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Initial Design of a Dilute Sulfuric Acid Pretreatment Process for Aspen Wood Chips

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ABSTRACT

A preliminary process design for dilute sulfuric acid pretreatment of aspen wood chips in order to obtain fermentable sugars has been prepared and subjected to an economic evaluation. The process design was prepared according to experimental data on the kinetics of dilute sulfuric acid prehydrolysis and particle size effects obtained in this study and our previous work. The initial economic evaluation shows woodchips are 56% of the cost of production, whereas the reactor is only 4%, and the comminution operation is just under 10%, indicating that the process economics are extremely vulnerable to feedstock costs and are thus yield-sensitive. Although chances for major cost improvements by modification of the reactor design and finding alternatives to dry milling of aspen chips to small (20–80 mesh) particles needed for acid penetration and enzymatic saccharification are not great, design improvements of the process will necessitate development of a cheaper acid resistant pretreatment reactor and a less energy intensive comminution system. Experimental results on effects of particle size on the dilute acid pretreatment design are presented.

Index Entries: Pretreatment; aspen; enzymatic hydrolysis; ethanol; xylan hydrolysis.

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INTRODUCTION

Prehydrolysis of aspen wood meal with dilute sulfuric acid at moderate temperatures (120–160°C) has been shown to be an effective pretreatment for increasing the digestibility of cellulose by a cellulase complex produced by the fungus *Trichoderma reesei* (*1,2*). The acid catalyst apparently creates pores in the lignin–hemicellullose shield that renders cellulose fibers accessible to cellulase enzymes (*3,4*). The pretreatment is effective in the solids concentration range of 10–40 wt% as long as the xylans are nearly completely hydrolyzed (*1,2*).

The economics of operating a dilute acid pretreatment facility improve with increasing weight-percent solids loading because of sizable savings in the consumption of steam and acid catalyst. Savings in steam consumption result from not having to heat the extra water to reaction temperature, and the savings in the acid consumption are caused by the proportionality of the hydrolysis rate to the acid concentration in the liquid phase (*5–9*). Thus, reducing the volume of the liquid phase decreases the amount of acid needed for biomass substrates with low ash content. The acid consumption can be decreased from approximately 7.4–2.6 wt% of H_2SO_4 on oven dry wood by increasing the solids concentration from 10–30 wt% in the pH range we have investigated previously (2). The acid consumption could be decreased further by operating at higher temperatures, but at the cost of additional steam consumption.

Dilute acid prehydrolysis, like any other pretreatment process that attempts to utilize whole trees as a feedstock, is actually a series of mechanical and chemical steps, and should be included in the family of chemomechanical pretreatments. The key steps in the process include the following: feedstock collection, transportation and storage, biomass handling, debarking and comminution, dilute acid prehydrolysis, and enzymatic saccharification/fermentation to ethanol or other fuels and chemicals. An optional liquid–solid separation step may be included after dilute acid prehydrolysis if separate processing of the xylose-rich liquid stream and the cellulose-rich solids stream is deemed desirable. However, such a separation is relatively costly and may compromise the sterility of the pretreated substrate, which is another advantage of dilute acid pretreatment with respect to subsequent biochemical processing.

Process development and economic analyses of hypothetical plants producing ethanol from hardwoods using dilute sulfuric acid with subsequent enzymatic hydrolysis have been reported (*10–12*). However, these reports deal with integrated pretreatment schemes of the lignocellulosic material expressly for ethanol production incorporating the separation of the soluble sugar stream. The current SERI study approaches the dilute acid pretreatment of aspen wood chips as an autonomous process capable of producing a sterile biomass slurry containing readily hydrolyzable C_5 and C_6 sugars for a generic biochemical conversion process.

Since the mechanical and chemical steps of dilute acid pretreatment interact in a complex manner, a preliminary process scheme was prepared for an economic parametric evaluation and to identify uncertainties in the process on which to focus additional research. Experimental process data have been obtained to support some selected areas of the proposed dilute acid pretreatment facility.

MATERIALS AND METHODS

Substrates and Other Materials

Air-dried, debarked aspen logs were coarsely chipped using a Brush Bandit mobile knife chipper (Foremost, Inc., Remus, MI) and subsequently milled in a rotary knife mill (model 10/12, Mitts and Merrill, Saginaw, MI) equipped with 1/16–1/2 in rejection screens. Further size reduction was accomplished via a laboratory knife mill (Wiley, model 4, equipped with a 2 mm rejection screen). The aspen wood particles were sized using separating screens on a vibrating screening machine (Model RX-24, W.S. Tyler, Inc., Mentor, OH). The particle size fractions used in subsequent experiments are specified in the text. The equipment and experimental procedures for determination of maximal mill throughput were described previously (*13*).

