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FINAL REPORT:

THE COST OF ETHANOL PRODUCTION FROM LIGNOCELLULOSIC BIOMASS -

A COMPARISON OF SELECTED ALTERNATIVE PROCESSES

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CHAPTER 1

REPORT SUMMARY

The purpose of this report is to compare the cost of selected alternative processes for the conversion of lignocellulosic biomass to ethanol. In turn, this information will be used by the ARS/USDA to guide the management of research and development programs in biomass conversion. The report will identify where the cost leverages are for the selected alternatives and what performance parameters need to be achieved to improve the economics.

Because of the early stage of the process development, most of the design information is based on laboratory or limited pilot plant data. As a result, a feasible design case is developed for each alternate process based on this preliminary information and then likely improvements are evaluated to see what potential the process has.

The process alternatives considered here are not exhaustive, but are selected on the basis of having a reasonable potential in improving the economics of producing ethanol from biomass. When other alternatives come under consideration, they should be evaluated by the same methodology used in this report to give fair comparisons of opportunities.

A generic plant design is developed for an annual production of 25 million gallons of anhydrous ethanol using corn stover as the model substrate at $30/dry ton. A plant capacity of 25 million gallon per year is a compromise between trying to design a large plant to get economy of scale and a small plant to keep transportation costs low by collecting biomass within a 50 mile radius of the plant.

Standard chemical engineering techniques are used to give first order estimates of the capital and operating costs. The cost estimate technique used is consistent with the prior design study on corn to ethanol (9). Following the format of the corn to ethanol plant, there are nine sections to the plant; feed preparation, pretreatment, hydrolysis, fermentation, distillation and dehydration, stillage evaporation, storage and denaturation, utilities, and enzyme production.

The process alternatives that are considered change the way the pretreatment, hydrolysis or fermentation are done.

Because of the extensive work done in the laboratory and pilot plant by the Tennessee Valley Authority (TVA), the concentrated acid hydrolysis process is the base case of this study. Since the glucose concentration from the hydrolysis section is 11.63%, this set the standard for the alternative processes. When the fermentor microorganism can tolerate the ethanol, the glucose is brought to 11.63% with a sugar evaporator.

There are three pretreatment alternatives considered: the AFEX process, the modified AFEX process (which is abbreviated as MAFEX), and the STAKETECH process. These all use enzymatic hydrolysis and so an enzyme production section is included in the plant. The STAKETECH is the only commercially available process among the alternative processes.

The final alternative is the Bioenergy process for the simultaneous fermentation of hexose and pentose. Thus, the generic plant design is simplified to one fermentation step. The Bioenergy process uses a genetically altered *E. coli*.

By imbedding each process alternative into a complete plant design, including a utility section with a boiler/tubogenerator and stillage evaporation section to handle the major effluent, one can evaluate the interactions among the plant sections and get the total cost of production of ethanol. This is the only fair way to compare process alternatives because a savings in one section of the plant, due to a given alternative, may require larger expenditures in another.

For each process alternative, a Microsoft Excel spreadsheet model was developed to give the material balance, energy requirements, equipment lists, capital and operating costs. The details of the design simulations are given in the Appendices.

A bottom line comparison of the cost of production per gallon of ethanol is given for the alternative processes in Table 1-1. The table does not give all the cost excursions that were studied, but rather gives a sample of the most important results. The table gives a brief summary of process parameters, the capital investment and unit costs. The column called net unit cost is due to sale of excess electricity, if any. For purposes of comparison, the last row in the tables gives the unit cost of ethanol from a dry mill corn plant at 50 million gallons/year capacity. The net cost in the corn plant is due to the sale of Distillers Dry Grains and Solubles (DDGS). If the corn to ethanol plant were down sized to 25 million gallons per year, the unit cost would be about 10¢/gal more than indicated. The raw material is corn stover at $30 per dry ton. Note: for every $10/ton change in the cost of corn stover, the cost of ethanol changes by about 10.5¢/gal.

It is clear that the concentrated acid process is not competitive. This is because just the cost of acid, lime and gypsum sludge disposal alone accounts for 50¢/gal. An alternative concentrated HCl hydrolysis process was identified (but was not quantitatively analyzed in this study) which can save a substantial part of the 50¢/gal cost. It is the process developed on the laboratory scale by Dr. Goldstein at North Carolina State University in which HCl is recovered and reused by distillation and electrodialysis. Realistic capital costs are not available at this time, but a qualitative judgement would expect the capital to be equal to or even less than that for the TVA process. Thus, critical evaluation of this process in the format of this study is recommended.

With a solids concentration of 10% in the hydrolysis reactor, the AFEX-10 process has a lower cost ($1.12/gal) than when the solids are 20% in the AFEX-20 process ($1.15/gal). This is because the cellulose conversion is assumed to be reduced from 98% to 80% as the solid concentration is increased. The STAKETECH process is similar in cost ($1.154/gal) to the AFEX-20 process. However, when comparing the STAKETECH process and the AFEX-10 process, where both operate on 10% solids in the hydrolysis, the AFEX-10 process has a lower cost by 3¢/gal. The major reason is the lower enzyme loading that is demonstrated for the AFEX process, 5 FPU/g instead of 15 FPU/g. When the enzyme loading is increased from 5 to 15 FPU/g the unit cost will go up about 8¢/gal.

Because of the lower capital, the MAFEX process with 80% cellulose conversion, has a cost about 4¢/gallon lower than the AFEX-10 process with 98% cellulose conversion. There is also a small energy credit due to the larger residue available for fuel.

Table 1-1
Summary of Process Alternatives
Unit Cost of Ethanol Production for 25 Million Gallon per Year from Corn Stover

Process Alternative	Major Design Parameters	Capital Investment $	Unit Cost* $/gal	Net Unit Cost* $/gal
TVA Base Case	Concentrated acid hydrolysis, 90% cellulose conversion, separate fermentation, 90% C_6 and 50% C_5 yield	81,157,000	1.708	1.632
AFEX-10	10% solids in enzymatic hydrolysis, 98% cellulose conversion, 5 FPU/g, separate fermentation, 90% C_6 and 50% C_5 yield	93,287,000	1.128	1.128
AFEX-20	20% solids in enzymatic hydrolysis, 80% cellulose conversion, 5 FPU/g, separate fermentation, 90% C_6 and 50% C_5 yield	94,354,00	1.154	1.153
MAFEX	10% solids in enzymatic hydrolysis, 80% cellulose conversion, 5 FPU/g, separate fermentation, 90% C_6 and 50% C_5 yield	84,128,000	1.087	1.077
STAKETECH	10% solids in enzymatic hydrolysis, 90% cellulose conversion, 15 FPU/g, separate fermentation, 90% C_6 and 50% C_5 yield	101,479,000	1.187	1.154
Bioenergy Case A with AFEX-10	10% solids in enzymatic hydrolysis, 98% cellulose conversion, no sugar evaporator, 5 FPU/g, combined fermentation, 95% C_6 and 50% C_5 yield in 58 hours	96,254,000	1.190	1.190

Table 1-1 (continued)
Summary of Process Alternatives
Unit Cost of Ethanol Production for 25 Million Gallon per Year from Corn Stover

Process Alternative	Major Design Parameters	Capital Investment $	Unit Cost* $/gal	Net Unit Cost* $/gal
Bioenergy Case A with MAFEX	10% solids in enzymatic hydrolysis, 80% cellulose conversion, no sugar evaporator, 5 FPU/g, combined fermentation, 95% C_6 and 50% C_5 yield in 58 hours	90,257,000	1.179	1.161
Bioenergy Case B with AFEX-10	10% solids in enzymatic hydrolysis, 98% cellulose conversion, no sugar evaporator, 5 FPU/g, combined fermentation, 95% C_6 and 95% C_5 yield in 58 hours	82,365,000	1.087	1.087
Bioenergy Case B with MAFEX	10% solids in enzymatic hydrolysis, 80% cellulose conversion, no sugar evaporator, 5 FPU/g, combined fermentation, 95% C_6 and 95% C_5 yield in 58 hours	77,860,000	1.030	1.030
Bioenergy Case B with AFEX-10 with Sugar Evaporator	10% solids in enzymatic hydrolysis, 98% cellulose conversion, 5 FPU/g, 11.6% glucose in combined fermentation, 95% C_6 and 95% C_5 yield in 58 hours	71,226,000	1.003	1.003
Bioenergy Case B with MAFEX with Sugar Evaporator	10% solids in enzymatic hydrolysis, 80% cellulose conversion, 5 FPU/g, 11.6% glucose in combined fermentation, 95% C_6 and 95% C_5 yield in 58 hours	63,422,000	0.941	0.941

Table 1-1 (continued) Summary of Process Alternatives Unit Cost of Ethanol Production for 25 Million Gallon per Year from Corn Stover				
Process Alternative	Major Design Parameters	Capital Investment $	Unit Cost* $/gal	Net Unit Cost* $/gal
MAFEX with *Pichia stipitis*	10% solids in enzymatic hydrolysis, 80% cellulose conversion, 5 FPU/g, 9.2% glucose in combined fermentation, 95% C_6 and 95% C_5 yield in 51 hours	64,293,000	0.947	0.947
Reference Case	Corn to ethanol (50 million gal/year) dry mill	118,060,759	1.570	1.154

*Operating cost includes raw materials, energy, labor for operations, maintenance and laboratory and capital recovery in 9 years (11.1% per year) insurance (1%) and maintenance materials (2.5% of capital). Corn stover at $30 per dry ton.

In the Bioenergy process, a genetically modified *E. coli* is used to simultaneously ferment glucose and xylose in a single fermentation step. When the best Bioenergy fermentation yields in Case B are coupled with the AFEX-10 pretreatment or the MAFEX pretreatment to prepare the hydrolyzate, there is a 4¢ to 6¢ per gallon saving over the respective AFEX-10 or MAFEX with a two stage fermentation. The xylose is not fully utilized so the capital is higher for either Case A Bioenergy process with one fermentation step than the corresponding AFEX-10 or MAFEX process with two fermentation steps. This is because the dilute ethanol in the Bioenergy process becomes even more dilute when all the sugars are used.

A "what if" scenario can be evaluated for the Bioenergy process when the glucose in the hydrolyzate is concentrated to the standard 11.63% as in the other alternatives. Under this condition, the microorganism must be tolerant to the corresponding higher ethanol concentration. The cost drops to 1.003 and 0.940 $/gal, respectively, when the AFEX-10 and MAFEX pretreatments are coupled to the Bioenergy process with a sugar evaporator.

A similar low cost of 94.7¢/gal is in fact achievable by using the yeast *Pichia stipitis* as shown in the end of Table 1-1. This yeast can handle 13% of total sugars with 9.2% glucose and complete the fermentation in 51 hours.

Clearly, the challenge in the Bioenergy process is to achieve higher alcohol tolerance.

From the results summarized in Table 1-1, there are some very encouraging prospects to develop a biomass to ethanol process that can certainly get the cost of production below that for corn to ethanol, and even get below 95¢/gallon with a focused development program. Note these production costs do not depend on heavy by-product credits, such as DDGS in a corn plant, or free or negative cost substrates. The detailed analysis in the report explores these processes in a parametric way to understand better the cost sensitive areas. Because a number of yields and times are assumed from interpreting a limit set of experimental data in each alternative, the verification of critical points are needed. The biggest uncertainty is in the solid/liquid handling - such issues are filtration rate, filter cake solids, washing rate of cake, soluble recovery from cakes. They are all assumed (in this study) to be at the same reasonable conservative level.

As a final point, one should realize that the uncertainties in solid/liquid handling are accounted for in all the process alternatives in the same way, so comparisons of cost among the alternatives are valid. It is just the absolute cost that is in doubt due to unknown engineering factors.

In contrast to the cost of production of ethanol from corn, which is based on actual operating plants, the cost of ethanol from biomass for all the alternatives is based on limited laboratory and pilot plant data, engineering judgement of what is reasonable to expect, and extrapolations by analogies of similar steps or operations that work in another context. As a result, we will comment on each alternative to identify the issues that are firm and those that can contribute to improved process economics, but need more work.

A wide range of lignocellulosic substrates have been shown to be very reactive to low cellulase loadings when pretreated by the AFEX process. Thus, there is little doubt that 5 FPU/g will give 95% or more conversion of the cellulose in 12 to 24 hours in 5% slurries. This low enzyme loading is a strong advantage of AFEX and lower than 5 FPU/g are expected to be effective. The issue is scale-up. The expectation is that the above rate and yield are possible in 10% slurries - a practical entry point for industrial enzymatic hydrolysis. While the pretreatment itself gives a high solids cake of biomass, the degree of the biomass dilution in the hydrolysis reactor is an engineering issue involving a process optimization. The effect of time, agitation energy, temperature, enzyme profile and enzyme reuse in the hydrolysis reactor with changes in the pretreatment parameters needs to be studied. Reduction in capital cost for equipment is expected after completing an innovative bench scale/pilot plant program. The separation of the lignin residue from the sugar in the hydrolyzate or in the stillage also needs to be studied to pick the most cost effective equipment. This can only be done after real samples of the hydrolyzate in adequate quantities are available.

In the MAFEX process, the laboratory data is only on forage grass. While these results show good yields of biomass to sugars, again at low enzyme loading (5 FPU/g and less), efficacy on a wide range of lignocellulosic substrates needs to be established. From what is known from other pretreatment processes, they generally work on a wide range of substrates - so there is every reason to expect that to be the case for MAFEX. Since MAFEX is simpler, less capital requiring pretreatment than the others, it may have the long term advantage. The process engineering and equipment optimization studies needed for the MAFEX process are similar to those for the AFEX process.

The Bioenergy process raises the issue of simultaneous fermentation of pentose and hextose - an inevitable consequence of using lignocellulosic biomass. There are real significant savings in capital and operating costs if the process parameters are biased in the direction of higher ethanol tolerance, shorter fermentation times, and complete, 90% or more, utilization of both hexose and pentose. Important developments by Dr. Ingram and co-workers have shown progress on all of these issues. More attention is needed on alcohol tolerance; and failing that, short time, complete sugar utilizing fermentation cycles are needed. The possibility of high cell density cultures for continuous fermentation should also be studied. Long term reproducibility and stability of the culture needs to be demonstrated.

The alternative with a special yeast, such as *Pichia stipitis* which has shown ethanol tolerance up to 7 vol % and good sugar utilization, is promising and needs to be studied on real hydrolyzates from the various alternative pretreatment hydrolysis processes to confirm the performance and establish reproducible results and stable culture properties.

CHAPTER 2

INTRODUCTION

A. Purpose of the Study

While ethanol production from corn is well established and growing, the growth in ethanol use as a blending component in gasoline or as a neat fuel is likely to accelerate as its role in reducing air pollution from internal combustion engines is established and accepted. Since the economics of ethanol production from corn are strongly coupled to the cost of corn and the by-products derived from it, a 100% or 200% increase in ethanol demand would increase the cost of ethanol from the current level because the by-product credits will not keep pace with such an increased in production.

As a long-term alternative, lignocellulosic biomass is a potential resource for ethanol production. Since there are no widely accepted, commercially demonstrated processes for ethanol from biomass, some more research and development is required to advance the commercial application of biomass conversion. The purpose of this report is to consider several promising process alternatives and to compare them on a common substrate and a common ethanol production capacity. The results will highlight the parameters that affect the cost and identify the opportunities that will benefit from a focused research and development program.

Since a great deal of research has been done over the last 20 years, much of what remains to be done is development and optimization of promising process concepts. The research should be focused on the cost sensitive areas. Naturally, this research and development activity is an iteration that is guided by how well it overcomes the economic hurdles. As improved performance parameters are achieved, their impact on the cost of ethanol should be evaluated and programs modified accordingly.

Given the limited time, this report is not an exhaustive study of all or most of the biomass to ethanol process alternatives. Rather, a few promising alternatives are considered in the context of a full plant design. Consequently, the economic interactions among the various sections of the plant will be understood. Since the overall plant is integrated, the economic impact of change in a given section cannot be appreciated or viewed in isolation. As more alternative processes are identified, the approach of this report can be used as a model to evaluate them.

B. Substrate

Although lignocellulosic biomass includes a wide variety of materials, the major components are cellulose, hemicellulose and lignin. Naturally, the relative amounts of these components vary from substrate to substrate, and from time to time in the same substrate. Since the technologies that are considered below apply to a wide variety of lignocellulosic biomass, the selection of a model substrate is a matter of convenience which will not limit the overall observations and conclusions of this study.

Corn stover is selected as the biomass because of the interest in a large available crop residue at USDA and because in the base case below it is used as the substrate for the TVA concentrated acid hydrolysis process.

The composition of the corn stover on a dry basis is 45% cellulose, 29% hemicellulose, 20% lignin and 6% other. The material arrives at the plant with 16% moisture. When under going hydrolysis, the cellulose (a homopolymer of glucan) is converted to glucose.

$$\underset{glucan}{[C_6(H_2O)_5]_n} + n\,H_2O \rightarrow n\,\underset{glucose}{C_6\,(H_2O)_6} \qquad (1)$$

Thus, 162 pounds of glucan gives 180 pounds of glucose using 18 pounds of water. In the real process, cellulose conversion is less than 100% of theory and it appears as a parameter in the mass balance tables in this report.

On the other hand, hemicellulose is a branched heteropolymer of pentosans with some hexans. When it is hydrolyzed, it gives a mixture of sugars of which over 80% are xylose. There are other minor sugars such as mannose, galactose and arabinose. The mannose and galatose, which are isomers of glucose, are generally fermented by the same organisms that utilize glucose while the arabinose, an isomer of xylose, is utilized by the xylose consuming micro-organisms. As a result, we make two simplifying assumptions in this study. One, mannose and galactose are considered as glucose, and arabinose as xylose. Two, all the hemicellulose will be considered xylan and is converted to xylose when hydrolyzed as follows:

$$\underset{xylan}{[C_5(H_2O)_4]_n} + n\,H_2O \rightarrow n\,\underset{xylose}{C_5(H_2O)_5} \qquad (2)$$

Thus, 132 pounds xylan give 150 pounds xylose using 18 pounds of water.

The way the process analysis is set up, it is possible to change the percent cellulose and hemicellulose and in effect alter the ratio of glucose to xylose produced by a substrate.

The lignin is taken to remain as an unconverted solid throughout the various process steps. It is recovered with the soluble sugars and enzymes and cells as a fuel to provide for the thermal and electrical needs of the plant.

C. Plant Sections

All of the process alternatives will be compared on an integrated plant design producing 25 million gallons of 199 proof (≥99.5 wt%) ethanol. To allow for annual shut down and maintenance, the plant will operate 24 hours per day for 330 days per year or 7920 hours per year. The costs are based on January 1993 with a Chemical Equipment Index (CEI) of 358.

The plant design will be laid out in the following sections:

Section 100:	Feed Preparation
Section 200:	Pretreatment
Section 300:	Hydrolysis
Section 400:	Fermentation
Section 500:	Distillation and Dehydration
Section 600:	Stillage Evaporation
Section 700:	Product Storage and Denaturation
Section 800:	Utilities
Section 900:	Enzyme Production

Note that, not every alternative will have Section 200 - Pretreatment or Section 900 - Enzyme Production. By having the utilities section on site, the thermal and electrical loads of the plant are met in part or completely by burning the stillage and lignin residues. Thus, there is no by-product to sell other than possibly the excess electricity from the cogeneration power plant. This type of complete plant design puts a high burden on the economics of an alternative process since we do not take advantage of any special consideration such as a steam source near the plant or the sale of animal feed by-products. By considering processes in the context of the whole plant, it gives a fair comparison of process alternatives.

D. Process Alternatives

The first biomass conversion process is based on the extensive laboratory and pilot plant work of the investigators at the Tennessee Valley Authority (TVA) on the concentrated sulfuric acid process for corn stover. The process will serve as the base case for this study because there are data available on the yield of sugar and ethanol and material balances. Moreover, tests on filtering biomass with screw presses and automated filter presses are available which give an insight on the very important solid/liquid separations that are part of all biomass conversion processes. The TVA process achieves 90% of the theoretical conversion of the cellulose to glucose and complete conversion of the hemicellulose to xylose. Separate fermentations on the neutralized hydrolyzate have demonstrated 90% of the theoretical conversion from glucose to ethanol with the regular yeast, *Saccharomyces cerevisiea* and 50% of theoretical yield from xylose with *Pachysolen tannophilus.*

The second process alternative is the AFEX process, a proprietary pretreatment process of the AFEX Corporation, Brenham, Texas. This is a pretreatment on ground biomass with ammonia under pressure. The ammonia alters the lignocellulose structure so that when the ammonia is removed, the biomass solids are very reactive to cellulases to produce a high yield of glucose (98% theory) and complete conversion of hemicellulose. Moreover, the enzyme loading is low, five filter paper units (FPU) per gram of substrate, compared to other enzymatic hydrolysis work. The plant design is based on the data from AFEX Corporation's laboratory studies and the authors' best engineering judgment to estimate the cost of production and capital investment.

As the third alternative, we have the MAFEX process, another proprietary pretreatment process of the AFEX Corporation. Not as much laboratory data are available as for the AFEX process, but a pilot plant scale demonstration did give 70% conversion of the potential carbohydrate with 5 FPU/g in 12 hours. The plant design is handled in a parametric way so sensitivity to assumptions can be evaluated.

The fourth alternative is the STAKETECH process, a proprietary continuous steam explosion pretreatment process from Stake Technology Ltd., Norval, Ontario. This process is being offered for commercial use by Stake Technology Ltd. A pilot plant process using the Stake pretreatment has been under testing by IFP in Souston, France to produce ethanol and enzyme from *Trichoderma reesei*. The yield parameters for the STAKETECH process are given by Stake Technology. They also supplied an estimate for a complete ethanol plant with pretreatment, fermentation, hydrolysis, distillation and enzyme production. However, they consider the overall plant design proprietary, so we did not have the details on each section of the plant. Nevertheless, we adapted the Stake pretreatment into the general plant scheme we present below. Later in this report, a check on the economics can be made by comparing the production cost we estimate versus the ones given by Stake Technology Ltd.

For the fifth alternate process, we evaluated the genetically modified microorganism by Dr. L. Ingram of Bioenergy International L.C., Gainesville, Florida. This patented microbe transferred the genes for pyruvate decarboxylase and alcohol dehydrogenase of the bacteria *Zymomonas mobilis* into *E. coli*. Consequently, the modified organism ferments both hexose and pentose. This development offers a way to compare the impact of one simultaneous fermentation of the mixed sugars from the hydrolysis of biomass with a separate yeast fermentations of the hexose and pentose.

Finally, as often happens, one learns of more alternatives that look interesting and should have been considered, but were not due to a lack of time or enough information. A recent detailed report was published by the former Bio-Hol Developments (12), Toronto, Ontario, on dilute acid hydrolysis using the Wenger single screw extruder for the hydrolysis of the hemicellulose and the plug flow reactor of St. Lawrence Reactors. Of greatest interest in this report are the fermentation yields reported for the yeast *Pichia stipitis* on the simultaneous utilization of glucose and xylose in the hydrolyzate. An evaluation of the economics of using *Pichia stipitis* will be given in the discussion of the Bioenergy process.

Another improvement is the use of bacteria such as *Zymomonas m.* (14, 15) for continuous fermentation of the hexose, in a few hours. Since the plant costs developed in this report change significantly when the fermentation time is dramatically reduced, there may be cases where a separate hexose fermentation is economic.

Recently, Ethanol International is trying to scale-up and market the HCl concentrated acid process of Dr. Irving Goldstein (13). Since the details are confidential, only very general information was available, so a detailed comparison with the cases presented in this report was not possible. Some implications of this process are considered under the discussion of the TVA process.

CHAPTER 3

METHODOLOGY

A. Overall Features of the Plant Design

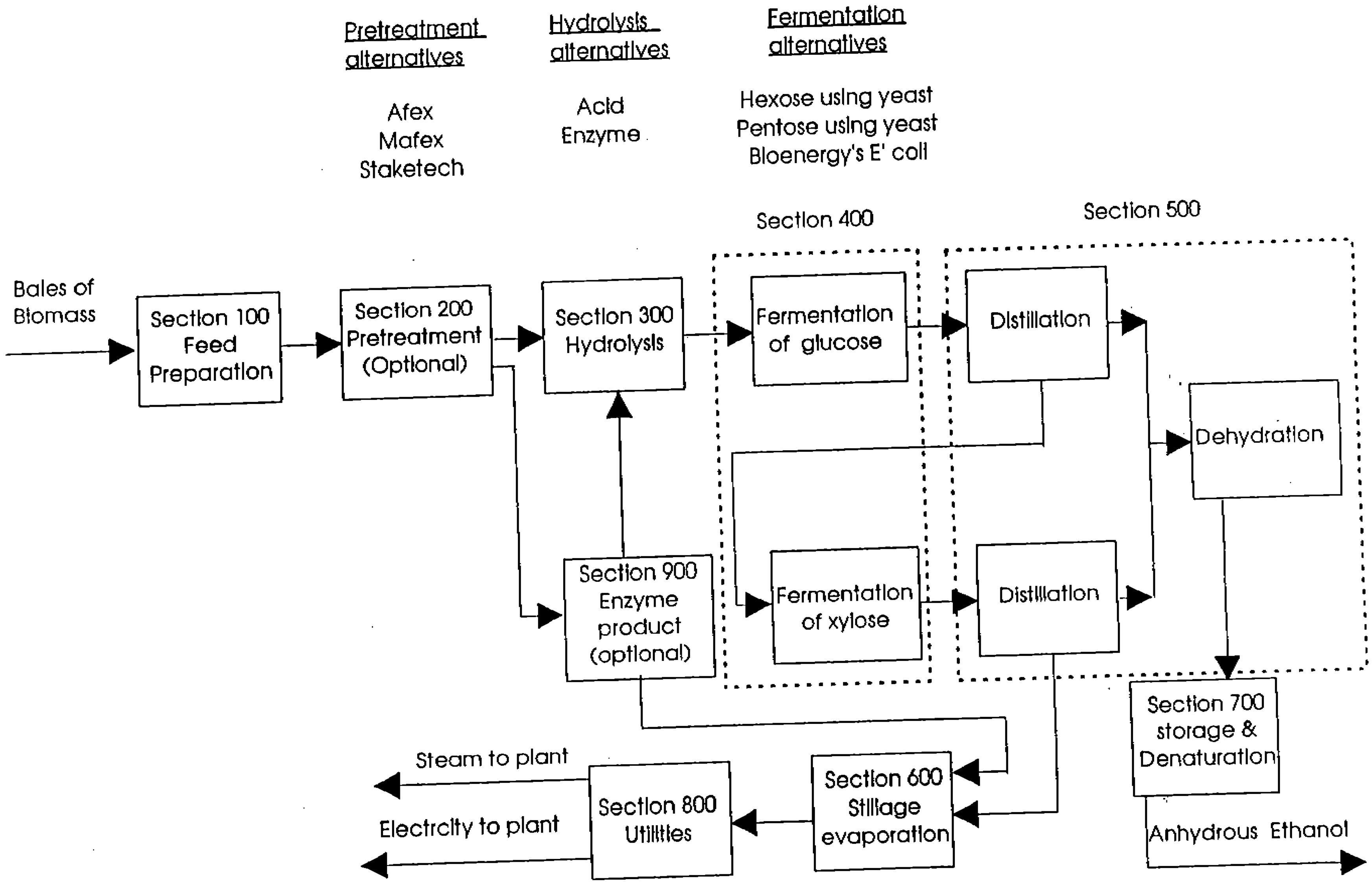
An overview of the generic plant design for a biomass to ethanol plant is given in Figure 3-1 that organizes the functions of the plant in nine sections. For the TVA acid hydrolysis process, Section 200 - Pretreatment and Section 900 - Enzyme Production are not needed. For the remaining alternate processes, all 9 sections are needed. A brief description of the functions of each section is given immediately below and more complete details are discussed under the chapters for each process alternative.

