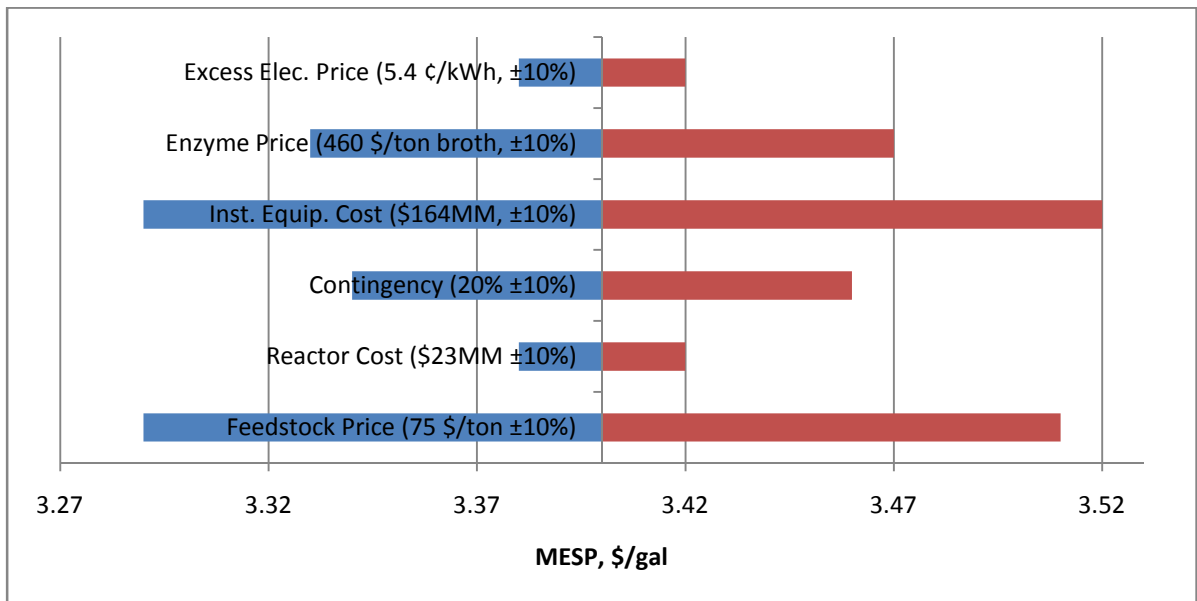


Figure 2.6 Sensitivity analysis results for the dilute acid pretreatment scenario with equal perturbations.

The feedstock price and installed equipment cost have the most significant impact on MESP. In Figure 2.4 the enzyme price has the largest range of MESP, which reflects the uncertainty in enzyme

price. However, Figure 2.6 shows that it has less weight as a parameter than feedstock price and installed equipment cost.

Pioneer Plant Risk Analysis

An analysis was conducted to estimate the increased risk associated with pioneer plant construction and operation due to equipment and process uncertainties. The analysis was performed using the method developed by the RAND Corporation which is described above. The results are presented in Table 2.14.

Table 2.14 Pioneer plant risk analysis results.

		Most Probable	Optimistic	Pessimistic
Dilute Acid Pretreatment	MESP	\$5.22/gal	\$4.19/gal	\$6.68/gal
	TCI	\$769MM	\$579MM	\$1067MM
Hot Water Pretreatment	MESP	\$6.63/gal	\$5.28/gal	\$8.34/gal
	TCI	\$699MM	\$512MM	\$950MM
Dilute Acid Pretreatment – Pilot Scale	MESP	\$5.51/gal	\$4.13/gal	\$7.15/gal
	TCI	\$794MM	\$550MM	\$1101MM

Recall from Table 2.12 that the MESP for the n^{th} plant are estimated to be \$3.40, \$3.60, and \$4.43 for the dilute acid, dilute acid pilot scale, and hot water pretreatment scenarios, respectively. For the most probable case, MESP for pioneer plants are estimated at 54%, 44%, and 50% higher than the n^{th} plant estimates for the respective scenarios. Very large increases in capital cost are estimated. For the most probable case the TCI is estimated to increase by 104%, 85%, and 94% above the n^{th} plant TCI for the dilute acid, dilute acid pilot, and hot water pretreatment scenarios, respectively.

The TCI and MESP for the pioneer plant analysis using pilot-scale conversions are lower than for the models using lab scale conversions for the optimistic case. This is because the pretreatment area was not included in the estimation of variables PCTNEW and NEWSTEPS, both of which take into account the process steps which employ new technology. The variable BALEQS, which is defined as the percentage of mass and energy balances that can be commercially verified, was also increased by 10% for the optimistic case under the assumption that pilot scale mass balances increase accuracy. The lower MESP reflects the impact of increased certainty by using data from larger scale experiments.

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 2.6 presents a plot of estimated ethanol price from seven previous studies as a function of feedstock price. The ethanol and feedstock prices were updated to 2007 dollars using the CPI. The solid line on the plot represents the MESP for the dilute acid pretreatment scenario using the model developed in this study as a function of feedstock price.

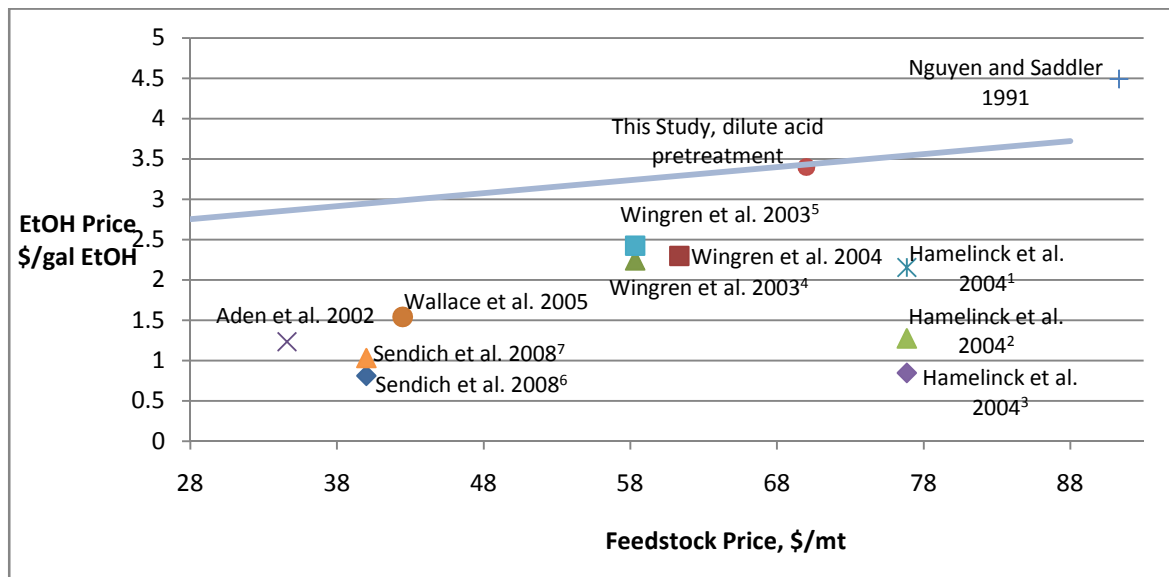


Figure 2.6 Cost estimations from previous techno-economic studies.

¹Short term technology—Simultaneous saccharification and fermentation (SSF)

²Middle term technology—Simultaneous saccharification and co-fermentation (SSCF)

³Long term technology—Consolidated bioprocessing (CBP)

⁴Separate hydrolysis and fermentation (SHF)

⁵SSF

⁶CBP

⁷SSCF

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 remain lower than the line derived from this study. The study by Hamelinck et al. 2004 represents a significant outlier from the apparent correlation between feedstock price and ethanol price. The three ethanol price estimates are for short (5 years from time of study), middle (10-15 yrs), and long term (20+ yrs) technology implementation. The short term estimate is closer 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 very similar to those in this study and the TCI, updated to 2007 dollars is nearly equal as well. The most significant difference from this study is the operating cost, which is approximately \$0.32/gal EtOH compared to \$1.68/gal EtOH (not including feedstock cost), partly due to lower costs for CSL, cellulase, and other raw materials. This accounts for most of the discrepancy between ethanol price estimates.

The ethanol price from the study published by Sendich et al. 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 (CBP), which is an advanced technology that is also modeled in the long term estimate from Hamelinck et. al. The higher ethanol price estimate of \$1.03 is from a model using simultaneous saccharification and co-fermentation (SSCF). SSCF is also a more advanced technology than was considered in this study, which results in a 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, and because enzyme cost is such a significant fraction of the MESP, it contributes significantly to the discrepancy between the current study and previous studies. For example, the enzyme prices used in Wingren et al. (2004) and Aden et al. (2002) are approximately 30% and 17% of the price used in this study, respectively.

Conclusions

Dilute acid pretreatment for cellulosic ethanol production is economically favorable to hot water pretreatment under the assumptions considered in this study and based on a 5-8 year time frame for initiating plant operation. It is estimated that the dilute acid pretreatment scenario MESP is \$3.40/gal EtOH compared to \$4.29/gal EtOH for the hot water pretreatment scenario. The most significant factor causing the higher MESP for the hot water pretreatment scenario is the lower ethanol production rate. The annual ethanol production rate for the dilute acid pretreatment scenario is 53.4 MMGal/yr compared and is 39.0 MMGal/yr for the hot water pretreatment scenario. The reduced ethanol production is a result of lower monosaccharide yields during pretreatment and subsequent saccharification, meaning less sugar is available for fermentation. However, a tradeoff exists between the higher capital cost of the pretreatment area for dilute acid pretreatment and the increased ethanol yields. The reactors and auxiliary equipment for dilute acid pretreatment is significantly more costly than the hot water pretreatment reactor. Additionally, the acid neutralization step required after dilute

acid pretreatment also increases costs. Although, it is apparent that the lower capital cost of hot water pretreatment equipment does not offset the effect of reduced monosaccharide yields.

Using reaction conversions from pilot scale dilute acid pretreatment experiments in the process model results in an increase in the MESP to \$3.60/gal EtOH. Clearly there is a level of risk involved with process scale-up and it is uncertain how further increases to commercial scale operation will impact production cost. To estimate the potential risk associated with process scale-up, pioneer plant risk analysis was conducted. Under the most probable assumptions for pioneer plant operation for the dilute acid pretreatment scenario the MESP is \$5.22/gal EtOH. The MESP for the optimistic and pessimistic assumptions are \$4.19 and \$6.68/gal EtOH, respectively. Additionally, the TCI for the most probable case is estimated to double from the cost of an n^{th} plant. Because of the large capital cost and the MESP being well above market prices for a pioneer plant, it may prove very difficult for the cellulosic ethanol industry to finance growth until a number of biotechnology barriers are broken.

Significant opportunities exist to reduce the MESP through biochemical technology breakthroughs. Enzyme cost in this study is assumed to be \$0.70/gal EtOH, representing a potential ethanol cost reduction by reducing enzyme production cost and increasing specific activity. In the dilute acid pretreatment model in this study 75.6% of xylose is converted to ethanol during fermentation and none of the other hemicellulose sugars are converted to ethanol. The development of organisms which can ferment xylose at conversions similar to those of glucose to ethanol, as well as other hemicellulose sugars also offers potential for reducing ethanol cost.