A cellulase preparation (Celluclast 1.5 L) produced by *Trichoderma reesei* was a gift of NOVO Industries Ltd. (Copenhagen, Denmark). The cellulase preparation was in liquid form, stabilized by the addition of glycerol. The activity of the enzyme preparation used in the assay with 10 mg/mL cellulose on Whatman #1 filter paper averaged 0.42 International Filter Paper Units (IFPU)/mL of assay mixture (*2,14*). The β-glucosidase activity of cellulase preparation, measured using *p*-nitrophenyl-β-D-glucopyranoside as a substrate (*15*), was approximately 0.049 IU/mL of assay mixture. The β-glucosidase enzyme in cellulase preparaton was supplemented with fungal β-glucosidase (Novozyme 188, NOVO Ltd., specific activity 132 IU/mL) at 0.07/IU/mL for total β-glucosidase activity of 0.12 IU/mL of assay mixture. Both stock enzyme preparations were stored at 4°C. The enzyme solutions retained their activities for over a year. The remaining chemicals, which were specified in our previous publications (*1,2*), were purchased from national laboratory supply houses (Sigma Chemical Co., Aldrich Chemical Co., and VWR Inc.).

Chemical Pretreatment and Analysis

The aspen particles were pretreated with dilute (0.5% v/v) sulfuric acid in a 2-gal stainless-steel reactor (Carpenter 20 Cb-3, Parr Co., Moline, IL) equipped with an impeller mixer and acid injection device (*16*). As a result of mixing limitations of impeller mixers with large particles, only low (5 wt%) solids slurries were investigated. The pretreatment experiments were conducted at 140 and 160°C. Reaction times are specified in the text and start when the slurries of wood in deionized water reach the desired reaction temprature and the acid is injected. The pretreated chips were exhaustively washed with distilled water to allow the diffusion of the hydrolyzed components out of the chips.

The experimental conditions for enzymatic hydrolysis and chemical analysis have been described previously (*1,2,17*). Glass beads used for shearing action of the wood particles were not present in the enzymatic digestion studies unless stated.

Process Engineering Design Assumptions and Calculations

The feed material to the plant was assumed to be commercial 1-in. aspen wood chips containing 50% moisture content and having the chemical composition described previously (*1,2*). The size of the pretreatment facility was based on a base case of producing 25 million gallons of absolute ethanol from the total potentially fermentable hexose and pentose streams produced from the pretreatment. The location of the plant was arbitrarily chosen as the southeastern United States for two reasons: There is a surplus of underutilized hard wood in that region of the country, and the weather is relatively mild year around, thus allowing the delivery of wood chips on a continuous basis and requiring a relatively inexpensive wood receiving and handling yard. The yield of hexoses was assumed to be 100%, whereas the pentoses were assumed to undergo 20% thermal destruction during the pretreatment. Although use of 100% of the hexose available in the feedstock for the calculation of yield of ethanol produced by the plant is admittedly very optimistic, this remains the goal of our research efforts. The overall recovery of sugars from the prehydrolysis was calculated at 98.6%. The fermentation yield of ethanol from sugars was 95% of theoretical, and the ethanol recovery yield was calculated at 99.5%.

A wood receiving and handling facility, a comminution step, and a dilute sulfuric acid pretreatment process were designed in detail from the best currently available information and research data (*1,2,10–13,18,19*). To estimate capital costs in a consistent manner, the ICARUS computer-aided cost estimating program was used. This program produces a detailed time and materials estimate with an accuracy of ±10% for a completely defined process (*20*). Vendor quotes for specific unique process equipment items were used whenever possible and supplied to the ICARUS program to assure quality of the estimate.

The costs of the feedstock, chemicals, process water, electricity, and steam were obtained from vendor quotes, SERI studies (*21*), and the *Chemical Marketing Reporter*. The energy balances were calculated assuming the feed chips, process water, and acid to be at 60°F. The enthalpy of the low pressure steam was obtained from steam tables, and the specific heat of bone-dry aspen wood was assumed to be 0.32 BTU/lb. The capital charges were calculated using a fixed charge rate (FCR) of 0.13 times the inside battery limits (ISBL) (*21*).

RESULTS AND DISCUSSION

Process Design Data

Our focus on the research and development of a dilute acid pretreatment process for a renewable form of biomass has been to produce a substrate for an enzymatic or microbial bioconversion process for liquid fuels, with the present emphasis on ethanol. We have shown previously (*1,2*) that for aspen wood meal with a high solids loading (up to 40 wt%), very rapid hydrolysis of the xylan fractions occurs with dilute sulfuric acid at 140 and 160°C with little concomitant destruction of the sugars. This prehydrolysis forms a necessary prerequisite step for subsequent enzymatic saccharification of cellulose. The capital cost, as well as the energy cost of the comminution step, would a priori seem relatively large in such a pretreatment process. Therefore, we have investigated the effects of size of the aspen wood chips as it relates to the rates of both enzymatic saccharification and xylan hydrolysis.

The enzymatic digestibility of cellulose in screened aspen particles pretreated with dilute acid to remove greater than 90% of the xylan (140°C, 60 min, 0.5% H_2SO_4 (v/v), 10% solids) was investigated first. All pretreated screened particles were chemically analyzed for xylan content. The xylan removal was found to average between 91–96% of the starting xylan in untreated wood. The results for long-term digestibility of cellulose in uniformly pretreated aspen wood particles are shown in Table 1.