The corn stover is brought to the plant in bales. The purpose of Section 100 - Feed Preparation, is to provide the desired particle size for the next process section by some type of milling operation.

There are three pretreatment process alternatives considered for Section 200 - namely AFEX, MAFEX and STAKETECH. The purpose of the pretreatment is to enhance the rate of enzymatic hydrolysis and increase the degree of conversion of the cellulosic materials to sugars.

For Section 300 - Hydrolysis, there is either concentrated acid hydrolysis from TVA or enzymatic hydrolysis. The cellulase complex of endo- and exo-glucanase in the extracellular broth of the fungi *Trichoderma reesei (T.r.)* is grown on site in Section 900 - Enzyme Production. Since *T.r.* is deficient in β-glucosidase which completes the hydrolysis of cellobiose to glucose, purchase of a small amount of β-glucosidase is included in the plant cost.

Figure 3 - 1. Overall Block Diagram of Generic Process for Conversion of Biomass to Ethanol



In the base case (the TVA process), the fermentation of the glucose and xylose in the hydrolyzate is done in two separate fermentors in Section 400 (see Figure 3-1). In the first fermentation step, the normal glucose utilizing yeast, *Saccharomyces cerevisiea* leaves the xylose in the resulting ethanol containing beer. The ethanol is removed in Section 500 in the overhead of stripper/rectifier column of the distillation section. The bottoms of the column, which contain the xylose, are returned to Section 400 for a second fermentation with the yeast *Pachysolen tannophilus*. Finally, the resulting beer is stripped of ethanol in the second stripper/rectifier column of Section 500. The still bottoms from this column which contain all the non-fermentable organic matter, such as lignin, cell mass and residual sugars in an aqueous dilute slurry are directed to Section 600. The overheads from both columns with ethanol near the azeotropic composition are combined to a dehydration column which uses a hydrocarbon entrainer to break the azeotrope.

In all the process alternatives, except the Bioenergy Case, the same combination of fermentation and distillation sections are used. For the Bioenergy Case there is a single fermentation step in Section 500 where both glucose and xylose are fermented to ethanol. The ethanol is then recovered in Section 500 with one stripper/rectifier column and one dehydration column. The still bottoms are directed to Section 600.

The dehydrated ethanol is denatured with gasoline and stored in Section 700 from where it is shipped to market.

The still bottoms are concentrated to a syrup in an evaporator to 50% total solids in Section 600. This provides a way to avoid a high degree of water pollution because of the high BOD of the stillage and it provides all or most of the energy to operate the entire plant.

In Section 800, Utilities, we group the support functions to operate the plant. There is the boiler which burns the evaporated stillage and produces high pressure, super heated steam. This is run through an extraction turbine to generate electricity and low pressure steam to serve the thermal loads of the plant. The other part of the utility section provides for the cooling tower, fire protection and electrical distribution.

B. Design Approach

Each alternative process is described and analyzed in a separate chapter of this report. However, those sections that are common to all cases are discussed in Section 3E of this chapter. Within each alternative process, there are a number of design simulation runs prepared on an Excel spreadsheet model, to explore the effect of changes in selected process parameters on the cost of production and capital investment. The run number will have two numbers separated by a dash; the first is the chapter number for the alternative process and the second is a particular run for a given set of parameters. A complete run generates up to 9 tables which are identified as follows:

Table 1	Mass Balance
Table 2	Energy Requirements and Production
Table 3	Equipment List for Section 100 - Feed Preparation
Table 4	Equipment List for Section 200 - Pretreatment (if needed)
Table 5	Equipment List for Section 300 - Hydrolysis
Table 6	Equipment Summary for Sections 400 through 800
Table 7	Equipment List for Section 900 - Enzyme Production (if needed)
Table 8	Fixed Capital Cost Estimate
Table 9	Operating Cost

A complete set of design simulation runs with these nine tables for each run is given in the Appendix. Note that the nine table numbers will be repeated for each run, but the tables will be distinguished by the run number preceding the table number.

A first order process design is prepared for each process alternative using standard chemical engineering techniques. First, an overall generic process flow diagram is prepared. Then, for each alternative, a process flow diagram is prepared for those sections unique to the given alternative to show the major items of equipment and the connection of the various streams. The stream numbers are three digits with the hundreds position coincident with the section in the plant. All corresponding streams in alternative cases have the same stream numbers. When a stream is listed in the mass balance table with two numbers, one refers to the number it has in the leaving section and the other refers to the number it has in the receiving section. Mass balances for the major plant streams are prepared and identified with the same stream numbers as in the flow diagram.

The component and total mass flows as well as the volumetric flows are given in the mass balance tables. The mass balance information is used to size the major pieces of equipment for the various sections and an equipment list is prepared for each section. In turn, the equipment size is used to estimate the purchased cost of the equipment. From the purchased equipment costs, one estimates the total capital for the installed equipment by using factors from prior plant design of the same type (1, 2). A summary of the installation factors is given in Table 3-1. An estimate is made of the thermal heating load of the plant and the mechanical energy needed in the form of electricity which is based on the throughput of the plant. The operating labor requirements is based on the experience of corn based ethanol plant of the same size.

Finally, all of the information on material, energy, labor and capital is used to estimate the cost of production. Then one can compare the cost of production of the various alternative processes in a fair way with a common design methodology. The key point here is that the costs are not exact in an absolute sense, but that the relative differences are real and reflect increases or decreases due to a particular alternative compared to the base case or to some other case.

The analysis of each process alternative is described in corresponding separate chapters. However, those aspects that are common to all the designs such as Sections 400, 500, 600, 700 and 800 are discussed below.

Table 3-1
Installation Factors Used to Convert Purchased Equipment Cost to Installed Capital Cost

Section 100	2.8
Section 200	
Packaged equipment	1.5
Other equipment	3.5
Section 300	
Packaged evaporators	1.5
Other	3.5
Section 400	3.5
Section 500	4.0
Section 600 - Packaged evaporators	1.5
Section 700	3.2
Section 800 - Packaged boiler	2.3
Section 900	3.5

C. Design Information

There are many aspects that are similar in a plant converting lignocellulosic biomass to ethanol to one converting corn to ethanol - especially in sections involving fermentation, distillation and dehydration, storage and denaturation and utilities. Thus, for these sections, the best information is based on detailed design studies published by Raphael Katzen Associates International (1). The Katzen design is for an ethanol plant from corn with a 50 million gallon per year capacity in 1978 which corresponds to a Chemical Equipment Index (CEI) of 219.

The costs are updated for inflation to 1993 using a CEI of 358. When equipment sizes are changed, the costs are changed by the capacity ratio raised to the exponential power of 0.6. Thus,

$$\textit{Cost for Capacity } 2 = \textit{Cost for Capacity } 1 \left(\frac{\textit{Capacity } 2}{\textit{Capacity } 1}\right)^{0.6} \tag{3}$$

When equipment items are increased in number, rather than size, the cost is directly proportional to the number.

For the TVA concentrated acid hydrolysis process, the design of the hydrolysis section is taken from a report by Moore and Barrier (2). The design base was modified for a plant of 12 million gallons of ethanol per year to 25 million gallons. Equipment costs for the acid hydrolysis section come from this report corrected for capacity changes and for inflation with a CEI of 324 in 1987 to CEI of 358 in 1993.

In the enzyme plant, the design is based on the report by Chem Systems (3) for a 25 million gallon ethanol plant using wood as a substrate. The plant prepared enough enzyme for a load of 15 FPU/g of substrate. The time base is for 1986 with a CEI of 323. The equipment items in the enzyme plant are adjusted for inflation to a CEI of 358 and scaled to capacity as needed.

Dr. Ernest Yu, Vice President of Stake Technology, provided the overall mass balance for the STAKETECH continuous steam explosion pretreatment process (4). The company gave the total capital for a complete ethanol plant. Since they preferred to offer a complete proprietary design to clients for ethanol production, we were not privy to the detailed information on each section. Thus, we requested the mass and energy consumption information and cost on just the pretreatment process for the Stake reactor system. We then used this information to interface the STAKETECH pretreatment into our generic design of the ethanol plant.

For the AFEX and MAFEX pretreatment processes, data on laboratory results were supplied by Mr. Earnest Stuart of the AFEX Corporation (26) and used to project yields and times for the pretreatment and hydrolysis step. The design of equipment such as pumps, tanks, agitators, filters and costs are based on the Katzen Associates report (1). For unique equipment, cost estimates are from vendor quotes.

In the Bioenergy case, only data in published reports from laboratory scale fermentations with genetically modified bacteria was used (17, 18). Thus, we designed the fermentors to achieve the same yields and concentrations in the plant as in the laboratory and used the Katzen (1) report as a basis for costing the equipment.

D. Energy Estimates

The major mechanical energy of the section of the plant are developed by estimating the horsepower for all the major pumps, agitators, conveyors, mills, blowers, etc. The pump and blower power is estimated from the volumetric flow rate and ΔP pressure differential. The flow rates come from the mass balance tables and the ΔP's are taken as 80 ft H_2O head in most cases and 40 ft head in recirculation pumps in the enzyme reactor. The agitator power is based on horsepower per 1000 gal of working volume. For thick slurries - 1.5 HP/1000 gal, for thin slurry or concentrated solution - 1.0 HP/1000 gal and for water like fluids - .33 HP/1000 gal. The power for filter presses, centrifuges, evaporator were taken from comparable items in the Katzen (1), Chem Systems (3) or TVA (2) reports and scaled linearly with throughput.

The major thermal loads of the plant are associated with pretreatments such as AFEX and STAKETECH, the distillation of the ethanol from the hexose and pentose fermentations, the evaporators in concentrating sugars for fermentation, and the stillage.

The distillation energy varies with the concentration of the ethanol in the beer or fermentor broth. The distillation is modelled as a binary water / ethanol separation. Thus, the energy for a typical stripper-rectifier with 29 theoretical stages and feed introduction on stage 15 from the top was estimated by adjusting the reflux ratio to achieve desired stages using a rigorous distillation model called RADFRAC on the proprietary P.C. program called MAX (ASPEN Technologies, Cambridge, Massachusetts). Figures 3-2 and 3-3 give the

Figure 3-2
Specific energy for distillation of hexose fermentation broth for 29 theoretical stages

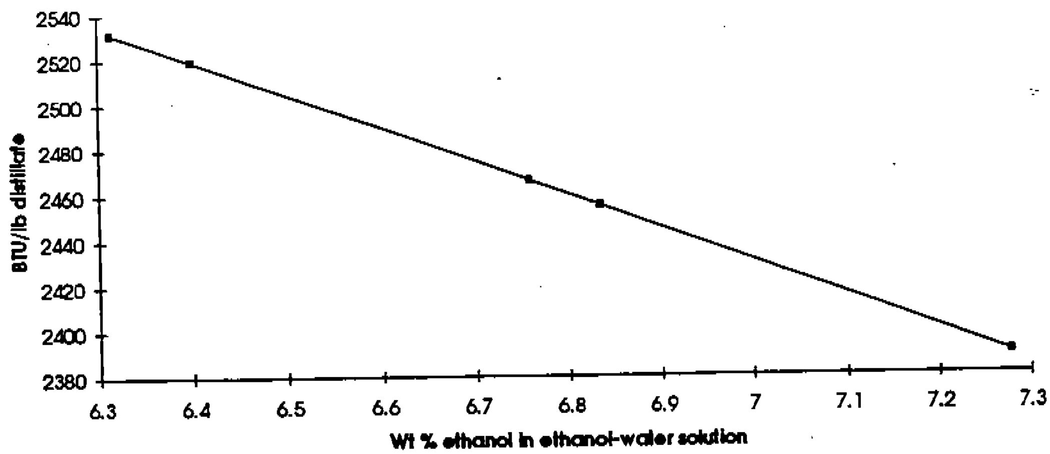
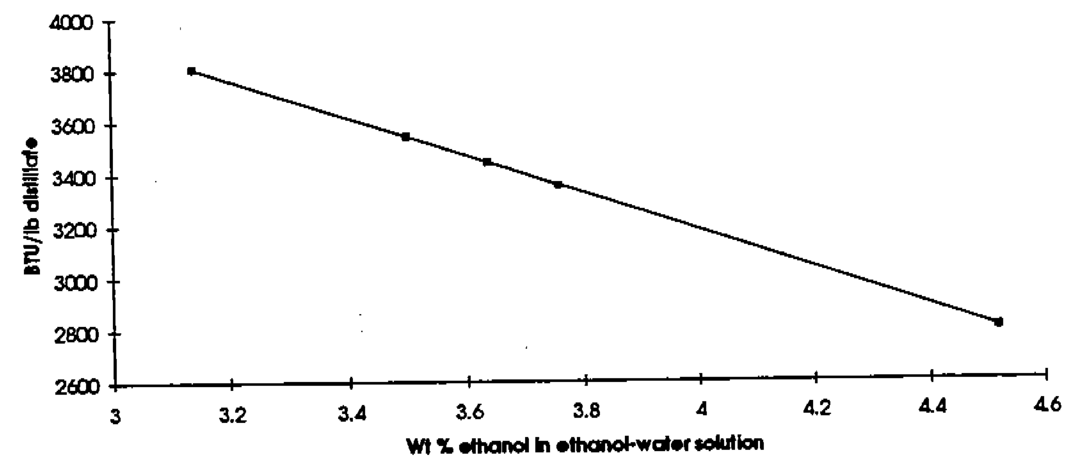


Figure 3-3
Specific energy for distillation of pentose fermentation broth for 29 theoretical stages



results of the reboiler duty per pound of distillate over the feed concentration range found in this study for the hexose and pentose fermentations, respectively. A least square line is fit to the data to get a model to estimate the energy for the distillation as a function of feed composition. This feed concentration in this graph is based on the binary system of ethanol in the water-ethanol mixture. In a given design, the binary concentration of ethanol is obtained from the corresponding mass balance table. Evaporator energy is estimated based on the water removed which is also obtained from the mass balance table and the steam efficiency of the multi staged evaporator of 6/1.

The thermal energy and electrical energy for each section of the plant is summarized in a table for each case (for examples, see Table 2 for any run in the Appendices). In the second part of this table, the heating value of the lignin and organic residue available from the stillage is calculated. This material is evaporated to a syrup of 50% total solids and burned. Thus, there is one pound of water per pound of organic fuel. The net heating energy of the fuel is also shown to account for the water evaporation in the boiler. The steam is generated in a set of 3 packaged Cleaver Brooks boilers modified to burn syrup. The steam is generated at 300 psi and 650°F and run through an extraction turbine. For the given boiler efficiency, 84% of the net energy of the fuel ends up as thermal energy to make steam. In turn, the high pressure steam passes through the turbine and for every 1000 pounds of stream, 138,000 BTU of electricity (or 40.43 KWH) are generated for a turbine exhaust of 50 psi. If the electrical load of the plant is less than that generated, the excess is exported to the power grid as the law allows at the avoided cost of new capacity of the local power company. If there is not enough thermal or electrical energy from the boiler-turbogenerator, natural gas or electricity is purchased to balance the plant load. This happens in cases where there is a high conversion of cellulose to glucose.

E. Common Process Sections

E1. Fermentation - Section 400

The fermentors in Section 400 are standard batch fermentors used in a corn to ethanol plant using a cycle time of 48 hours. A typical process flow diagram is given in Figure 3-4 which is taken from the Katzen report (1) for 50 million gallons/year capacity of ethanol. It is quite a simple layout. There are 16 fermentors, with 250,000 gallons capacity, which time share a number of cooling loops with recirculation pumps and heat exchangers since the major heat load of each fermentation cycle is during the first 12 hours. The CO_2 from all the fermentors is scrubbed with water to recover any ethanol that has evaporated with the CO_2.

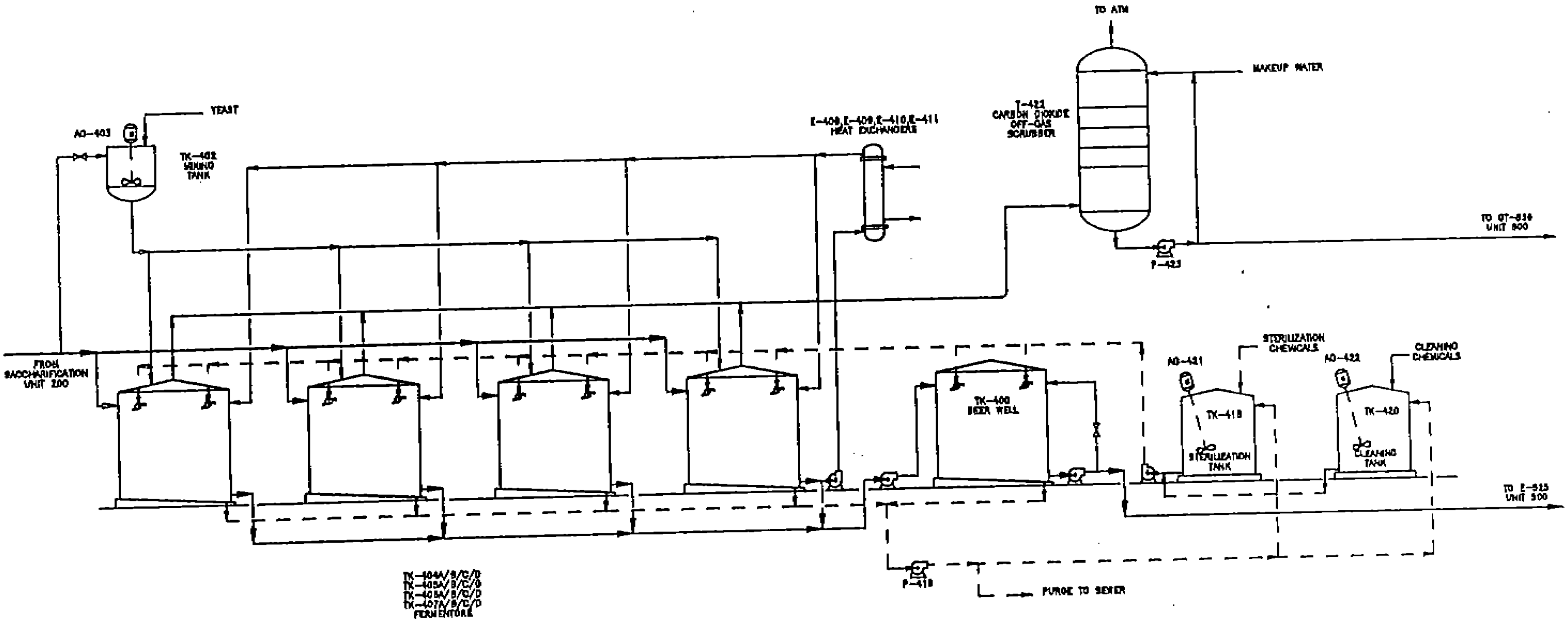


Figure 3-4: FERMENTATION SECTION FOR ETHANOL PRODUCTION KATZEN DESIGN (1)

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Finally, there is a common surge tank or beer well to level out the batch cycles into a continuous flow to the distillation section. The cleaning and sterilization tanks and pumps service all the fermentors. When regular yeast (*S. c.*) are used they are purchased from a supplier. In this design, the yeast are not recycled.

This basic configuration will be used for all case studies in this report. For separate hexose and pentose fermentations, a separate train of fermentors will be added which share the common CO_s scrubber and cleaning and sterilizations tanks.

Rather than get into laying out the individual pieces of equipment in Section 400, we developed the following cost equation which is a function of capacity to handle all the cases in this report. The purchased cost for equipment in the fermentation section in the Katzen report in today's dollars is \$5,346,000 for a 50 million gal/y plant. Of this, \$4,826,000 is related to the 16 fermentors, recirculation pumps and heat exchangers and the balance \$520,000 is for the common CO_2 scrubber and cleaning tanks.

The fermentor volume is scaled up to 310,000 gallons to allow an 80% working volume of 250,000 gal. The cost per fermentor with cooling loop is:

$$\frac{4{,}826{,}000}{16}\left(\frac{310{,}000}{250{,}000}\right)^{0.6} = \$343{,}180 \textit{ per unit} \tag{4}$$

The remaining common equipment will be scaled back to 25 million gallon service or:

$$520{,}000\left(\frac{25}{50}\right)^{0.6} = \$343{,}000 \tag{5}$$

As we consider different cases in this study, the number of fermentors changes because of the time of the fermentation and the concentration of the ethanol. Since all cases are adjusted to produce the same annual production of ethanol, cost equation is:

$$\frac{\textit{Purchased Equipment}}{\textit{Cost of Section } 400} = \frac{\textit{Number of Hextose}}{\textit{Fermentor}} * (343{,}180) + \frac{\textit{Number of Pentose}}{\textit{Fermentor}} * (343{,}180) + 343{,}000 \qquad (6)$$

To obtain the installed cost of the equipment in the fermentation, the purchased equipment cost is multiplied by the factor of 3.5 which is consistent with the Katzen design.

E2. Distillation and Dehydration - Section 500

Distillation of ethanol from fermentation broth or beer is fairly energy intensive and is about equal to the energy needed in the dehydration step to make anhydrous ethanol. To avoid having to provide energy in the form of steam to both distillation and dehydration, a number of energy savings designs are available. In the Katzen report (1), a patented pressure staged process (7) is used in which the beer still is operated at above atmospheric pressure so the overhead vapors are condensed in the reboiler of the dehydration column. There is also the need to preheat the feed to the distillation column to the high column temperature. This is done by extensive heat exchange with the very hot still bottoms. The complexity of such a continuous plant can be appreciated by referring to Figure 3-5 from the Katzen report (1). This design will be scaled by the 6/10 power to adjust for capacity changes in this report.

The purchased cost of the equipment in the distillation and dehydration section for a plant with a capacity of 50 million gallons per year is $3,344,000 in today's dollars. At this size, the liquid beer feed rate is 547,250 lb/h and the distillate feed rate to the dehydration column at 93.5 wt% ethanol is 41,917 lb/h.

Again, we will not get into the details of the design of Section 500, but will use an overall approach to adjust the cost for the feed rates. A schematic flow diagram for the fermentation section (400) and distillation and dehydration section (500) is given in Figure 3-6.