Appendix 2.A Dilute Acid Pretreatment Process Flow Diagrams

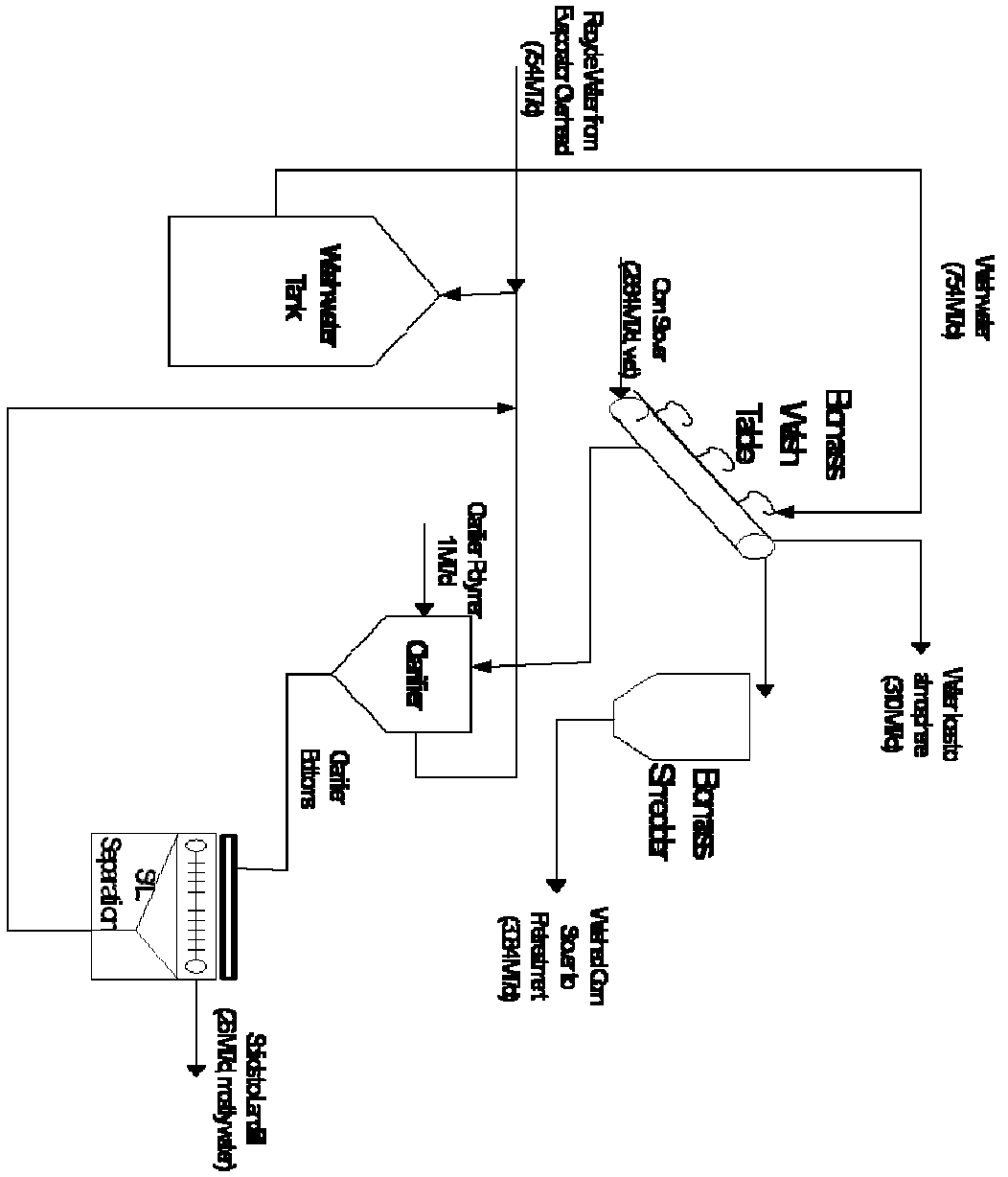


Figure 2.A.1 Feedstock washing and shredding area

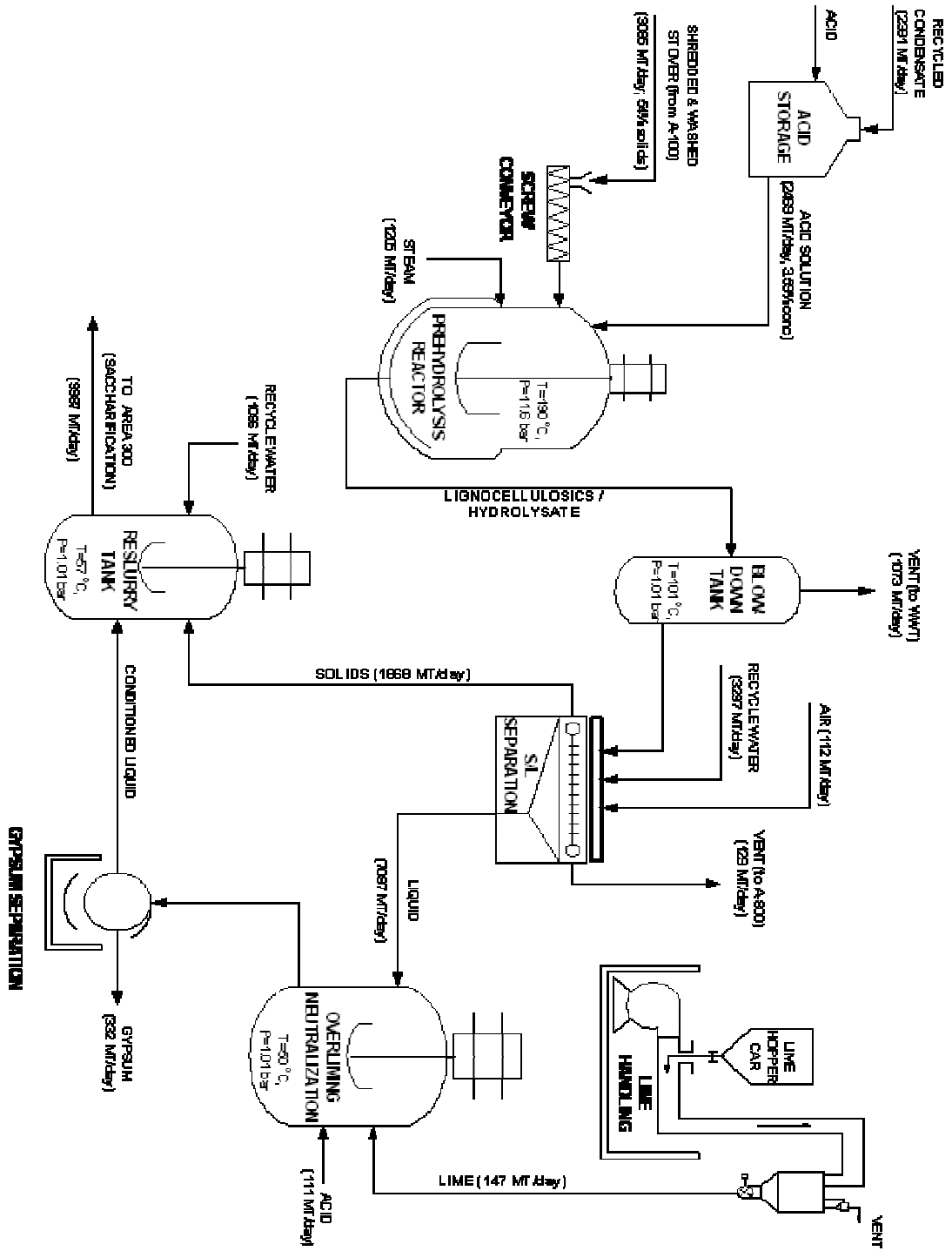


Figure 2.A.2 Dilute acid pretreatment area

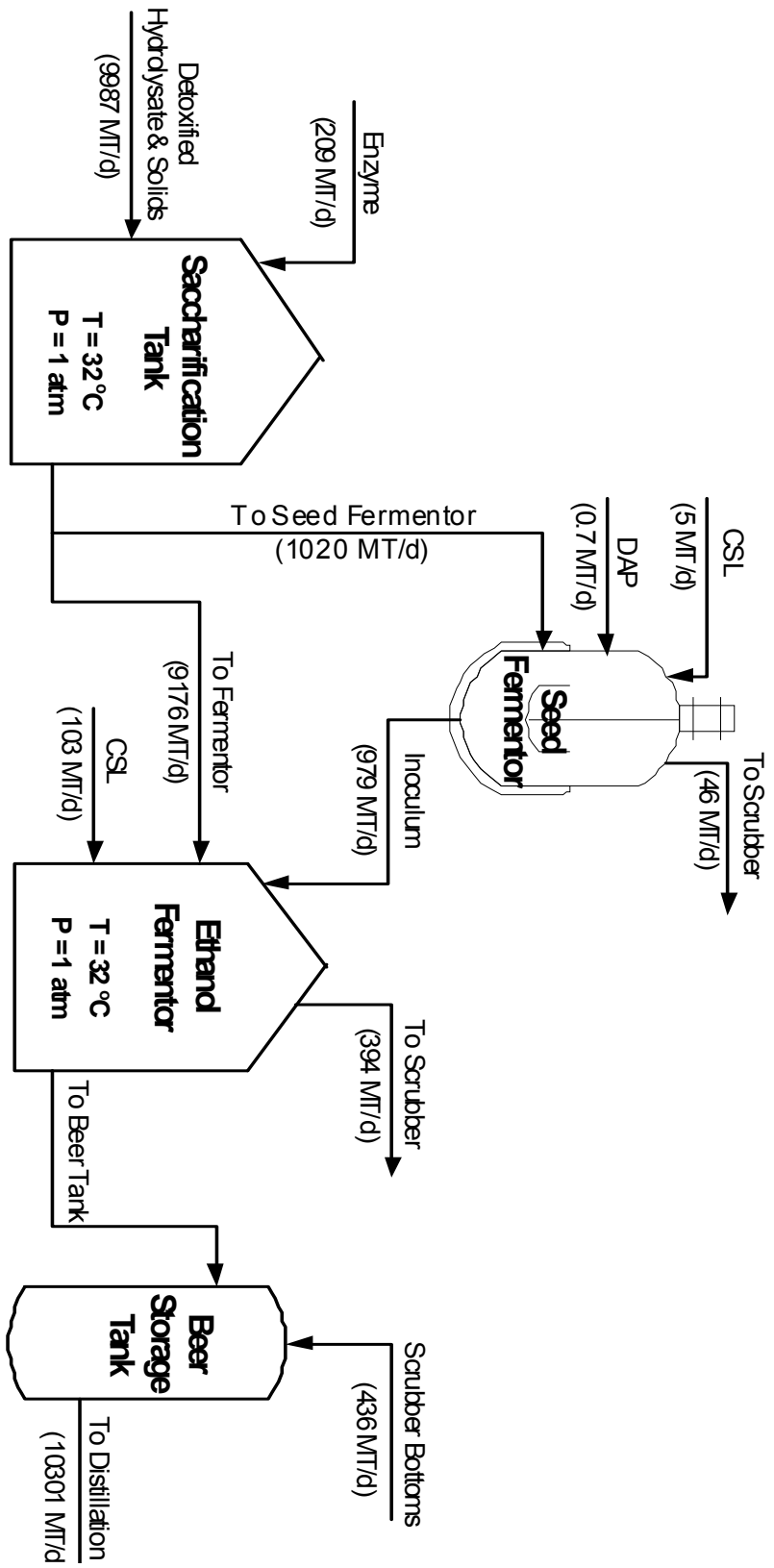


Figure 2.A.3 Saccharification and fermentation area

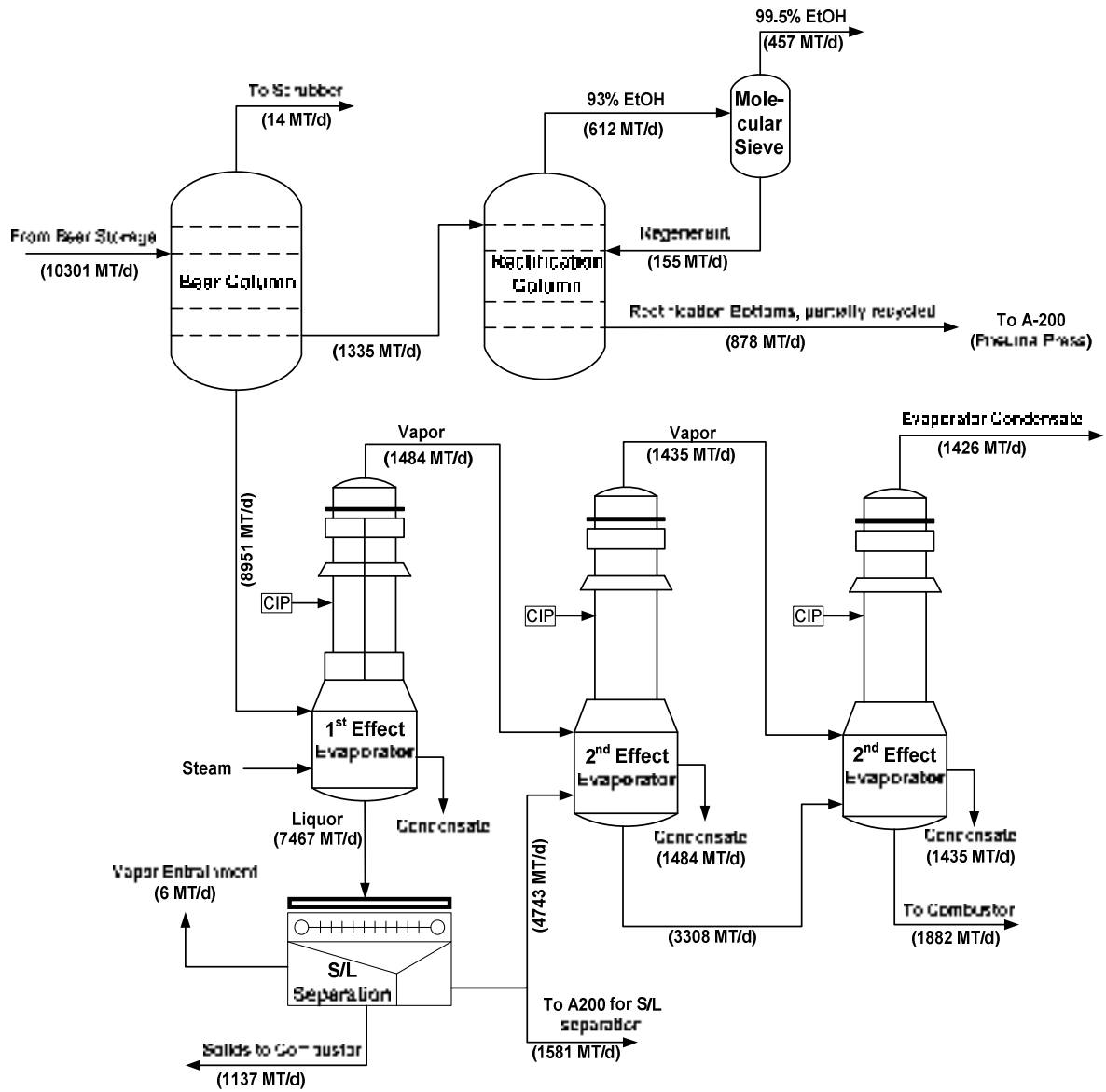


Figure 2.A.4 Ethanol separation and solid/liquid separation

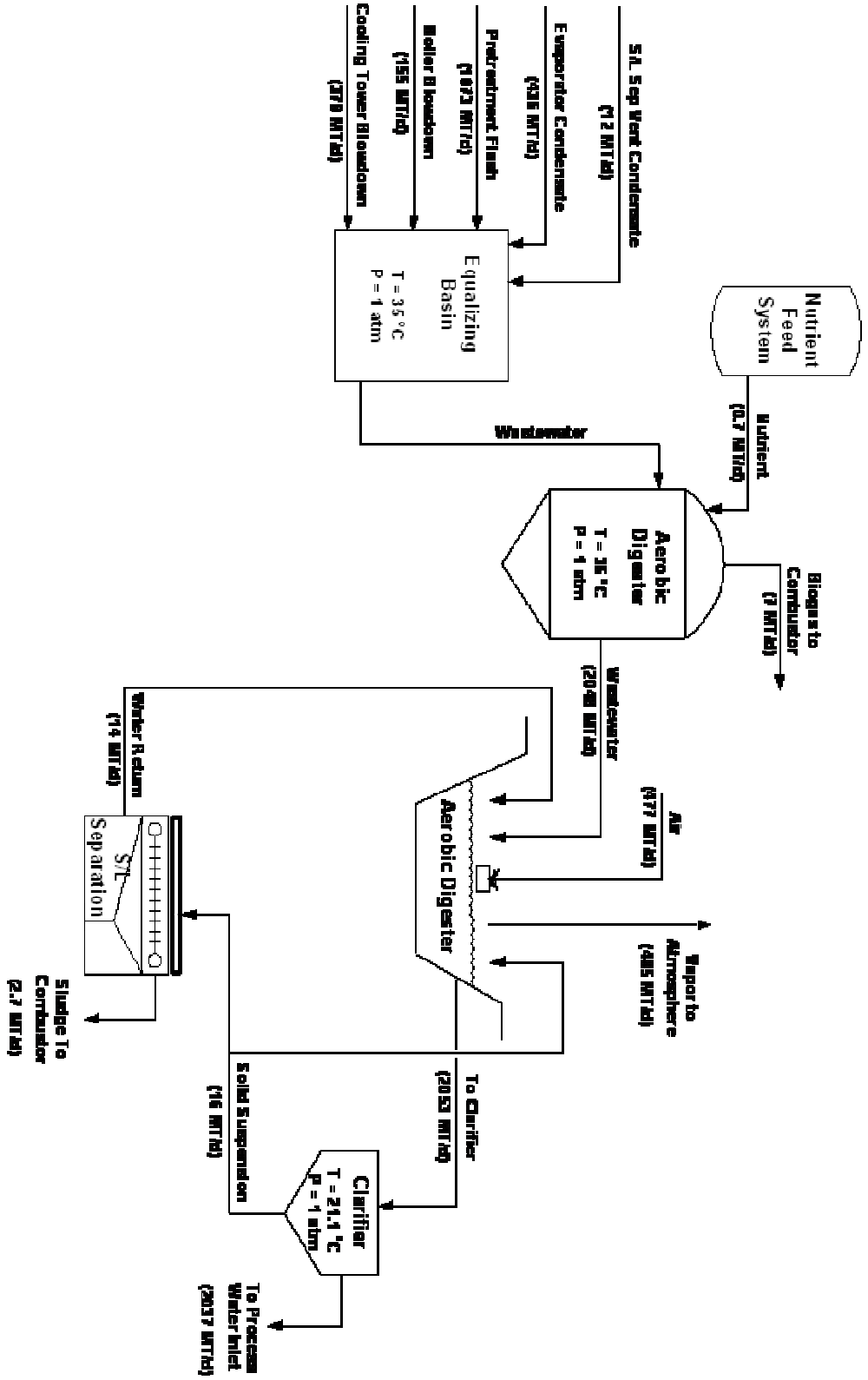


Figure 2.A.5 Wastewater treatment area

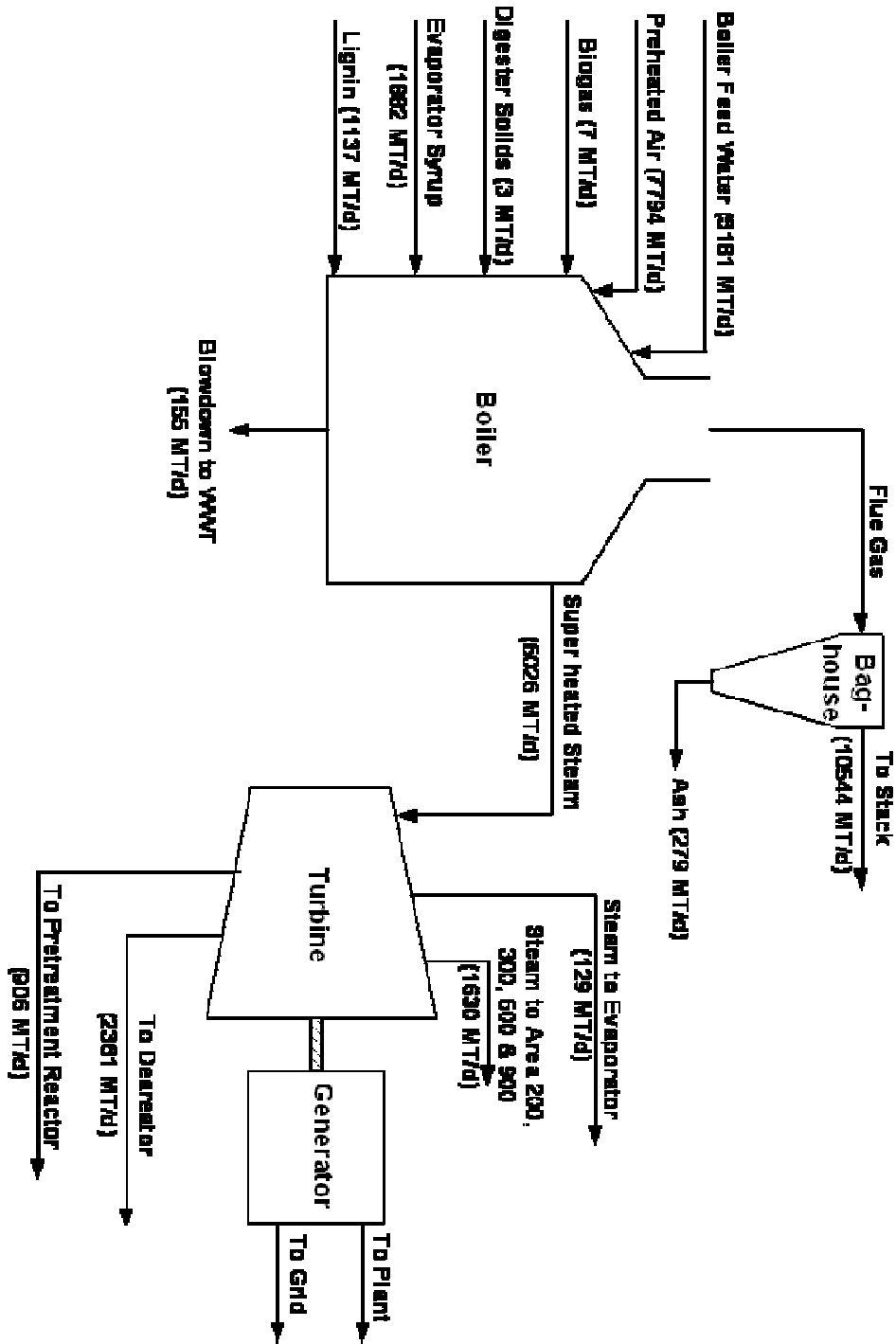


Figure 2.A.6 By-product combustion and heat and power generation area

Appendix 2.B Hot Water Pretreatment Process Flow Diagrams

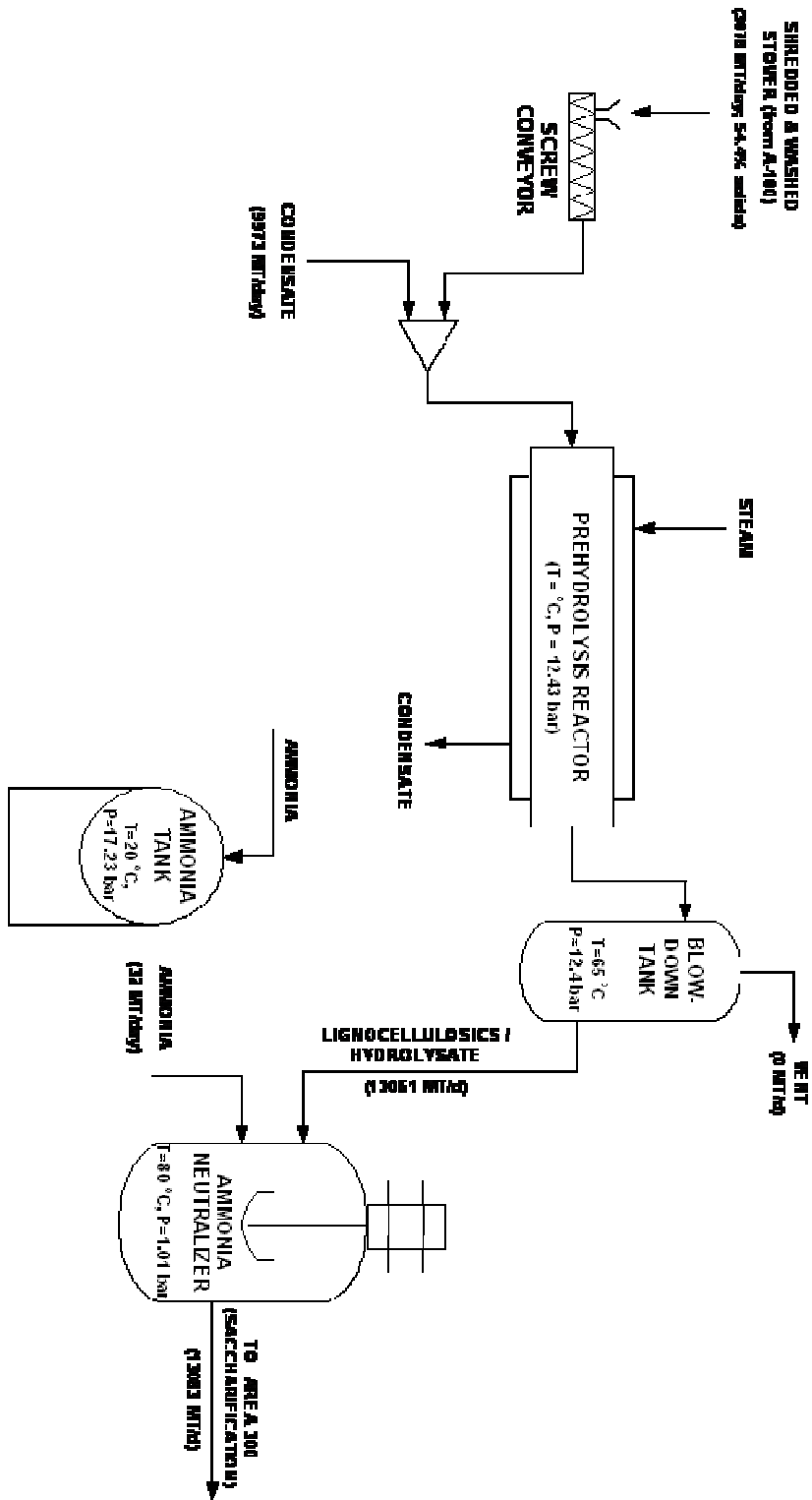


Figure 2.B.1 Hot water pretreatment area

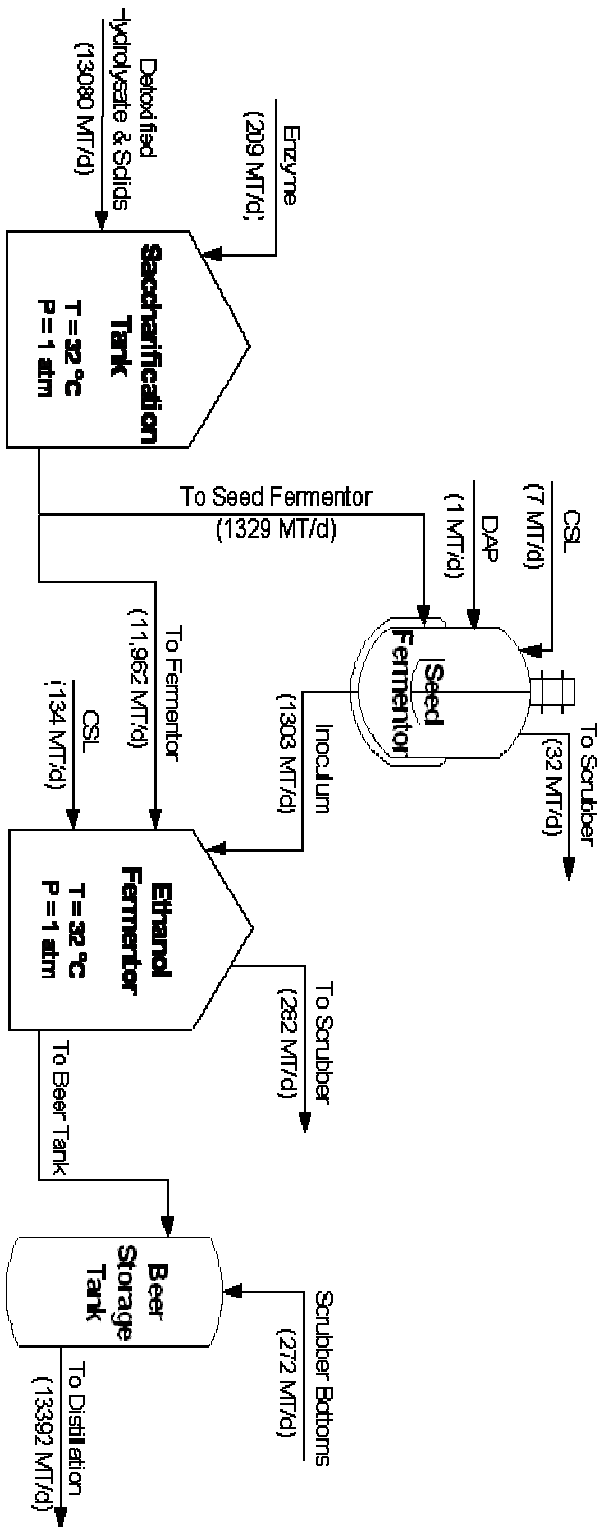


Figure 2.B.2 Saccharification and fermentation area

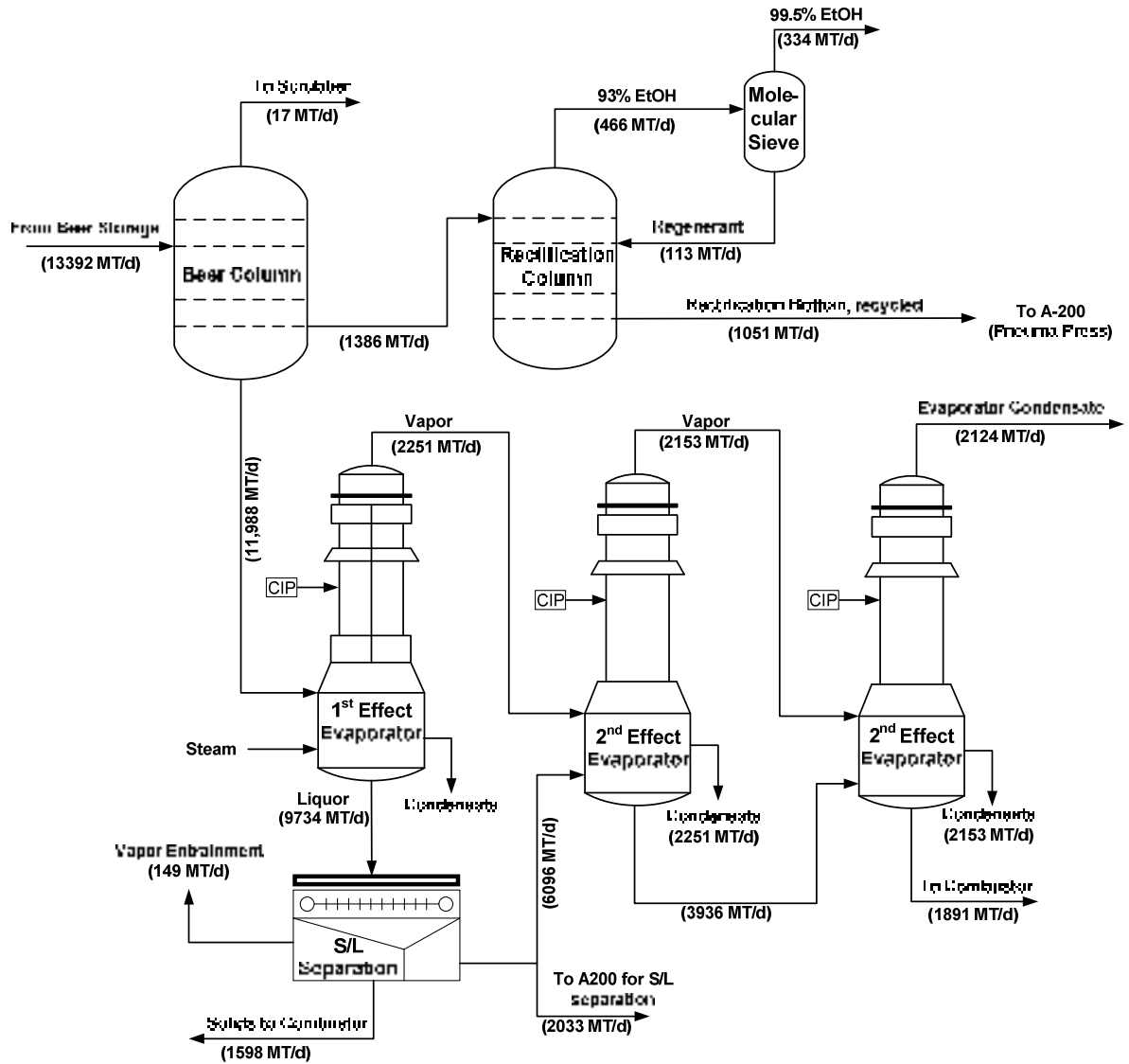


Figure 2.B.1 Ethanol distillation and solids separation area

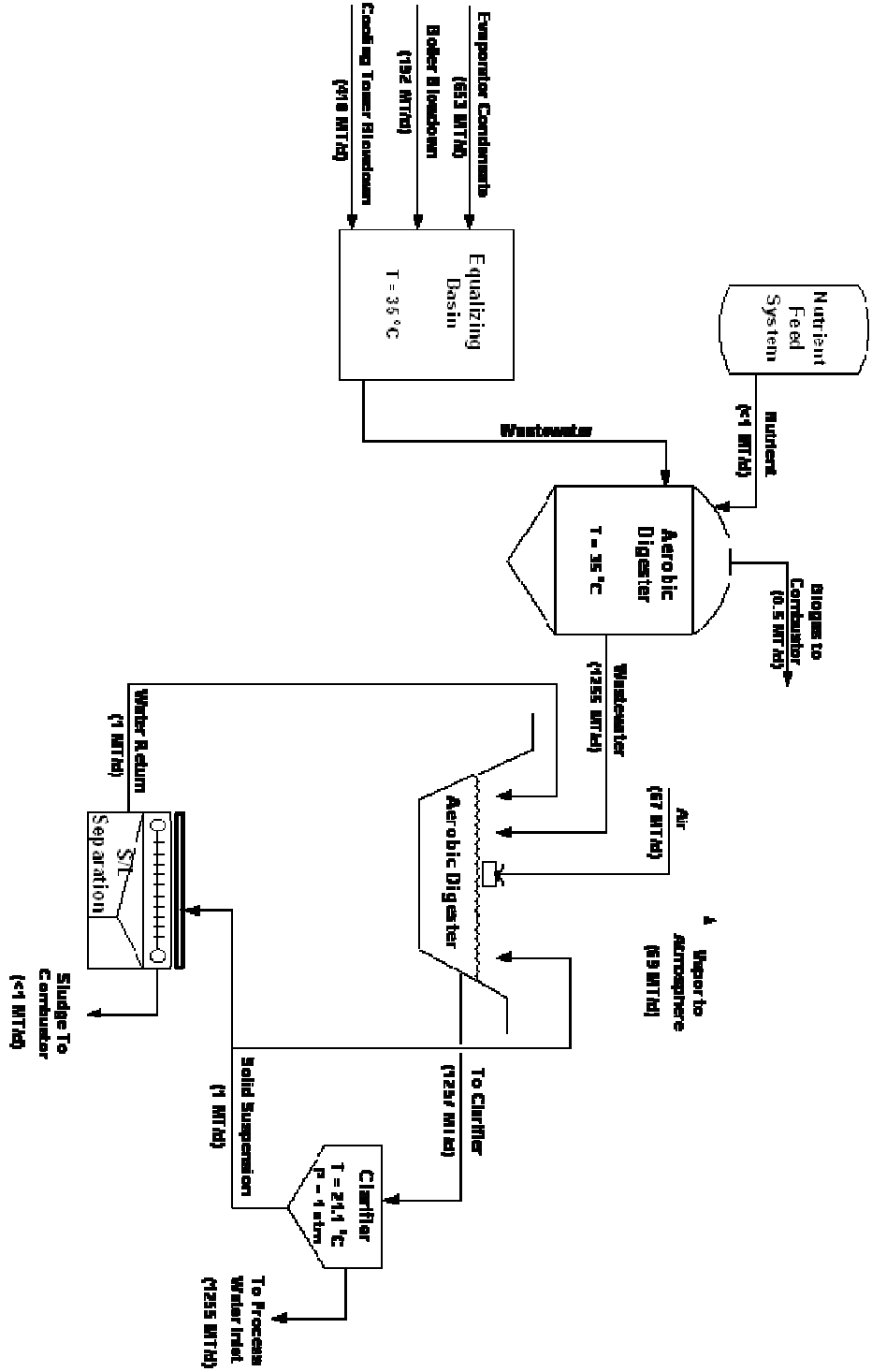


Figure 2.B.4 Wastewater treatment area

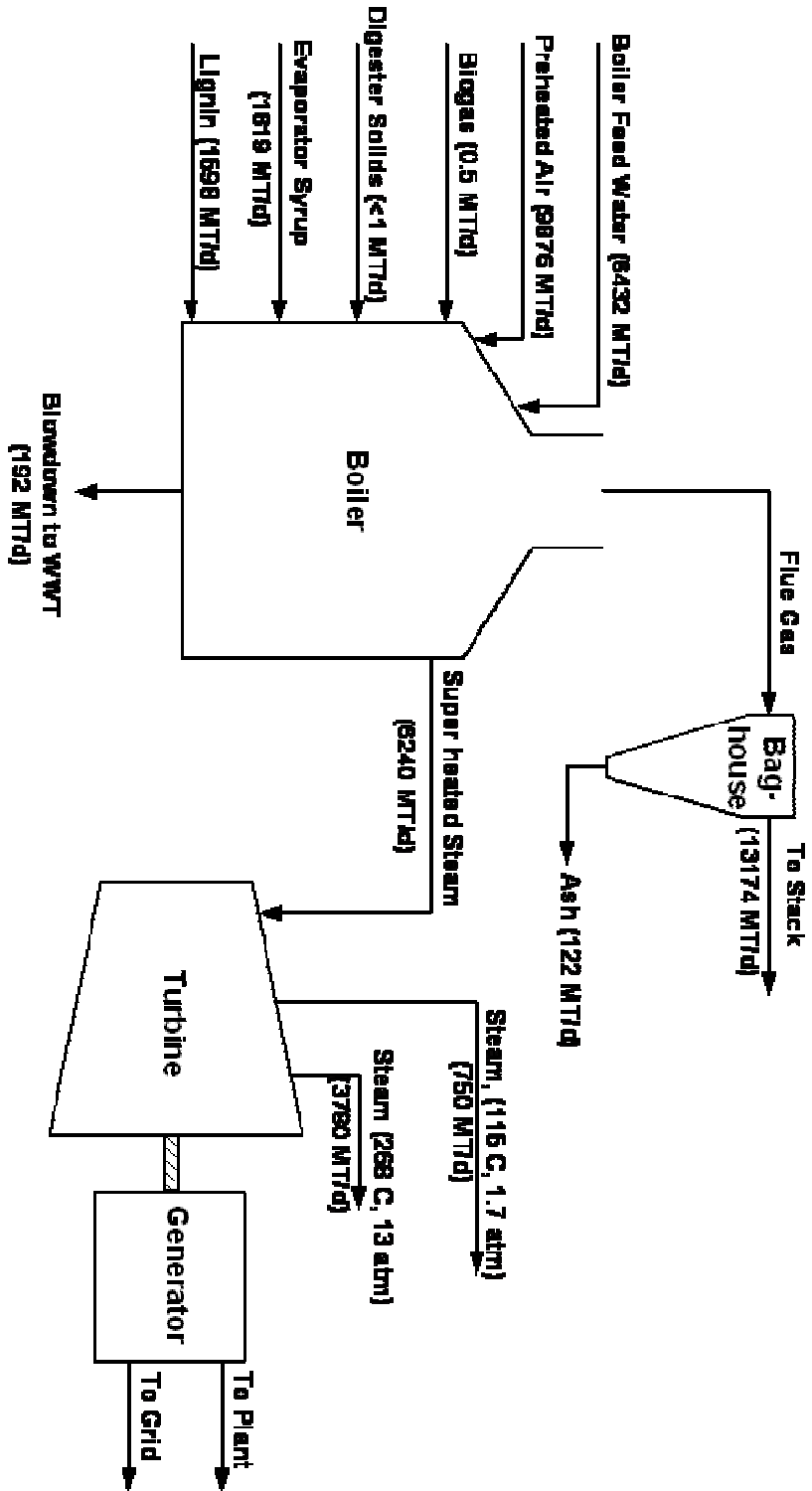


Figure 2.B.5 Heat and power generation area

Appendix 2.C Energy Flows in Heat, Power, and Mass Streams

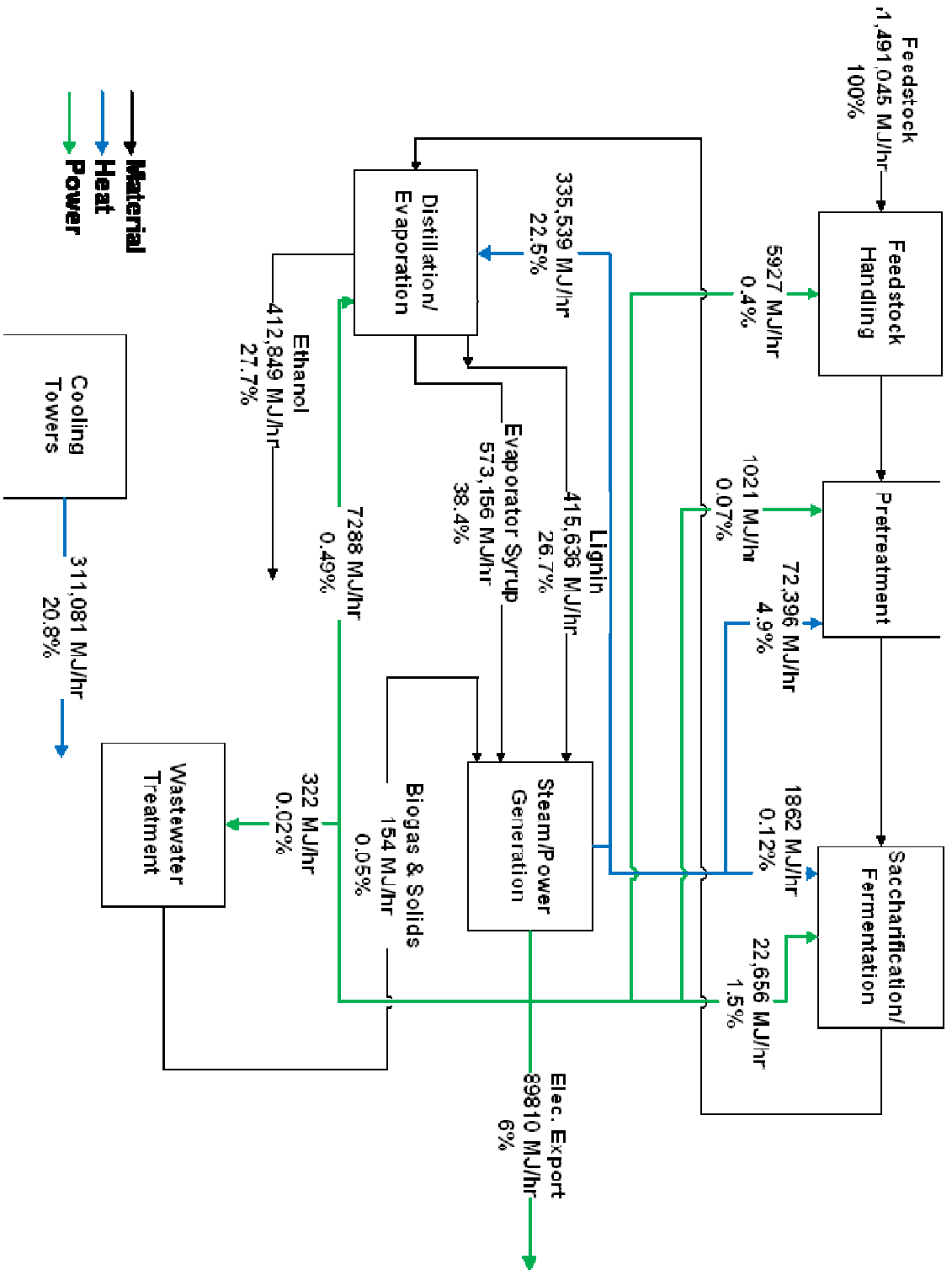


Figure 2.C.1 Energy flow for the hot water pretreatment scenario

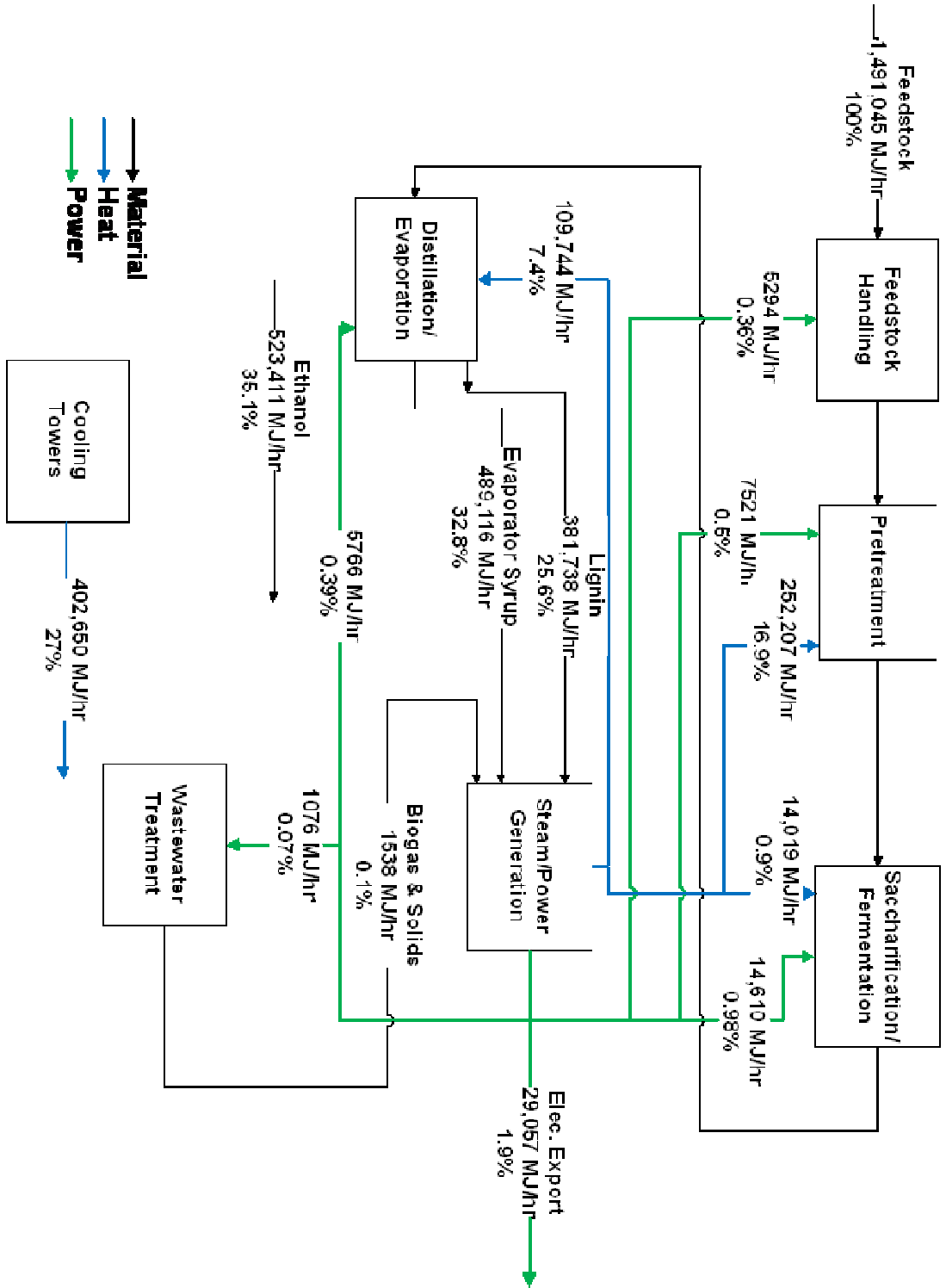


Figure 2.C.2 Energy flow chart for the dilute acid pretreatment scenario

Appendix 2.D Pretreatment Sensitivity Parameters and Results

Table 2.D.1 Dilute acid pretreatment sensitivity values

		Sensitivity		Production (MM Gal/Yr)	MESP	
		Parameter	Values		(\$/Gal)	Change (%)*
Dilute Acid Pretreatment						
2007 EVD Scenario		Base Case		53.4	3.40	
Pretreat- ment	High 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 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
Sacchar- ification	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

Table 2.D.2 Hot water pretreatment sensitivity parameters

		Sensitivity		Production (MM Gal/Yr)	MESP	
		Parameter	Values		(\$/Gal)	Change (%)*
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
		Cell to Glucose (% conv.)	2	39.1	4.29	0
		Xyln to Olig (% 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
		Cell to Glucose olig (% conv)	2.5	40.0	4.19	-2.3
		Xyln to Xylose olig (% conv)	25	42.9	3.92	-8.4
Saccharification	High Scenarios	Xyln to Xylose (% conv)	63	39.6	4.23	-1.4
	Low Scenarios	Cell to Glucose (% conv)	65	29.7	5.61	30.7

Appendix 2.E Equipment Lists and Cost

Table 2.E.1 Dilute acid pretreatment scenario equipment list