Table 1
Enzymatic Digestibility of the Cellulose in Aspen Wood Solid Residues After Acid Pretreatment at 140°C for 1 h as a Function of Mesh Size and Shearing Action During the Enzymatic Digestion

	Weight percent anhydroglucose released from total glucan present in the pretreated residues	
Mesh size range, in.	Glass beads present in reaction vessel	No glass beads present in reaction vessel
3/8–1/2	62	52
1/4–3/8	77	76
5 Mesh–1/4	84	78
5–8	84	77
8–12	77	78
12–16	86	84
16–20	85	82
20–35	83	76
35–60	89	81
60–80	86	79
Alpha–cellulose control	95	94

The experimental protocol used two approaches. In the first system, the wood particles were gently tumbled in the presence of the cellulase enzyme complex and several glass beads (6 mm in diam) to simulate mild abrasive action during the digestion process. In the second system, the wood particles were slowly tumbled with enzyme and buffer alone. The results from both systems indicate that a clear drop in digestibility occurs when the particle size of pretreated aspen wood reaches 1/4–3/8 in. size, and the additional grinding action in this system provides slightly higher cellulose digestion.

The results of the kinetic experiments were more revealing (Fig. 1). The most rapid digestion was obtained for particles in the 60–80, 40–60, and 40–20 mesh size ranges, and rates became very slow for the particles larger than 5 mesh. Since the enzymatic hydrolysis of cellulose is by far the slowest step in the overall conversion of cellulose in biomass to soluble sugars, the results in Fig. 1 show that the optimal particle size of the

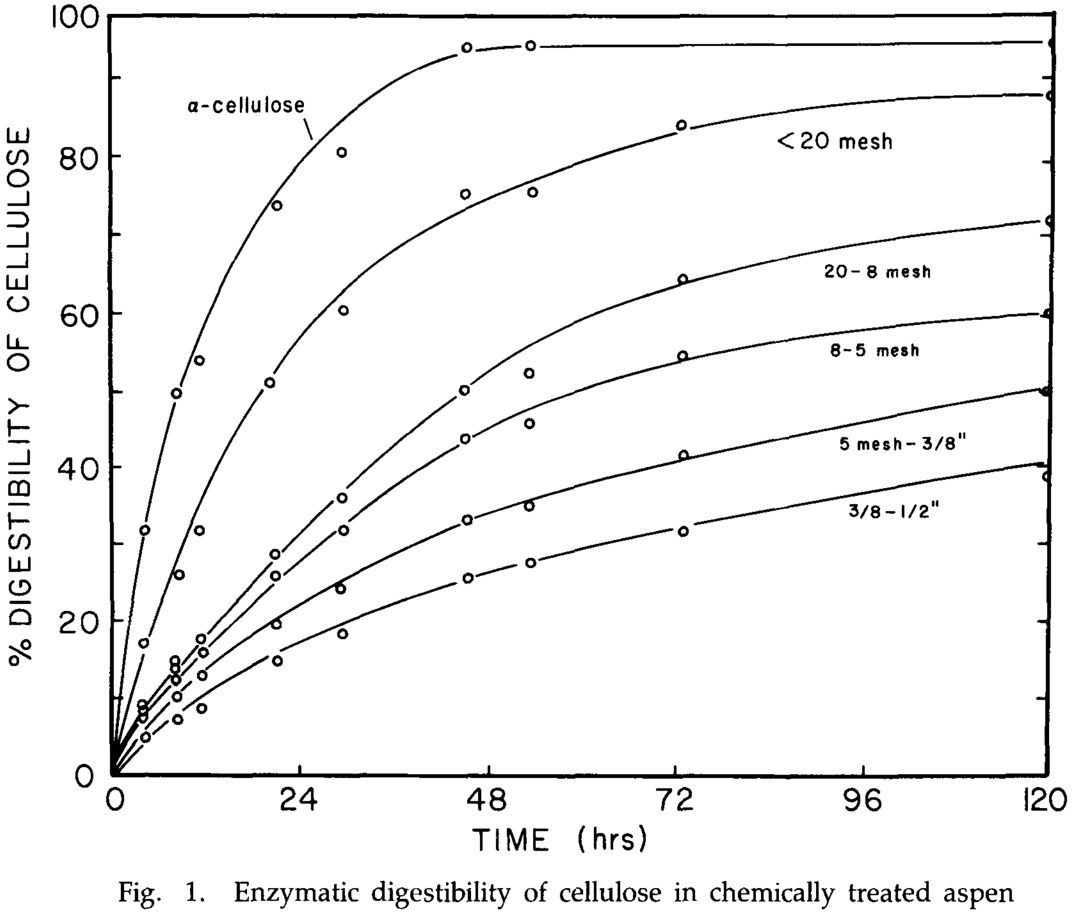

Fig. 1. Enzymatic digestibility of cellulose in chemically treated aspen wood (140°C/1 h, 0.5% H_2SO_4 v/v, 10% solids) using NOVO Celluclast 1.5 L cellulase supplemented with beta-glucosidase as a function of particle size compared to an α-cellulose control.

pretreated aspen wood for enzymatic hydrolysis is approximately 20 mesh or less.

The situation is similar but less severe in the dilute acid pretreatment step. The preliminary data shown in Fig. 2 and 3 indicate that the reaction time needed for high (above 90%) removal of xylan is 2–3 times longer for particles in the 1/2–1 in. range than for sawdust (20–80 mesh). Since the xylan removal correlates highly with enzymatic digestibility of cellulose in pretreated aspen wood (*1,2*), the reaction conditions always have to be adjusted to achieve the necessary degree of hydrolysis of xylan.