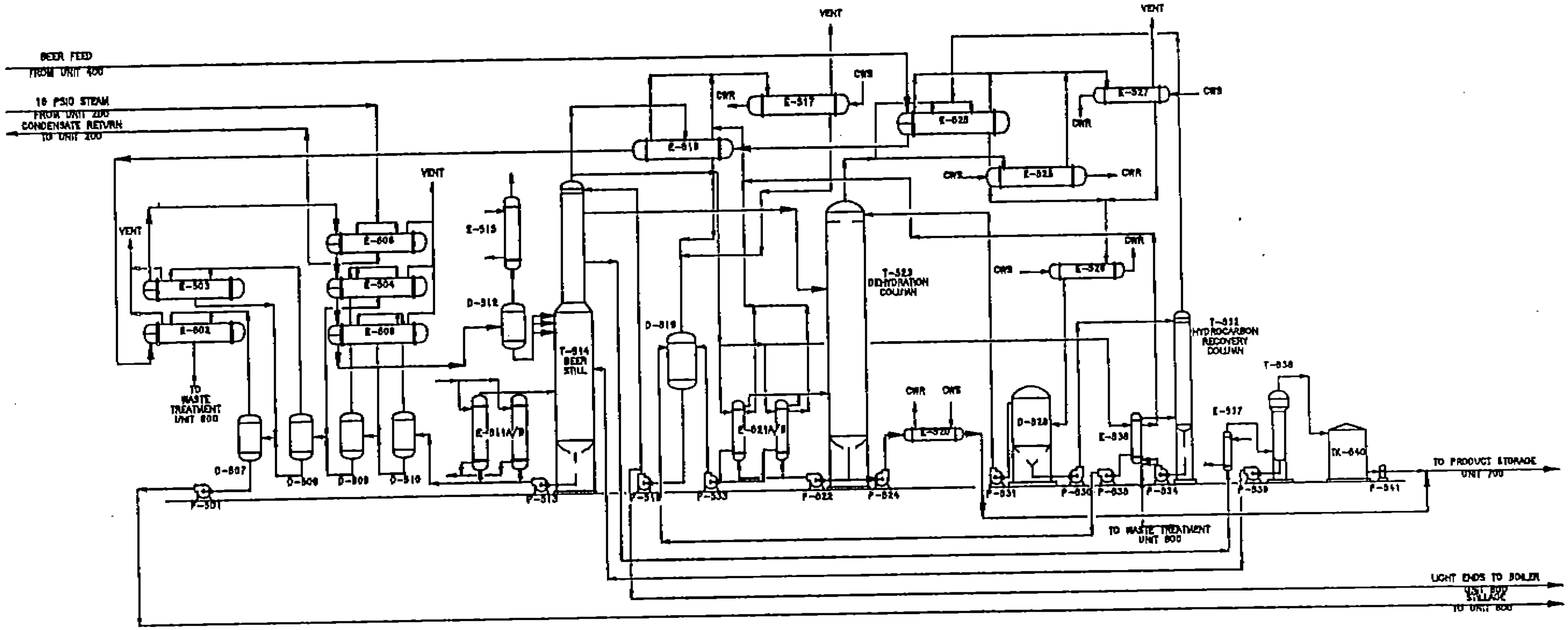

Figure 3-5: DISTILLATION AND DEHYDRATION SECTION FOR ETHANOL PRODUCTION KATZEN DESIGN (1)

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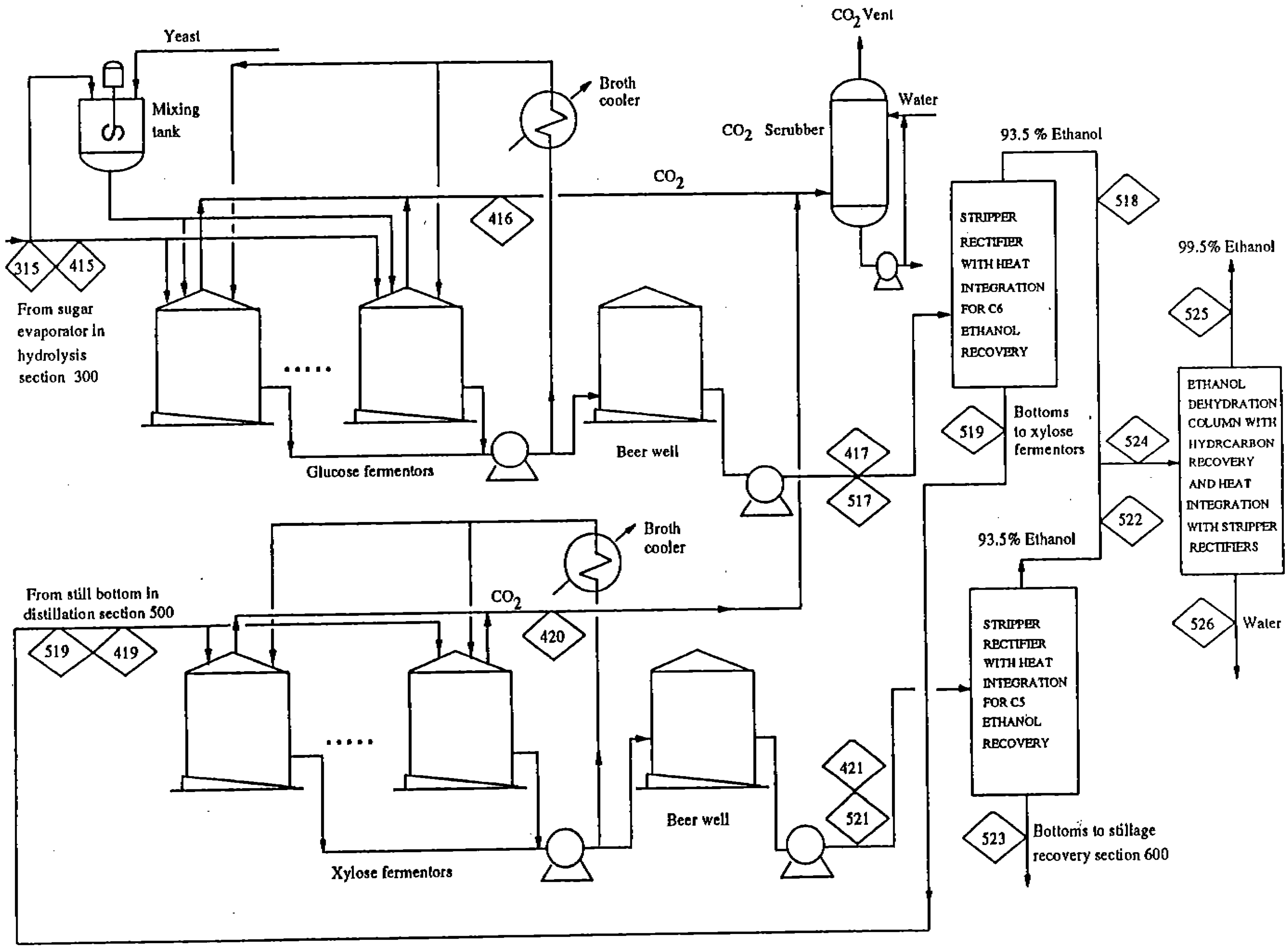
Figure 3-6 Fermentation Section 400/ Distillation-Dehydration Section 500



When there are two fermentors, there are two beer stills (stripper-rectifiers) and both of these feed in one dehydration column. We will use the same design concept of pressure staging and break the above cost into that allocated for the beer still and the preheating of feed ($1,850,000) and that allocated to dehydration and recovery of the hydrocarbon entrainer ($1,493,000). When the adjustments are made for capacity in our case studies to the reference capacity of the Katzen design, we developed the following cost equations:

$$\begin{matrix}\textit{Cost of Beer Still and}\\ \textit{Supporting Equipment}\end{matrix} = \$1,850,000\left(\frac{\textit{liquid feed lb/h}}{547,250\ \textit{lb/h}}\right)^{0.6} \tag{7}$$

$$\begin{matrix}\textit{Cost of Dehydration Column}\\ \textit{and Supporting Equipment}\end{matrix} = \$1,493,500\left(\frac{\textit{distillate lb/h}}{41,917\ \textit{lb/h}}\right)^{0.6} \tag{8}$$

When we have two beer stills, the total purchased cost of Section 500 is obtained with equation (7) applied to the liquid feed rates in stream numbers 417 and 421 from the mass balance table, respectfully. Then equation (8) is applied to the combined distillate in stream 524. When we have just one beer still, equation (7) is applied once to the feed into the still and equation (8) to the distillate. The final installed cost is 4 times the purchased cost, which is larger than the 3.5 installation factor used on other sections due to the large number of heat exchangers and pumps to heat integrate the units properly which leads to more field piping and installation costs.

E3. Stillage Evaporation - Section 600

The bottoms from the beer still contain only 245 ppm ethanol in an essential water slurry containing unused sugars, unhydrolyzed cellulose, lignin, cell mass and other components from the lignocellulosic biomass and fermentation nutrients. The stillage, as it is called, stream 523, is combined with the cell purge, stream 916, of the enzyme plant. The water is evaporated until it is equal to the total mass of suspended and dissolved solids to give a syrup at 50% total solids. This is used as the fuel in the packaged boilers in Section 800.

While in corn based ethanol plants the stillage is sold for animal feed due to its 20% protein content, this material is about 50% lignin in the biomass based ethanol plants. Thus, its value as animal feed is less likely and in order not to distort the economic analysis with wishful by-product credits, we decided to pay for the cost of evaporation and use the concentrated waste of the process as fuel.

The evaporator is a six stage multiple effect evaporator which comes as a packaged plant. The cost is taken from the Chem Systems report (3). There are two evaporators used in that report; one with a water evaporation rate of 328,000 lb/h and the other with a rate of 344,000 lb/h. Since the capacities are about the same, we took an average of these estimates and corrected to a CEI of 358. The evaporator cost equation is scaled by the 6/10 law for water rates other than 336,000 lb/h.

$$\textit{Purchased Evaporator Cost} = 2{,}680{,}000\left(\frac{358}{323}\right)\left(\frac{\textit{Water Evaporated lb/h}}{336{,}000}\right)^{0.6} \quad (9)$$

The assumption is that all the solids can be handled in the evaporator. If this is not possible, the suspended solids (mostly lignin) must be removed before the evaporation. The cost for removal of the lignin is considered in Chapter 5, Section C, Table 5-4.

In any case, the water balance is closed on the plant by the stillage evaporation and it will be an expensive equipment investment. Improvements in solid/liquid separation and concentration of residues have not attracted research and development money, but they are just as important as work on pretreatments, organisms or enzymes in reducing the cost of the plant.

E4. Product Storage and Denaturation - Section 700

The product storage and denaturation section is taken directly from the Katzen report (1) and reduced to 1/2 the size for a $25 million gallon plant. It consists of a storage tank for a 30 day supply with pumps and blend tanks to denature the 199 proof ethanol with 5% gasoline. Since this section does not change in the various alternatives, the purchased cost is fixed at $510,800 in today's dollars with an installation factor of 3.2 for an installed cost of $1,634,500.

E5. Boiler and Turbogenerator - Section 800

For the cheapest boiler cost we use package boilers. A quotation for a set of 3 packaged, high pressure boilers from Cleaver-Brooks (6) for a total capacity of 210,000 pounds of steam per hour was $3.8 million when fired by natural gas. When modified to burn an organic syrup at 50% solids, the cost is adjusted to $4.2 million. The installation factor suggested by Cleaver-Brooks is 2.3 to get an installed capital cost.

Since each case has a different steam demand, this basic reference quote is scaled up or down by the 6/10 law according to steam capacity.

For the turbogenerator, the cost is based on a recently (1993) installed cost of a unit at Michigan State University Simon Power Plant. The installed cost of a 24 MW turbogenerator is $5,830,000. Again, the cost will be adjusted for the MW capacity by the 6/10 law.

Thus, the installed cost equation for boiler and turbogenerator is:

$$\begin{matrix}\textit{Installed}\\ \textit{Cost}\end{matrix} = 4{,}200{,}000 \; x \left(\frac{\textit{Steam Required lb/h}}{210{,}000 \textit{ lb/h}}\right)^{.6} x \; 2.3 + 5{,}830{,}000 \left(\frac{\textit{KW required}}{24{,}000 \textit{ KW}}\right)^{0.6} \qquad (10)$$

The burning of the concentrated syrup may raise some technical and air pollution questions. There are analogous streams being burned successfully in the paper industry such as black liquor and in the Madison Process for acid hydrolysis of wood during WWII. There is low ash and very low sulfur in the organic residue of biomass conversion plant. For a definitive answer, one has to make the proper tests and evaluation from a pilot plant.

E6. Enzyme Production - Section 900

In starch hydrolysis, the alpha amylase and glucoamylase are not produced on the plant site because they are commodity enzymes available at low costs. For example, in a 25 million gallon per year ethanol plant, the annual cost for starch hydrolyzing enzymes is of the order of $600,000 or 2.4¢/gal. This represents about 225,000 pounds of enzyme protein per year.

Unfortunately, cellulase enzyme are not produced and sold at this type of cost. Moreover, depending on enzyme loading between 4 to 12 million pounds of cellulase protein are needed per year for 25 million gallons of ethanol. This order of magnitude difference between enzymes needed in starch and cellulosic hydrolysis is due in part by a difference in specific activity of the enzyme and in part by the difference in substrate accessibility. Moreover, commercial cellulases sell for $5 to $6 per pound of enzyme solution (20% solids), a cost far too high for large scale use in an ethanol plant. Thus, the only realistic way to estimate the cost of producing ethanol from biomass when enzymatic hydrolysis is used, is to produce the cellulase on site and estimate the cost of the enzyme plant. With this type of analysis, one can set a limit at what price commercial cellulase has to be before one can avoid the investment for a cellulase plant.

While the cellulase producing companies may have better economics for cellulase production, we will use a published design by Chem Systems (3). Their design is based on laboratory and pilot plant work of the U.S. Army National Laboratory in the 1970's. By using this design, we believe we are putting an upper boundary on the enzyme cost. It is not the most advanced design, but is a conservative design that will work.

The plant flow sheet for Section 900 is given in Figure 3-7. The production occurs in a series of batch aerobic fermentors using a part of the pretreated biomass as growth substrate. The fermentation time is 288 hours to achieve a broth titer with 15 FPU/ml or 22.5 mg/ml cellulase protein in the broth (0.667 FPU/mg cellulase). The major nutrient is corn steep liquor which is added at 759 lb CSL per 6179 lb of growth substrate carbohydrate (cellulose, hemicellulose and sugar). Special circumstances where cellulase at 40 FPU/ml is obtained from lactose (8) from cheese whey are available, but they are too limited for general economic analysis of biomass process alternatives.

When the batch is finished, it is separated from the fungal cell mass (*Trichoderma reesei*) by two centrifuges in series with repulping and washing steps. About half of the fungal cells are returned to inoculate the next fermentor and the balance is processed with the stillage in the evaporation section (600). The total cycle time per fermentor which includes cleaning, sterilization, loading, production, and discharge is 312 hours.

The original Chem Systems (3) design has an enzyme plant to produce the required cellulase for a loading of 15 FPU/g of substrate in a 25 million gallon per year ethanol plant. The fermentors have a total volume of 220,000 gallon with a working volume of 85% of the total. Heat is removed by a sidestream loop through a heat exchanger. We adjust the number of fermentors to meet the enzyme requirement for a given alternative and scale the rest of the equipment such as holding tanks, centrifuges, transfer pumps by the 6/10 law. The purchased equipment cost is multiplied by a factor of 3.5 to get the installed cost of Section 900.

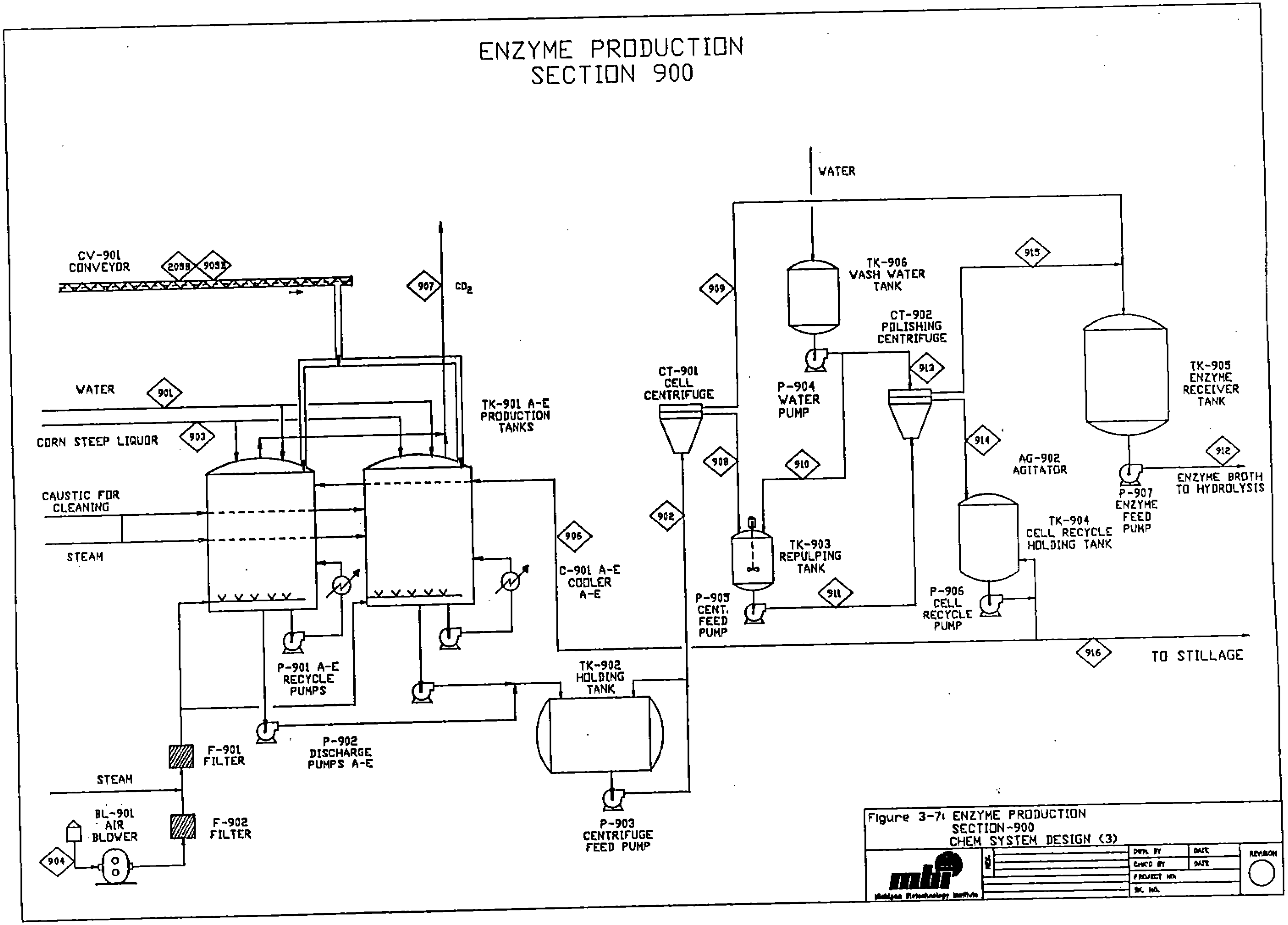


Figure 3-7: ENZYME PRODUCTION SECTION-900 CHEM SYSTEM DESIGN (3)

CHAPTER 4

TVA CONCENTRATE ACID HYDROLYSIS - BASE CASE

The TVA report (2) gives a design for 12 million gallons ethanol per year using corn stover as the substrate. The material balances for the various steps, based on their laboratory and pilot plant work, were used to scale up the plant to produce 25 million gallons. The detailed process flowsheets for Section 100 - Feed Preparation and Section 300 - Hydrolysis, are given in Figures 4-1 and 4-2 for the concentrated acid hydrolysis process. The flowsheet for Section 400 - Fermentation and Section 500 - Distillation is given in Figures 3-6, which is common for all the alternatives. An overall view of the integration of the hydrolysis section with the rest of the process sections of the TVA plant design are given in Figure 4-3 which conforms to the generic diagram of Figure 3-1.

A. Process Description

Section 100 - Feed Preparation

Corn stover is brought to the plant via truck and rail cars in a baled form. The bales are stacked by forklift trucks outside and covered with plastic sheeting. The management of the bales is on a first-in/first-out basis. The bales are processed through a tub grinder, which is a large open tub with a hammer mill type rotor which reduces the material to pass through a 4-mesh screen (4 wires per inch). The power consumption is reasonable at 325 HP per 40 T/h throughput (22). The shredded biomass is conveyed to a surge bin from which it is metered to the hydrolysis process.

FEED PREPARATION
SECTION 100

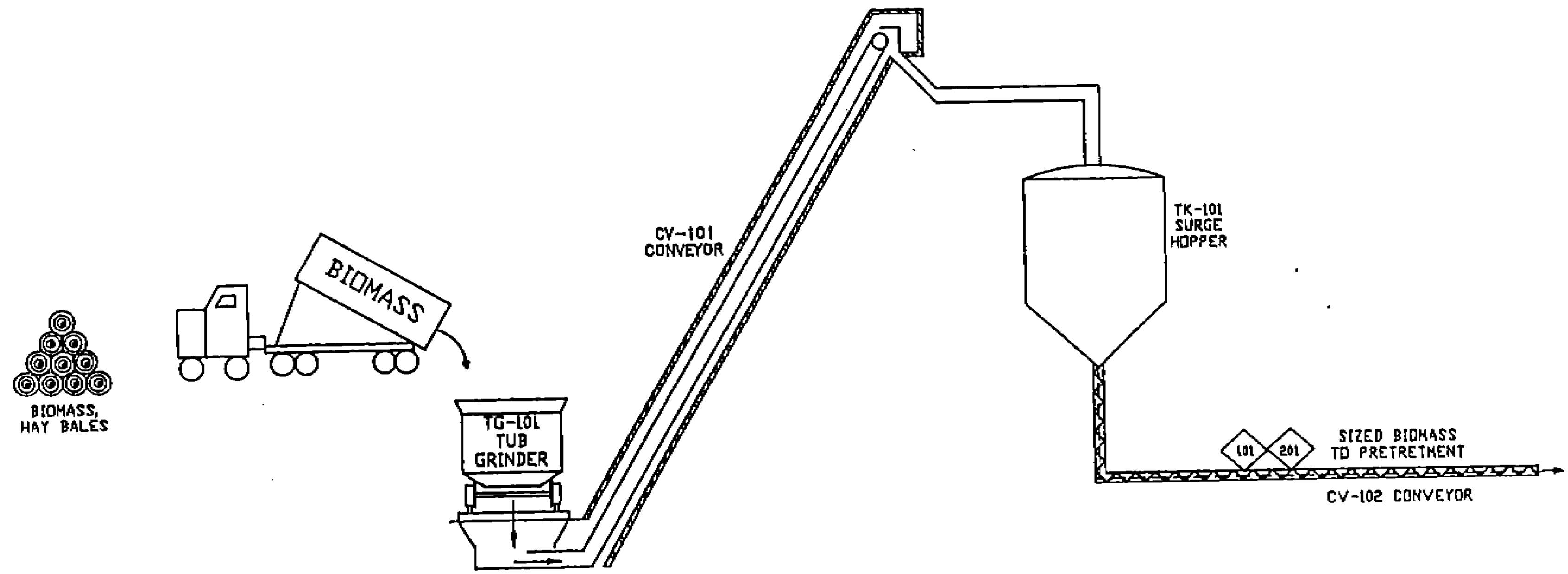


Figure 4-1: FEED PREPARATION SECTION-100

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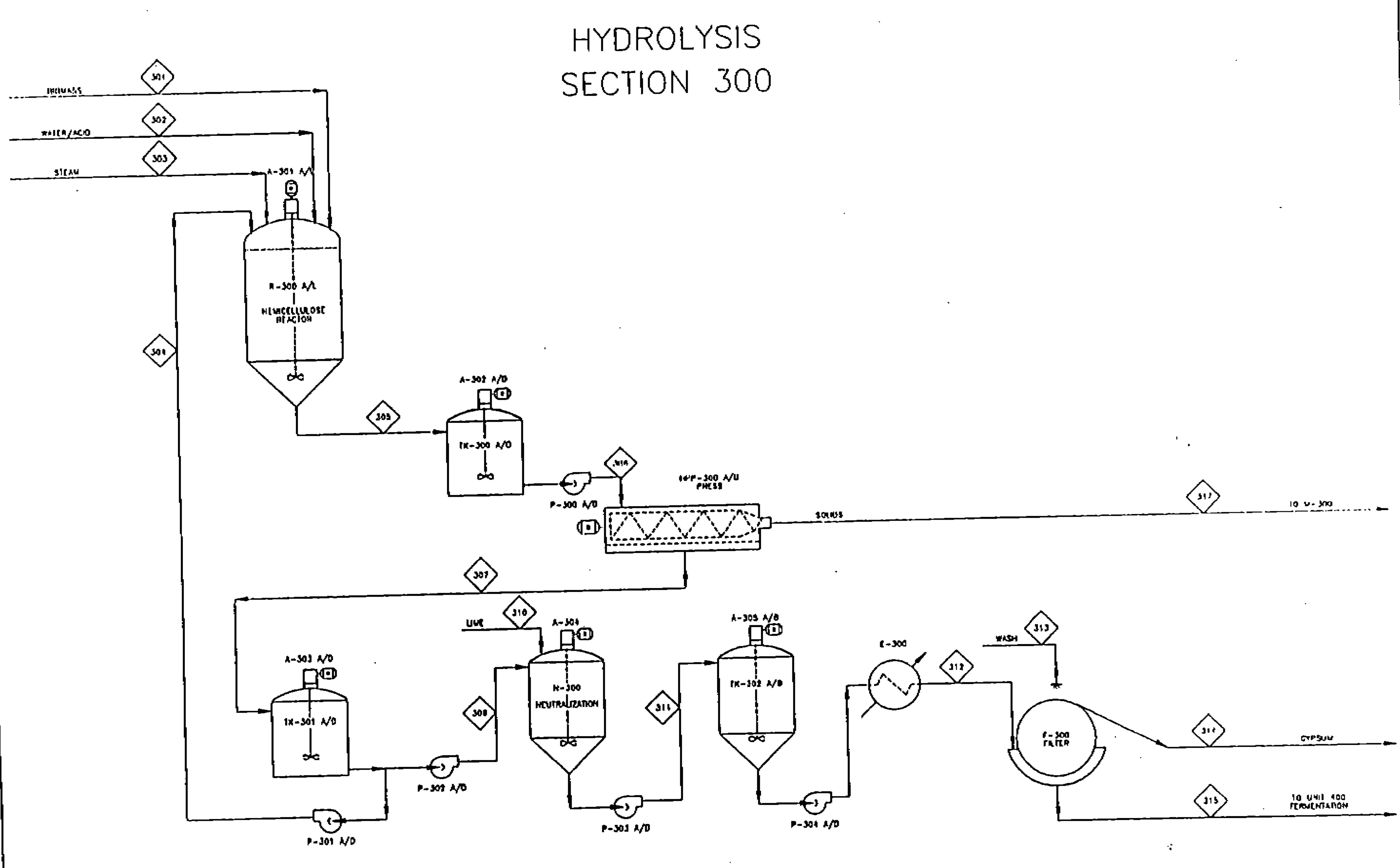

Figure 4-2a: TVA CONCENTRATED ACID HYDROLYSIS (2)

HYDROLYSIS
SECTION 300

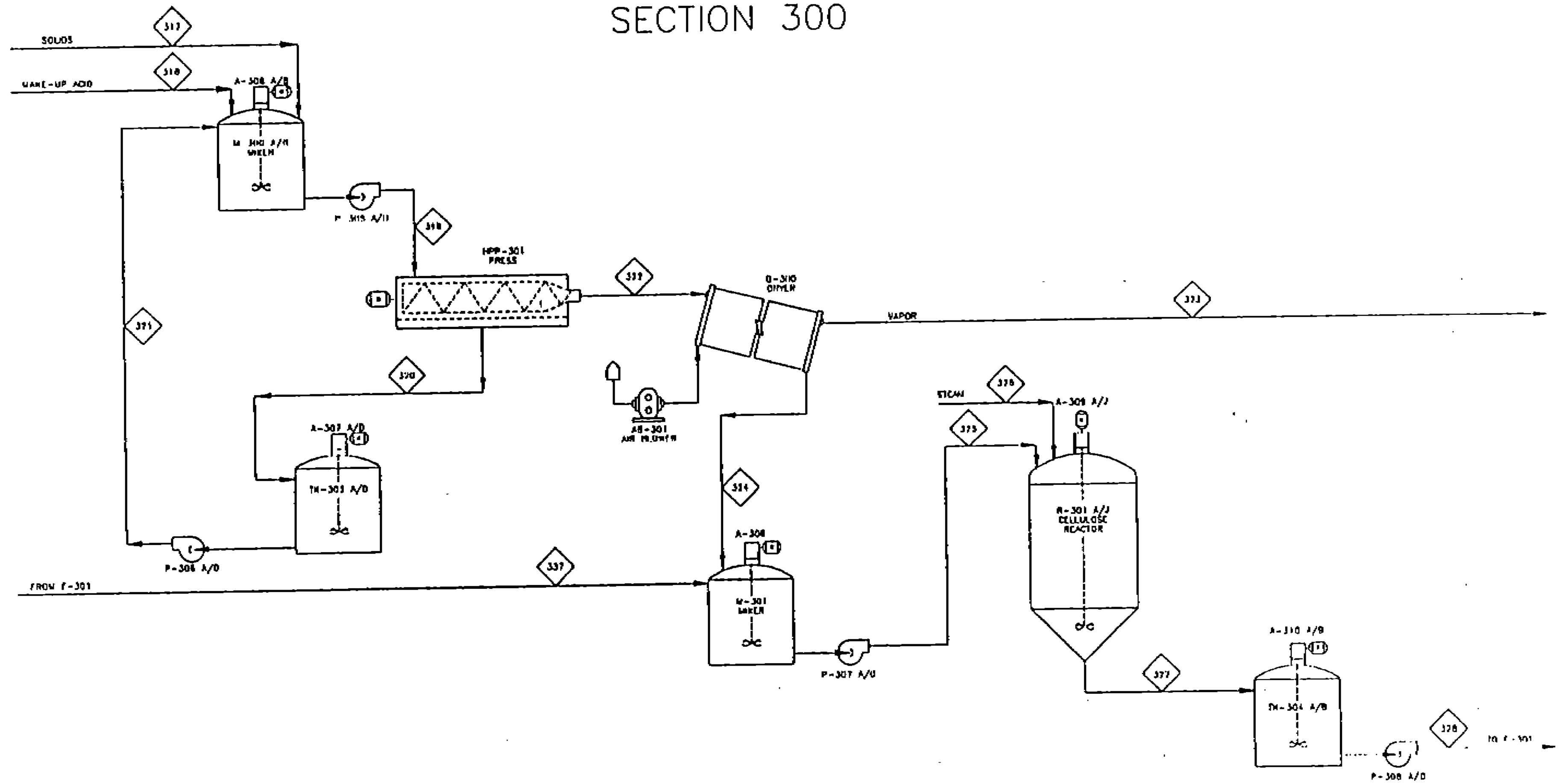

Figure 4-28: TVA CONCENTRATED ACID HYDROLYSIS (2)