Equipment ID	Number Required	Spares Nos	Equipment Name	Scaling Exponent	Installation Factor	Installed Cost in 2007\$
C-101	2		Bale Transport Conveyor	0.6	1.62	\$1,862,521
C-102	2		Bale Unwrapping Conveyor	0.6	1.19	\$513,056
C-103	1		Belt Press Discharge Conveyor	0.6	1.89	\$135,809
C-104	4		Shredder Feed Conveyor	0.6	1.38	\$475,978
M-101	2		Truck Scales	0.6	2.47	\$241,380
M-102	4	1	Truck Unloading Forklift	1	1	\$135,982
M-103	4		Bale Moving Forklift	1	1	\$108,785
M-104	2		Corn Stover Wash Table	0.6	2.39	\$714,426
M-105	4		Shredder	0.6	1.38	\$2,395,754
M-106	1		Concrete Feedstock-Storage Slab	1	2.2	\$1,497,976
M-107	1		Polymer Feed System	0.6	2.28	\$98,300
P-101	2	1	Wash Table Pump	0.79	3.87	\$341,732
P-102	2	1	Wash Water Pump	0.79	5.19	\$343,719
P-103	1	1	Clarifier Underflow Pump	0.79	13.41	\$236,828
P-104	1	1	Clarified Water Pump	0.79	7.07	\$312,151
P-105	1	1	Belt Press Sump Pump	0.79	2.92	\$163,301
S-101	1		Clarifier Thickener	0.6	1.51	\$292,959
S-102	1		Belt Press	0.6	1.25	\$179,641
S-103	1		Magnetic Separator	0.6	1.3	\$19,537
T-101	1		Wash Water Tank	0.51	2.8	\$198,945
T-102	1		Clarifier Thickener Tank	0.51	3.04	\$583,192
A100					1.81	\$10,851,970
A-201	1		In-line Sulfuric Acid Mixer	0.48	1	\$3,479
A-205	1		Hydrolyzate Mix Tank Agitator	0.51	1.2	\$47,517
A-209	1		Overliming Tank Agitator	0.51	1.3	\$47,313
A-224	1		Reacidification Tank Agitator	0.51	1.2	\$144,827
A-232	1		Reslurrying Tank Agitator	0.51	1.2	\$63,335
C-201	1		Hydrolyzate Screw Conveyor	0.78	1.3	\$109,217
C-202	1		Hydrolysate Washed Solids Belt Conveyor	0.76	1.45	\$136,643
C-225	1		Lime Solids Feeder		1.3	\$6,892
H-200	1		Hydrolyzate Cooler	0.51	2.1	\$137,997

Equip- ment ID	Number Required	Spare Nos	Equipment Name	Scaling Expo- nent	Install- ation Factor	Installed Cost in 2007\$
H-201	2	1	Beer Column Feed Economizer	0.68	2.1	\$1,117,823
H-205	1		Pneumapress Vent Condensor	0.68	2.1	\$55,825
H-244	2	1	Waste Vapor Condensor	0.68	2.1	\$293,532
M-202	3		Prehydrolysis/Screw Feeder/Reactor	0.6	2.29	\$22,992,607
P-201	1	1	Sulfuric Acid Pump	0.79	2.8	\$131,112
P-205	2	1	Pneumapress Feed Pump	0.79	3.34	\$180,563
P-209	1	1	Overlimed Hydrolyzate Pump	0.79	2.8	\$129,980
P-211	1	1	Primary Filtrate Pump	0.79	3.56	\$349,590
P-213	1	1	Wash Filtrate Pump	0.79	2.71	\$370,096
P-222	1	1	Filtered Hydrolyzate Pump	0.79	2.8	\$131,099
P-223	1		Lime Unloading Blower	0.5	1.4	\$300,600
P-224	2	1	Saccharification Feed Pump	0.7	2.8	\$771,348
P-239	1	1	Reacidified Liquor Pump	0.79	2.8	\$132,631
S-205	3		Pneumapress Filter	0.6	1.05	\$5,985,662
S-222	1		Hydroclone & Rotary Drum Filter	0.39	1.4	\$456,634