The problem is compounded by a decrease in bulk packing density of wood as the particle size is increased. Experimental results for aspen wood are shown in Fig. 4. The decrease in particle size from the 1/2–1 in. range to sawdust (20–80 mesh) allows an approximately 44% increase in bulk packing density of aspen wood in the pretreatment reactor.

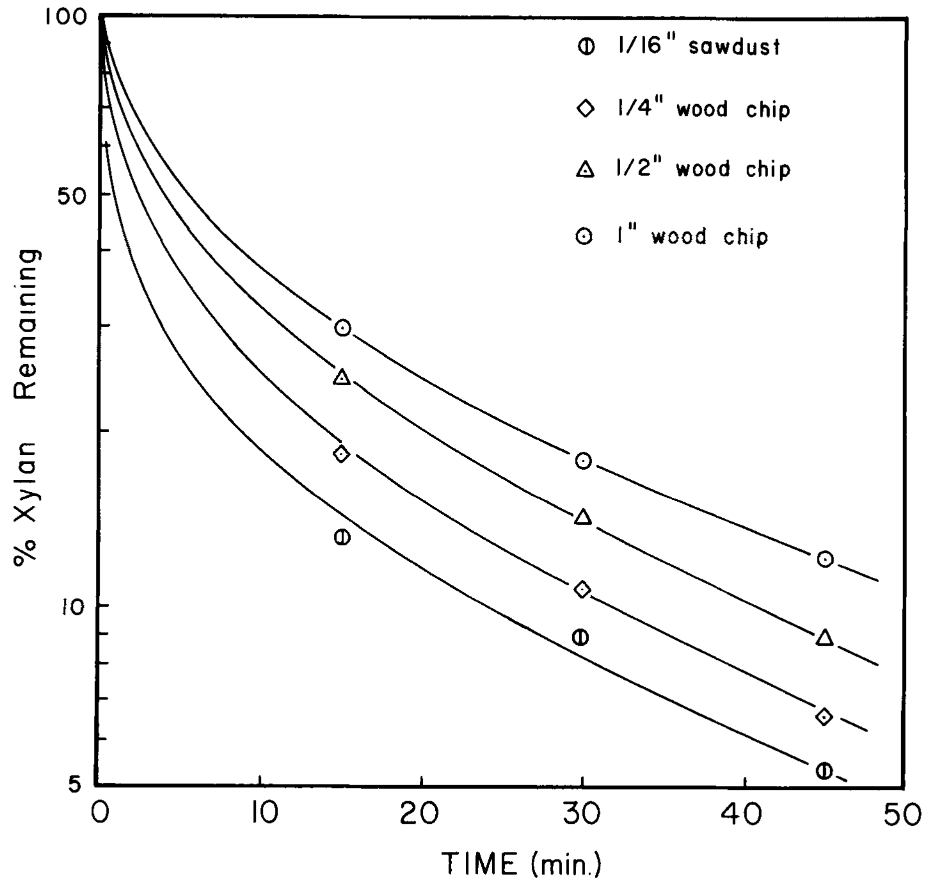

Fig. 2. Anhydroxylose remaining in solid residues versus time of dilute acid hydrolysis (0.5% v/v H_2SO_4) of aspen wood (5 wt% solids) at 140°C using four different wood chip sizes.