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HYDROLYSIS
SECTION 300

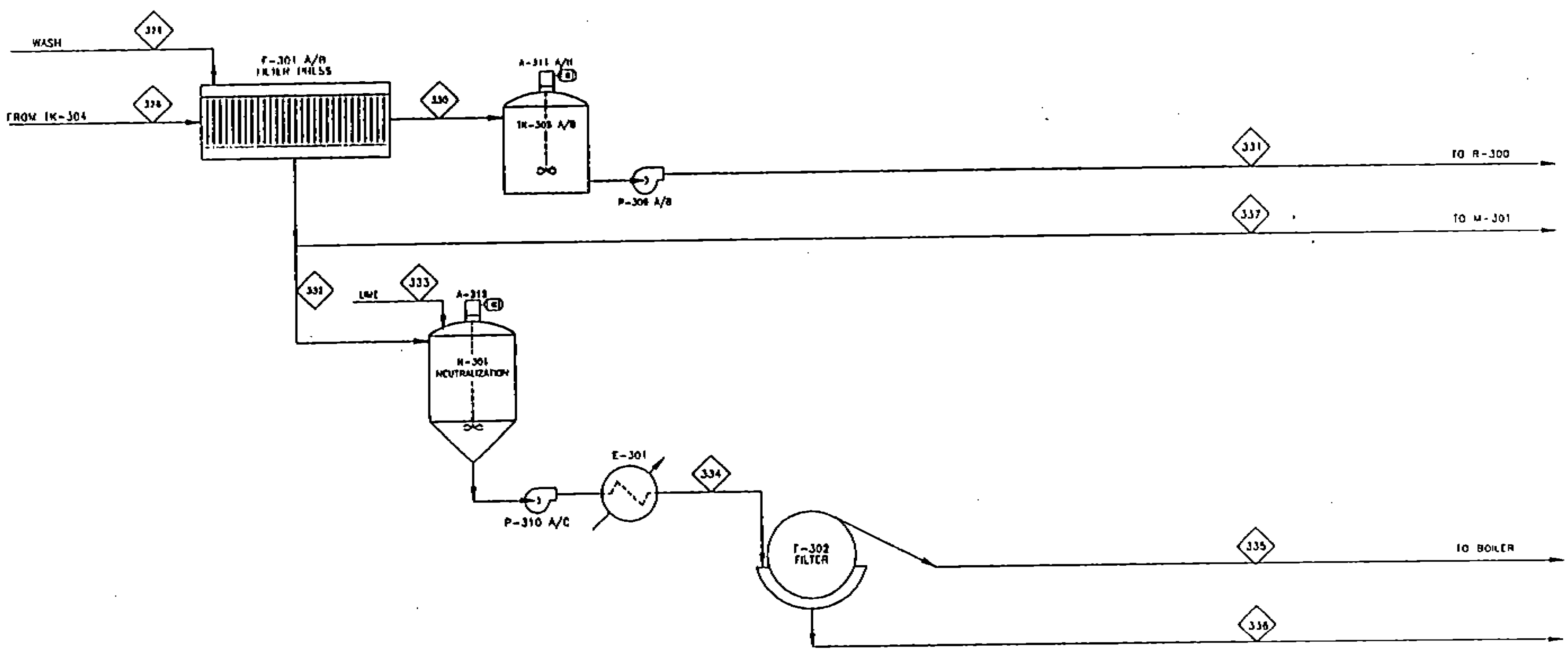

Figure 4-20: TVA CONCENTRATED ACID HYDROLYSIS (2)

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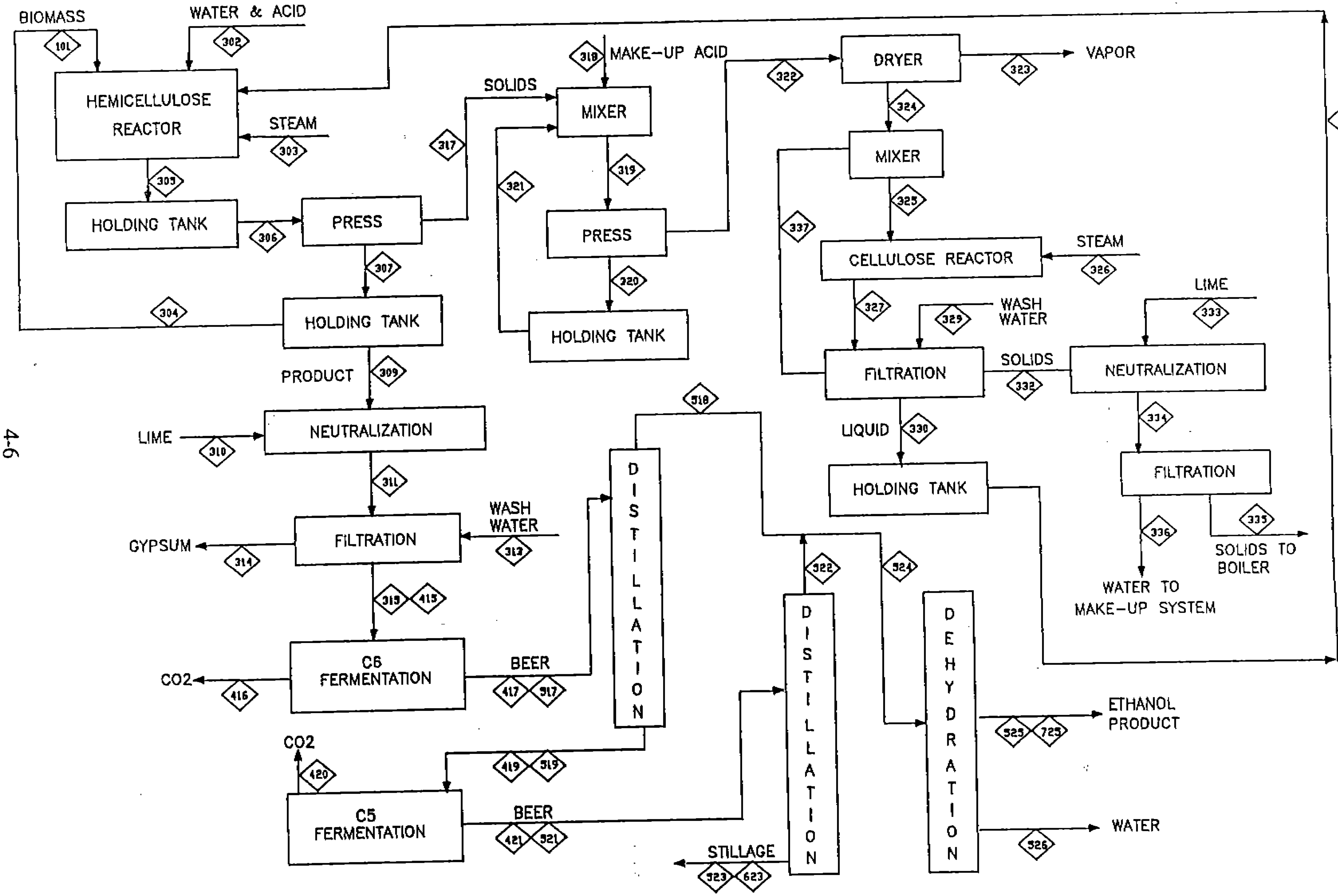
Figure 4-3: TVA Concentrated Acid Hydrolysis Process Diagram



Section 200 - Acid Hydrolysis

The hydrolysis section is divided into several parts. First the biomass is mixed with recycled dilute acids and sugars to hydrolyze the hemicellulose in hemicellulose reactor R-300. The slurry is mixable at a liquid to solid ratio of 10:1. The acid concentration is about 7.6 wt%. The reaction temperature is 100°C for 2 h and results in 80% of the hemicellulose being hydrolyzed. Live steam is injected to maintain the temperature. Because of the recycle of acidified sugar streams from the cellulose reactor, no new acid addition occurs at R-300. The slurry is then held in a holding tank from which it is pumped to a reciprocating screw press (HPP-300). The press delivers about a 50% solids cake which passes on to the mixing vessel M-300 where fresh acid is added for the preparation of the concentrated acid hydrolysis steps down the line in the drier D-300 and reactor R-301.

The liquid effluent from the press is collected in another holding tank (TK-301). About 63% of the acidified sugar is recycled back to R-300 to provide a way to build up the pentose concentration and provide the liquid to prepare more slurry. The remaining flow from TK-301 exits from the acid hydrolysis reaction section of the plant and goes to the neutralizer N-300. The hexose sugars from the cellulose hydrolysis are included in this stream because of the recycle from R-301 back to R-300.

Lime is used in 10% excess to insure the neutralization of the H_2SO_4. Gypsum ($CaSO_4$) slurry is formed in the sugar solution. Since $CaSO_4$ is less soluble as the temperature increases, the neutralization is done on a hot solution. The gypsum is removed by a rotary vacuum filter with a wash cycle to reduce the sugar loss in the gypsum cake. The clarified effluent from the tank is ready for the fermentation section.

In order to get complete penetration of the acid into the biomass particles, the liquid to solid ratio in mixer M-300 is increased to 10 by using the recycle stream 321. The liquid to solid ratio is reduced in the second high pressure press HPP-301. The acid concentration in the liquid phase is about 33.5%. In order to get the acid to 75%, sufficient water is evaporated from the press cake in the oven dryer D-300.

During this drying step the concentrated acid completes the hydrolysis of the biomass cellulose and hemicellulose, similar to the laboratory quantitative saccharification procedure. In concentrated acid all the carbohydrates are converted to low molecular weight oligomers and the lignin with ash remains insoluble.

In order to convert the oligomers to monomer sugars, a post-hydrolysis is completed in reactor R-301 at lower acid concentration. The dilution comes from the wash water in the subsequent filtration step in F-301. The hydrolysis in R-301 is completed at 100°C for 4 hours. All the remaining hemicellulose and 90% of the cellulose are hydrolyzed. Again, live steam is used to maintain the temperature. The separation of the unreacted lignin is accomplished in an automatic filter press with a wash cycle. Part of the liquid effluent is directed back to mixer M-301 to provide about a 5:1 liquid to solid ratio to facilitate the flow and mixing in R-301. The remaining acidified effluent liquid now rich in hexose and pentose is recycled back to R-300. The lignin cake is neutralized in N-301 with lime and dewatered in a rotary vacuum filter F-302. The wet cake at 10% moisture is sent to the power plant to be the major fuel for operating the plant. The dilute sugar solution from the press is lost to the process and sent to the waste water plant.

Section 400 - Fermentation

Since the biomass has both cellulose (45%) and hemicellulose (29%) (see the composition in column one of Table 1 in Run 4-1 in Appendix 4), both hexose (C_6 or glucose equivalent) and pentose (C_5 or xylose equivalent) are present in the solution stream 315/415 and sent to Section 400 - Fermentation. The respective sugar concentrations are 11.63% glucose and 9.00% xylose. At this point we used a conventional batch yeast (*Saccharomyces cerevisiae*) fermentation for the glucose as used in a typical Katzen (1) design for a corn based ethanol plant with a 48-hour cycle time (1). There are 6 fermentors with 310,000 gallon design capacity. The working volume is 85% of the design volume. The fermentation yield is 90% of theory which gives 5.63 wt% ethanol in the beer stream 417/517. While this is lower than the typical 7.5 to 10 wt% ethanol developed in the corn based fermentation plants, the ethanol concentration is high enough to be recovered in a conventional distillation plant.

Since xylose is not fermented with *S. cerevisiae*, the xylose is left unconverted in the beer from the C_6 fermentation which is sent to beer still. The beer still is a set of stripper/rectifier columns that strips the ethanol from the water phase at the bottom of the stripper column to about 245 ppm in stream 419/519. The bottom flow containing xylose, residual glucose and yeast is sent to the pentose (C_5) fermentor. The overhead from the rectifier (stream 518) is 93.5 wt% ethanol, which is near the azeotropic composition (95 wt%).

In the TVA design, a pentose utilization yeast, *Pachysolen tannophilus*, is used in the C_5 fermentor. This is also a batch process with an 84-hour cycle using 9 fermentors with 310,000 gallon capacity. The ethanol yield from xylose is taken at 50% of theory (2). Note that the remaining glucose is also utilized in this fermentation. The concentration of ethanol in the beer is 3.37 wt% in stream 421/521. This is on the lower edge of practice for the distillation. Since there are laboratory results on *P. tannophilus* that have given a 70% yield which will increase the ethanol concentration in the beer (2), 50% represents the least ethanol yield from xylose that can be achieved. If both C_6 and C_5 sugars are fermented to 90% of theory, the beer would have about 10 wt% ethanol from the TVA hydrolyzate.

Section 500 - Distillation

Since the fermentation is carried out separately for the C_6 and C_5 sugars, the ethanol recovery is also done for both beers. Thus the capital is more than if only one beer still (stripper/rectifying column) is needed. The first beer still receives steam 417/517 from the C_6 fermentor and operates at a reflux ratio of 4.0. The reboiler duty is about 2,470 BTU/lb of distillate (see Figure 3-2) when the feed is 5.63 wt% ethanol..

The second beer still receives the beer from the C_5 fermentor stream 421/521 and operates at reflux ratio of 5.0, because of the lower feed concentration (3.37%). The reboiler duty is about 3,265 BTU/lb of distillate (see Figure 3-3). The overhead from both beer stills (streams 518 and 522) is sent to one solvent dehydration column using gasoline as the entraining solvent to break the azeotrope of ethanol and water. The hydrocarbon is recovered in a separate column and recycled to the dehydration column. In this study it was not possible to design the distillation section in detail with all the needed heat exchangers and support equipment which are used to heat integrate the beer stills, dehydration column and hydrocarbon recovery column. However, conservative capital cost estimates for the distillation section are made based on the discussion in Chapter 3, Section E2.

Section 600 - Stillage Recovery

In the TVA process the sugars from the hydrolysis plant are free of solids. The lignin is removed at filter press F-302 while the sugars are the effluent from the neutralization filter F-300. Thus the only solids in the stillage are the combined cell mass of the yeasts developed in the C_6 and C_5 fermentations. However, the stillage is evaporated to 50% total dissolved solids in a multi-effect evaporator to recover both the organic matter in the cells and the remaining xylose in the stillage. The syrup is burned along with the lignin cake in the boiler.

B. Process Analysis

The results of process simulation Run 4-1 are given in Appendix 4 (Tables 1 through 9). The material flows for the major process streams are given in Run 4-1, Table 1 where the stream numbers are identified in Figures 4-1, 4-2 and 4-3. The major energy loads in each section are given in Table 2 for the electricity for motive power of pumps, agitators, conveyer, etc. and the thermal energy for process heating. Also given in Run 4-1, Table 2 is the thermal energy available from the lignin and stillage. With a boiler combustion efficiency of 84% and a turbogenerator efficiency of 90%, both the thermal and electrical loads of the plant can be met. In fact, there is an excess of electricity available for export to the electric grid.

The major items of equipment for the feed preparation and acid hydrolysis sections (100 and 300) are given in Run 4-1, Tables 3 and 5 with the purchased cost in the first quarter 1993. For the TVA case, the cost of the fermentation (400), distillation (500), stillage (600), product storage (700) and utilities (800) are scaled down in Run 4-1, Table 6 to the required capacity as discussed in Chapter 3, Section E. A summary of capital investment for each section is given in Run 4-1, Table 8. The total capital investment is about $81,150,000. This corresponds to an investment of $3.40 per annual gallon of capacity which is higher than the $2 to $3 for a corresponding corn plant.

With the material, the energy and the capital requirements, the annual operating cost for the plant is given in Run 4-1, Table 9. Also shown is the cost per gallon of ethanol. The overall cost is 1.708 $/gal but with the credit for electricity the net cost is 1.632 $/gal. Note that the corn stover contributes about 38¢/gal whereas the acid and lime add 21 and 12¢/gal more. The disposal of the gypsum costs 33¢/gal, a real burden for this plant. This is a cost that is difficult to estimate but the $20/ton cost to remove wet gypsum cake is on the low side and will just go up over time. Labor costs run 18¢/gal and the capital related costs are 47¢/gal.

For a basis of comparison recall that in a plant of 50 M gal/yr of ethanol capacity using corn, the cost of ethanol is 1.57$/gal without DDGS credits. (9)

The operating cost sheet in Table 9 shows only the cost of manufacturing. There is no profit included. The capital cost is recovered in nine years or 11.1% per year. Since the return on the investment is a policy decision for a given business, it can be different for different parties. As a result, we left this issue aside and can at least compare process alternatives on their cost of production. Moreover, the costs generated in this report are on the same basis as those of the MBI report on ethanol from corn done for the USDA in July 1992 (9). For those who are interested to see the effect of capital recovery with interest (profit), refer to Table 4-1. Here, the contribution due to capital on the production cost per gallon of ethanol is given for $10 million investment increments for a variety of interest rates and years. The zero interest rate is the case used in this report to just recover the capital.

While the concentrated sulfuric acid process demonstrated at TVA is not competitive, the concentrate acid process in another form may be. As a case in point, the new concentrated HCl process developed by Dr. Irving Goldstein (13) has some significant advantages over the TVA process. Concentrated HCl (41 wt%) will give 90% more conversion of the cellulose and 100% of the hemicellulose as in the TVA case, but the HCl is truly recycled. Since HCl is volatile, the major amount of HCl is recovered by distillation, a concept well demonstrated in the Bergius process (21). The new aspect is to complete the recovery of remaining HCl by electrodialysis. The hydrolysis in concentrated acid is complete in 10 minutes and a post hydrolysis to convert the resulting oligomers takes one hour. A qualitative look at the equipment for the concentrated HCl versus the TVA process suggests that the capital cost will be lower with the exception of the electrolysis equipment. The latter costs are hard to pin down without realistic, pilot plant scale work. The membrane-flux and membrane life set the capital cost for this equipment.

Table 4-1
The Cost per Gallon of Ethanol Due to Capital Recovery with Interest per $10 Million Incremental Investment

	Interest Rate (%)				
Years to Recover Investment	**0**	**8**	**10**	**12**	**15**
9	4.44	6.40	6.95	7.50	8.38
10	4.00	5.96	6.51	7.08	7.97
15	2.67	4.67	5.26	5.87	6.84

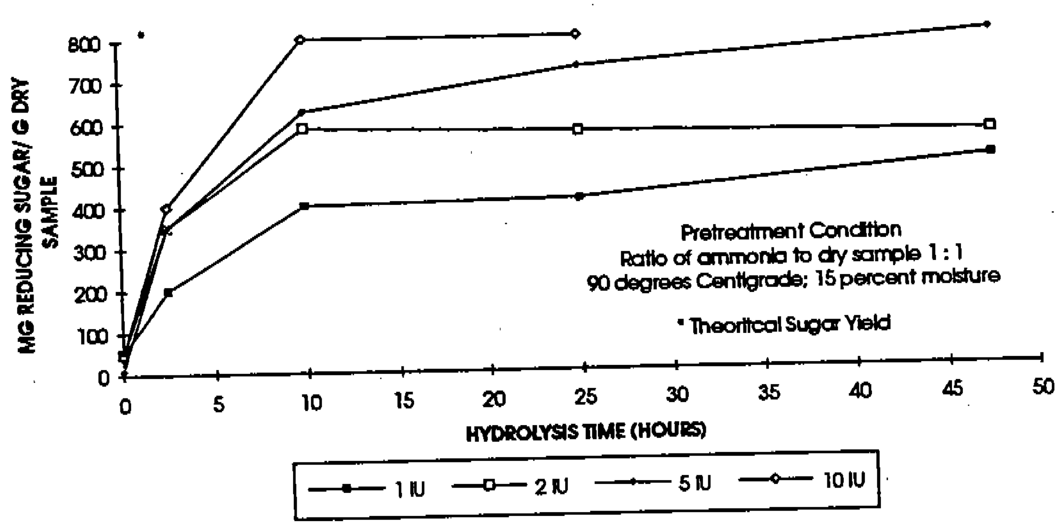
Even if the capital costs are similar, the concentrated HCl process will save about 50¢/gal in acid, lime and disposal costs. If the total capital costs are lower, there is a further saving. Thus the cost could come in $1.00/gal range as some of the other alternatives discuss in the next chapters. As a result, the concentrated HCl process should be given a careful evaluation and supported for further development if merited.

CHAPTER 5

AFEX PROCESS WITH ENZYMATIC HYDROLYSIS

The AFEX process is a proprietary process of the AFEX Corporation, Brenham, Texas which uses ammonia to pretreat lignocellulosic biomass. The ammonia swells and decrystallizes the cellulose/hemicellulose so that the material is very accessible to cellulase. Since the ammonia treatment is done from room temperature to 90°C, there are no thermal decomposition products such as furfural, hydroxmethyl furfural, levulinic acid, formic acid, etc. formed with this pretreatment as is the case of the better known steam explosion and dilute acid pretreatments. Also, the protein which can be 10 to 20% in grasses is not degraded in the AFEX process. This gives the process a wide range of substrates for which it can be used.

FIGURE 5-1 : SUGAR YIELD BY CELLULASE HYDROLYSIS OF AFEX PRETREATED SWITCHGRASS FOR VARIOUS ENZYME LOADINGS, IU/g



The laboratory data from AFEX is on Bermuda grass and switch grass. The process has been demonstrated at the lab scale where ground "as is" biomass is mixed with anhydrous ammonia in a pressure vessel. After a specified holding time at a given temperature, the contents of the vessel are suddenly opened causing the liquid ammonia to explode or flash off the biomass. The solids are dried of ammonia and evaluated in a cellulase assay to determine the conversion rate and the extent of potential sugar conversion. Typical enzymatic hydrolysis curves of AFEX pretreated hay are shown in Figure 5-1. Note that it is possible to get essentially complete conversion (98% of theory) in 24 hours of hydrolysis with a cellulase loading of only 5 IU/g of substrate. Naturally at 10 IU/g the rate is even faster. The most significant observation is that 1 and 2 IU/g give 50 to 70% conversion in 24 hours or less. These results are unique to AFEX. Other studies on pretreatments using steam explosion or dilute acid require 10, 15, 20 or more IU/g to get practical conversions in 24 hours.

In order to evaluate the economic potential of the AFEX process, some engineering judgement is needed to design a process with high recovery and reuse of the ammonia. Since the details of the process are proprietary, the flow diagram and equipment list will be aggregated to show overall flows and capital cost. Also the hydrolysis experience observed for AFEX pretreated switch grass is taken to apply to AFEX pretreated corn stover. From the author's past experience on steam explosion and dilute acid hydrolysis pretreatments, it was found that similar yields of sugars were obtained from a variety of agricultural residues when applying a given pretreatment. By staying with corn stover in the base case and the other alternatives, we avoid changes in substrate flow and cost in the plant design due to changes in substrate composition.

B. Process Description

The AFEX process uses, where possible, similar sections as in the base case design. This is done in order to be able to see specific impacts of the alternative process. Thus, the fermentation, distillation, stillage recovery, product storage, and utilities sections are the same design as in the base case except the capacities may be changed to accommodate any change in mass flows. All streams are adjusted so the ethanol is 25 million gallons per year or 21,269 lb/h.

Since there is an important effect on the process design and economics as the solids concentration is increased in the enzymatic hydrolysis, two versions of the enzymatic hydrolysis section are presented: one for 10% solids (see Run 5-4) and another for 20% solids (see Run 5-9). The process flowsheets are given in Figures 5-2 through 5-5 for the initial AFEX pretreatment with 10% solids in the enzymatic hydrolysis and is referred to as AFEX-10. Then, the modified hydrolysis section with 20% solids is given in Figure 5-6 and is referred to as AFEX-20.

Section 100 - Feed Preparation

The feed preparation section, shown in Figure 5-2, uses a similar front end as in the TVA case with a tub grinder to give the initial particle size reduction. In order to achieve the desired 40 mesh sized particles of biomass, an attrition mill is added to process the material leaving the surge bin. The attrition mill, or burr mill as it is also called, uses about 1/3 the power of a hammer mill for the same capacity. Two mills are needed, each processing 25 tons per hour with the consumption of 50 hp (23). The particle size is controlled by a sieving screen or sifter with the larger particles recycled to the mill. The mill clearance can be adjusted while in operation in order to produce the correct particle size range.

Section 200 - Pretreatment

The metered ground biomass is mixed with the ammonia in a pressurized system. The ammonia treated biomass is held for 30 min at 90°F and 165 psi. As a result, the lignocellulosic cell walls of the biomass are swollen and completely decrystallized. The effect is that the biomass is very accessible to cellulases.