S-227	1		LimeDust Vent Baghouse	1	1.5	\$732,886
T-201	1		Sulfuric Acid Tank	0.71	1.4	\$34,995
T-203	1		Blowdown Tank	0.93	1.2	\$108,624
T-205	1		Hydrolyzate Mixing Tank	0.71	1.2	\$54,420
T-209	1		Overliming Tank	0.71	1.4	\$205,654
T-211	1		Primary Filtrate Tank	0.71	2.45	\$131,394
T-213	1		Wash Filtrate Tank	0.71	3.68	\$90,562
T-220	1		Lime Storage Bin	0.46	1.3	\$370,983
T-224	1		Reacidification Tank	0.51	1.2	\$328,304
T-232	1		Slurrying Tank	0.71	1.2	\$81,188
A200					1.84	\$36,235,330
A-300	12		Ethanol Fermentor Agitator		1.2	\$390,002
A-301	1		Seed Hold Tank Agitator	0.51	1.2	\$20,484
A-304	2		4th Seed Vessel Agitator	0.51	1.2	\$37,716
A-305	2		5th Seed Vessel Agitator	0.51	1.2	\$33,751
A-306	2		Beer Surge Tank Agitator	0.51	1.2	\$167,381
A-310	30		Saccharification Tank Agitator		1.2	\$975,006
F-300	6		Ethanol Fermentor		1.2	\$4,791,884
F-301	2		1st Seed Fermentor		2.8	\$111,904

Equip-ment ID	Number Required	Spares Nos	Equipment Name	Scaling Expo-ment	Install-ation Factor	Installed Cost in 2007\$
F-302	2		2nd Seed Fermentor		2.8	\$248,168
F-303	2		3rd Seed Fermentor		2.8	\$617,376
F-304	2		4th Seed Fermentor	0.93	1.2	\$126,081
F-305	2		5th Seed Fermentor	0.51	1.2	\$471,004
H-300	6	1	Fermentation Cooler	0.78	2.1	\$132,463
H-301	1	1	Hydrolyzate Heater	0.68	2.1	\$162,939
H-302	3		Saccharified Slurry Cooler	0.78	2.1	\$0
H-304	1		4th Seed Fermentor Coil	0.83	1.2	\$9,815
H-305	1		5th Seed Fermentor Coil	0.98	1.2	\$62,326
H-310	15	1	Saccharification Cooler	0.78	2.1	\$10,286
P-300	6	1	Fermentation Recirc/Transfer Pump	0.79	2.8	\$355,875
P-301	1	1	Seed Hold Transfer Pump	0.7	1.4	\$82,457
P-302	2		Seed Transfer Pump	0.7	1.4	\$200,951
P-306	1	1	Beer Transfer Pump	0.79	2.8	\$144,484
P-310	15	1	Saccharification Recirc/Transfer Pump	0.79	2.8	\$26,461
T-301	1		Seed Hold Tank	0.51	1.2	\$258,450
T-306	1		Beer Storage Tank	0.71	1.2	\$418,181
T-310	15		Saccharification Tank		1.2	\$11,979,711
A300					1.3	\$21,835,156
A400						\$0
A-530	1		Recycled Water Tank Agitator	0.51	1.3	\$12,720
C-501	1		Lignin Wet Cake Screw	0.78	1.4	\$33,897
D-501	1		Beer Column	0.68	2.1	\$1,505,174
D-502	1		Rectification Column	0.68	2.1	\$1,621,105
E-501	2		1st Effect Evaporation	0.68	2.1	\$4,062,339
E-502	1		2nd Effect Evaporation	0.68	2.1	\$1,624,839
E-503	2		3rd Effect Evaporation	0.68	2.1	\$3,249,677
H-501	1	1	Beer Column Reboiler	0.68	2.1	\$1,002,181
H-502	1		Rectification Column Reboiler	0.68	2.1	\$83,876
H-504	1		Beer Column Condenser	0.68	2.1	\$49,892
H-505	1		Start-up Rect. Column Condenser	0.68	2.1	\$259,542
H-512	1	1	Beer Column Feed Interchanger	0.68	2.1	\$110,462