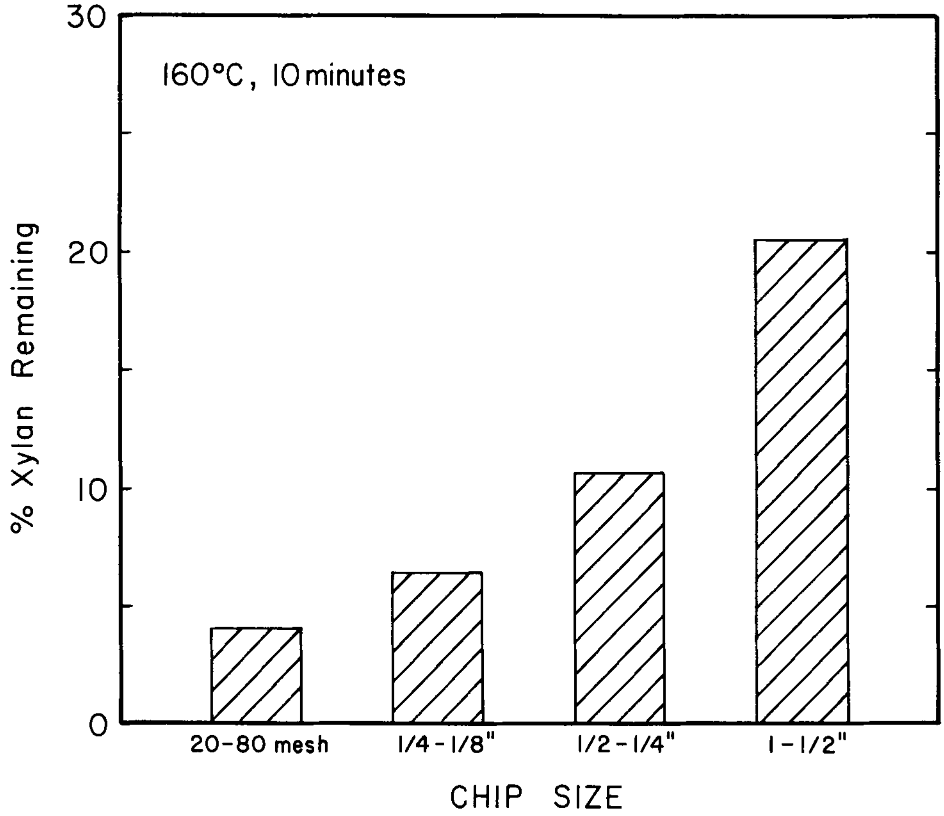

Fig. 3. Anhydroxylose remaining in chemically pretreated aspen wood (0.5% H_2SO_4 v/v, 160°C/10 min, 5 wt% solids) as a function of particle size.

The negative effects of large particle size are somewhat offset by a large decrease in electrical power consumption needed for milling of aspen chips to smaller particles (*13*) and by increased throughput of knife mills (Fig. 5) needed for this operation. However, the large pretreated particles would have to be mechanically disintegrated to smaller particles for subsequent enzymatic hydrolysis to increase the rate of cellulose hydrolysis (Fig. 1). Since dilute acid pretreatment under conditions (140–160°C) we have investigated does not provide sufficient pressure drop for chip disintegration by rapid decompression, and since experimental data on comminution of wet, pretreated aspen chips are lacking at the present time, the preliminary process design is based on initial comminution of dry, debarked wood chips and chemical pretreatment of small (approx. 20 mesh) particles.

Overall Process Description

The simplified flow diagram of the process is shown in Fig. 6. Fresh aspen chips are delivered by truck, debarked, and cleaned through the shearing action of an auger cleaner (Morbark Industries, Inc., Winn, MI), and stored for no longer than one month. The debarker-cleaner is not

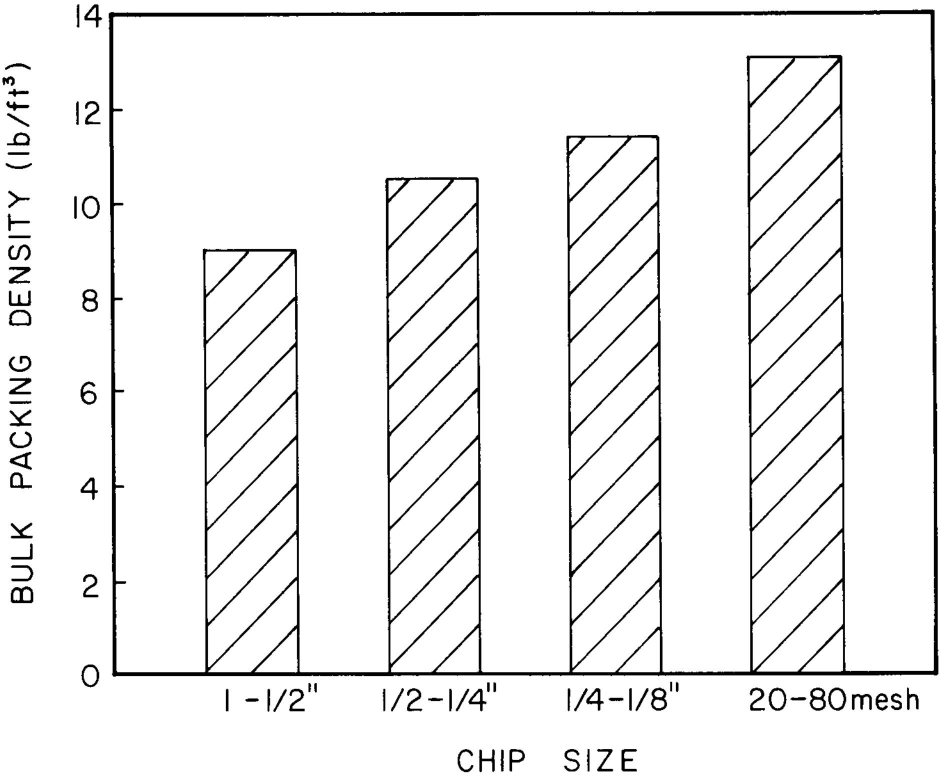

Fig. 4. Bulk packing density of oven dried aspen wood (45°C) as a function of chip size.

shown in Fig. 6, but is included in the detailed process design, as are several other minor pieces of process equipment. Chips are then transferred to a knife mill equipped with a 1/16 in screen to produce particles primarily in the 20–80 mesh range (*13*).

After milling, the wood particles are fed to a continuous Pandia digester (Shartle-Pandia Division, Middletown, OH) via a feed hopper and pressurized screw feeder. The continuous prehydrolysis is based on the Pandia continuous pulping digester that has been used for similar operations in the pulp and paper industry (22). All equipment to be in contact with hot, dilute sulfuric acid is assumed to be constructed from Carpenter 20 Cb-3 stainless steel, which exhibits high corrosion resistance under these conditions. The use of much cheaper 316 stainless steel was avoided because, in our experience, it suffers from high corrosion rates in hot, dilute sulfuric acid.