It is not known if the reactivity of cellulose can be made to approach that of starch by proper pretreatment. However, a real long-term opportunity exists in the optimization of the AFEX process with respect to particle size, ammonia to solid level, time, temperature and energy input. All these variables affect the rate and yield of biomass conversion to sugar and

FEED PREPARATION
SECTION 100

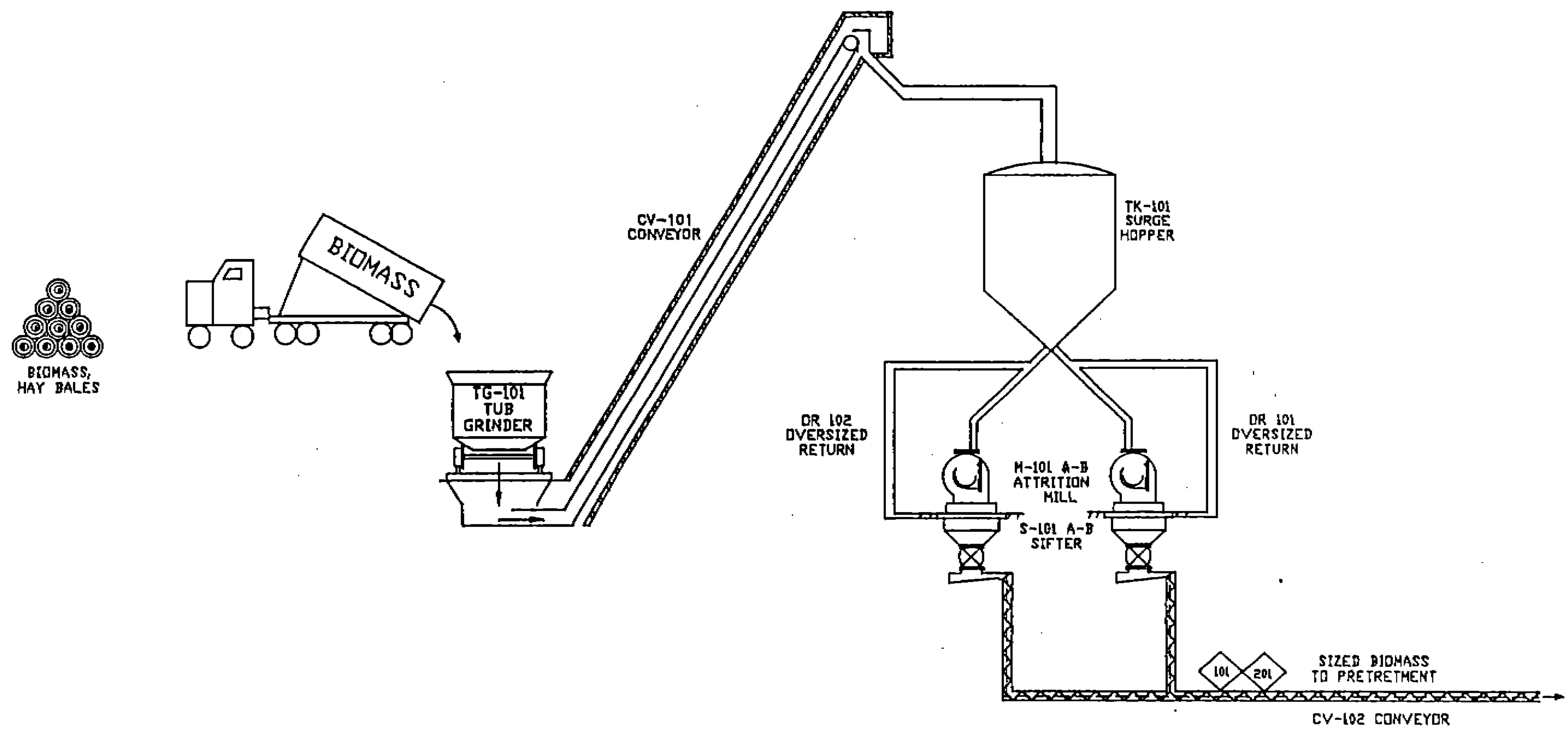


Figure 5-2: FEED PREPARATION SECTION-100
AFEX PROCESS

AFEX PRETREATMENT
SECTION 200

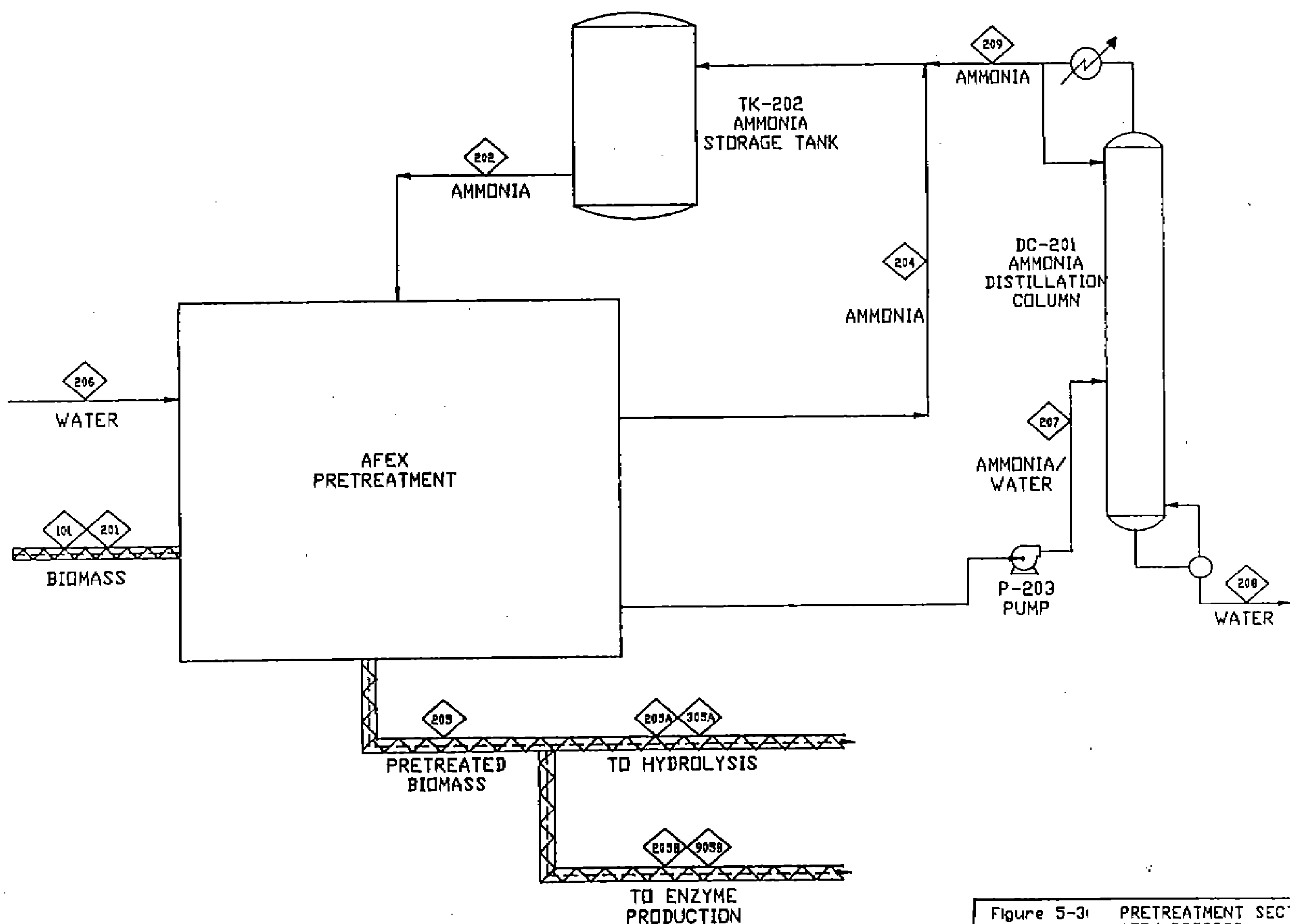


Figure 5-3 PRETREATMENT SECTION-20
AFEX PROCESS

mbi Michigan Biotechnology Institute

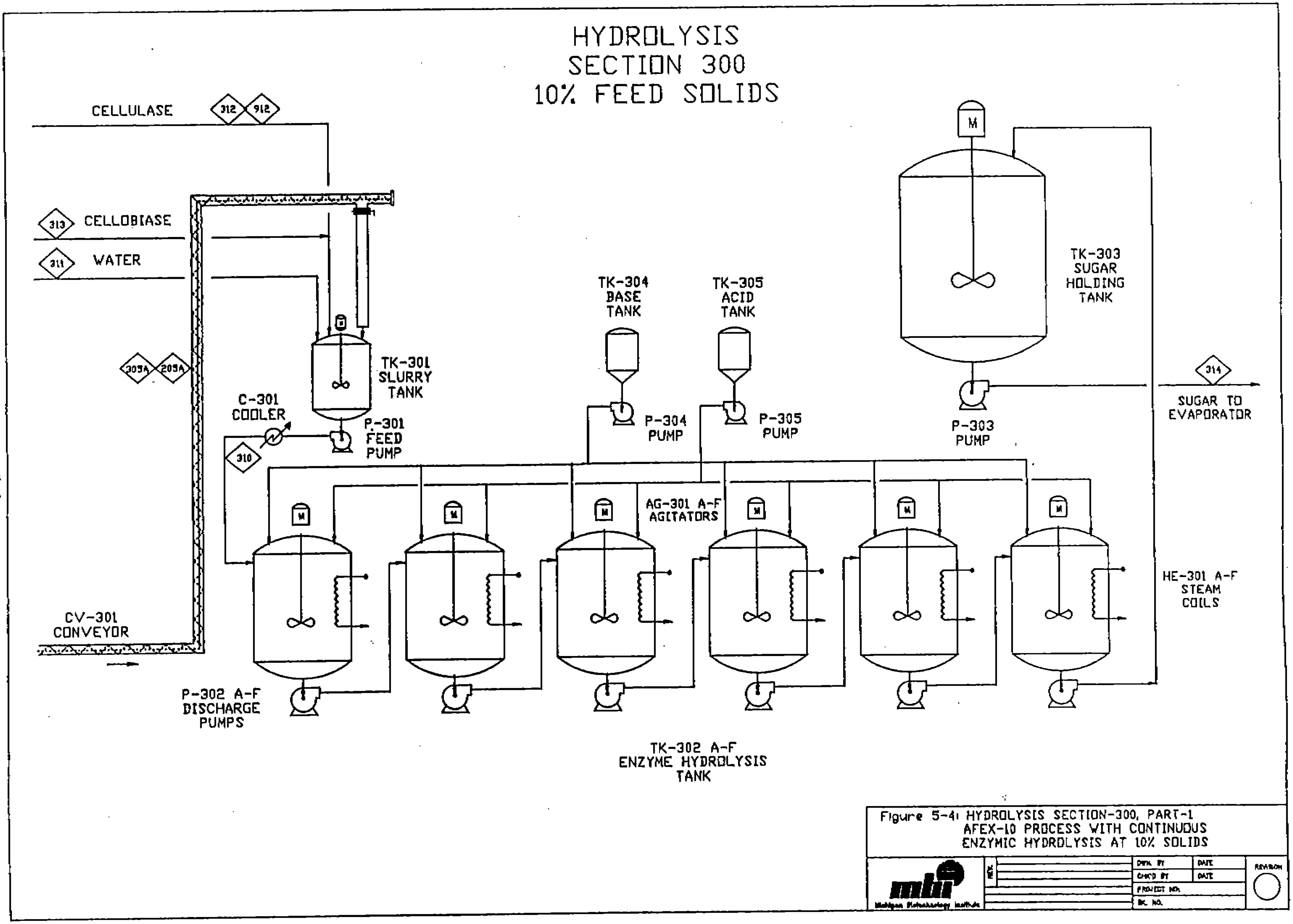


Figure 5-4: HYDROLYSIS SECTION-300, PART-1
AFEX-10 PROCESS WITH CONTINUOUS
ENZYMIC HYDROLYSIS AT 10% SOLIDS

HYDROLYSIS
SECTION 300
MULTI-EFFECT EVAPORATOR
(PACKAGE BY VENDOR)

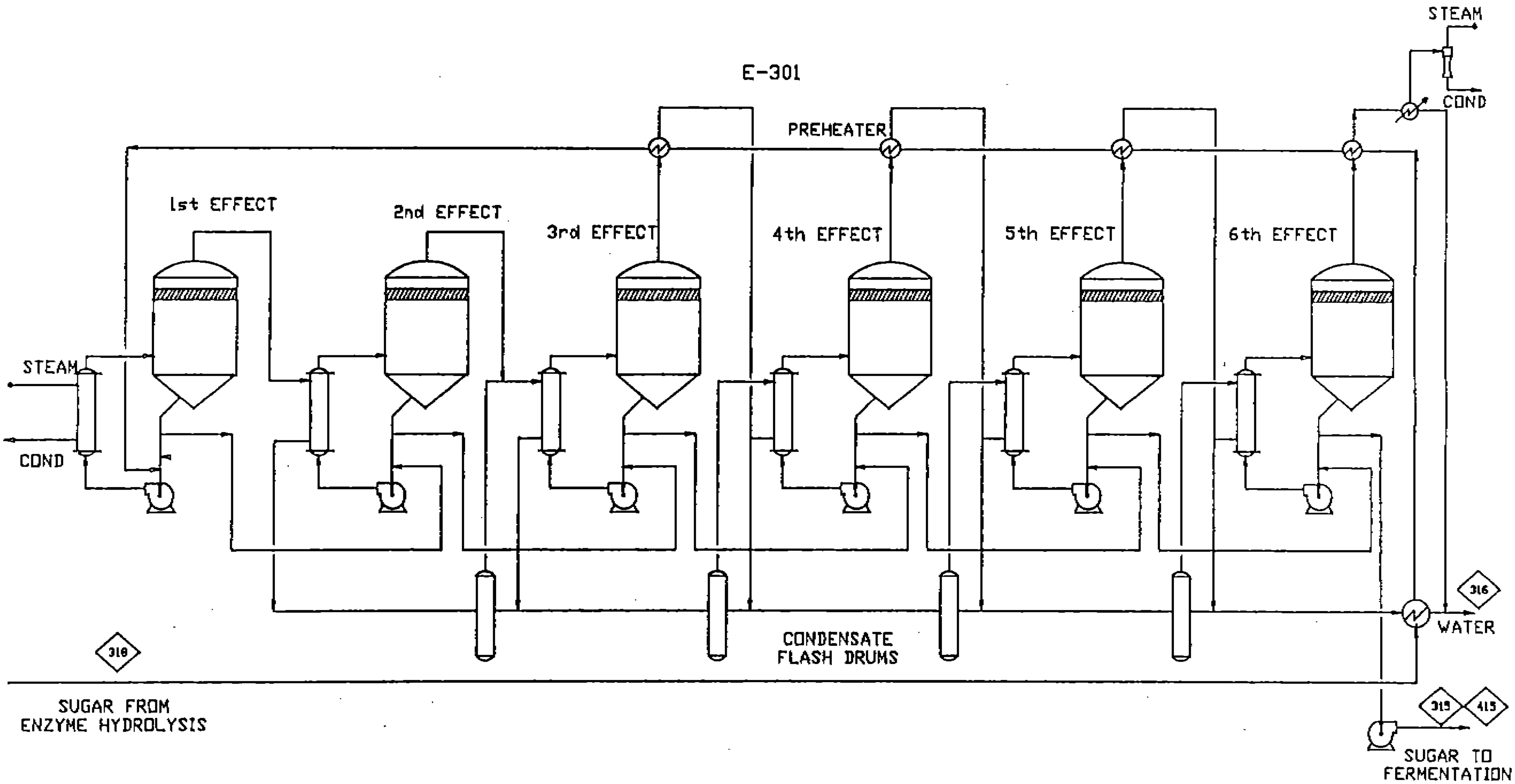


Figure 5-5: HYDROLYSIS SECTION-300 PART-2
MULTI-EFFECT SUGAR EVAPORATOR

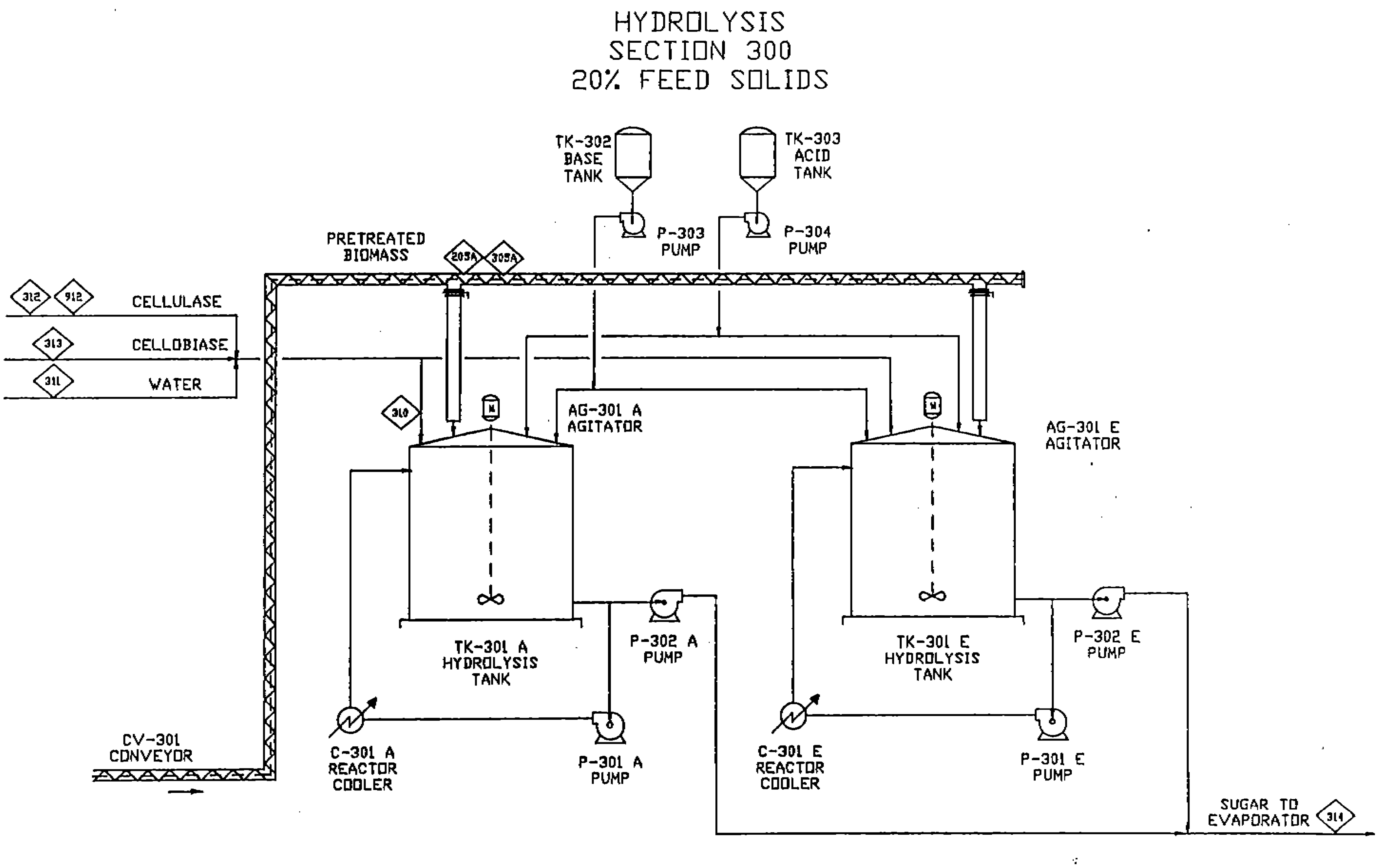


Figure 5-6: HYDROLYSIS SECTION-300 FEED BATCH
AFEX-20 PROCESS WITH ENZYMIC HYDROLYSIS
AT 20% SOLIDS

influence the lowest enzyme loading. For this design we use the demonstrated enzyme loading of 5 IU/g.

In order to recover the ammonia, most ammonia is recovered directly and the residual ammonia is washed from the biomass with water. In the design the ammonia recovery is set at 99%. Later we will see what the cost impact is if this is less than 99%.

The washed, pretreated biomass is then conveyed as a 33% solid paste to the hydrolysis section. The remaining residual ammonia is used in the fermentation downstream.

As a result of washing the biomass with water, an aqueous ammonia solution is formed. The ammonia is easily recovered in a distillation column operated at a pressure high enough so that the overhead vapor ammonia is condensed to liquid at the storage temperature. This recovery unit uses standard industrial practice (16). The ammonia recovery in the mass balance is 99%. This leaves about 422 lb/h of ammonia in the water washed biomass leaving Section 200 in stream 205.

Section 300 - Hydrolysis

The hydrolysis of the cellulose and hemicellulose in the pretreated biomass is achieved in the hydrolysis reactor. The enzyme loading for the cellulase is at 5 IU (international units or filter paper units of activity FPU) per gram of solid biomass. This will give 98% conversion of the cellulose to glucose in 24 hours at 45 to 50°C, pH 4.8. In order to achieve this yield the biomass slurry is prepared to be 10% solids. Note: in Run 5-4, Table 1 cellulase protein required is 483 lb/h stream (312) per 63,697 lb/h of solids feed to the hydrolysis reactor (stream 205A). This is the AFEX-10 version of the hydrolysis process.

In the second version, AFEX-20, 20% solids are used in the hydrolysis. However, it is known from other pretreatments (10) that as the solids concentration goes up at a given enzyme/solids ratio, the rate of hydrolysis is reduced. Thus time is extended to 36 hours and the conversion of the cellulose to glucose is reduced to 80%. In both versions the hemicellulose is completely converted to its xylose equivalent. Finally, the cellulase, which is a mixture of endo- and exo-glucanases and cellobiase has to be enriched with extra cellobiase. In this design the cellulase is produced from a slip stream of the pretreated biomass in Section 900. The cellobiase, which is about 1/5 the mass of cellulase (stream 313), is purchased from commercial suppliers.

In the AFEX-10 hydrolysis process, the pretreated biomass is prepared in a mixing tank where enzymes and water are added to give 10% slurry. Since this is a pumpable fluid, the slurry is pumped through a series of continuously stirred reactors for a total hold up of 24 hours. The pH is maintained at 4.8 by addition of acid or base. The hydrolyzate is stored in the sugar holding tank for further processing in the fermentation Section 400. The glucose concentration is 4.9% which is lower than the target of 11.63% glucose achieved in the TVA base case. The xylose concentration is 3.28%. Thus a sugar concentration step is added which uses a multi-effect evaporator, a standard packaged design, to give 11.63% glucose concentration. By adding the evaporator, we essentially keep the fermentation and distillation Sections 400 and 500 similar in size to the TVA base case. Thus, we can make a fairer comparison, between alternatives. It is possible that an overall process optimization would not set the glucose concentration to 11.63%, but this is beyond the scope of a first order design analysis.

In the AFEX-20 hydrolysis process, the hydrolysis operation is a series of fed batch units, phased in such a way as to give continuous supply of sugars when using the sugar holding tank as a buffer. See Figure 5-6 for the flowsheet modifications for Section 300. Since it is not possible to stir a 20% slurry of biomass, the hydrolysis tanks are started up with 10% slurry and all the enzymes. In effect the enzyme loading is twice as high initially than in the continuous process. This promotes rapid hydrolysis and permits the gradual addition of the remaining biomass over the reaction cycle so the final solids added are equivalent to 20% solids based on the batch mass. The batches are held for 36 hours at 45 to 50°C, pH 4.8. The final sugar concentrations are 8.00% glucose and 6.59% xylose. The xylose/glucose ratio is higher in this case because the hemicellulose is completely hydrolyzed in both process versions while the glucose yield is 80% in the 20% solids case and 98% in the 10% solids case. The glucose yield penalty in the 20% slurry needs to be explored in the pilot plant to establish the actual trade-offs.

In order to be able to interface with the fermentation and distillation process of the base case, the glucose is concentrated again by a multiple effect evaporator to 11.63%. Naturally the size of the evaporator is smaller than with 10% solids.

Because the percent of cellulose converted to glucose in AFEX-20 is less than AFEX-10, more biomass is treated in the plant to achieve the 25 million gallons of ethanol per year.

Section 400 - Fermentation

There are no significant design changes in the fermentation section over the base case except that a bit more ethanol is produced in the C_5 fermentor relative to the C_6 fermentor because of the shift in xylose/glucose ratio.

Section 500 - Distillation and Dehydration

This section is the same as in the base case.

Section 600 - Stillage Evaporation

For AFEX-10, the stillage evaporator is about the same size as in the base case, but the capacity is reduced in AFEX-20. This is because with 80% conversion, there are more solids in the stillages than in AFEX-10. Thus, less water is evaporated to get a 50% syrup which is used in the boiler as fuel.

Section 900 - Enzyme Production

For both AFEX-10 and AFEX-20, the cellulase is produced as described in Chapter 3, Section E6 and follows the flow diagram in Figure 3-7. The enzyme plant is a bit larger in the AFEX-20 version because more biomass is used in that case due to the 80% conversion limit on the cellulose.

C. Process Analysis

Sixteen simulation runs are given in Appendix 5 for the AFEX process which explores a range of process parameters. The basic AFEX-10 case is given in Run 5-4 and the basic AFEX-20 case is given in Run 5-9.

The energy produced from the combustion of evaporated stillage with lignin in AFEX-10 does not meet the thermal and electrical energy needs of the plant (see Run 5-4, Table 2). As a result, natural gas and electricity are purchased to balance the respective energy deficits. In effect, with the 100% conversion of hemicellulose and 98% conversion of cellulose, there is not enough residue left for fuel. Thus some extra fuel is needed. In contrast, for AFEX-20, with only 80% cellulose conversion, there is energy in excess to meet the needs of the plant (see Run 5-9, Table 2). Thus the surplus is used to generate electricity for export.

One of the highest capital cost sections for the AFEX process is the pretreatment - Section 200 ($18 to $20 million). There is an engineering design opportunity here to reduce the capital cost by optimum equipment selection.

The other section with high capital cost is the fermentation ($19,220,000). This is because of the separate C_6 and C_5 fermentation. The distillation cost is also relatively high because of two beer stills, over $12 million. Later an analysis of the Bioenergy fermentation process (Chapter 8) will show the impact of having only one fermentor and beer still.

From the capital cost and material costs, the operating costs are developed for AFEX-10 in Run 5-4, Table 9 and for AFEX-20 in Run 5-9, Table 9. Due to higher yield on cellulose, AFEX-10 gives a lower net cost ($1.128/gal) compared to the AFEX-20 ($1.153/gal). These costs are very good in comparison to the total cost of $1.57/gal for ethanol from corn in the MBI report (9). When DDGS credit is applied to corn, the cost is \$1.154/gal. Moreover, the corn-based ethanol costs are for a plant twice the size as the biomass based plant. If it were on the same capacity, the corn-based ethanol costs would be higher by about 11¢/gal.

One concern of the AFEX process is how good the ammonia recovery has to be. While the design runs in Appendix 5 are based on 99% recovery in Section 200, the effect on the operating cost of a range of recoveries is given in Table 5-1. As the recovery decreases from 99% to 95%, the costs increase 3.3¢/gal. Since 95% or better recovery seems within engineering reality (16), the ammonia recovery is not a primary area of economic leverage, particularly when the ammonia that is not recovered in the pretreatment leaves with the hydrolyzate. This ammonia is used by the fermentation microorganisms and so will displace some nutrient costs.

Table 5-1
Sensitivity of Operating Cost to Ammonia Recovery AFEX Process with Enzymatic Hydrolysis at 10% Solids - Run 5-4 Corn Stover to Ethanol, 25 Million Gallons per Year

Ammonia Recovery (%)	Operating Cost (\$/gal)
99	1.128
98	1.135
97	1.143
96	1.150
95	1.158

Since there is a need to buy extra cellobiase to get the most effective hydrolysis rates of cellulose, the cost of cellobiase is fairly important. First we do not know for sure what level is reasonably required and what levels are just too much. It is a very complex issue that is best answered by producing the cellulase on the substrate for which it is to be used. Given the relative amounts of the various isoenzymes produced in the cellulase, the level of cellobiase can be determined to maximize the rate of hydrolysis. This is not an issue solely related to the AFEX process but has to be addressed in any enzymatic hydrolysis process. The cost impact of the cellobiase is presented in Table 5-2. A change from \$1.00/lb to \$2.50/lb increases the ethanol cost by about 4¢/gal.