Equip- ment ID	Number Required	Spares Nos	Equipment Name	Scaling Expo- nent	Install- ation Factor	Installed Cost in 2007\$
H-517	1	1	Evaporator Condenser	0.68	2.1	\$860,240
M-503	1		Molecular Sieve (9 pieces)	0.7	1	\$3,461,120
P-501	1	1	Beer Column Bottoms Pump	0.79	2.8	\$368,970
P-503	1	1	Beer Column Reflux Pump	0.79	2.8	\$5,490
P-504	1	1	Rectification Column Bottoms Pump	0.79	2.8	\$41,787
P-505	1	1	Rectification Column Reflux Pump	0.79	2.8	\$37,883
P-511	2	1	1st Effect Pump	0.79	2.8	\$245,432
P-512	1	1	2nd Effect Pump	0.79	2.8	\$146,736
P-513	2	1	3rd Effect Pump	0.79	2.8	\$134,608
P-514	1	1	Evaporator Condensate Pump	0.79	2.8	\$114,582
P-515	1		Scrubber Bottoms Pump	0.79	2.8	\$12,040
P-530	1	1	Recycled Water Pump	0.79	2.8	\$109,306
S-505	4		Pneumapress Filter	0.6	1.04	\$6,376,193
T-503	1		Beer Column Reflux Drum	0.93	2.1	\$16,288
T-505	1		Rectification Column Reflux Drum	0.72	2.1	\$135,944
T-512	1		Vent Scrubber	0.78	2.1	\$285,071
T-514	1		Evaporator Condensate Drum	0.93	2.1	\$115,034
T-530	1		Recycled Water Tank	0.745	1.4	\$36,495
A500					1.51	\$26,118,926
A-602	1		Equalization Basin Agitator	0.51	1.2	\$31,005
A-606	1		Anaerobic Agitator	0.51	1.2	\$34,562
A-608	16		Aerobic Lagoon Agitator	0.51	1.4	\$336,908
C-614	1		Aerobic Sludge Screw	0.78	1.4	\$2,000
H-602	1		Anaerobic Digester Feed Cooler	0.74	2.1	\$191,017
M-604	1		Nutrient Feed System		2.58	\$109,278
M-606	1		Biogas Emergency Flare	0.6	1.68	\$12,737
M-612	1		Filter Precoat System		1.4	\$5,665
P-602	1	1	Anaerobic Reactor Feed Pump	0.79	2.8	\$46,587
P-606	1	1	Aerobic Digester Feed Pump	0.79	2.8	\$44,055
P-608	1		Aerobic Sludge Recycle Pump	0.79	1.4	\$3,848
P-610	1		Aerobic Sludge Pump	0.79	1.4	\$3,848
P-611	1	1	Aerobic Digestion Outlet Pump	0.79	2.8	\$43,766

Equip- ment ID	Number Required	Spares Nos	Equipment Name	Scaling Expo- nent	Install- ation Factor	Installed Cost in 2007\$
P-614	1	1	Sludge Filtrate Recycle Pump	0.79	2.8	\$8,474
P-616	1	1	Treated Water Pump	0.79	2.8	\$44,178
S-600	1		Bar Screen	0.3	1.2	\$162,338
S-614	1		Belt Filter Press	0.72	1.8	\$368,203
T-602	1		Equalization Basin	0.51	1.42	\$449,694
T-606	1		Anaerobic Digester	0.51	1.04	\$864,309
T-608	1		Aerobic Digester	1	1	\$393,578
T-610	1		Clarifier	0.51	1.96	\$310,050
A600					1.36	\$3,466,097
A-701	1		Denaturant In-line Mixer	0.48	1	\$2,608
A-720	1		CSL Storage Tank Agitator	0.51	1.2	\$6,645
A-760	1		CSL/DAP Day Tank Agitator	0.51	1.2	\$37,060
C-755	1		DAP Solids Feeder		1.3	\$6,892
P-701	2	1	Ethanol Product Pump	0.79	2.8	\$87,472
P-703	1	1	Sulfuric Acid Pump	0.79	2.8	\$218,520
P-704	1	1	Firewater Pump	0.79	2.8	\$143,059
P-710	1	1	Gasoline Pump	0.79	2.8	\$31,139
P-720	1	1	CSL Pump	0.79	2.8	\$124,937
P-750	1	1	Cellulase Pump	0.79	2.8	\$169,900
P-755	1		DAP Unloading Blower	0.5	1.4	\$39,511
P-760	1	1	CSL/DAP Pump	0.79	2.8	\$124,937
S-755	1		DAP Vent Baghouse	1	1.5	\$3,565
T-701	2		Ethanol Product Storage Tank	0.51	1.4	\$639,755
T-703	1		Sulfuric Acid Storage Tank	0.51	1.2	\$158,168
T-704	1		Firewater Storage Tank	0.51	1.4	\$320,472
T-709	1		Propane Storage Tank	0.72	1.4	\$47,719
T-710	1		Gasoline Storage Tank	0.51	1.4	\$77,841
T-720	1		CSL Storage Tank	0.79	1.4	\$312,699
T-750	2		Cellulase Storage Tank	0.79	1.4	\$448,617
T-755	1		DAP Storage Bin	0.44	1.3	\$28,054
T-760	1		CSL/DAP Day Tank	0.79	1.4	\$140,866
A700					1.6	\$3,170,437
H-801	1		Burner Combustion Air Preheater	0.6	1.5	\$1,507,493

Equip- ment ID	Number Required	Spares Nos	Equipment Name	Scaling Expo- nent	Installa- tion Factor	Installed Cost in 2007\$
H-811	1		BFW Preheater	0.68	2.1	\$110,325
M-803	1		Fluidized Bed Combustion Reactor	0.75	1.3	\$30,271,886
M-804	1		Combustion Gas Baghouse	0.58	1.5	\$4,141,941
M-811	1		Turbine/Generator	0.71	1.5	\$16,413,244
M-820	1		Hot Process Water Softener System	0.82	1.3	\$2,120,311
M-830	1		Hydrazine Addition Pkg.	0.6	1	\$26,150
M-832	1		Ammonia Addition Pkg	0.6	1	\$26,150
M-834	1		Phosphate Addition Pkg.	0.6	1	\$26,150
P-804	2		Condensate Pump	0.79	2.8	\$101,233
P-811	2		Turbine Condensate Pump	0.79	2.8	\$81,525
P-824	2		Deaerator Feed Pump	0.79	2.8	\$51,853
P-826	5		BFW Pump	0.79	2.8	\$463,890
P-828	2		Blowdown Pump	0.79	2.8	\$38,189
P-830	1		Hydrazine Transfer Pump	0.79	2.8	\$19,955
T-804	1		Condensate Collection Tank	0.71	1.4	\$9,339
T-824	1		Condensate Surge Drum	0.72	1.7	\$103,397
T-826	1		Deaerator	0.72	2.8	\$535,911
T-828	1		Blowdown Flash Drum	0.72	2.8	\$34,684
T-830	1		Hydrazine Drum	0.93	1.7	\$27,083
A800					1.4	\$56,110,709
M-902	1		Cooling Tower System	0.78	1.2	\$2,048,054
M-904	2	1	Plant Air Compressor	0.34	1.3	\$1,509,336
M-910	1		CIP System	0.6	1.2	\$157,225
P-902	1	1	Cooling Water Pump	0.79	2.8	\$1,866,538
P-912	1	1	Make-up Water Pump	0.79	2.8	\$73,506
P-914	2	1	Process Water Circulating Pump	0.79	2.8	\$97,975
S-904	1	1	Instrument Air Dryer	0.6	1.3	\$43,558
T-902	3		Prehydrolysis Filter Air Receiver	0.72	1.2	\$75,098
T-904	1		Plant Air Receiver	0.72	1.3	\$17,673
T-905	4		Product Recovery Filter Air Receiver	0.72	1.2	\$111,806
T-914	1		Process Water Tank	0.51	1.4	\$315,082
A900					1.5	\$6,315,850
					1.50	\$164,104,477

Table 2.E.2 Hot water pretreatment equipment list

Equipment Number	Number Required	Number Spares	Equipment Name	Scaling Exponent	Installation Factor	Installed Cost in 2007\$
C-101	2		Bale Transport Conveyor	0.6	1.62	\$1,862,521
C-102	2		Bale Unwrapping Conveyor	0.6	1.19	\$513,056
C-103	1		Belt Press Discharge Conveyor	0.6	1.89	\$135,809
C-104	4		Shredder Feed Conveyor	0.6	1.38	\$475,978
M-101	2		Truck Scales	0.6	2.47	\$241,380
M-102	4	1	Truck Unloading Forklift	1	1	\$135,982
M-103	4		Bale Moving Forklift	1	1	\$108,785
M-104	2		Corn Stover Wash Table	0.6	2.39	\$714,426
M-105	4		Shredder	0.6	1.38	\$2,395,754
M-106	1		Concrete Feedstock-Storage Slab	1	2.2	\$1,497,976
M-107	1		Polymer Feed System	0.6	2.28	\$98,300
P-101	2	1	Wash Table Pump	0.79	3.87	\$341,732
P-102	2	1	Wash Water Pump	0.79	5.19	\$343,719
P-103	1	1	Clarifier Underflow Pump	0.79	13.41	\$236,828
P-104	1	1	Clarified Water Pump	0.79	7.07	\$312,151
P-105	1	1	Belt Press Sump Pump	0.79	2.92	\$163,301
S-101	1		Clarifier Thickener	0.6	1.51	\$292,959
S-102	1		Belt Press	0.6	1.25	\$179,641
S-103	1		Magnetic Separator	0.6	1.3	\$19,537
T-101	1		Wash Water Tank	0.51	2.8	\$198,945
T-102	1		Clarifier Thickener Tank	0.51	3.04	\$583,192
A100					1.81	\$10,851,970
A-200	1	0	Mix Tank Agitator	0.51	1.2	\$102,050
A-201	1	0	Flash Tank Agitator	0.51	1.2	\$71,487
H-200	1	0	Pretreatment Cooler	0.59	1.53	\$270,192
H-202	1	0	Pretreatment Cross Exchanger	0.59	1.53	\$3,155,279
H-203	1	0	Pretreatment Trim Heater	0.59	1.53	\$1,134,417
P-200	1	1	Pretreatment Feed Pump	0.7	2.8	\$624,993
P-201	1	1	Fermentation Feed Pump	0.7	2.8	\$624,993
R-200	1	0	Pretreatment Reactor	0.78	2.1	\$311,417
T-200	1	0	Mix Tank	0.71	2.1	\$208,881
T-201	1	0	Flash Tank	0.71	2.1	\$208,881
A200					1.71	\$6,712,589

A-300	16		Ethanol Fermentor Agitator		1.2	\$520,003
A-301	1		Seed Hold Tank Agitator	0.51	1.2	\$23,708
A-304	2		4th Seed Vessel Agitator	0.51	1.2	\$43,653
A-305	2		5th Seed Vessel Agitator	0.51	1.2	\$39,064
A-306	2		Beer Surge Tank Agitator	0.51	1.2	\$191,343
A-310	40		Saccharification Tank Agitator		1.2	\$1,300,008
F-300	8		Ethanol Fermentor		1.2	\$6,389,179
F-301	2		1st Seed Fermentor		2.8	\$111,904
F-302	2		2nd Seed Fermentor		2.8	\$248,168
F-303	2		3rd Seed Fermentor		2.8	\$617,376
F-304	2		4th Seed Fermentor	0.93	1.2	\$164,593
F-305	2		5th Seed Fermentor	0.51	1.2	\$545,139
H-300	8	1	Fermentation Cooler	0.78	2.1	\$127,424
H-301	1	1	Hydrolyzate Heater	0.68	2.1	\$21,868
H-302	3		Saccharified Slurry Cooler	0.78	2.1	\$0
H-304	1		4th Seed Fermentor Coil	0.83	1.2	\$9,351
H-305	1		5th Seed Fermentor Coil	0.98	1.2	\$58,864
H-310	20	1	Saccharification Cooler	0.78	2.1	\$593,458
P-300	8	1	Fermentation Recirc/Transfer Pump	0.79	2.8	\$341,065
P-301	1	1	Seed Hold Transfer Pump	0.7	1.4	\$100,776
P-302	2		Seed Transfer Pump	0.7	1.4	\$245,598
P-306	1	1	Beer Transfer Pump	0.79	2.8	\$177,759
P-310	20	1	Saccharification Recirc/Transfer Pump	0.79	2.8	\$1,602,597
T-301	1		Seed Hold Tank	0.51	1.2	\$299,130
T-306	1		Beer Storage Tank	0.71	1.2	\$503,801
T-310	20		Saccharification Tank		1.2	\$15,972,948
A300					1.3	\$30,248,776
A-530	1		Recycled Water Tank Agitator	0.51	1.3	\$14,459
C-501	1		Lignin Wet Cake Screw	0.78	1.4	\$44,201
D-501	1		Beer Column	0.68	2.1	\$1,846,808
D-502	1		Rectification Column	0.68	2.1	\$1,627,838
E-501	2		1st Effect Evaporation	0.68	2.1	\$5,468,760
E-502	1		2nd Effect Evaporation	0.68	2.1	\$2,187,373
E-503	2		3rd Effect Evaporation	0.68	2.1	\$4,374,747
H-201	2	1	Beer Column Feed Economizer	0.68	2.1	\$625,829

H-501	1	1	Beer Column Reboiler	0.68	2.1	\$1,133,517
H-502	1		Rectification Column Reboiler	0.68	2.1	\$86,040
H-504	1		Beer Column Condenser	0.68	2.1	\$56,180
H-505	1		Start-up Rect. Column Condenser	0.68	2.1	\$266,237
H-512	1	1	Beer Column Feed Interchanger	0.68	2.1	\$136,461
H-517	1	1	Evaporator Condenser	0.68	2.1	\$1,136,912
M-503	1		Molecular Sieve (9 pieces)	0.7	1	\$2,778,195
P-501	1	1	Beer Column Bottoms Pump	0.79	2.8	\$437,234
P-503	1	1	Beer Column Reflux Pump	0.79	2.8	\$6,302
P-504	1	1	Rectification Column Bottoms Pump	0.79	2.8	\$48,209
P-505	1	1	Rectification Column Reflux Pump	0.79	2.8	\$39,021
P-511	2	1	1st Effect Pump	0.79	2.8	\$302,621
P-512	1	1	2nd Effect Pump	0.79	2.8	\$168,626
P-513	2	1	3rd Effect Pump	0.79	2.8	\$131,116
P-514	1	1	Evaporator Condensate Pump	0.79	2.8	\$158,090
P-515	1		Scrubber Bottoms Pump	0.79	2.8	\$8,300
P-530	1	1	Recycled Water Pump	0.79	2.8	\$133,302
S-505	4		Pneumapress Filter	0.6	1.04	\$7,056,264
T-503	1		Beer Column Reflux Drum	0.93	2.1	\$19,159
T-505	1		Rectification Column Reflux Drum	0.72	2.1	\$139,660
T-512	1		Vent Scrubber	0.78	2.1	\$223,306
T-514	1		Evaporator Condensate Drum	0.93	2.1	\$168,031
T-530	1		Recycled Water Tank	0.745	1.4	\$44,006
A500					1.58	\$30,866,801
A-602	1		Equalization Basin Agitator	0.51	1.2	\$24,138
A-606	1		Anaerobic Agitator	0.51	1.2	\$10,240
A-608	16		Aerobic Lagoon Agitator	0.51	1.4	\$124,207
C-614	1		Aerobic Sludge Screw	0.78	1.4	\$284
H-602	1		Anaerobic Digester Feed Cooler	0.74	2.1	\$163,214
M-604	1		Nutrient Feed System		2.58	\$109,278
M-606	1		Biogas Emergency Flare	0.6	1.68	\$2,576
M-612	1		Filter Precoat System		1.4	\$5,665
P-602	1	1	Anaerobic Reactor Feed Pump	0.79	2.8	\$31,611
P-606	1	1	Aerobic Digester Feed Pump	0.79	2.8	\$29,957
P-608	1		Aerobic Sludge Recycle Pump	0.79	1.4	\$530
P-610	1		Aerobic Sludge Pump	0.79	1.4	\$530