The continuous digester is composed of two parts connected by a rotary pressure valve. The wood in the first section is impregnated with hot dilute sulfuric acid, preheated by it and also deaerated. The flows of wood, hot dilute sulfuric acid, and additional steam are adjusted for the mixture to reach approximately 95°C at the end of the first section, and residence time of solids in the first section is assumed to be 10 min. The

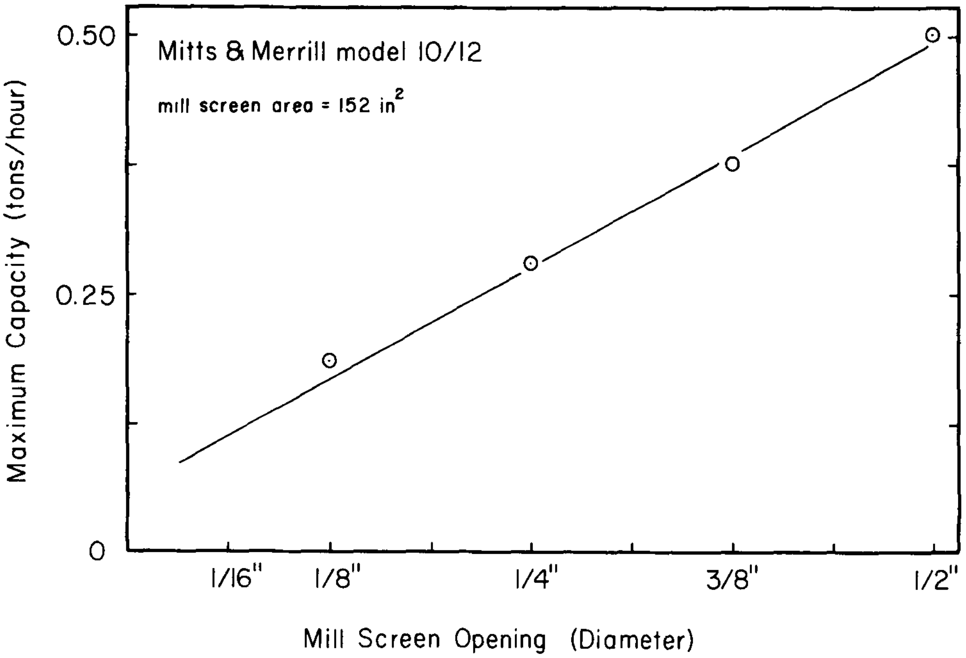

Fig. 5. Maximal feed rates of aspen wood chips through a Mitts and Merrill knife mill with a fixed size mill screen (152 in.^{2}) as a function of chip size.

impregnated hot wood is then transferred to the second section where it is rapidly heated by steam injection to 160°C. The residence time of solids in this section is also assumed to be 10 min to conform with our previous kinetic data (*1,2*).

The initial concentration of solids in the digester is calculated to be 30 wt%. The concentration of solids at the exit is, of course, lower since approximately one-third of the aspen wood is dissolved in the pretreatment process. The hot pretreated mixture is discharged into a blowdown tank where it is rapidly cooled to 95°C by evaporation of entrained water. Rapid cooling of the pretreated wood will minimize the degradation of xylose to furfural which is detrimental to subsequent saccharification/fermentation steps. The vapors are purified by passage through a vent scrubber to remove small amounts of acetic acid, furfural and other volatile organics produced by the pretreatment process. The hot acidic solids are then transported to the neutralization blender where they are mixed with limestone powder. Carbon dioxide produced by neutralization will be vented to the atmosphere. It should be noted that pretreated solids are kept hot after the pretreatment to preserve their sterility for subsequent saccharification/fermentation.

Discussion of the Process Design and Economics

This preliminary process design and economic analysis had two major goals. First, a relatively detailed equipment list combined with the