Table 5-2
Sensitivity of Operating Cost to Cellobiase Cost AFEX Process with Enzymatic Hydrolysis at 10% Solids - Run 5-4 Corn Stover to Ethanol, 25 Million Gallons per Year

Cellobiase Cost (\$/lb)	Operating Cost (\$/gal)
1.00	1.128
1.50	1.141
2.00	1.153
2.50	1.166

There is a need for detailed laboratory work on the interaction of pretreatment optimization with respect to enzyme loading of cellulase as well as cellobiase in order to get a firm impact on the cost of ethanol from biomass.

Because the cost for cellulase is not directly on the operating cost sheet, it is not clear how much the cellulase costs. Yet, one needs to know the cost of cellulase to know when it pays to buy it. Also, it is interesting to evaluate the operating cost for the AFEX-10 process at less than 5 FPU/g since some laboratory work indicates that less than 5 FPU/g is effective. A series of simulations for AFEX-10 were prepared for 98% cellulose conversion in a 24 hour hydrolysis time for 15, 10, 5, 3, 1 and 0 FPU/g under Run 5-1 through Run 5-6 in Appendix 5. The zero level will give the cost of ethanol production with no investment, material and energy used in cellulase production. Then one can calculate the price one can afford to pay for the cellulase to equal the operating cost with an enzyme plant on site. The 10 and 15 FPU/g loading in Runs 5-5 and 5-6 are included to appreciate the cost of enzyme when more cellulase is needed as in the STAKETECH process in Chapter 6.

A summary of the simulation results is given in Table 5-3. Note that as the FPU/g is decreased from 5 FPU/g to 0, the operating cost decreases from \$1.128/gal to \$1.065/gal - difference of 6¢/gallon. Given the amount of cellulase protein needed per hour for 5 FPU, one can calculate the annual protein needed (483 lb/h x 24 h/d x 330 d/y = 3,825,360 lb/y). In turn, the 6¢/gal times 25 million gallons or \$1,500,000 per year could buy this cellulase protein at 39.2¢/lb to maintain the \$1.128/gal ethanol product cost. The marginal cost of cellulase is shown on the last line of Table 5-3 for the other cellulase loadings. Notice the marginal cost is 81.4¢/lb at 1 FPU/g due to the smaller production level.

Naturally, when more than 5 FPU/g are needed the marginal cost for enzyme continuous to decrease, but the contribution of cellulase cost in the ethanol cost increases to 13.8¢/gal at 15 FPU/g. Clearly, the enzyme loading is a very significant variable in the overall economics of ethanol production. One of the advantages of AFEX pretreatment is the 5 FPU/g or less can give high conversion of cellulose with 24 hours of hydrolysis. That is almost an 8¢/gal ethanol advantage over a 15 FPU/g loading.

Recall that in the second fermentation for xylose with *Pachysolin tannophilus* the parameters are 84 h and xylose fermentation yield of 50%. This is typical of most performances from the TVA experience. However, TVA did report some laboratory results (2) with *P. t.* that gave 70% yield. In order to evaluate the impact of the yield increase, a simulation was done in Run 5-7 in the Appendix. The key results are summarized in Table 5-4 and compared to Run 5-4, the basic AFEX-10 design in the same table. Because of the lower feed rate, the capital is reduced by about \$4.5 million. The increased yield of ethanol from xylose reduces the operating cost by 3.3¢/gal.

Table 5-3
Summary of AFEX-10 Process Design Simulation Runs - Vary Enzyme Loading, FPU/g at 98% Cellulose Conversion

Run Number	5-6	5-5	5-4	5-3	5-2	5-1
FPU/g	15	10	5	3	1	0
Feed Rate lb/h	86,330	83,032	79,734	78,415	77,096	76,436
Cellulase Rate lb/h	1,450	967	483	290	97	0
% Substrate to Enzyme Production	11.46	7.94	4.13	2.52	0.856	0
Capital Cost per Section						
100	650,714	640,841	630,800	626,737	622,646	620,591
200	19,239,850	18,632,350	18,126,100	17,923,600	17,721,100	17,619,850
300	12,711,170	12,689,076	12,666,944	12,658,080	12,649,210	12,644,772
400	19,216,947	19,216,947	19,216,947	19,216,947	19,216,947	19,216,947
500	12,678,658	12,683,103	12,687,547	12,689,324	12,691,101	12,691,990
600	3,043,037	3,013,524	2,983,816	2,971,878	2,959,907	2,953,909
700	1,634,560	1,634,560	1,634,560	1,634,560	1,634,560	1,634,560
boiler/generator 800	12,337,903	12,083,735	11,825,949	11,721,779	11,616,988	11,564,355
non-boiler 800	9,151,700	9,151,700	9,151,700	9,151,700	9,151,700	9,151,700
900	10,888,600	7,411,818	4,362,519	2,829,634	1,159,720	0
Total	101,553,148	97,157,655	93,286,881	91,424,238	89,423,879	88,098,674
Total Operating Cost $/gal	1.206	1.165	1.128	1.111	1.093	1.068
Operating Cost Due to Cellulase $/gal	.138	.097	.060	.042	.025	–
Equivalent Price of Cellulase $/lb	.300	.317	.392	.457	.814	–

In the generic plant design the hydrolyzate has fine suspended particles of lignin which come from the biomass. These solids do not interfere with the fermentation or beer still and so they are left in the hydrolyzate. However, the suspended solids may be a problem in the sugar evaporator and stillage evaporator. If the lignin has to be removed after the hydrolysis is completed in Section 300, additional capital and operating costs are incurred. This issue requires pilot plant studies on evaporators and solid/liquid separation options to define the best approach.

Table 5-4
Summary of AFEX-10 Process Design Simulation Runs - Vary Xylose Fermentation Yield and Solids Removal Prior to Sugar Evaporation 5 FPU/g and 80% Cellulose Conversion

Run Number		5-4	5-7	5-8
Feed Rate lb/h		79,734	72,427	80,311
Cellulose Rate lb/h		483	439	487
Glucose Fermentation Yield %		90	90	90
Xylose Fermentation Yield %		50	70	50
Lignin Solids Removed Prior to Sugar Evaporation		no	no	yes
Capital Cost per Section				
	100	630,800	607,941	632,567
	200	18,126,100	16,911,100	18,227,350
	300	12,666,944	12,163,051	19,738,124
	400	19,216,947	19,216,947	19,216,947
	500	12,687,547	12,176,514	13,082,247
	600	2,983,816	2,856,153	3,249,583
	700	1,634,560	1,634,560	1,634,560
	boiler/generator 800	11,825,949	10,264,497	11,941,572
	non-boiler 800	9,151,700	9,151,700	9,151,700
	900	4,362,519	3,721,471	4,369,146
Total		93,286,881	88,703,934	101,243,794
Operating Cost $/gal		1.128	1.095	1.187
Net Operating Cost $/gal		1.128	1.095	1.187

For now we will estimate the cost of adding to Section 300 the solid/liquid separation equipment shown in the flow diagram in Figure 5-7. The solids are concentrated in the centrifuge and the overflow is clarified by a rotary drum filter. The solids stream from the centrifuge, the drum filter and the enzyme cell purge are further dewatered in a dewater press. This scheme is essentially adapted from the Chem Systems design (3). The AFEX-10 with lignin removal is evaluated in Run 5-8. The modification of the equipment for Section 300 can be seen in Table 5, Run 5-8 in the appendix. The solid recovery equipment adds a purchased cost of about $2,000,000 and an installed cost of about $7,000,000 to Section 300. To compare Run 5-8 with the basic AFEX-10 design in Run 5-4, we include the key results in the last column of Table 5-4. The operating cost increases from $1.128/gal to $1.187/gal - about 6¢/gal.

In this study, none of the cases have lignin removal. Thus, the comparison of cost is valid to see effects of other design parameters. Clearly, the way the solids are handled will effect the absolute cost of ethanol products. The design in Figure 5-7 is technical over kill and so represents the worst case. However, careful engineering development work on solid/liquid separations in any biomass process should receive high priority.

Some preliminary data on the hydrolysis of AFEX pretreated grass indicates that the hydrolysis can achieve higher than 80% cellulose conversion when operating in a fed batch mode with effectively 15 to 20% solids loading in the slurry. To appreciate the economic advantage of increasing the cellulose conversion for AFEX-20, a series of simulations were run for 80%, 85%, 90% and 95% conversion for hydrolysis times of 24 hours and 36 hours. The full set of simulation runs are given in Appendix 5 under Runs 5-9 through 5-16. A summary of the results is given in Table 5-5. The operating cost is reduced by about 4.6¢/gal when the cellulose conversion is increased from 80% to 90% in a 36 hour hydrolysis and about 5.0¢/gal in a 24 hour hydrolysis.

LIGNIN SEPARATION
SECTION 300

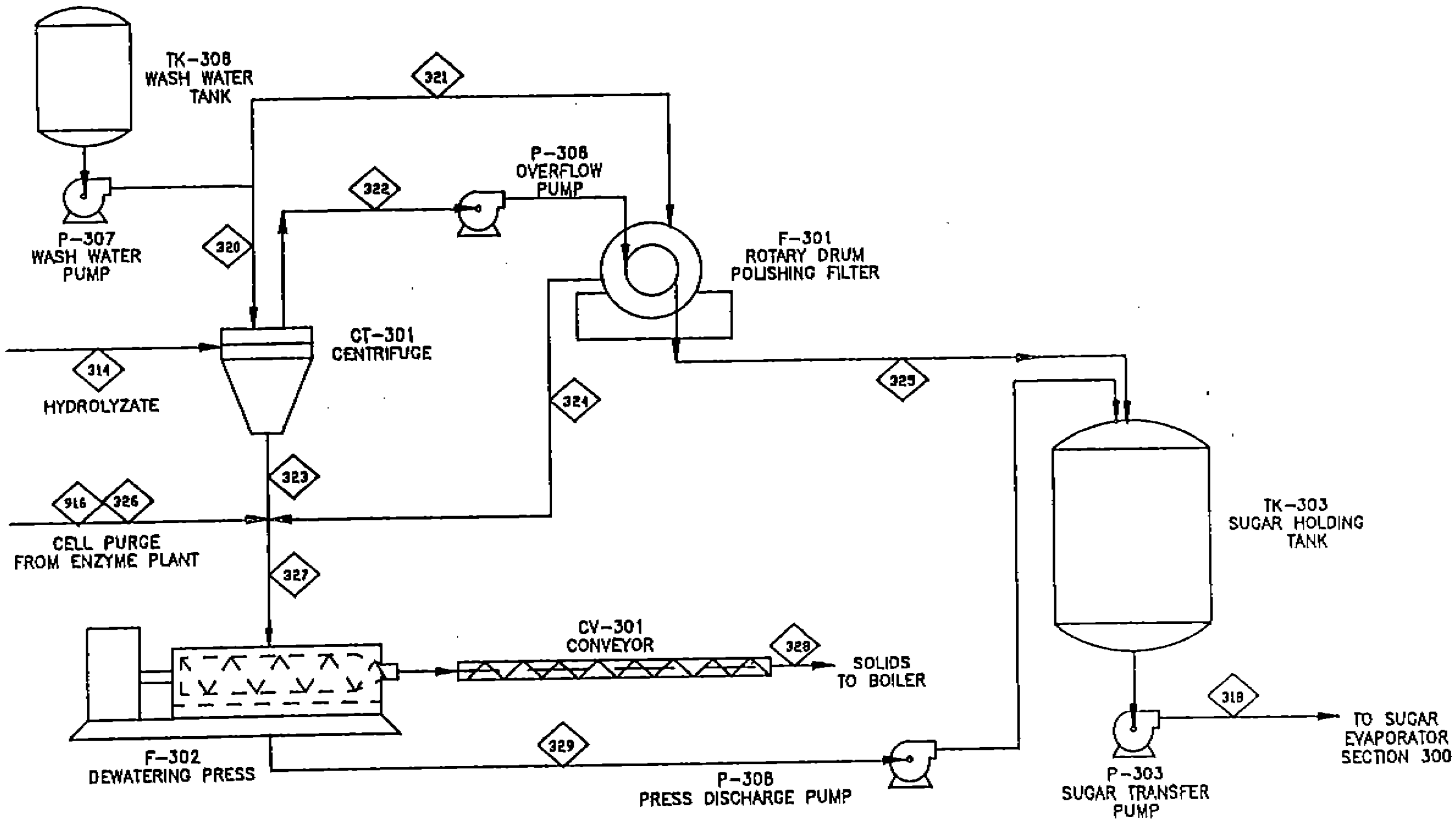


Figure 5-7: LIGNIN SEPARATION FROM HYDROLYZATE IN HYDROLYSIS SECTION-300

Michigan Biotechnology Institute

Table 5-5
Summary of AFEX-20 Process Design Simulation Runs - Vary the Cellulose Conversion and Hydrolysis Time at 5 FPU/g

Run Number	5-9	5-10	5-11	5-12
Cellulose Conversion %	80	85	90	95
Feed Rate lb/h	92,414	88,505	84,912	81,600
Cellulose Rate lb/h	560	536	515	495
Hydrolysis Time, h	36	36	36	36
Total Capital	94,353,735	92,343,197	90,224,462	89,009,401
Operating Cost $/gal	1.154	1.130	1.108	1.091
Net Operating Cost $/gal	1.153	1.130	1.108	1.091

Run Number	5-13	5-14	5-15	5-16
Cellulose Conversion %	80	85	90	95
Feed Rate lb/h	92,412	88,505	84,912	81,600
Cellulose Rate lb/h	560	536	515	495
Hydrolysis Time, h	24	24	24	24
Total Capital	92,763,716	90,753,179	89,429,452	88,214,392
Operating Cost $/gal	1.145	1.116	1.095	1.079
Net Operating Cost $/gal	1.135	1.112	1.095	1.079

CHAPTER 6

MAFEX PRETREATMENT PROCESS

A. Background

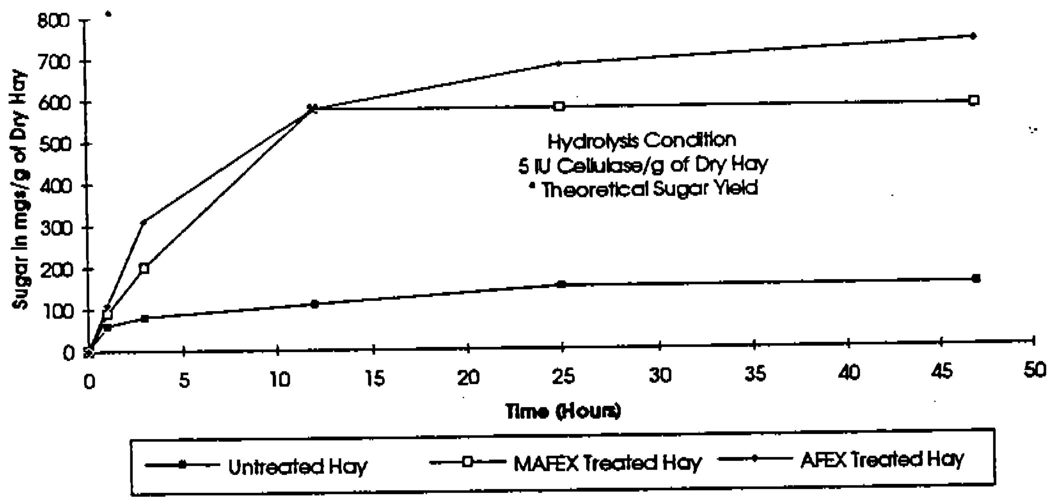
A second pretreatment process for lignocellulosic biomass is the modified AFEX (or MAFEX) process, another proprietary process of the AFEX Corporation, Brenham, Texas. A patent application for this process has been submitted.

The amount of data on MAFEX is more limited than on AFEX. There were some laboratory scale pretreatment tests done at a vendor's facilities and a brief pilot plant demonstration of the process was done at MBI in November of 1992 on Bermuda grass. These results are encouraging enough to warrant its consideration in this report and will be the basis of the plant design. Certain assumptions are made to give some measure of the economic potential of this process.

The biomass is ground to pass through a 40 mesh screen and and fed to the pretreatment process. Initial pretreatment tests were conducted at a vendor laboratory. The samples were shipped back to AFEX Corporation for enzymatic hydrolysis with cellulase (Genencore Cytolase 300) 5 FPU/g substrate in a 5% biomass slurry. The pretreated samples for none (the control), low, medium and high levels of pretreatment gave the following total reducing sugars per grams of dry biomass: 150, 350, 540 and 570 mg, respectively after 24 hours of hydrolysis. The complete hydrolysis of the cellulose and hemicellulose is estimated to be 800 mg/g. Note that at 570 mg/g, the yield is 71% of theory.

A second pretreatment was run at MBI with a 25 pound sample of Coastal Bermuda grass, which was hammer milled to pass through a 40 mesh screen. A slurry with 5% pretreated biomass solids was prepared in a sodium acetate buffer at pH 4.8. Cellulase (Genencor Cytolase 300) was added at the level of 5 FPU/g of dry substrate and the slurry

FIGURE 6-1 : AFEX AND MAFEX PRETREATMENTS OF HAY versus UNTREATED HAY



hydrolyzed at 48 to 50°C. Aliquot samples were analyzed for sugar at various times. The results of this run are given in Figure 6-1 along with the hydrolysis of the control (no treatment) and a prior run of the same material on an AFEX pretreatment. The rate of hydrolysis of the MAFEX treated hay was essentially similar to the AFEX material during the first 12 hours. At that point, the MAFEX curve becomes flat - showing no further increase. Since the slurry was not sterilized in this test, there was microbial contamination (as noted by a drop in pH). Thus, the slurry was heated in the hydrolysis reactor to 80°C to try to pasteurize the batch after 12 hours of hydrolysis. Thus, the most likely reason for no further hydrolysis after 12 hours is due to the thermal inactivation of the cellulase. In any case, the achieved yield at 12 hours is 564 mg sugars/g of dry biomass or 70% of theory.

The above results are very encouraging since the MAFEX process design is quite simple as discussed below. Naturally, more work needs to be done to confirm the relationship between the level of pretreatment, enzyme loading and range of applicability to various lignocellulosic materials. The design will consider the 70% conversion yield of the cellulose as the minimum achievable and assume that at least 80% can be achieved with further process optimization. The effect of yield will be explored in a parametric way. It is also assumed that when the hydrolysis slurry is 10% solids, that 80% yield will be achieved in 24 hours. By analogy on acid pretreatments of wood, the work of Allen, et. al. (10)

showed that the yield of hydrolysis does not drop off significantly with increased solids concentration until after 10% solids. Thus, we should expect the effect of slurry concentration up to 10% not to hinder the yield of the hydrolysis.

B. Process Description

The MAFEX process is an alternative pretreatment process that fits into the generic plant design of Figure 3-1. Thus, the MAFEX pretreatment (Section 200) is designed to interface with the remaining plant sections as in the AFEX design in Chapter 5. Only the relative sizes of the sections may change due to difference in flow rates. The plant flows will be set to produce 25,000,000 gallons of ethanol per year or 21,269 lb/h in the mass balance tables. A total of seven design runs are prepared in Appendix 6 for the MAFEX process to explore the change in cost with various process parameters.

Section 100 - Feed Preparation

The feed preparations section uses the same equipment as for the AFEX process flow diagram given in Figure 5-2. The mills are adjusted in size (and cost) to accommodate the biomass mass flow given in stream 101 in the mass balance tables for each design simulation run given in Appendix 6.

Section 200 - Pretreatment

The basic design case with MAFEX will be for 80% yield in 24 hours of hydrolysis using 5 FPU/g. Because the MAFEX process is proprietary, the flow diagram for Section 200 (Pretreatment) given in Figure 6-2 shows the pretreatment equipment in one box. The pretreated biomass is stored in slurry feed tank TK-201 to provide substrate for hydrolysis (Section 200) and a small stream, about 4% of the total biomass, for the enzyme production (Section 900). In order to keep the solids concentration in the enzyme plant to the desired level, a centrifuge is used to enrich the solids in the slurry in Stream 205B (Figure 6-2).

Section 300 - Hydrolysis

As in the AFEX-10 case, the pretreated biomass solids are hydrolyzed in a continuous series of hydrolysis reactor tanks (see Figure 5-4) at 10% solids with 5 FPU/g in 24 hours to 80% conversion of the cellulose and 100% conversion of the hemicellulose. The various process parameters are given in the mass balance (Table 1) with each MAFEX design run in Appendix 6. The glucose concentration is brought to 11.63 wt% by the multi-effect evaporation E-301 as in the other design cases.

Sections 400, 500, 600, 700, 800

As shown in Table 6 of each design run in Appendix 6, these sections follow the generic design which uses a separate fermentation for glucose and xylose and separate stripping and rectifying of the beers. The sizes and costs of the sections are adjusted to accommodate the required mass flows to achieve the ethanol production goal.

Section 900 - Enzyme Production

While the flowsheet for Section 900 follows the generic design in Figure 3-7, the equipment list in Run 6-2, Table 7 shows 6 production tanks are used instead of 5 in the basic AFEX-10 design. The major reason for this is that the MAFEX pretreatment requires more feed than the AFEX-10 design. Thus, at 5 FPU/g the cellulase needed is 560 lb/h instead of 483 lb/h.

C. Process Analysis

The fixed capital investment for each section for the basic MAFEX design is summarized in Run 6-2, Table 8. With a total capital of $84,127,895, this pretreatment requires about $9.2 million less than the AFEX-10 pretreatment (Run 5-4). This reflects the simpler equipment requirements of MAFEX.

MAFEX
PRETREATMENT
SECTION 200

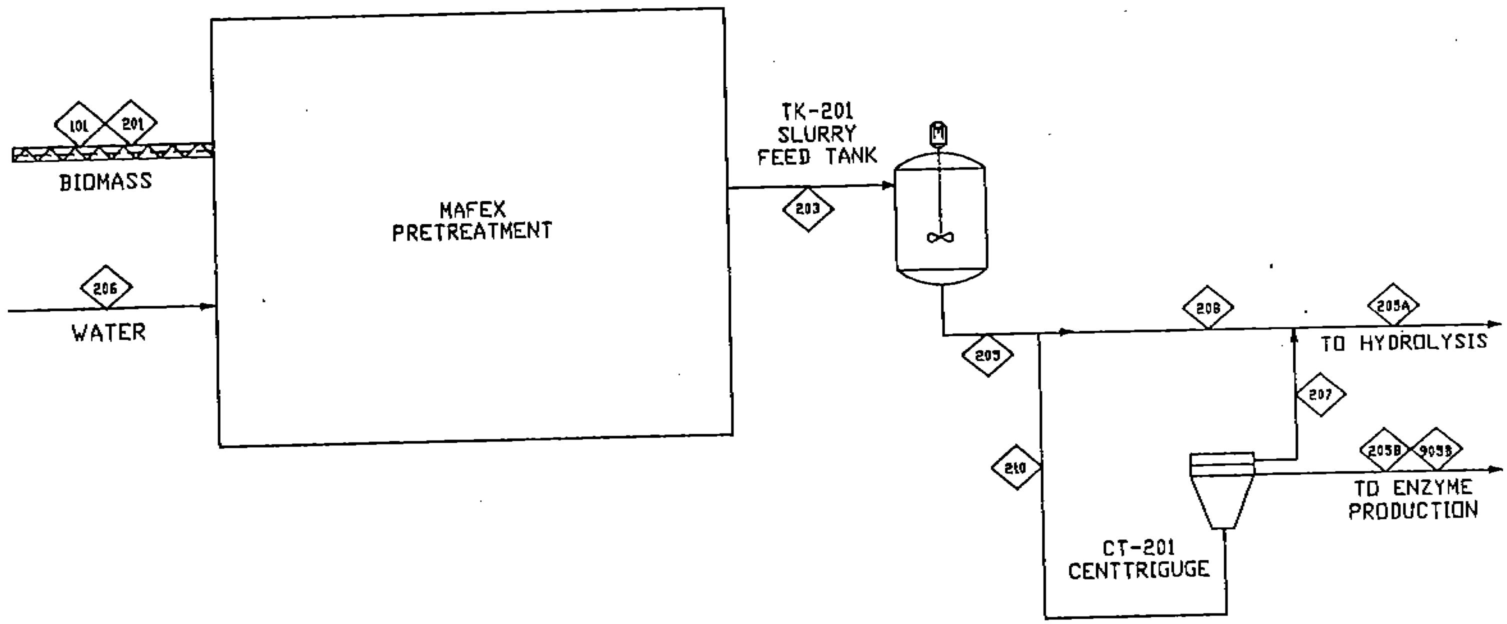


Figure 6-2: PRETREATENT SECTION-200
MAFEX PROCESS

A summary of the energy requirements by the various sections and the thermal and electrical load produced by the power plant is given in Run 6-2, Table 2. While the electrical load and thermal load are lower for MAFEX over AFEX-10 (Run 5-4), the major observation is that for an 80% cellulose conversion to glucose, there is a big increase in the available energy from the evaporated stillage - an increase from 189 to 256 million BTU/h. Thus, the MAFEX process results in an export of electricity and no purchase of natural gas.

For the bottom line, we go to Table 9 in Run 6-2 to get the operating cost of the basic MAFEX process. The cost is 1.087 $/gal without the electricity credit and with the credit it is 1.077¢/gal - about 5¢/gal less than the AFEX-10 basic design in Table 9 in Run 5-4.

In order to see the impact of the cellulose conversion on the capital and operating cost, Runs 6-1, 6-2, 6-3 and 6-4 in Appendix 6, correspond to cellulose conversion of 70%, 80%, 85% and 90% with 5 FPU/g enzyme loading with the usual 24 hour hydrolysis. Recall 70% conversion has been demonstrated in the laboratory while at least 80% is assumed to be possible by proper pretreatment optimization. If further improvements prove possible in the pretreatment, the 85% and 90% give the upper bounds in conversion. The key results for these runs are given in Table 6-1.

The capital cost decreases as the cellulose conversion in the enzymatic hydrolysis increases from 70% to 90% by about $5 million. The raw material flow decreases from 101,000 lb/h to 84,900 lb/h. Notice that in the 70%, 80% and 85% conversion is there an energy credit where the total operating cost is greater than the net operating cost. As we have seen in AFEX-10, when the yield is high there is less energy available for the boiler and so there is no excess power generated by the tubogenerator.

There is a decrease in total operating cost of 9¢/gal in going from 70% to 90% conversion and 4.1¢/gal in going from 80% to 90%.