P-611	1	1	Aerobic Digestion Outlet Pump	0.79	2.8	\$29,697
P-614	1	1	Sludge Filtrate Recycle Pump	0.79	2.8	\$1,166
P-616	1	1	Treated Water Pump	0.79	2.8	\$30,140
S-600	1		Bar Screen	0.3	1.2	\$140,106
S-614	1		Belt Filter Press	0.72	1.8	\$90,007
T-602	1		Equalization Basin	0.51	1.42	\$350,093
T-606	1		Anaerobic Digester	0.51	1.04	\$256,075
T-608	1		Aerobic Digester	1	1	\$241,580
T-610	1		Clarifier	0.51	1.96	\$241,717
A600					1.44	\$1,882,810
A-701	1		Denaturant In-line Mixer	0.48	1	\$2,243
A-720	1		CSL Storage Tank Agitator	0.51	1.2	\$7,615
A-760	1		CSL/DAP Day Tank Agitator	0.51	1.2	\$42,469
C-755	1		DAP Solids Feeder		1.3	\$6,892
P-701	2	1	Ethanol Product Pump	0.79	2.8	\$68,256
P-703	1	1	Sulfuric Acid Pump	0.79	2.8	\$0
P-704	1	1	Firewater Pump	0.79	2.8	\$111,632
P-710	1	1	Gasoline Pump	0.79	2.8	\$24,296
P-720	1	1	CSL Pump	0.79	2.8	\$154,294
P-750	1	1	Cellulase Pump	0.79	2.8	\$169,896
P-755	1		DAP Unloading Blower	0.5	1.4	\$45,304
P-760	1	1	CSL/DAP Pump	0.79	2.8	\$154,294
S-755	1		DAP Vent Baghouse	1	1.5	\$4,687
T-701	2		Ethanol Product Storage Tank	0.51	1.4	\$545,091
T-703	1		Sulfuric Acid Storage Tank	0.51	1.2	\$0
T-704	1		Firewater Storage Tank	0.51	1.4	\$273,051
T-706	1		Ammonia Storage Tank	0.72	1.4	\$521,845
T-709	1		Propane Storage Tank	0.72	1.4	\$38,064
T-710	1		Gasoline Storage Tank	0.51	1.4	\$66,318
T-720	1		CSL Storage Tank	0.79	1.4	\$386,173
T-750	2		Cellulase Storage Tank	0.79	1.4	\$448,608
T-755	1		DAP Storage Bin	0.44	1.3	\$31,643
T-760	1		CSL/DAP Day Tank	0.79	1.4	\$173,965
A700					1.6	\$3,276,638
H-801	1		Burner Combustion Air Preheater	0.6	1.5	\$1,755,061

H-811	1		BFW Preheater Fluidized Bed	0.68	2.1	\$127,920
M-803	1		Combustion Reactor Combustion Gas	0.75	1.3	\$35,609,795
M-804	1		Baghouse	0.58	1.5	\$4,666,919
M-811	1		Turbine/Generator	0.71	1.5	\$19,140,921
M-820	1		Hot Process Water Softener System	0.82	1.3	\$2,659,912
M-830	1		Hydrazine Addition Pkg.	0.6	1	\$29,778
M-832	1		Ammonia Addition Pkg	0.6	1	\$29,778
M-834	1		Phosphate Addition Pkg.	0.6	1	\$29,778
P-804	2		Condensate Pump	0.79	2.8	\$161,668
P-811	2		Turbine Condensate Pump	0.79	2.8	\$56,534
P-824	2		Deaerator Feed Pump	0.79	2.8	\$64,512
P-826	5		BFW Pump	0.79	2.8	\$550,435
P-828	2		Blowdown Pump	0.79	2.8	\$45,314
P-830	1		Hydrazine Transfer Pump	0.79	2.8	\$23,677
T-804	1		Condensate Collection Tank	0.71	1.4	\$14,225
T-824	1		Condensate Surge Drum	0.72	1.7	\$126,173
T-826	1		Deaerator	0.72	2.8	\$626,327
T-828	1		Blowdown Flash Drum	0.72	2.8	\$40,536
T-830	1		Hydrazine Drum	0.93	1.7	\$33,125
A800					1.4	\$65,792,387
M-902	1		Cooling Tower System	0.78	1.2	\$2,168,496
M-904	2	1	Plant Air Compressor	0.34	1.3	\$1,509,336
M-910	1		CIP System	0.6	1.2	\$157,225
P-902	1	1	Cooling Water Pump	0.79	2.8	\$1,977,754
P-912	1	1	Make-up Water Pump Process Water	0.79	2.8	\$78,110
P-914	2	1	Circulating Pump	0.79	2.8	\$160,536
S-904	1	1	Instrument Air Dryer Prehydrolysis Filter	0.6	1.3	\$43,558
T-902	3		Air Receiver	0.72	1.2	\$0
T-904	1		Plant Air Receiver	0.72	1.3	\$17,673
T-905	4		Product Recovery Filter Air Receiver	0.72	1.2	\$111,806
T-914	1		Process Water Tank	0.51	1.4	\$433,380
					1.5	\$6,657,873

Appendix 2.F Cost Summaries

Table 2.F.1 Cost summary for the dilute acid pretreatment scenario

Ethanol Production Process Engineering Analysis			
UCR Dilute Acid - Corn Stover, Current Case Dilute Acid Prehydrolysis with Saccharification and Co-Fermentation All Values in 2007\$			
Minimum Ethanol Selling Price	\$3.40		
Ethanol Production (MM Gal. / Year)	53.4	Ethanol at 68°F	
Ethanol Yield (Gal / Dry US Ton Feedstock)	69.2		
Feedstock Cost \$/Dry US Ton	\$75		
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	\$25,400,000	Biomass to Boiler	0.0
Neutralization/Conditioning	\$10,800,000	CSL	16.0
Saccharification & Fermentation	\$21,800,000	Cellulase	69.5
Distillation and Solids Recovery	\$26,100,000	Other Raw Materials	17.8
Wastewater Treatment	\$3,500,000	Waste Disposal	12.7
Storage	\$3,200,000	Electricity	-21.9
Boiler/Turbogenerator	\$56,100,000	Fixed Costs	18.5
Utilities	\$6,300,000	Capital Depreciation	30.5
Total Installed Equipment Cost	\$164,100,000	Average Income Tax	26.7
		Average Return on Investment	62.2
Added Costs	\$211,800,000	Operating Costs (\$/yr)	
(% of TCI)	56%	Feedstock	\$57,900,000
Working Capital	49,030,000	Biomass to Boiler	\$0
Total Capital Investment	\$375,900,000	CSL	\$8,500,000
		Cellulase	\$37,100,000
Installed Equipment Cost/Annual Gallon	\$3.07	Other Raw Matl. Costs	\$9,500,000
Total Project Investment/Annual Gallon	\$7.04	Waste Disposal	\$6,800,000
		Electricity	\$11,700,000
Capital Charge Factor	0.170	Fixed Costs	\$9,900,000
Denatured Fuel Prod. (MMgal / yr)	55.9	Capital Depreciation	\$16,300,000
Denatured Fuel Min. Sales Price	\$3.29	Average Income Tax	\$14,300,000
Denaturant Cost (\$/gal denaturant)	\$0.739	Average Return on Investment	\$33,300,000
Maximum Yields (100% of Theoretical)		Excess Electricity (KWH/gal)	4.06
Ethanol Production (MM Gal/yr)	82.5	Plant Electricity Use (KWH/gal)	2.23
Theoretical Yield (Gal/ton)	106.9	Plant Steam Use (kg steam/gal)	17.5
Current Yield (Actual/Theoretical)	65%	Boiler Feed -- LHV (Btu/lb)	2,209
		Boiler Feed -- Water Fraction	0.542

Table 2.F.2 Cost summary for the dilute acid pretreatment scenario using pilot scale parameters

Ethanol Production Process Engineering Analysis			
UCR Dilute Acid - Pretreatment yields based on NREL FY08 SOT Dilute Acid Prehydrolysis with Saccharification and Co-Fermentation All Values in 2007\$			
Minimum Ethanol Selling Price	\$3.60		
Ethanol Production (MM Gal. / Year)	50.8	Ethanol at 68°F	
Ethanol Yield (Gal / Dry US Ton Feedstock)	65.8		
Feedstock Cost \$/Dry US Ton	\$75		
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	\$25,200,000	Biomass to Boiler	0.0
Neutralization/Conditioning	\$12,800,000	CSL	16.5
Saccharification & Fermentation	\$21,800,000	Cellulase	73.1
Distillation and Solids Recovery	\$25,700,000	Other Raw Materials	18.8
Wastewater Treatment	\$5,800,000	Waste Disposal	13.0
Storage	\$3,100,000	Electricity	-24.8
Boiler/Turbogenerator	\$57,600,000	Fixed Costs	19.8
Utilities	\$6,600,000	Capital Depreciation	33.3
Total Installed Equipment Cost	\$169,400,000	Average Income Tax	29.0
		Average Return on Investment	67.5
Added Costs	\$219,100,000	Operating Costs (\$/yr)	
(% of TCI)	56%	Feedstock	\$57,900,000
Working Capital	50,670,000	Biomass to Boiler	\$0
Total Capital Investment	\$388,500,000	CSL	\$8,400,000
		Cellulase	\$37,100,000
Installed Equipment Cost/Annual Gallon	\$3.34	Other Raw Matl. Costs	\$9,600,000
Total Project Investment/Annual Gallon	\$7.65	Waste Disposal	\$6,600,000
			-
Capital Charge Factor	0.169	Electricity	\$12,600,000
Denatured Fuel Prod. (MMgal / yr)	53.1	Fixed Costs	\$10,100,000
Denatured Fuel Min. Sales Price	\$3.47	Capital Depreciation	\$16,900,000
Denaturant Cost (\$/gal denaturant)	\$0.739	Average Income Tax	\$14,700,000
		Average Return on Investment	\$34,200,000
Maximum Yields (100% of Theoretical)		Excess Electricity (KWH/gal)	4.58
Ethanol Production (MM Gal/yr)	82.5	Plant Electricity Use (KWH/gal)	2.39
Theoretical Yield (Gal/ton)	106.9	Plant Steam Use (kg steam/gal)	
Current Yield (Actual/Theoretical)	62%	18.2	
		Boiler Feed -- LHV (Btu/lb)	2,286
		Boiler Feed -- Water Fraction	0.539

Table 2.F.3 Cost summary for the hot water pretreatment scenario

Ethanol Production Process Engineering Analysis			
Hot Water - Corn Stover, Current Case Hot Water Prehydrolysis with Saccharification and Co-Fermentation All Values in 2007\$			
Minimum Ethanol Selling Price	\$4.29		
Ethanol Production (MM Gal. / Year)	39.0	Ethanol at 68°F	
Ethanol Yield (Gal / Dry US Ton Feedstock)	50.6		
Feedstock Cost \$/Dry US Ton	\$75		
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	148.4
Pretreatment	\$6,700,000	Biomass to Boiler	0.0
Neutralization/Conditioning	\$0	CSL	28.5
Saccharification & Fermentation	\$30,200,000	Cellulase	95.1
Distillation and Solids Recovery	\$30,900,000	Other Raw Materials	5.1
Wastewater Treatment	\$1,900,000	Waste Disposal	3.5
Storage	\$3,300,000	Electricity	-29.0
Boiler/Turbogenerator	\$65,800,000	Fixed Costs	24.5
Utilities	\$6,700,000	Capital Depreciation	36.4
Total Installed Equipment Cost	\$156,300,000	Average Income Tax	32.9
		Average Return on Investment	83.7
Added Costs	\$128,000,000	Operating Costs (\$/yr)	
(% of TCI)	45%	Feedstock	\$57,900,000
Working Capital	42,600,000	Biomass to Boiler	\$0
Total Capital Investment	\$284,300,000	CSL	\$11,100,000
		Cellulase	\$37,100,000
Installed Equipment Cost/Annual Gallon	\$4.00	Other Raw Matl. Costs	\$2,000,000
Total Project Investment/Annual Gallon	\$9.25	Waste Disposal	\$1,400,000
Capital Charge Factor	0.210	Electricity	-\$11,300,000
Denatured Fuel Prod. (MMgal / yr)	40.9	Fixed Costs	\$9,600,000
Denatured Fuel Min. Sales Price	\$4.13	Capital Depreciation	\$14,200,000
Denaturant Cost (\$/gal denaturant)	\$0.739	Average Income Tax	\$12,800,000
		Average Return on Investment	\$32,700,000
Maximum Yields (100% of Theoretical)		Excess Electricity (KWH/gal)	5.37
Ethanol Production (MM Gal/yr)	82.6	Plant Electricity Use (KWH/gal)	3.30
Theoretical Yield (Gal/ton)	106.9	Plant Steam Use (kg steam/gal)	40.7
Current Yield (Actual/Theoretical)	47%	Boiler Feed -- LHV (Btu/lb)	2,354
		Boiler Feed -- Water Fraction	0.529
		Specific Operating Conditions	
		Saccharification Time (days)	5.0
		Conversion Cellulose --> Glucose	0.8997
		Fermentation Time (days)	2.0

Chapter 3 Comparison of Cellulase Enzyme Production Schemes as Alternatives to Purchasing Enzymes

Introduction

The saccharification of cellulose and hemicelluloses to monosaccharides is a critical step in the biochemical production of ethanol from lignocellulosic feedstocks. Some biomass pretreatment technologies, using strong acid catalysts, hydrolyze a significant fraction of hemicelluloses, while alkali pretreatments typically hydrolyze only a small portion of hemicelluloses to monomers. However, in both cases, only a small fraction of cellulose is hydrolyzed to monomers. It is possible to hydrolyze a significant portion of cellulose through a single concentrated acid—typically H_2SO_4 —treatment or a second stage of dilute acid treatment following pretreatment. The use of acids presents a number of difficulties in downstream processing such as the need for neutralizing the biomass slurry before fermentation and the undesired production of fermentation inhibitors such as furfural and hydroxy-methylfurfural through sugar dehydration. Concentrated acid hydrolysis is also likely to require acid recycle for economical production (16). Therefore, post-pretreatment hydrolysis of polysaccharides using cellulase and hemicellulase enzymes offers some advantages in processing. However, enzymes are commonly cited as one of the most significant expenses in cellulosic ethanol production, and are viewed as a critical research area for economically viable production (17). Table 3.1 presents results of past techno-economic studies in regards to the cost of enzymes or enzyme production. While plant size, process assumptions, technologies, and feedstocks vary considerably between these studies, it is clear that enzymes are a significant factor in the cost of producing ethanol.

As an alternative to purchasing enzymes, the production of enzymes at the ethanol plant may offer a number of cost advantages. Currently, industrial production of cellulase relies on high value substrates such as lactose. On-site enzyme production could utilize a fraction of the pretreated biomass—a relatively low value feedstock. Stabilizers are commonly added to enzyme broth to mitigate the problem of decreasing enzyme viability during storage. Also, purchased enzyme broth is concentrated to reduce bulk during transportation. Because of immediate use of enzymes produced on-site, broth concentration and stabilizer addition are not necessary. The co-location of utilities for both enzyme and ethanol production may also offer economy-of-scale advantages.

The economic competitiveness of on-site enzyme production with purchasing enzymes is likely to vary with the plant scale. This is due to fixed operating costs such as labor, overhead and maintenance which do not scale linearly with plant size, as well as the economy of scale associated

Table 3.1 Enzyme costs in previous lignocellulosic ethanol production models

Reference	Feedstock	Pretreatment	Scale	Enzyme Production- Percent of Operating Cost	Enzyme Production Area-% of Total Capital Cost	Enzyme Production- % of Total Production Cost*	Enzyme Cost, \$/gal EIOH
<i>On-Site Enzyme Models</i>							
Nygren and Sauter, 1991	Aspen Wood	Steam	500 dry MT/day	-	58	16	-
Woolley et al., 1999	Poplar	Dilute Acid	2000 ton/day	-	11.5	18	--
<i>Purchased Enzyme Models</i>							
Wingren et al., 2003	Softwood	SO ₂ Steam	587 MT/day	9-18	--	--	0.21-0.36
Aden et al., 2002	Com Stover	Dilute Acid	2000 ton/day	22.2	-	9	0.12
Fan and Lynd, 2007	Paper Sludge	-	15 dry MT/day	12.2	--	--	0.17
J. Fortman, Ch. 2	Com Stover	Dilute Acid	2000 MT/day	-	-	204	0.69
J. Fortman, Ch. 2	Com Stover	Hot Water	2000 MT/day	-	--	22.2	0.95

*This is different from % of operating cost in that it includes capital recovery

with increasing enzyme production equipment capacity. Zhuang et al. found that the cost of cellulase production for a stand-alone production facility showed some correlation with plant scale (18). A ninefold increase in plant scale resulted in a 36% decrease in unit cost of cellulase.

As a second alternative, production of enzymes at a central plant producing both ethanol and excess enzymes for export to satellite ethanol plants may also offer economic advantages. If the satellite plants are located within a relatively close proximity to the central plant so that regular enzyme shipments are received, stabilizers may not be needed. Increasing the scale of enzyme production to provide for multiple ethanol plants takes advantage of economies of scale in enzyme production.

This study analyzes the two alternative scenarios for supplying enzymes mentioned above. For the first scenario the relationship between plant scale and the economic competitiveness of on-site enzyme production as an alternative to purchasing enzymes is examined. For the second scenario the number of satellite plants is varied in order to assess the impact on the Minimum Ethanol Selling Price (MESP) as well as the economic competitiveness with purchasing enzymes.

Research efforts directed at reducing the cost of enzymes in bio-refining include increasing enzyme volumetric productivity, the use of cheaper substrates, enhancing enzyme stability for specific processes, producing enzymes with greater specific activity, and enzyme recycling (19) (20) (17). The US DOE-EERE Biomass Program estimates the current cost of enzymes to be \$0.10-0.25/gal EtOH with a goal of reducing the cost tenfold (21). Regardless of advancements in these areas, it is likely that a relationship will continue to exist between plant scale and the competitiveness of on-site enzyme production with purchasing enzymes. Although, that relationship may need to be re-evaluated upon future breakthroughs.

Because of the relatively near-term increases in renewable fuel mandates enacted in the US (3), it is important to evaluate renewable fuel production schemes in terms of the current state of technology in order to both set benchmarks and compare different production schemes proposed for short term commercialization. The parameters used in this analysis represent current publicly available data and the results are not meant to reflect the potential production cost reductions or future cost targets.