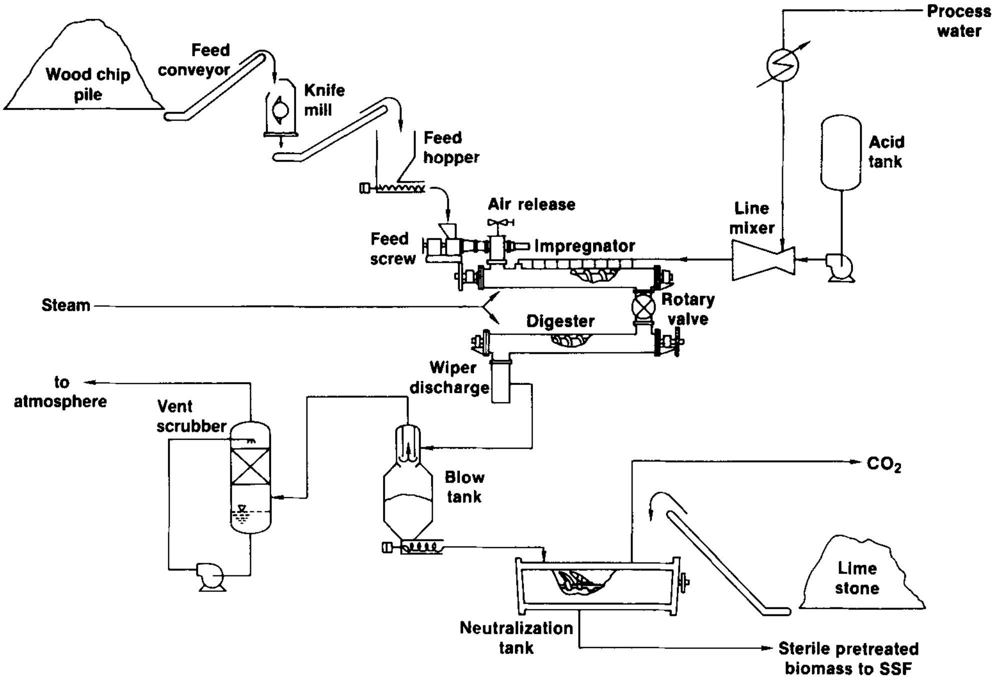

Fig. 6. A simplified schematic process flow diagram of a dilute sulfuric acid pretreatment for aspen wood meal.

overall mass and energy balance was needed to yield a model to begin a parametric study of the key process variables. Second, if the process economics seem encouraging, then areas of uncertainty in the process need to be explored and characterized through further research before pilot studies can be initiated.

The estimated cost of production per pound of total sugar produced by the process is seen in Table 2. The cost of operating the plant (before byproduct and steam credits) is approximately 4.50¢/lb of fermentable sugar (C_5 and C_6) or $0.621/gal ethanol. The major cost is that of the wood feedstock (2.54 ¢/lb, or 56% of the total production cost) with capital charges and utility costs primarily contributing to the remaining cost of production. Thus, the overriding parameters that affect the economics of sugar production from aspen wood chips via dilute acid prehydrolysis are feedstock costs and yield of the sugars produced. The total cost of sugar is somewhat deceiving because the pretreatment produces only approximately one-third of the sugar in water soluble, monomeric form immediately available for fermentation. The bulk of the sugar is tied up in solid lignocellulosic residue and has to be released at additional expense by enzymatic hydrolysis of cellulose. The cost of enzymatic hydrolysis was not included because it is process-specific whereas the design goal of the pretreatment process was that it be generic in nature. Some saccharification/fermentation processes will probably have to rely on saccharification of cellulose by separately produced cellulase enzymes. An example of such a process is ethanol production, where yeast

Table 2
Estimated Cost of Production for Dilute Sulfuric Acid Pretreatment[a]

CAPITAL COST SUMMARY $\$10^6$

Inside battery limits (ISBL)	11.2
Offsites (20% ISBL)	2.2
Total Fixed Investment (TFI)	13.4
Working capital (10% of TFI)	1.3
Total utilized investment	14.7

PRODUCTION COST SUMMARY

	Units/yr	Price, \$/unit	Annual cost, $\$10^3$	Ferm. sugars C_5 & C_6 cents/lb	\$/gal Ethanol
Raw materials					
Wood chips-feedstock, tons (dry)	208,000	42.00	8,770	2.54	
Sulfuric acid, tons	5,359	65.00	349	0.10	
Calcium carbonate, tons	5,468	34.00	186	0.05	
Total cost			9,305	2.69	0.372

Utilities					
Power (KWH)	29,500,000	0.05	1,474		
Process water (10^3 gal)	34,900	0.60	21		
Steam, 150 psi (10^3 lb)	229,000	4.00	916		
Total cost			2,411	0.70	0.096
Operating costs					
Operators (3/shift, $26,000/man)			328		
Supervision			35		
Maintenance material and labor (6% of ISBL)			672		
Total costs			1,035	0.30	0.041
Overhead expenses					
Direct overhead (45% labor + supervision)			163		
General plant overhead (65% operating costs)			673		
Insurance, taxes (1.5% of TFI)			201		
Total expenses			1,037	0.30	0.041
Total cost of production			13,788	4.00	0.551
Capital Charges (FCR = 0.13)			1,742	0.50	0.070
Net cost of production			15,530	4.50	0.621

[a] Basis: US Southeast, 1st quarter 1987
Capacity: 345 Million lb/yr potentially fermentable sugars (C_5 and C_6)
Operating: 8,000 h/yr for 25 million gal/yr ethanol

and bacteria which efficiently ferment sugars to ethanol either do not produce cellulase enzymes at all, or, if improved by genetic engineering, secrete them in insufficient amounts.

Other potential processes, i.e., anerobic digestion and single-cell protein production, usually employ cellulolytic microorganisms. The cost of cellulase in these processes is then buried in the overall process costs. Even in the processes relying on the separate production of cellulase enzymes, the rapid evolution of both enzymes and microorganisms producing them makes the estimates of enzyme costs highly inaccurate. In addition, the enzyme loading and saccharification/fermentation times can vary widely according to process configurations and conditions employed. Therefore, they have to be investigated and evaluated as a part of the overall bioconversion processes, which are closely coupled to the pretreatment step yet distinct from it.