The impact of the enzyme loading is explored over a range of 5, 3, 1 and 0 FPU/g in Runs 6-2, 6-7, 6-6, and 6-5, respectively, when the cellulose conversion is 80%. A summary of the results is given below in Table 6-2. Note that the operating cost due to enzyme production is 6.6¢/gal at 5 FPU/g, 4.8¢/gal at 3 FPU/g and 2.9¢/gal at 1 FPU/g. These results are very similar to the effect of enzyme loading in the AFEX-10 study in Table 5-3.

Table 6-1
Summary of MAFEX Process Design Simulation Runs - Vary Cellulose Conversion at 5 FPU/g

Run Number	6-1	6-2	6-3	6-4
Cellulose Conversion, %	70	80	85	90
Feed Rate lb/h	101,371	92,415	88,505	84,931
Cellulase Rate lb/h	614	560	536	515
Capital Cost per Section				
100	694,004	668,569	657,158	646,495
200	5,560,515	5,551,291	5,547,153	5,543,287
300	14,687,709	13,877,773	13,509,553	13,162,280
400	19,216,947	19,216,947	19,216,947	19,216,947
500	11,777,353	12,164,223	12,329,311	12,479,113
600	2,327,242	2,616,090	2,735,671	2,842,553
700	1,634,560	1,634,560	1,634,560	1,634,560
boiler/generator 800	15,701,315	14,168,383	13,463,583	12,793,600
non-boiler 800	9,151,700	9,151,700	9,151,700	9,151,700
900	5,173,550	5,078,358	4,461,367	4,421,380
Total	85,924,895	84,127,895	82,707,004	81,891,915
Total Operating Cost ($/gal ethanol)	1.136	1.087	1.062	1.046
Net Operating Cost ($/gal ethanol)	1.109	1.077	1.060	1.046

Naturally, a combination of increased conversion and lower enzyme loading could give as a first approximation, additive cost savings. For example, going from 80% conversion with 5 FPU/g to 85% with 3 FPU/g would save about 1.8¢/gal plus 1.5¢/gal or 3.3¢/gal.

Table 6-2
Summary of MAFEX Process Design Simulation Runs - Vary Enzyme Loading, FPU/g at 80% Cellulose Conversion

Run Number	6-2	6-7	6-6	6-5
FPU/g	5	3	1	0
Feed Rate lb/h	92,415	90,886	89,357	88,592
Cellulase Rate lb/h	560	336	112	0
% Substrate to Enzyme Production	4.13	2.52	.858	
Capital Cost per Section				
100	668,569	664,130	659,661	657,414
200	5,551,291	5,508,627	5,451,205	5,389,671
300	13,877,773	13,877,789	13,877,805	13,877,806
400	19,216,947	19,216,947	19,216,947	19,216,947
500	12,164,223	12,164,236	12,164,249	12,164,249
600	2,616,090	2,598,917	2,581,668	2,573,013
700	1,634,560	1,634,560	1,634,560	1,634,560
boiler/generator 800	14,168,383	14,078,593	13,988,420	13,943,178
non-boiler 800	9,151,700	9,151,700	9,151,700	9,151,700
900	5,078,358	3,507,046	1,781,985	0
Total	84,127,895	82,402,546	80,508,200	78,608,538
Total Operating Cost $/gal	1.087	1.069	1.050	1.021
Net Operating Cost $/gal	1.077	1.059	1.040	1.011
Operating Cost Due to Cellulase $/lb	0.066	0.048	0.029	
Equivalent Price of Cellulase $/lb	0.372	0.451	0.817	

CHAPTER 7

STAKETECH STEAM EXPLOSION PRETREATMENT

A. Background

One of the few pretreatment technologies that is commercially available for biomass is the STAKETECH steam explosion process by Stake Technology Limited, Norval, Ontario. The initial market was directed to prepare cattle feed from Aspen wood chips by steam explosion. Steam explosion involves heating wood chips or any other chopped lignocellulosic material with high pressure live steam, holding for several minutes and releasing the pressure to the atmosphere. The effect is that the high temperature promotes the deacetylation and auto hydrolysis of the hemicellulose, melts the lignin and disrupts the chips into a mush. The resulting biomass has the major part of the hemicellulose solubilized and residual solids are more accessible to hydrolytic enzymes to produce sugars. Thus, the steam exploded biomass could be a ruminant feed or a fermentation substrate.

Stake Technology Limited has developed a patented, continuous steam explosion process that takes chopped biomass and feeds it into a high pressure reactor tube with a specially designed screw feeder which forms a dynamic seal against the high pressure steam in the reactor. The solids are moved through the high pressure reactor tube with an auger and discharged through a specially designed exit value and exploded into a flash tank to recover the disintegrated, pretreated biomass and steam, enriched with volatile organics that are formed such as furfural.

The increased accessibility of steam pretreated wood has been widely demonstrated and is explained by redistributing the lignin into droplets within the biomass and increasing the pore size distribution available to cellulase. In this regard, steam explosion (auto hydrolysis) is very similar to dilute acid pretreatment in a flow reactor.

The integration of the STAKETECH pretreatment into a plant to convert corn stover to ethanol by enzymatic hydrolysis and fermentation is considered proprietary by Stake Technology Limited. As a result, they would prefer to provide the cost of a complete design for the ethanol plant without going into the details of the process integration. In order to provide input into this study, Dr. Ernest Yu, Vice President of Technology, Stake Technology Limited, was willing to give information on just the cost of the STAKETECH equipment for the pretreatment and identify the major performance parameters for biomass pretreatment so this pretreatment can be put into our generic design. This may not put STAKETECH in the same light as their own design, but as an overall reality check, we can compare the cost for the plant provided by Stake Technology Limited with the one developed below.

B. Process Description

Within the generic plant design, the STAKETECH pretreatment process is used in Section 200. The other sections of the plant are adjusted only in capacity so that the plant produces 25 million gallons of ethanol from corn stover.

Section 100

The feed preparation is the same as in the TVA concentrated acid case in Figure 4-1. The tub grinder is more than adequate to provide 4 mesh particles into the continuous feeder of the STAKETECH equipment.

Section 200

The STAKETECH continuous steam explosion pretreatment equipment comes packaged on a skid. One unit can handle 8 dry tons of biomass per hour (4). The unit is complete with feed mechanism, reactor tube, discharge mechanism, drive motors, controls and instrumentation to operate the unit. The flow diagram for Section 200 - Pretreatment is given in Figure 7-1. Five parallel units are required to meet the mass flows. Since the STAKETECH digestors are packaged units, the installation factor is taken as 1.5. In addition,

STAKETECH PRETREATMENT
SECTION 200

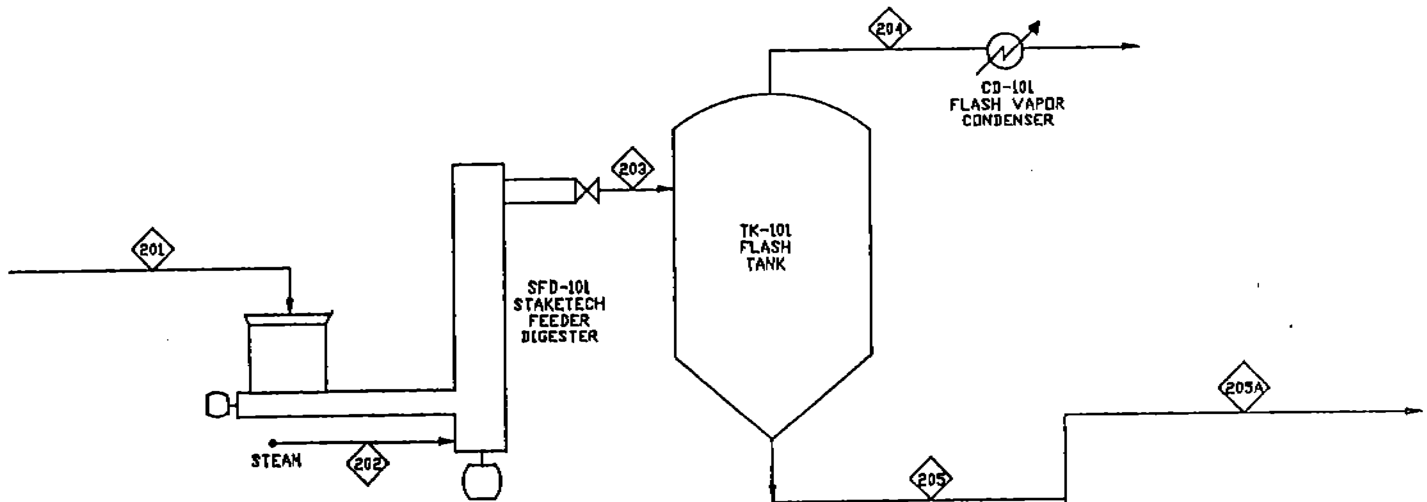


Figure 7-1: PRETREATMENT SECTION-200
STAKETECH PROCESS

5 flash tanks and a common steam condenser are added to capture the exploded biomass and disengage it from the steam. The installation factor for these items is the normal 3.5. Thus, the total installed cost for Section 200 is $18,670,000 (see Run 7-1, Table 4).

The exact operating pressure and time in the reactor are proprietary to Stake Technology, and are chosen to minimize the production of furfural - a thermal decomposition product of xylose which is formed by the auto hydrolysis of the hemicellulose. A typical operating pressure is 350 psi. The steam required is given in stream 202 in Run 7-1, Table 1 and is calculated to provide the sensible heat to raise the temperature of the corn stover with 16% moisture to the steam saturation temperature of the reactor with a 90% steam efficiency. When the pressure is released, steam flashes to atmospheric pressure in the flash tanks. These are designed to disengage the flash steam from the solid, pretreated biomass. The amount of flash steam is calculated in stream 204 by assuming an adiabatic flash process.

During the pretreatment time in the reactor, 80% of the pentosans (hemicellulose) remain as pentosans and 20% is assumed to be converted to xylose. There is in reality, a certain fraction ranging from 5% to over 80% of this xylose than can be converted to furfural by using higher pressure (saturation temperature) and longer time in the reactor. In this analysis, we assume the ideal case with no furfural production and as a result only xylose is formed from the pentosan hydrolysis. See stream 203 in the mass balance tables in Appendix 7. The sizing of the various sections and cost estimates are not materially affected by a few percent xylose converted to furfural. Also, over half of the furfural is removed from the biomass with the steam during the flash. If several pressure stages are used, the furfural remaining in the biomass is reduced roughly by half on each flash. Usually furfural causes inhibition of the yeast in the fermentation if above 1 g/l in the broth. If furfural removal has to be more complete than a single flash step, more capital than estimated for this plant design will be needed.

Section 300

The hydrolysis section follows the same design as Figure 5-4 for the AFEX-10 pretreatment. Based on published data on the hydrolysis of steam exploded wood (11), the enzyme loading used in the initial STAKETECH design is 15 FPU/g in Run 7-1. A second process simulation, Run 7-2, is prepared for an enzyme loading at 10 FPU/g. This should bracket the range expected to be needed in steam exploded corn stover. In the STAKETECH process, the cellulosic conversion to glucose is 90% of theory and the hemicellulose conversion to xylose is 100%.

Sections 400, 500, 600, 700, 800

As in the prior cases, Sections 400 through 800 follow the generic plant design and the size of the sections are based on the mass balance. Thus, the ethanol yield is 90% of theory from glucose in the first fermentation step using *S. cerevisiae* with 6 fermentors and 50% of theory from xylose in the second using *P. tannophilus*, with 9 fermentors (see Run 7-1, Table 6).

Section 900 - Enzyme Production

While the enzyme production section flowsheet is the same as the generic design in Figure 3-7, the size is about 3 times larger than in the prior pretreatments, because the enzyme loading in the hydrolysis section is 15 FPU/g instead of 5. Thus, the number of production tanks given in Run 7-1, Table 7 is 14. When the enzyme loading is reduced to 10 FPU/g the number of production tanks is reduced to 9 (see Run 7-2, Table 7).

C. Process Analysis

Three process simulation runs were prepared in Appendix 7, Run 7-1, 7-2 and 7-3, corresponding to an enzyme loading of 15, 10, and 0 FPU/g. The last run is done in order to get an evaluation of the enzyme cost as has been done in Chapter 5 and 6. The key results of these runs are summarized in Table 7-1.

Table 7-1
Summary of STAKETECH Process Design Simulation Runs - Vary Enzyme Loading, FPU/g at 90% Cellulose Conversion

Run Number	7-1	7-2	7-3
FPU/g	15	10	0
Feed Rate lb/h	91,221	87,948	81,400
Cellulose Rate lb/h	1,454	970	0
% Substrate to Enzyme Product	10.7	7.45	
Capital Cost per Section			
100	467,529	461,615	449,518
200	18,672,875	18,672,875	18,672,875
300	12,775,064	12,753,184	12,709,325
400	19,216,947	19,216,947	19,216,947
500	12,465,672	12,470,155	12,479,137
600	2,898,276	2,867,492	2,805,254
700	1,634,560	1,634,560	1,634,560
boiler/generator 800	13,301,946	13,061,507	12,571,539
non-boiler 800	9,151,700	9,151,700	9,151,700
900	10,894,192	7,434,447	0
Total	101,478,762	97,724,482	89,690,854
Total Operating Cost $/gal	1.187	1.148	1.053
Net Operating Cost $/gal	1.154	1.115	1.019
Operating Cost due to Cellulose $/gal	.135	.096	
Equivalent Price of Cellulase $/lb	.293	.312	

The total capital with the STAKETECH pretreatment is $101,480,000 for the 15 FPU/g case and $97,725,000 for the 10 FPU/g case. The difference in cost (about $3.5 million) is due to the smaller enzyme plant in the latter case. Because the biomass feed in stream 101 to the plant in the 15 FPU/g case is a bit higher than the 10 FPU/g case (91,200 lb/h versus 87,900 lb/h), the energy flows in the plant are correspondingly higher when compared in Table 2 for Runs 7-1 and 7-2, respectively.

When all of these changes are considered in the operating cost estimates, the cost of production per gallon is $1.187 in the 15 FPU/g case and $1.148 in the 10 FPU/g case - a difference of about 4¢/gal. When the credit for export of excess electricity is considered, the costs are $1.154 and $1.115 per gallon, respectively.

In the STAKETECH design using 15 FPU/g, our capital cost estimate is $101,480,000. In contrast, the capital cost estimate supplied by Stake Technology Limited for an ethanol plant using corn stover with a capacity of 25,000,000 gallons of ethanol is $95,600,000. On closer inspection, the Stake Technology estimate does not include an on-site power plant. As a result, steam and electricity are purchased, which raises the operating cost. The operating costs supplied by Stake Technology Limited are given in Table 7-2 for utilities, labor, maintenance and materials. When we add the costs for corn stover and annualized capital cost at 11.1% and 1% for insurance, the total cost of production is $31,754,000 per year or $1.270/gal. This is about 9¢/gal more than given in Table 7-1 for 15 FPU/g. The major difference is due to the purchase of energy. To illustrate this point, consider adding the boiler and turbogenerator to the Stake Technology design for about $13,000,000; this would add about 7.6¢/gal to the operating cost, but it would save 18.8¢/gal now spent on steam and electricity for a savings of about 11¢/gal.

Given all of these considerations, the overall estimate of the STAKETECH process in our generic plant design is reasonable and does allow comparisons with the other alternatives.

Table 7-2
STAKE Technology's Cost Estimate for Production of 25 million gallons Ethanol per Year from Corn Stover

	Canadian Dollars	U.S. Dollars	$/gallon
Total Capital	119,590,000	95,600,000	
Chemicals for Ethanol	1,970,400	1,576,320	.063
Chemicals for Enzyme	384,800	307,840	.012
Labor	2,750,000	2,200,000	.088
Maintenance	2,989,250	2,391,400	.096
Utilities			
Steam @ 8.80C$/1000 kg	3,207,424	2,565,939	.103
Electricity @ 0.035 C$/KWH	2,642,500	2,114,000	.085
Subtotal	13,944,374	11,155,500	
Corn Stover 602,064,222 lb/y @ 30$/T		9,030,963	.361
Capital Charges 11.1% / Insurance 1.0%		11,567,600	.463
Total Annual Operating Cost		31,754,063	1.270

In pretreatments at high temperatures as in STAKETECH or dilute acid hydrolysis such as in BIOHOL, there are always by-products of sugar decomposition such as furfural, hydroxymethyl furfural, levulinic acid, and formic acid as well as acetic acid from deacetylation of hemicellulose and some phenolics from lignin. These by-products vary with the intensity of the pretreatment where higher temperature and/or longer time of pretreatment increase the quantity of the by-products.

Since both yeast and bacteria fermentation of the hydrolyzate are inhibited by one, some or all of these by-products, some type of hydrolyzate conditioning is required such as adjustment of the pH to greater than 10 by lime addition, then removal of precipitated materials or steam stripping. Since STAKETECH controls the time/temperature profile very carefully, the by-products are as low as they get. No provision is made in the cost estimate of the STAKETECH for the hydrolyzate conditioning, but some extra cost of the order of

2¢/gal for conditioning is likely. Thus, the operating cost estimate in this report are the best case scenario.

The one area for improvement in the STAKETECH is to reduce the capital in the pretreatment area - one of the largest components in the capital cost. With the design of a larger reactor unit, one can capture some economy of scale. For example, if a unit handling 20 t/h were to cost $\$2,400,000\ (20/8)^{0.6} = \$4,150,000$ instead of one unit with 8 t/h capacity costing $2,400,000, the cost of Section 200 would be about $13,000,000 instead of $18,670,000. This would lower the operating cost by about 3.25¢/gal.

Naturally, the cost of ethanol production by STAKETECH can benefit from the combined glucose and xylose fermentation as discussed in Chapter 8 - Bioenergy.

CHAPTER 8

BIOENERGY PROCESS FOR FERMENTATION OF HEXOSE AND PENTOSE

A. Background

Any lignocellulosic material will give both hexose and pentose in the hydrolysis of the biomass. Although separate fermentations of these sugars are part of the generic plant design, it appears reasonable to expect cost savings if both hexose and pentose can be fermented together in one fermentation step. Dr. Lonnie Ingram at the University of Florida has focused on this problem in recent years. As a result, U.S. Patent No. 5,000,000 of March 19, 1991 was issued to Ingram and co-workers. This patent teaches the construction of a unique portable operon for ethanol production, which consists of alcohol dehydrogenase II and pyruvate decarboxylase genes from *Zymomonas mobilis*, which is implanted in a host cell such as *E. coli*, *Erwina* or *Klebsiella*. These hosts have the advantage of being able to use both hexose and pentose. This cloned system enhances ethanol production by diverting pyruvate to ethanol during growth under either aerobic or anaerobic conditions. Thus, lactose, glucose, xylose, arabinose, galactose and mannose can be converted to ethanol without the co-production of organic acids.

Bioenergy International of Gainesville, Florida has the proprietary rights to Dr. Ingram's technology. At this point, the production of ethanol from glucose and xylose has been on the laboratory scale and the results have been published. The information used to design the Bioenergy Case comes from these sources and personal discussions with Dr. Lonnie Ingram and Dr. John Gerber of Bioenergy International. While 5 gallon pilot plant fermentations on paper hydrolyzates are under way, the latest results have not been disclosed.

The two papers by Dr. Ingram and co-workers which serve our purpose best are ones that deal with the fermentation of corn cob and corn fiber acid hydrolyzate (17) and with pine wood hydrolyzates (18) using the modified *E. coli* KO11. These substrates are more realistic than pure hydrolyzed cellulose substrates such as Sigma-cell or pure glucose/xylose mixtures used in the other Ingram papers.

Since dilute acid hydrolysis (as well as steam explosion) preparation of the lignocellulosic biomass introduces sugar decomposition products such as furfural, levulinic and formic acid as well as acetic from deacetylation of hemicellulose, the hydrolyzates have to be conditioned by lime addition to increase the pH above 10, held for a period of a half hour to an hour, and then neutralized to a pH of 6.8 to get rid of these inhibitory by-products of the pretreatment. By using the AFEX or MAFEX pretreatment, no thermal decomposition products of sugar are formed and so it is assumed that no pH adjustment, and conditioning of these hydrolyzates are needed.

In the pine hydrolyzate, the sugar solution contained about 5 g/l galactose, 30 g/l mannose and 20 g/l glucose for a total of about 55 g/l hexose with 15 g/l xylose and 3 g/l arabinose for a total of 18 g/l pentose (18). It was supplemented with 20 g/l corn steep liquor to provide the nitrogen, phosphorous and trace minerals for the fermentation.

Similarly, the corn fiber hydrolyzate contained about 55 g/l glucose with 13 g/l xylose and 8 g/l arabinose for a total of 21 g/l pentose. It was supplemented with 5 g/l yeast extract and 10 g/l casamino acid. Generally, the use of CSL is desired as a cheaper media than yeast extract, casamino acid or trypton used in prior laboratory work with the modified *E. coli* KO11. Dr. Ingram has found that CSL at 20 g/l is equivalent to the more expensive nutrients.

The fermentation is started by growing the organism overnight (12 hours) at 30°C, pH 6.0 on a pure sugar growth substrate in a seed fermentor to develop enough cells to give about 0.36 g dry cell/l in the starting fermentor broth. The fermentation is carried out at 30°C, pH 6.0 for 48 hours or more under anaerobic conditions. From the material balance point of view, all the substrate diverted to inocula preparation is added to the fermentor, so the final ethanol yield is based on the total carbohydrate sent to the main fermentor and seed fermentor.

There are a range of results achieved from the corn fiber and pine hydrolyzate fermentations. In order to assess the economic impact of a range of outcomes, several cases will be evaluated below to bracket the performance range to be expected. Several assumptions are made to complete a plant design using Bioenergy's *E. coli* KO11 that are reasonable, based on discussions with Dr. Ingram.

1. The hydrolyzate prepared from corn stover in the AFEX or MAFEX process will perform similarly to the corn fiber and pine hydrolyzates.
2. CSL will be the only other nutrient required and 7 g CSL per liter of broth will be adequate.
3. The inocula can be grown on the hydrolyzate so pure sugars are not needed.
4. The total sugar concentration can be about 80 g/l and will give about 40 g/l ethanol. The microorganism really slows down above 40 g/l ethanol.
5. When the AFEX or MAFEX pretreated corn stover is used in the hydrolysis section (Section 300) at 10% solids, the total concentration of glucose and xylose is in the 70 to 80 g/l range. Thus, the hydrolyzate is used directly in the Bioenergy fermentation.
6. The hydrolyzate does not need to be conditioned by lime addition.

When 10% solids are used in the hydrolysis section (Section 300) of the AFEX-10 plant design, the glucose concentration is about 50 g/l and the xylose is about 34 g/l in stream 314. This sugar profile is similar to the 55 g/l hexose and about 50% more than the 18 to 21 g/l pentose in the hydrolyzates studied by Ingram. The yield/time profile is not known without actual experiments on this corn stover hydrolyzate prepared by the AFEX-10 process. However, three cases will be considered to bracket the yield time behavior that may be expected based on prior work (17, 18).

Case A. Batch fermentation time 48 h at 30°C, pH 6.0 with 95% of theory yield of ethanol from glucose (that is .95 x .51 = .485 g ethanol/g glucose supplied) and 50% of theory yield of ethanol from xylose (.50 x .51 = .255 g ethanol/g xylose). Total cycle time with loading, discharge, cleaning and sterilization is 58 hours.

Case B. Same time, temperature and pH as above with 95% of theory yield of ethanol from glucose and 95% of theory yield of ethanol from xylose.

Case C. Fermentation time 24 h at 30°C, pH 6.0 with 95% of theory yield of ethanol from glucose and 67% of theory yield of ethanol from xylose. Total fermentor cycle is 34 hours.

Case A reflects the observation that in some cases when glucose is the major sugar and the xylose level is relatively high, the xylose is not utilized until the glucose is essentially gone after 24 hours and that half of the xylose is still left after 48 hours. However, when the hydrolyzate is supplemented with 1 g/l Na_2SO_3, the xylose utilization rate is improved so that it is essentially gone by 48 hours (18) while the glucose is still used within the first 24 hours - this will be Case B. Finally, in Case C with 1 g/l Na_2SO_3 and the fermentation time reduced to 24 hours, the glucose is essentially used as before in Case B while 2/3 of the xylose is used (.67 x .51 = .34 g ethanol/g xylose) for ethanol production. The function of Na_2SO_3 is not known, but it is most likely needed to counteract the impurities from the acid pretreatment. It is not expected to have this problem with AFEX or MAFEX pretreatment.

Since the MAFEX pretreatment has a lower cost than the AFEX pretreatment, we also considered a series of Bioenergy fermentations (Cases A, B, and C) analogous to those above where the hydrolyzate is prepared via the MAFEX process. Now the glucose concentration in the hydrolysis section (Section 300) in stream 314 is about 41 g/l glucose and 34 g/l xylose. The reason for the relatively lower glucose concentration in the MAFEX case is because the conversion of the cellulose to glucose is set at 80% instead of 98% as in the AFEX-10. Both pretreatment processes give 100% conversion of the hemicellulose to xylose.

B. Process Description

Section 100 - Feed Preparation

This section will use the same flowsheet as in Figure 5-2 and capacity will be adjusted to meet the mass flow of the substrate for the various design cases.

Section 200 - Pretreatment

When the Bioenergy fermentation is coupled with the AFEX-10 pretreatment, the flow diagram is the same as in Section 200 in Figure 5-3. When it is coupled with the MAFEX pretreatment, the flow diagram is the same as in Section 200, Figure 6-2.

Section 300 - Hydrolysis

The pretreated corn stover is converted to sugar in a 10% solid slurry by enzymatic hydrolysis using 5 FPU/g following the flowsheet in Figure 5-4. However, in contrast to the prior hydrolysis with AFEX-10 or MAFEX pretreatment, the sugars in the hydrolyzate (Stream 314) are not concentrated with a sugar evaporator in preparation for fermentation.

The reason for using the glucose and xylose at the concentrations developed in the hydrolysis reactor is due to the low alcohol tolerance of the *E. coli* KO11 of about 40 g/l ethanol.

The capacity of the equipment is adjusted to meet the required mass flows.

Sections 400, 500, 600, 700, 800

The batch fermentor design that is used for the modified *E. coli* follows the design used in the prior cases using yeast. The process flow diagram given in Figure 8-1 uses a series of parallel fermentor tanks with a cooling loop and CO_2 scrubber. The beer is stored in the beer well after the batch is finished to free the tank for cleaning and reloading. Since both the hexose and pentose are fermented in one vessel, there is only one beer still with the conventional Katzen (1) design.