Methods

On-site Enzyme Production at Various Plant Scales

Bioethanol production and enzyme production are modeled using Aspen Plus. Two models are developed for each plant scale—one with on-site enzyme production and one without. The models are based on a previous model developed by the National Renewable Energy Laboratory (NREL) (12) (22). The pretreatment method modeled was dilute sulfuric acid treatment, along with Separate Hydrolysis and Fermentation (SHF). The process steps modeled include feedstock handling and washing, pretreatment, enzyme production, saccharification, fermentation, distillation and solids separation, wastewater treatment, and co-product combustion for combined heat and power generation. A detailed description of each process area may be found in Chapter 2. Figure 3.A.1 in Appendix 3.A shows a process flow diagram for the enzyme production area. The configuration of the remainder of the plant is the same as those in Appendix 2.A. However, the flow rates deviate from those shown in the process diagrams as plant scale is varied. The process parameters and yields for pretreatment are shown in Tables 3.2 and 3.3.

Table 3.1 Pretreatment process parameters

Parameters	
Acid Concentration (wt%)	1.9
Acid Loading (g acid/g dry biomass)	0.0443
Total Solids (wt%)	0.2959
Temperature (°C)	190
Pressure (atm)	11.4

Table 3.2 Dilute acid pretreatment conversions (13)

Cellulose to Glucose	0.063
Xylan to Oligomers	0.027
Xylan to Xylose	0.825
Mannan to Oligomers	0.027
Mannan to Mannose	0.825
Galactan to Oligomers	0.027
Galactan to Galactose	0.825
Arabinan to Oligomers	0.027
Arabinan to Arabinose	0.825

During enzymatic saccharification 91.1% of the cellulose remaining after pretreatment is converted to glucose, and 57.13% of xylan is converted to xylose (13). None of the glucose

oligomers or xylose oligomers are assumed to be converted to monomers. The conversion of glucose to ethanol during fermentation is 95%, and 75.6% of xylose is converted to ethanol. Hemicellulose sugars other than xylose are not converted to ethanol.

Enzyme production with the organism *Trichoderma reesei* occurs with pretreated biomass as the carbon source, with 9.2% of the pretreated biomass being diverted to the enzyme production area. Eleven production vessels are sequenced so that at any time one is filling, one is being sterilized, one is draining, and eight are in production. The seed is produced in three trains, each with three reactors. Each reactor in the seed train produces 5% seed for the subsequent reactor. Corn steep liquor and micronutrients are added to the cellulase production vessels and seed vessels. Ammonia is also added for pH control and provides additional nitrogen. Cellulase and seed production occur at 28°C. A residence time of 160 hours is used for cellulase production. The reactors are sparged with air at a rate of 0.577 vvm (volume of air at STP per unit reactor volume per minute). A list of simplified reactions as modeled and the conversions for *T. reesei* seed and cellulase production is shown in Table 3.4.

Table 3.3 *T. reesei* seed production and cellulase production reaction and conversions

Reaction	Saccharide Conversion
<i>T. Reesei</i> Seed Production	
2 Glucose + 7.452 O₂ + NH₃ → 9.935 H₂O + 7.652 CO₂ + 4.348 Cell Mass	1.0
2 Xylose + 5.452 O₂ + NH₃ → 7.935 H₂O + 5.652 CO₂ + 4.348 Cell Mass	1.0
2 Cellulose + 7.452 O₂ + NH₃ → 7.935 H₂O + 7.652 CO₂ + 4.348 Cell Mass	1.0
2 Xylan + 5.452 O₂ + NH₃ → 5.935 H₂O + 5.652 CO₂ + 4.348 Cell Mass	1.0
Enzyme Production	
2 Glucose + 7.452 O₂ + NH₃ → 9.935 H₂O + 7.652 CO₂ + 4.348 Cell Mass	0.47
2 Xylose + 5.452 O₂ + NH₃ → 7.935 H₂O + 5.652 CO₂ + 4.348 Cell Mass	0.47
2 Glucose + 8.459 O₂ + NH₃ → 10.793 H₂O + 8.552 CO₂ + 3.448 Enzyme	0.53
2 Xylose + 6.459 O₂ + NH₃ → 8.793 H₂O + 6.552 CO₂ + 3.448 Enzyme	0.53
2 Cellulose + 8.459 O₂ + NH₃ → 8.793 H₂O + 8.552 CO₂ + 3.448 Enzyme	1.0
2 Xylan + 6.459 O₂ + NH₃ → 6.793 H₂O + 6.552 CO₂ + 3.448 Enzyme	1.0

The cellulase requirement in saccharification is 31.3 mg enzyme/g cellulose in the untreated feed and the specific activity of the enzymes is assumed to be 600 FPU/g protein. The enzyme yield is

145.7 FPU/g cellulose plus glucose and 157.2 FPU/g xylan plus xylose. A more detailed description of the enzyme production scheme can be found in Wooley et al (1999).

Process models were developed for plants scales ranging from 500 to 3000 MT/day (tpd) for both the on-site enzyme scenario and the purchased enzyme scenario. Stream flow results from the Aspen Plus process model are used in the discounted cash flow analysis to calculate raw material costs, as well as equipment size and capital costs. Because the capital cost of most equipment does not scale linearly with size, Equation 3.1 is used to estimate equipment costs as plant scale varies.

$$\text{Equation 3.1} \quad C_{\text{new}} = C_0 \left(\frac{S_{\text{new}}}{S_0} \right)^n$$

where C_{new} is the scaled cost, C_0 is the original quoted cost, S_{new} is the value for the sizing attribute at the desired scale, S_0 is the value of the sizing attribute for the original price quote, and n is the scaling exponent (23). The price quotes are inflated to FY 2007 prices using the Chemical Engineering Purchased Equipment Index (24). The scaled capital cost is multiplied by an installation factor to estimate the installed cost. The original price quotes, installation factors, and scaling exponents were the same as those used in previous techno-economic studies from NREL (22) (12), and the original publications may be referenced to find the vendors who provided equipment quotes. The Total Capital Investment (TCI) was calculated using the same cost factors and methods described in Chapter 2.

Raw material costs were updated to FY 2007 estimates from those used in previous NREL studies (22) (12) using the Industrial Inorganic Chemical Index from the US Bureau of Labor Statistics (BLS) (25). Labor costs were updated to FY 2007 values using the BLS Labor Index for Chemical Production Workers (26).

Fixed operating costs include salaries, overhead, maintenance, and insurance. Overhead is estimated as 60% of salaries, maintenance is 2% of installed equipment cost, and insurance is 1.5% of the fixed capital investment (15). Total salaries are adjusted for plant scale using a scaling exponent of 0.25 (15), with the corn stover feed rate used as the scaling attribute.

The discounted cash flow analysis is performed to find the Minimum Ethanol Selling Price (MESP). The MESP was found by iterating the value of the ethanol price so that the net present value of the project is zero. The major economic assumptions used in the discounted cash flow analysis are shown in Table 3.5.

Table 3.4 Discounted cash flow analysis assumptions

Economic Parameter	
Return on Investment	10%
Project Life	20 yrs
Income Tax Rate	39%
General Plant Depreciation Period	7 yrs
Steam/Elec. Generation Depreciation Period	20 yrs
Base Case Feedstock Price	\$75/mt

*Total Installed Equipment Cost, **Total Capital Investment (equipment plus indirect costs)

The MACRS depreciation method is used and the heat and power plant is depreciated at a different rate than the rest of the plant. The plant is operated at full capacity for 350 days per year.

In one scenario the delivered feedstock price remains constant over the range of plant scales. In a second scenario the feedstock cost is varied to account for changes in transportation cost as the average transportation distance changes with plant scale. Assumptions are made regarding the size of the collection area needed to provide the plant with the necessary corn stover for continuous operation. These assumptions are shown in Table 3.6.

Table 3.5 Corn stover collection area and transport cost assumptions

Average Corn Yield	150 bu/ac
Harvest Index ()	1
Percent of Land as Cropland	90%
Percent of Cropland with Corn	50%
Percent of Stover Collected	25%
Tortuosity Factor (27)	1.5
Transport Cost (27)	\$0.71/ton-mile

The assumptions are meant as a general representation of stover availability in the Corn Belt region. However, significant spatial variation in stover availability exists throughout the Corn Belt due to such factors as corn yield and land slope. Using the assumptions in Table 3.6 the required stover collection area is calculated. The average transportation distance for a circular collection area is estimated using Equation 3.2 (27)

Equation 3.2

$$\bar{r} = \frac{2}{3} \tau \sqrt{\frac{A}{\pi}}$$

Where \bar{r} is the average transportation distance, τ is the tortuosity, and A is the collection area. Based on the average transportation distance, the average transportation cost per ton is estimated using a cost of \$0.71/ton-mile. A cost of \$75/ton, including delivery cost, is assumed for the 2000 tpd plant. The delivered biomass cost per ton is increased for plants larger than 2000 tpd and decreased for smaller plants according to the estimated average delivery cost associated with each scale. Table 3.7 shows the delivered corn stover cost over the range of plant scales.

Table 3.6 Delivered corn stover cost with varying plant scale

Plant Scale, tpd	Delivered Stover Cost, \$/ton
500	65.29
1000	69.31
1500	72.40
2000	75.00
2500	77.29
3000	79.36

The range of plant scales chosen for this study is on the lower end of most estimates for optimal plant scales estimated for future bioethanol production scenarios. The lower range is used here because the difficulties of acquisition, logistics, and storage associated with collecting large amounts of biomass may necessitate that the first plants to be built are smaller than the estimated optimal size. Previous estimates of optimal plant size range from 3800 to 8000 tpd (28) (22) (29). The range of plant scales chosen in this study is 500 to 3000 tpd.

Central Enzyme Production for Distribution to Satellite Plants

Two process models are used for this scenario. The first process model simulates the plant producing both enzymes and ethanol. This process model is identical to the on-site enzyme model described above, with the exceptions that additional biomass slurry from the pretreatment area is diverted to the enzyme production area, and the enzyme stream for the satellite plants is concentrated prior to transportation. The water removed from the enzyme stream during concentration is treated in the wastewater treatment area. The amount of slurry diverted is adequate to provide feedstock for enzyme production for the satellite plants. The second process model is for the satellite plants producing only ethanol. This model is identical to the dilute acid pretreatment model described in

Chapter 2, which receives purchased enzymes from off-site. Two cases are analyzed with the first case assuming the central plant supplies enzymes to two satellite plants and the second assuming there are four satellite plants. Both the central plant and satellite plant are assumed to receive 2000 tpd of corn stover.

The enzyme broth concentration is modeled as vacuum filtration described by Knutsen and Davis (19), where the resulting filter cake retains most of the enzymes due to cellulase's strong adsorption affinity to spent corn stover hydrolyzate. The filter cake is assumed to contain 60% moisture, with the solids being primarily lignin, insoluble cellulose and hemicellulose components not consumed by *T. reesei*, cellulase, and *T. reesei* cell mass. Because of the short transportation distance and limited storage duration, no enzyme stabilizers are added to the filter cake. The cost of the filtration unit is also taken from Knutsen and Davis. The capital cost of the filtration unit which processes 100,000 kg/hr of slurry is \$2,110,000 (2004\$). The cost is updated to 2007 dollars and the cost is scaled using Equation 3.1 with a scaling exponent of 0.7. An installation factor of 2.5 is assumed. The total installed cost for the unit (2007\$) is \$12,074,000.

The number of enzyme production vessels is scaled linearly with the enzyme production requirement rather than assuming an increase in the volume of the vessels. This is done because of the potential mass transfer issues that can arise with large scale bioreactors. The bioreactors are assumed to be 1000 m³ which is the same as the NREL study on which this model is based.

The enzyme transportation cost from the central plant to the satellite plants is estimated at \$0.31/ton-mile. This is based on the Bureau of Transportation Statistics data for truck freight revenue per ton-mile (30). However, data is unavailable after 2001, so it is further updated to 2007 dollars using the Consumer Price Index. Using the assumptions listed in Table 3.6 the transportation distance between two adjacent plants with circular collection areas is 54.3 miles. It is assumed that one truck carries 40 tons of enzyme cake per trip. The mass of enzyme cake needed by each satellite plant is 140 ton/day, requiring four trips.

The equipment costing methods, operating costs, and economic assumptions are identical to those described for the first scenario above. The total operating and capital costs from both process models are combined into a single spreadsheet and the MESP is calculated for all the plants in aggregate. For example, for the case with one central plant and two satellite plants, the combined MESP for all three plants was calculated. The Equivalent Purchased Enzyme Cost (EPEC, \$/gal EtOH produced) and the Equivalent Purchased Enzyme Price (EPEP, \$/kg protein) are calculated for both cases by adjusting the purchased enzyme cost for a stand-alone 2000 tpd plant so that the MESP is equal to that of the two alternative scenarios. As with the first study discussed above, the EPEC and EPEP

represent the enzyme price below which it is more economically favorable to purchase enzymes for a stand-alone plant, rather than produce enzymes under one of the enzyme production schemes described in this study.

Results and Discussion

Economic Competitiveness of On-site Enzyme Production with Purchased Enzymes

A clear trend exists between the economic competitiveness of on-site enzyme production with purchasing enzymes and plant scale. As expected, on-site enzyme production is more competitive as scale increases. However, the increase in cost advantage of on-site enzyme production diminishes as scale increases. To display this relationship, the price of purchased enzyme was adjusted so that the MESP of the purchased enzyme model was the same as the MESP of the on-site enzyme model. This was done for each plant scale. This EPEP is plotted vs. scale in Figure 3.1. Enzyme cost is sometimes reported in relevant literature in terms of \$/gallon of ethanol produced. While enzyme cost in these units is subjective to the process type and parameters used, it can be useful for viewing trends in this study, and for cautious comparisons with other studies. Figure 3.2 shows the EPEC per gallon of ethanol produced as a function of plant scale.

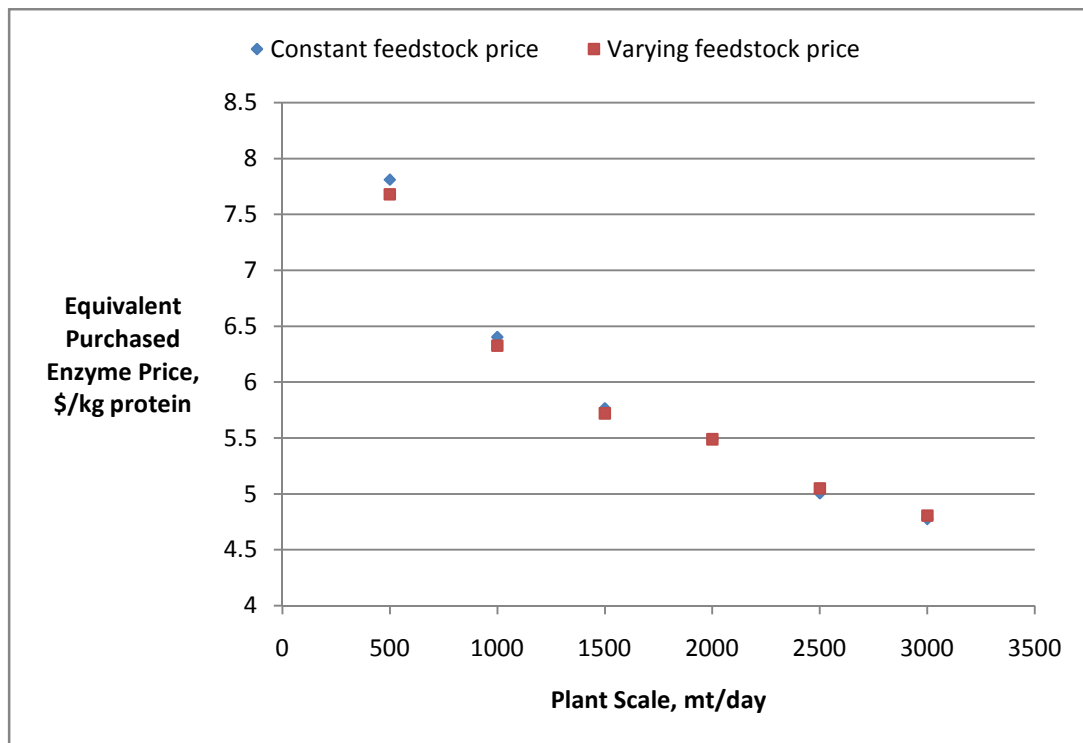


Figure 3.1 Equivalent purchased enzyme price with varying plant scale

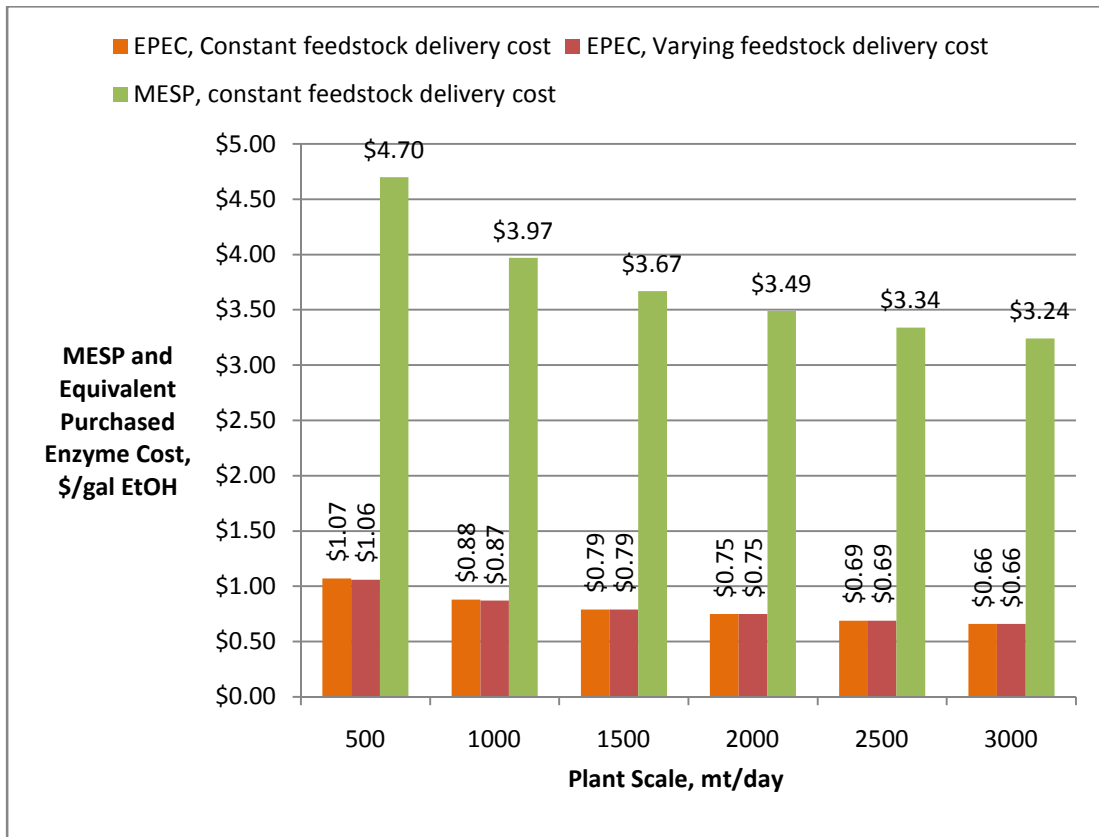


Figure 3.2 Equivalent purchased enzyme cost per gallon of ethanol with varying plant scale

It is clear from Figure 3.2 that with smaller scale plants, on-site enzyme production is competitive with purchasing enzymes only at very high enzyme prices. The diminishing advantage of larger scale can be seen in Figure 3.1, where a six-fold increase in scale (500 to 3000 mt/day) results in a decrease of the EPEP of 38.8%.