Table 3 lists the major sections of pretreatment equipment in the process in terms of capital cost and energy consumption. The reactor's installed cost comprises over 35% of the entire plant's capital cost with the knife mill contributing just under 6%. However, the capital charges are only accountable for 11% of the total cost of production (Table 2); thus improvements in the design of the pretreatment reactor and comminution systems will not improve the economics significantly. The energy consumption of the comminution step is about 92% of the entire plant's electrical energy needs. Therefore, the pretreatment plant is highly vulnerable to the cost of electricity. This vulnerability can be decreased by development of alternate comminution schemes that are less energy intensive than dry milling.

For instance, aspen wood becomes very friable after dilute acid pretreatment, providing an opportunity for development of additional

Table 3
Capital Cost and Energy Consumption of the Major Sections of Process Equipment for Pretreatment

Equipment	Installed capital investment, \$ Millions	Electricity, KWH $\times 10^{-6}$	Steam, $lb_m \times 10^{-6}$
Wood handling and storage	3.69	0.9	0
Knife mill	0.65	27.1	0
Prehydrolysis reactor	4.01	1.1	229
Blowdown tank and neutralization mixer	1.10	0.3	0
Conveyors, pumps, storage tanks, and misc.	1.75	0.1	0
Total:	11.20	29.5	229

systems for mechanical disintegration of pretreated chips either before or during enzymatic hydrolysis of cellulose in pretreated biomass. Such developments will have to be coupled to the cheaper design of the pretreatment reactor. The installed cost of the continuous digester made from Carpenter Cb 20-3 is approximately 12–18 times higher than the cost of the similar reactor made from carbon steel. The large span in the material costs indicates that the evaluation of cheaper construction materials that retain corrosion resistance at relatively moderate pressures and temperatures would have a great impact on the overall capital cost. The development of different, less expensive reactor configurations which could process wood particles in the size range of 1/8–1/2 in. at high concentration of solids would have a similar impact.

Although the present study is based on a mild temperature dilute acid pretreatment of biomass, other pretreatment methods such as steam explosion have received a great deal of attention (*21,23,24*). Acid processes may have a slight advantage because their lower operating temperatures generate lower quantities of degradation products, which may, in turn, inhibit downstream processes. Engineering designs prepared to date (*24*) do not show one method having a major capital or operating cost advantage over another.

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REFERENCES

1. Grohmann, K., Torget, R., and Himmel, M. (1985), *Biotech. Bioeng. Symp.* **15,** 59.
2. Grohmann, K., Torget, R., and Himmel, M. (1986), *Biotech. Bioeng. Symp.* **17,** 135.
3. Grethlein, H. E. (1985), *Bio/Technology* **3,** 155.
4. Sawabe, O., Gakkaishi, M. (1980), **26,** 641.
5. Kabayashi, T., and Sakai, Y. (1956), *Bull. Agric. Chem. Soc. Jpn.* **20,** 1.
6. Cahels, D. R., Lee, Y. Y., and Chambers, R. P. (1983), *Biotechnol. Bioeng.* **25,** 3.
7. Bhandari, N., MacDonald, D. G., and Bakhshi, N. N. (1984), *Biotechnol. Bioeng.* **26,** 320.
8. Maloney, M. T., Chapman, T. W., and Baker, A. J. (1985), *Biotechnol. Bioeng.* **27,** 355.
9. Kim, S. B., and Lee, Y. Y. (1987), *Biotech. Bioeng. Symp.* **17,** 71.
10. Wilke, C. R., and Blanch, H. W. (1985), *Ann. Rep. Solar Energy Res. Inst.*
11. Chem Systems, Inc., SERI Subcontract XX-2-02205-1 (1983).
12. Arthur D. Little, Inc., NYSERDA Report 85-9 (1985).

13. Himmel, M., Tucker, M., Baker, J., Rivard, C., Oh, K., and Grohmann, K. (1985), *Bioeng. Symp.* **15,** 39.
14. Mandels, M., Andreotti, R., and Roche, C. (1976), *Biotech. Bioeng. Symp.* **6,** 17.
15. Sternberg, D., Vigaykumar, P., and Reese, E. T. (1977), *Can. J. Microbiol.* **23,** 139.
16. Himmel, M. E. (1986), *Biotechnol. Bioeng.* **28,** 126.
17. Grohmann, K., Himmel, M., Rivard, C., Tucker, M., Baker, J., Torget, R., and Graboski, M. (1984), *Biotech. Bioeng. Symp.* **14,** 137.
18. Badger Engineers, Inc. "Economic Feasibility Study of an Acid-Based Ethanol Plant," SERI Subcontract ZX-3-030-96-2, 1984.
19. Stone and Webster Engineering Corp., "Economic Feasibility Study of an Enzyme-Based Ethanol Plant," SERI Subcontract ZX-3-03097-1, 1985.
20. ICARUS Corp. COST System User's Manual, Rochville, MD (1984).
21. Wright, J. D., Power, A. J., and Douglas, L. J. (1986), *Biotech. Bioeng. Symp.* **17,** 285.
22. Markham, L. D., and Courchepe, C. E. (1980), *Proc. 1980 Pulping Conf.*, TAPPI Press, Atlanta, GA.
23. Iotech Corporation Ltd., Final Report for Contract No. IB-1-9343-1, (1982).
24. "Fuel Alcohol Technical and Economic Evaluation," draft, Solar Energy Research Institute, Golden, CO, 1986.