Since the time of the fermentation cycle and the concentration of ethanol changes in the various runs, the number of fermentors is changed to maintain the desired 25 million gallons per year of ethanol product in each case. In turn, the size of the stripper/rectifier increases as the concentration of the ethanol in the beer is decreased because more water is handled for the same final amount of ethanol. Similarly, the stillage evaporator duty is dependent on the concentration of the stillage. Thus, for each run the cost developed in each section of the plant reflects the changes in equipment number or size. Finally, the size of the boiler and turbogenerator is governed by the residual solids left from the hydrolysis and fermentation.

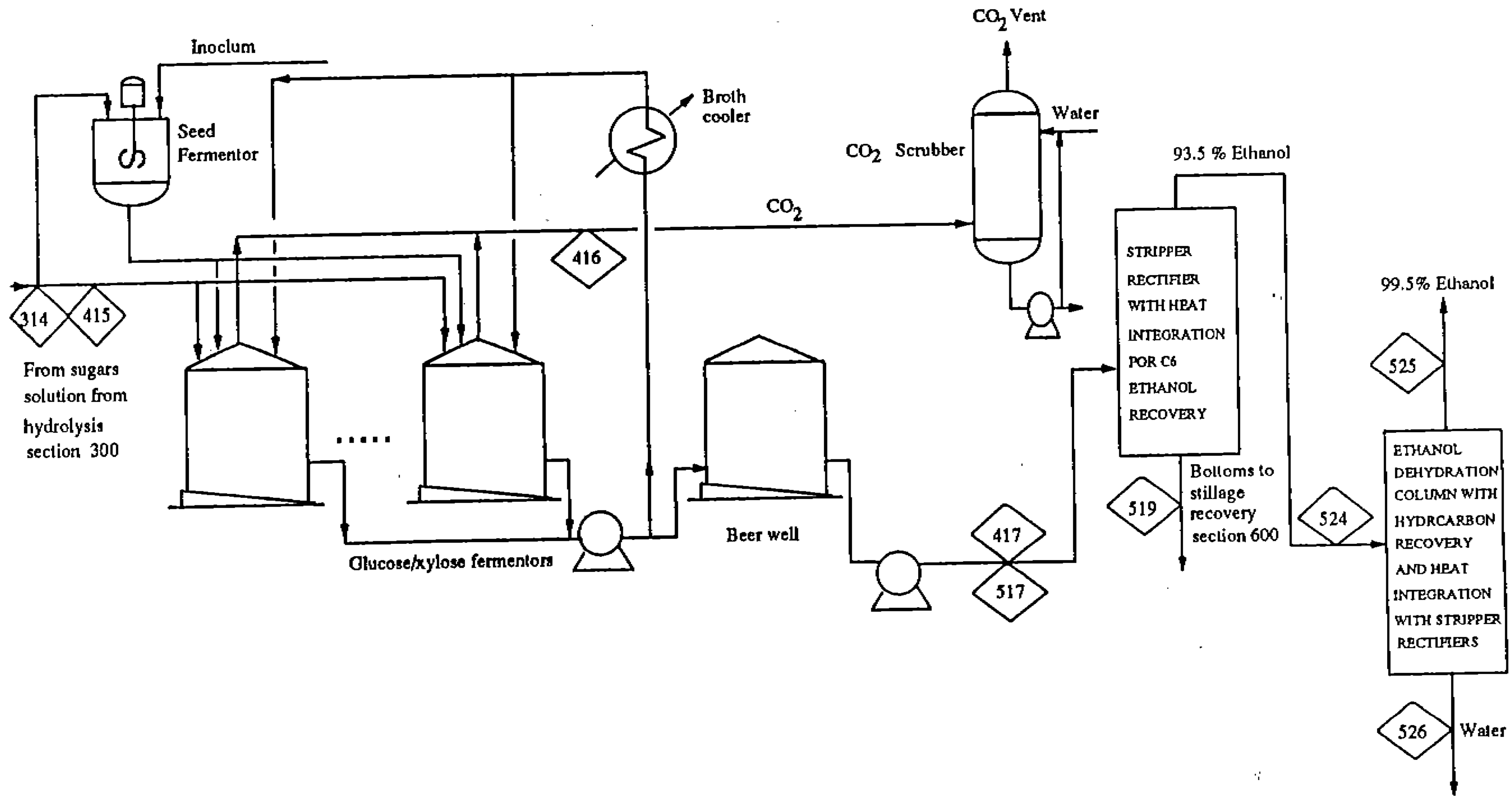
Section 900 - Enzyme Production

The enzyme plant follows the generic design and produces enzymes on a side stream of pretreated corn stover as shown in Figure 3-7. While the enzyme loading is 5 FPU/g in all cases considered, the actual amount of enzyme varies slightly from case to case according to the substrate flow through the plant.

C. Process Analysis

In the initial evaluation of the Bioenergy's fermentation, corn steep liquor (CSL) usage was set at 25 g per liter of broth which was used in the laboratory studies on pine hydrolyzate (18). However, with this level of CSL, the cost it contributes to the final ethanol amounts to 13.9¢/gallon - nearly half the cost attributed to corn stover. In discussions with Dr. Ingram, it was realized that 10 g/l also works in the laboratory. The final level of 7 g/l used in this report was chosen based on providing the nitrogen for the *E. coli* cell mass of 3 g/l that is developed in the fermentation process. As a result, the cost due to CSL is now 4.2¢/gal (see Table 9 in Run 8-1 in the appendix). While this is a reasonable cost impact of

Figure 8-1 Fermentation Section 400/ Distillation-Dehydration Section 500 For Bioenergy Fermentation



CSL, it does point to the need to optimize the non-carbon nutrient cost in all cases of ethanol from biomass. No doubt, some nutrient cost can be saved by recycling a certain fraction of the cells or stillage.

The mass balances, energy, equipment lists, capital and operating costs for plant designs that use the AFEX-10 pretreatment and a single fermentation step with Bioenergy's *E. coli* KO11 as described in Figure 8-1 are given in Tables 1 through 9 of Run 8-1, 8-2, and 8-3 for Cases A, B, and C, respectively in Appendix 8. Note: the process parameters such as hydrolysis yield, fermentation yield, FPU/g, hydrolysis time and fermentation time are given in the top rows of the mass balance tables for each case. The corresponding Cases A, B, and C for the MAFEX pretreatment with the Bioenergy fermentations are given in Runs 8-6, 8-7, and 8-8, respectively. A summary of key results for all of these runs are given in Tables 8-1 and 8-2.

The final investment for the AFEX-10 pretreatment with the Bioenergy single fermentation for glucose and xylose is $96,250,000 for Case A (first column in Table 8-1) and the operating cost is $1.190/gallon. Recall that the AFEX-10 process in Run 5-4 (last column in Table 8-1), where two separate fermentation steps are used, has a cost of $1.128/gallon with a $93,290,000 investment. So why is a single fermentation process more expensive than one with two steps? By comparing the capital cost in corresponding sections for the two processes, it is clear why; the lower cost in the hydrolysis section (Section 300) which is due to eliminating the sugar evaporator is exceeded by higher costs in other sections. For example: there are 18 fermentors in the Bioenergy Case A due to the lower sugar concentration (5.3% versus 11.63%) whereas there as only 15 fermentors in the AFEX-10 process with a two-stage fermentation. The lower sugar concentration also reduces the ethanol concentration to the beer still. Thus, for Section 500, one beer still in Bioenergy Case A costs almost as much as two in the AFEX-10. Moreover, the lower concentration also increases the capital significantly for the stillage evaporator in Section 600.

Table 8-1
Summary AFEX/Bioenergy Process Design Simulation Runs - Enzymatic Hydrolysis at 10% Solids in 24 hours with 5 FPU/g

Run Number	8-1	8-2	8-3	8-4	8-5	5-4
Case	A	B	C	B w/ Evap	A w/Solid Sep	AFEX-10
Feed Rate lb/h	83,212	67,257	76,377	67,034	83,818	79,734
Cellulase Rate lb/h	504	408	463	406	508	483
Sugar Evaporator Duty lb H_2O/h	--	--	--	311,483	--	370,489
Stillage Evaporator Duty lb H_2O/h	564,829	464,337	521,730	151,183	595,425	172,233
Number of Fermentors	18	14	10	6	18	6+9
Capital Cost per Section						
100	641,384	591,263	620,406	590,470	643,208	630,800
200	18,733,600	16,101,100	17,619,850	16,101,100	18,834,850	18,121,100
300	8,052,965	7,528,650	7,833,511	11,777,935	15,284,620	12,666,944
400	22,820,327	18,015,820	13,211,313	8,406,806	22,820,327	19,216,947
500	12,032,979	11,035,518	11,615,860	7,930,982	12,194,570	12,687,547
600	6,084,898	5,410,069	5,801,904	2,759,340	6,280,572	2,983,816
700	1,634,560	1,634,560	1,634,560	1,634,560	1,634,560	1,634,560
boiler/generator 800	12,698,899	9,237,624	11,301,917	9,217,729	12,814,435	11,825,949
non-boiler 800	9,151,700	9,151,700	9,151,700	9,151,700	9,151,700	9,151,700
900	4,402,213	3,658,758	4,323,542	3,655,775	4,409,052	4,362,519
Total	96,253,526	82,365,062	83,114,563	71,226,397	104,067,893	93,286,881
Total Operating Cost $/gal	1.190	1.087	1.104	1.003	1.249	1.128
Net Operating Cost $/gal	1.190	1.087	1.104	1.003	1.249	1.128

In order to see the cost impact of changing the fermentation yield on xylose to 95%, Case B was run for the AFEX/Bioenergy process. With the higher yield in Case B than in Case A, less biomass is required, 67,257 lb/h instead of 83,212 lb/h which results in a reduction of capital in all sections. In the fermentation section (Section 400), there are now

14 fermentors. Thus, the capital is reduced to $82,365,000 and the operating cost drops to $1.087/gallon for Case B. Clearly, a yield increase on the xylose from 50% to 95% has a large impact on the cost of ethanol - about 10¢/gal.

Finally, in Case C, the batch Bioenergy fermentation time is reduced from 58 hours to 34 hours, with the fermentation yield on glucose at 95% and yield on xylose at 67%. The capital for Case C is similar to B, where the decreased cost in Section 400 (now has 10 fermentors) is counter-balanced in the other sections because of the higher biomass feed rate, 76,377 lb/h. The operating cost is $1.104/gal, intermediate between Cases A and B. Note that in Case C there are 10 fermentors instead of 18 as in Case A or 14 as in Case B. This is due primarily by the 24 hour fermentation time instead of 48 hours.

It if should be necessary to remove the lignin solids from the hydrolyzate, a solid/liquid separation step as shown in Figure 5-7 can be added. The modified AFEX-10 process with enzymatic hydrolysis and Bioenergy fermentation, Case A is evaluated in Run 8-5 with lignin removal (see stream 328 in Run 8-5, Table 1). The capital for Run 8-5 is $104,000,000 and the operating cost is 1.249 $/gal (see Table 8-1). Compare this to the original Case A in Run 8-1 where the capital is $96,250,000 and the operating cost is 1.190 $/gal. The difference in operating cost, 5.9¢/gal is about the same as in the analogous comparison in Chapter 5, Table 5-4 for the AFEX-10 with a two stage fermentation.

Since the MAFEX pretreatment gave lower costs than the AFEX-10 pretreatment when using the two stage gluclose/xylose fermentation, we evaluated the MAFEX / Bioenergy process combination for the three cases (A, B and C). A summary of key results for Runs 8-6 through 8-9 are given in Table 8-2.

Table 8-2
Summary MAFEX/Bioenergy Process Design Simulation Runs - Enzymatic Hydrolysis at 10% Solids in 24 hours with 5 FPU/g

Run Number	8-6	8-7	8-8	8-9
Case	A	B	C	B w/ Evap
Feed Rate lb/h	96,377	75,627	87,326	75,282
Cellulase Rate lb/h	584	458	529	456
Sugar Evaporator Duty lb H_2O/h	–	–	--	396,349
Stillage Evaporator Duty lb H_2O/h	649,737	518,488	592,484	119,766
Number of Fermentors	20	16	11	6
Capital Cost per Section				
100	679,983	618,058	653,677	616,975
200	5,555,414	5,532,974	5,545,891	5,532,582
300	8,456,276	7,808,952	8,181,556	12,717,441
400	25,222,581	20,418,074	14,412,440	8,406,806
500	12,751,155	11,528,142	12,232,576	7,608,761
600	6,618,288	5,780,246	6,261,951	2,399,407
700	1,634,560	1,634,560	1,634,560	1,634,560
boiler/generator 800	15,065,880	11,073,389	13,422,642	11,043,035
non-boiler 800	9,151,700	9,151,700	9,151,700	9,151,700
900	5,120,908	4,314,737	4,448,311	4,310,675
Total	90,256,700	77,860,833	75,945,304	63,421,942
Total Operating Cost $/gal	1.179	1.030	1.052	0.941
Net Operating Cost $/gal	1.161	1.030	1.051	0.941

The capital for Cases A, B and C are $90,260,000, $77,860,000, and $75,950,000, respectively. With the MAFEX / Bioenergy process there is the same general trend in capital and operating costs with Cases A, B, and C as with the AFEX / Bioenergy process. As expected, the corresponding costs of each case is lower with the MAFEX pretreatment even though the cellulose conversion is only 80% instead of 98%. The higher substrate cost is compensated by lower capital and energy costs.

The real limitation of Bioenergy's fermentation is in the low concentration of about 50 g/l glucose and 30 g/l xylose due to the ethanol tolerance of 40 g/l. However, if the genetically altered *E. coli* or the new work on *Klebsiella oxytoca* M5A1 (19) or *Erwinia carotovora* SR38 or *Erwinia chrysanthemi* EC16 (20) can improve the alcohol tolerance to allow the hydrolyzate to be concentrated by the sugar evaporator to a glucose concentration of 11.63% as in the yeast fermentation, a very significant reduction in operating cost can be realized.

This becomes apparent when the AFEX / Bioenergy process is modified by adding the sugar evaporator into the flowsheet as shown in Figure 5-4 when Case B is redone in Run 8-4. By adding about $4.2 million for the sugar evaporator in Section 300, the total plant capital can be reduced by about $11 million (see Column 8-4 in Table 8-1). The operating cost is $1.003/gal. This is about 8.4¢/gallon lower than the corresponding case without the sugar evaporator. Note: the big change is the number of fermentors by adding the sugar evaporator. In Case B the number goes from 14 to 6.

When Case B in the MAFEX / Bioenergy process is modified with a sugar evaporator, the capital is reduced to $63,420,000 and the operating cost is $0.941/gal (see the last column of Table 8-2). Note that the number of fermentors was reduced from 16 to 6.

The take-home message is there is a real economic advantage in increasing the ethanol concentration in the fermentor, because it not only reduces the volume of the fermentor, which is the dominant cost of the fermentation section (Section 400), but also the volume (and cost) of the following beer still and stillage evaporator. In turn, the energy is less for distillation and stillage evaporation. While the work on genetically altered bacteria is very encouraging, it has not achieved the alcohol tolerance to be an economic winner over a two stage yeast fermentation.

Another way to decrease the cost of the fermentor is to decrease the cycle time for a batch. If the ethanol concentration is maintained while the cycle time is reduced, there will be no changes in size and cost of the beer still and stillage evaporator.

In fact, the process analysis of the fermentation step developed in this report can be adapted to any microorganism. Once certain parameters such as cycle time, fermentation yields from glucose and xylose, and sugar concentration in the fermentor are defined, the cost is determined for the fermentation.

For example in the Bio-Hol report (12), *Pichia stipitis-R* yeast are used to ferment both the hexose and pentose in an acid hydrolyzate at reasonably high starting concentrations of sugars (92 g/l glucose and 38 g/l xylose). Up to 7 vol% ethanol is developed in the broth with 95 to 99% utilization of the sugars. With proper initial inoculum the fermentation is completed in 41 hours. The nutrients are supplied with 0.2 g/l urea and 0.1 g/l $(NH_4)_2H\,PO_4$, commercial grade fertilizer. Alternatively, 5 g/l corn steep liquor can be used. In order to see what the capital and operating costs are for a plant design using a one stage fermentation of glucose and xylose with *Pichia stipitis-R*, we ran additional simulation runs with the sugar evaporator set to concentrate the hydrolyzate to 9.2 wt% glucose in Section 300. Fermentation yields for both glucose and xylose are set at 95% and the fermentation cycle is set at 51 hours to allow for 41 hours of fermentation and 10 hours for turnaround. The hydrolysis section is left as before with an enzyme loading of 5 FPU/g at 10% solid in a 24 hour continuous reactor. The pretreatment is AFEX-10 in Run 8-10 and MAFEX in Run 8-11. The results are given in Appendix 8 and a summary of key results are presented in Table 8-3. The costs with *Pichia* yeast are similar to the cost projected for the Bioenergy Case B with a sugar evaporator for a higher ethanol tolerant microorganism.

Table 8-3
Summary *Pichia stipitis* Process Design Simulation with AFEX or MAFEX Enzymatic Hydrolysis at 10% Solids in 24 hours with 5 FPU/g

Run Number		8-10	8-11
Pretreatment		AFEX-10	MAFEX
Feed Rate lb/h		67,080	75,329
Cellulase Rate lb/h		407	457
Sugar Evaporator Duty lb H_2O/h		252,061	341,929
Stillage Evaporator Duty lb H_2O/h		210,924	174,512
Number of Fermentors		7	6
Capital Cost per Section			
	100	590,621	617,124
	200	16,101,100	5,532,636
	300	11,271,720	12,301,791
	400	9,607,933	8,406,806
	500	8,627,222	8,282,120
	600	3,369,600	3,007,440
	700	1,634,560	1,634,560
	boiler/generator 800	9,221,528	11,047,207
	non-boiler 800	9,151,700	9,151,700
	900	3,656,345	4,311,234
Total		75,232,328	64,292,623
Total Operating Cost $/gal		1.021	0.947
Net Operating Cost $/gal		1.021	0.947

CHAPTER 9

DISCUSSION AND CONCLUSIONS

The rationale for having a boiler with a turbogenerator on site is that this investment results in having a cost for steam and electricity that is less than the corresponding purchased price. In effect, we can calculate the marginal cost of the energy in the form of steam and electricity in our designs.

The boiler is sized to burn at least all the residual fuel produced from the stillage evaporation. If the thermal load of the plant is larger than the residual fuel can provide, extra fuel is purchased. The necessary investment is in the boiler and stillage evaporator which produces the residual fuel. For example, in the AFEX-10 design in Run 5-4, Table 2, the thermal load on the plant is 202.9 x 10^6 BTU/h which is equivalent to 202,900 lb steam/h. Using the first term of equation 10 in Chapter 3, Section E.5, with an installation factor of 2.3 for the boiler, the installed investment for the boiler is $9,070,000. In addition, the installed cost of the stillage evaporator (Run 5-4 in Table 8) is $2,980,000. So the total investment to produce steam is $12,050,000.

Taking the 11.1% capital return in 9 years plus the 2.5% for insurance and maintenance, the annual cost for steam due to capital is $1,747,000 which produces 189 x 10^6 BTU/h from the residual fuel or 1.49 x 10^6 MBTU/y. Thus, the unit cost per million BTU of steam is $1.17. This is cheaper than buying natural gas or other fuels at $1.75/M BTU (with boiler efficiency of 84%, the effective cost is $2.08/MBTU). Clearly, it pays to make the steam on site. In fact if one did not use the evaporated stillage as fuel, the plant would incur additional cost for waste disposal.

For Run 5-4, the electricity generated is 6881 KW. From the second part of equation 10 for the turbogenerator, in Chapter 3, Section E.5, we see the installed capital for the turbogenerator is $2,750,000. The annual cost for this incremental capital is $398,750. Given that annual electricity produced is 6881 KWH/h x 24 h/d x 332 d/h = 54.5 x 10^6 KWH, we get a unit cost of electricity of 0.73¢/KWH. It is clear that the marginal cost of electricity is significantly less than purchased power at 4¢/KWH.

By having a complete plant with utilities, enzyme production and stillage evaporation, it is possible to see the advantage or disadvantage of a given process alternative for pretreatment, hydrolysis, or fermentation in a global way. This avoids selecting a candidate process that has a well recognized local improvement in one section of the plant only to realize later that the true bottom line of the plant - the operating cost to produce ethanol - was not reduced as much as was expected or even worse it went up.

Because the amount of cellulase in terms of actual enzyme protein is relatively high for cellulose hydrolysis compared to starch, we put an enzyme plant on site regardless of the FPU/g needed - 10 to 15 FPU/g as in the case of the STAKETECH process, or 5 FPU/g or less in the AFEX process. The marginal cost of cellulase is in the range of 30¢ to 81¢ per pound of cellulase protein depending on the size of the enzyme plant. Clearly, there is no commercial source of cellulase for anything in this price range. Thus, it always pays to have an enzyme plant on site. However, there is a great deal of cost leverage in reducing the cellulase use. In going from 15 FPU/g to 1 FPU/g the ethanol unit cost can be reduced by about 11¢/gal.

Both AFEX and MAFEX have shown good conversions at 5 FPU/g in 24 hours or less of hydrolysis. This is just the beginning as an optimized pretreatment with enzyme optimization can be expected to reduce the enzyme loading. There are several issues to consider. First the enzyme profile in endo- and exo-glucanases (as well as the many iso enzymes for each type) that gives the fastest hydrolysis rate needs to be understood for each type or degree of pretreatment. Moreover, the best cellulase "cocktail" is expected when the enzyme is grown on the same pretreated substrate as that used in the hydrolysis section. Thus, all laboratory tests with cellulase from vendors, who grow the cellulase on quite a different substrate are not able to even test this hypothesis. Eventually, one needs to make cellulase on the pretreated substrate of interest. The established cellulase companies can play a key role here because they have a wide variety of cellulase producing strains and microorganisms and proprietary know how on how to produce large quantities of cellulase as cheaply as possible.

A related issue is the recycling and reuse of enzymes that in effect cuts the net enzyme usage.

Since both the AFEX process and the STAKETECH process produce a high solids biomass (30 to 60% solids), it would be possible to have a high solids hydrolysis step. In this report, we mostly used 10% solids in the hydrolysis because that is a reasonable extrapolation of the information from laboratory studies at generally lower concentrations. If the conversion of cellulose to glucose could be at 95% and the solids concentration increased from 10% to 20%, it could save about 5¢/gal ethanol. A study on high solid hydrolysis with periodic retreatment of the biomass or with other effective mixing equipment should be on the development agenda.

Since the STAKETECH process is commercially available and proprietary, we can say little about how to improve it. It has solved an important problem in taking coarse chopped biomass directly into a high pressure reactor. The size of the STAKETECH reactor limits its economic application on large plants. A larger reactor will offer some economy of scale. The disadvantage of the STAKETECH process is in the higher enzyme loading and organic by-products in the hydrolyzate. Naturally, in those applications where it is desired to make furfural from the xylan rather than ethanol, the STAKETECH process would be ideal. The residual cellulosic residue would be used for a glucose to ethanol product.

In all the processes, improved solid/liquid handling and separations are key issues that affect the cost of production. For example, in the AFEX process, the cost of the pretreatment is in the range of $18 to $20 million - a good part of it in solid/liquid separation equipment. Another example is the removal of lignin solids from the hydrolyzate.

Up to \$7 million in capital investment may be added to a plant to handle the solid/liquid separation problem. By supporting some carefully thought out, creative engineering development work, there is a high probability of making significant reductions in the capital cost of the plant. Recall that every \$10 million reduction in capital translates in a reduction of about 4.4¢/gallon in ethanol.

Of all the pretreatment alternatives, the MAFEX process has the greatest potential. It is simple and effective on forages. Although it is expected that the MAFEX process will be effective on a wide range of lignocellulosic biomass, this point has to be demonstrated. Engineering development work is needed to define the slurry concentration, best equipment configuration and energy requirements to give the optimum pretreatment process.

In the event that MAFEX is useful only on forages, the AFEX process, which has been shown to work on a wide range of substrates, can be developed. It is a more complex pretreatment process due to handling ammonia. However, the cost of ethanol production is only about 4¢/gal more.

If the fermentation for the glucose and xylose are combined in one fermentation step, some further cost savings are possible. Because of the 40% limit on ethanol tolerance, the current Bioenergy process operates at too low a sugar concentration to have a strong competitive advantage over separate fermentations at higher sugar concentrations. If the work in progress on genetically modified bacteria can achieve 6% to 7% ethanol tolerance, the cost of ethanol production can get into the range of 0.94 \$/gal to 1.00 \$/gal when coupled with the MAFEX or AFEX pretreatment. The reason for the large economic impact when the ethanol concentration is raised is that it not only reduces the size of the fermentor, but also reduces the size of the distillation column in the beer still and the stillage evaporator.

In direct competition with the Bioenergy approach are yeasts with pentose as well as hexose utilization capability. For example, the alcohol tolerance for *Pichia stipitis* is in the range of 6 to 7%. Published work shows good pentose utilization and reasonable rates.

Clearly, a focused development on pentose utilizing microorganism will add real value to biomass conversion to ethanol and should have high priority.

Aside from alcohol tolerance, the size of the fermentation section, and therefore the capital cost, can be significantly reduced by considering continuous fermentation schemes with high cell density to give short fermentation times. In line with reducing process reactor volumes, one can also consider simultaneous saccharification and fermentation (SSF) which may save \$8 to \$12 million in capital investment. The key to success here is to have a microorganism for fermentation and an enzyme for cellulose hydrolysis that are compatible at an operating temperature over 40°C to get high rates of conversion and utilization.

Finally, it is useful to compare the general trend of the biomass to ethanol alternatives with the corn to ethanol economics.

The capital cost for a dry mill-ethanol plant from corn with a capacity of 50 million gallons per year is about \$118,000,000 or \$2.36 per annual gallon of capacity. When this type of plant is scaled down to a capacity of 25 million gallons per year, the investment is about \$77,800,000 which costs \$3.10 per annual gallon of capacity. These capital investments per annual gallon serve as interesting bench marks by which to judge the biomass to ethanol plant design. The AFEX-10 process investment is \$3.73/gal where as the MAFEX is \$3.36/gal. These are a bit higher than the corresponding corn to ethanol plants with the same capacity. However, when the best case, Bioenergy (Case B) with MAFEX, considered, the specific investment is \$3.11/gal. The lowest investment is \$2.57/gal with MAFEX and a single fermentation using *Pichia stipitis*. Thus, with a reasonable development program, the specific investment in a biomass to ethanol plant can be significantly lower than a corn to ethanol plant of the same capacity.

It is noteworthy, that the operating cost for most of the biomass process alternatives are below \$1.15/gal which is the cost of production for ethanol from corn with DDGS credit. Furthermore, the cost may get as low as 94¢/gal with the best pretreatment coupled with a good simultaneous pentose/glucose fermentation with a high ethanol tolerant microorganism.

With almost a 20¢/gal potential advantage over corn, there is a real incentive to put together a coherent, innovative development program to move biomass conversion into commercial reality.

CHAPTER 10

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