Figure 3.2 shows the MESP for the constant feedstock delivery cost scenario. At the 500 tpd plant scale the enzyme cost represents 22.5% of the MESP, and 20.4% of the MESP at 3000 tpd. These are slightly higher than the 16% and 18% of production cost attributed to enzyme production that were reported by Nguyen and Sadler (31) and Wooley et al (12).

For the case in which the delivered feedstock cost varies as a function of plant scale, there is very little difference from the case with constant feedstock delivery cost. For the 500 tpd plant the EPEC is \$1.06/gal EtOH compared to \$1.07/gal EtOH for the constant feedstock delivery price scenario. At the 3000 tpd scale the EPEC is \$0.66/gal EtOH for both scenarios.

Due to differences in scaling exponents between the enzyme production area and the remainder of the plant the installed capital cost for enzyme production equipment as a percentage of total capital cost decreases with increasing plant scale. This trend is shown in Figure 3.3. A significant

contributor to this is the relatively low scaling exponents associated with the cellulase fermentation vessels, fermentation vessel agitators, and air compressor unit, which comprise the three most significant capital costs of the enzyme production equipment. The scaling exponents for these units are 0.6, 0.5, and 0.34, respectively.

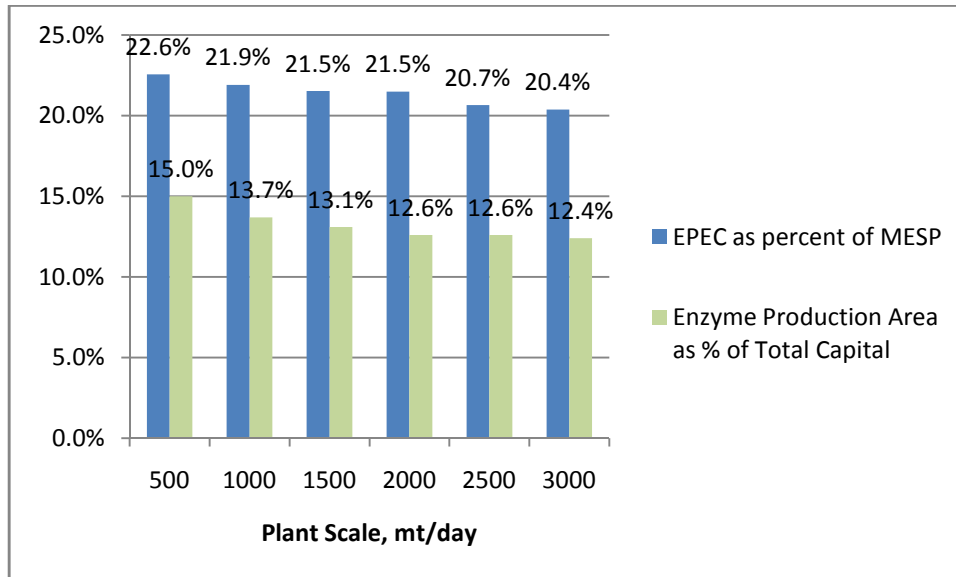


Figure 3.3 Installed capital cost of the enzyme production area as a percentage of total plant installed capital cost and EPEC as a percentage of MESP

This trend is also apparent in the decrease in EPEC as a percentage of MESP as the plant scale increases. At the 500 tpd scale the EPEC is 22.6% of MESP, while that falls to 20.4% at the 3000 tpd scale.

Fixed operating costs are also a factor in the decreasing EPEC which occurs with increasing plant scale. The fixed operating costs do not increase linearly with plant scale as shown in Figure 3.4.

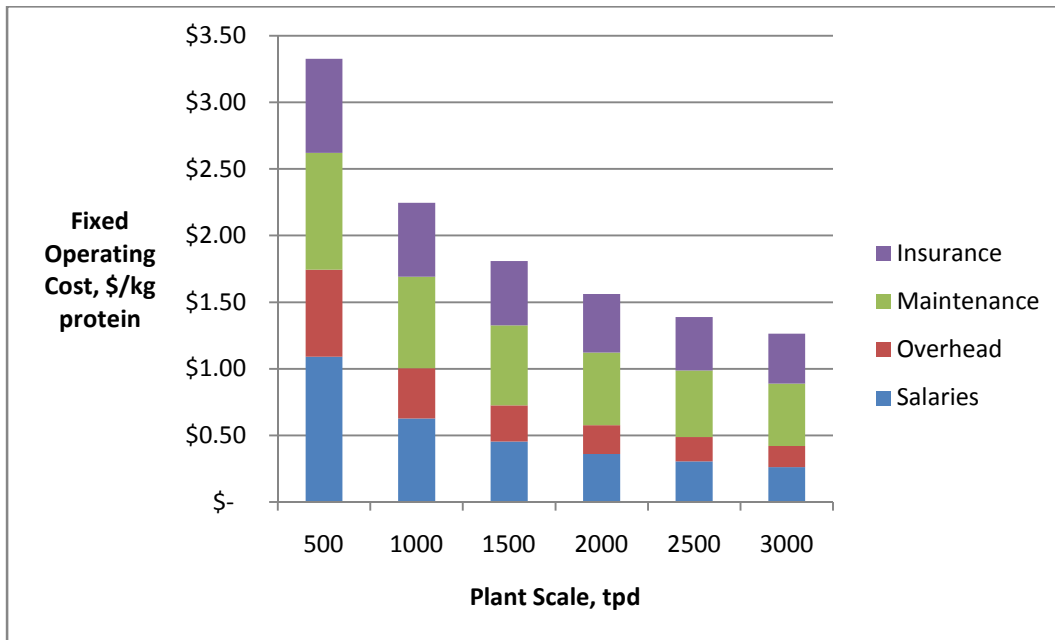


Figure 3.4 Fixed operating cost per kg of enzyme produced

The fixed operating costs shown in Figure 3.4 are for the complete plant and not only the enzyme production area, because some cost components such as management salaries cannot be clearly distributed between the plant sections. However, the trend clearly shows that fixed operating costs are a significant factor in the cost advantages of on-site enzyme production at larger scales. Salaries and overhead are the largest contributors to this trend, with a 76% reduction in their normalized costs as scale increases from 500 to 3000 tpd. This is due to the scaling exponent of 0.25 used to scale salaries with plant size.

Centralized Enzyme Production for Distribution to Satellite Plants

Production of both ethanol and excess enzymes for distribution to satellite plants is more favorable economically than on-site enzyme production at plant scales smaller than 2500 tpd for the plants modeled in the study discussed above. Results from the cases with two and four satellite plants are presented in Table 3.8.

Table 3.7 Aggregated MESP and enzyme cost

	EPEC, \$/gal EtOH	EPEP, \$/kg protein	MESP, \$/gal	FCI, \$/Annual Gal EtOH	Annual Operating Cost, \$/gal EtOH
2 Satellite Plants	\$0.71	\$5.15	\$3.38	\$7.34	\$1.98
4 Satellite Plants	\$0.69	\$5.02	\$3.36	\$7.28	\$1.97

The aggregated MESP—the combined MESP for both the central plant and satellite plants—for the cases with two and four satellite plants are \$3.38 and \$3.36/gal EtOH, respectively, compared with \$3.49/gal EtOH for the stand-alone 2000 tpd plant with on-site enzyme production that was discussed in the previous section. The difference in MESP is not significant and is within the $\pm 30\%$ accuracy for this level of economic study. The EPECs for the cases with two and four satellite plants are \$0.71 and \$0.69/gal EtOH, respectively, meaning that below those enzyme prices it is more favorable to purchase enzymes and operate a stand-alone 2000 tpd plant.

A significant contributor to the lower MESP for the central enzyme production scenario than for the 2000 tpd stand-alone plant with on-site enzyme production is the significantly lower TCI. Table 3.8 shows that by increasing the number of satellite plants—and thus increasing enzyme production at the central plant—the TCI per gallon of ethanol production capacity decreases. Furthermore, the TCI for a stand-alone 2000 tpd plant with on-site enzyme production is only \$8.03/gal of EtOH capacity, which is higher than the two scenarios shown in Table 3.8. This trend is due largely to the economy of scale with the enzyme production equipment. Table 3.9 shows the installed equipment cost for each process area.

Table 3.8 Equipment cost for all process area (\$/gal of annual EtOH production capacity)

	Stand-alone 2000 tpd Plant w/On- site Enzyme Production	Central Enzyme Production w/2 Satellite Plants	Central Enzyme Production w/4 Satellite Plants
Feedstock Handling	\$0.23	\$0.23	\$0.22
Pretreatment	\$0.77	\$0.75	\$0.74
Enzyme Production	\$0.50	\$0.32	\$0.33
Saccharification/Fermentation	\$0.46	\$0.43	\$0.42
Distillation/Solids Recovery	\$0.54	\$0.50	\$0.50
Wastewater Treatment	\$0.07	\$0.12	\$0.12
Chemical Storage	\$0.05	\$0.06	\$0.06
Cogeneration	\$1.22	\$1.13	\$1.11
Utilities	\$0.14	\$0.14	\$0.13

The largest difference in capital cost between the stand-alone plant and the central enzyme production scenarios is the enzyme production area, which is \$0.50/gal EtOH compared with \$0.32 and \$0.33/gal EtOH for the cases with two and four satellite plants, respectively.

The annual operating costs for the two cases shown in Table 3.8 are slightly lower than the stand-alone 2000 tpd on-site enzyme production scenario, which is \$2.02/gal EtOH. The estimated enzyme transportation cost from the central plant to the satellite plants is only \$0.01/gal EtOH.

Conclusions

Plant scale plays a significant role in the economic competitiveness of on-site enzyme production with purchasing enzymes for cellulosic ethanol production. At the 500 tpd scale the cost of purchased enzyme must be greater than \$7.81/kg protein (\$1.07/gal EtOH) in order for on-site enzyme production to be more economically advantageous than purchasing enzymes, while that cost reduces to \$4.77/kg protein (\$0.66/gal EtOH) at the 3000 tpd scale. Within the range of plant scales studied, the scenario in which feedstock cost varies—due to delivery cost increases at larger scales—does not differ significantly in terms of enzyme cost from the scenario with a constant feedstock cost.

The production of both ethanol and excess enzymes at a central plant for distribution to satellite plants does not offer significant economic advantages over a stand-alone ethanol plant with on-site enzyme production. For the case with one central plant and two satellite plants—each receiving 2000 tpd of biomass—the EPEC is \$0.71/gal EtOH, and is \$0.69/gal EtOH for the case with four satellite plants. These EPECs are only slightly lower than a 2000 tpd stand-alone plant with on-site enzyme production. Therefore, the central enzyme production scheme is not a short term financial game-changer, but may offer advantages in longer term industry optimization.

This analysis shows that even with a significant reduction in the cost of enzyme production, on-site production may not be a cost effective scenario at smaller plant scales in the near term. For example, a 50% reduction in enzyme production cost—either through production improvements or improvements in enzyme specific activity—would still require a purchased enzyme cost greater than \$0.54/gal EtOH at the 500 tpd scale in order for on-site enzyme production to be economically advantageous. This represents a considerably higher cost than the 2006 DOE estimates of \$0.10-0.25/gal EtOH. However, even at the 3000 tpd plant scale the purchased enzyme cost must be higher than \$0.66/gal EtOH in order for on-site enzyme production to be economically advantageous, which is also considerably higher than the DOE estimate. This may indicate that on-site enzyme production may only be viable at scales much larger than 3000 tpd. There also may be significant differences between the assumptions made in this study and those in the DOE enzyme cost estimate. Differences in enzyme production yield or production rate assumptions or in ethanol production assumptions would result in different enzyme cost estimates. For example, higher fermentation conversions of sugars to ethanol would cause the normalized enzyme cost in \$/gal EtOH to be lower.

If the DOE goal of a tenfold reduction in enzyme cost is achieved the EPEC at the 500 tpd scale would be \$0.105/gal EtOH, representing only 2.8% of ethanol production cost, and only \$0.066/gal EtOH at the 3000 tpd plant—approximately 2.5% of ethanol production cost. While these costs are still higher than the \$0.01-0.025/gal EtOH goals set by the DOE, they may be low enough for economic viability of on-site enzyme production at those scales.

Appendix 3.A Process Flow Diagram for On-site Enzyme Production

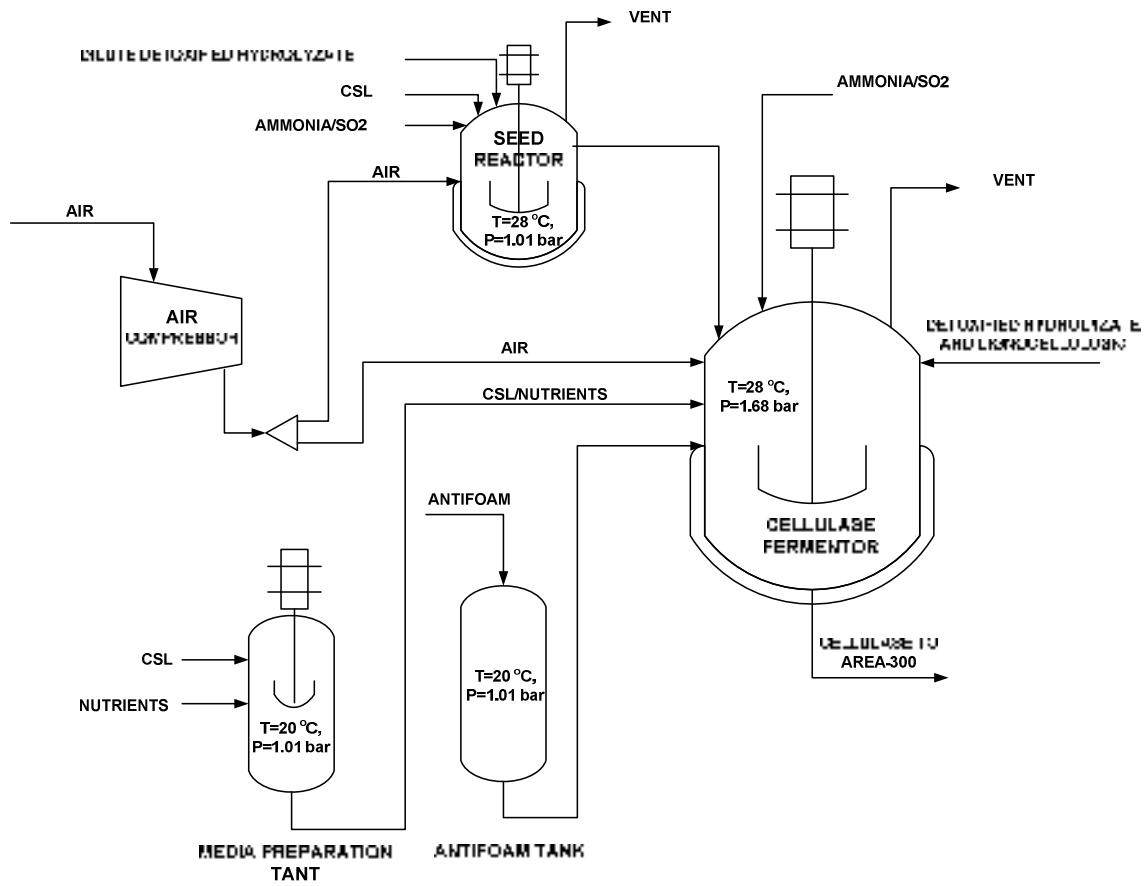


Figure 3.A.1 Enzyme production area

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