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# **Production of Butyl Solvents from Lignocellulose: An Economic Analysis**

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## PRODUCTION OF BUTYL SOLVENTS FROM LIGNOCELLULOSE: AN ECONOMIC ANALYSIS

### Introduction

This paper describes a conceptual process to convert wood and other lignocellulosics to butanol, isopropanol, and ethanol. The concept is based primarily on the well-known butyl solvents fermentation in which *Clostridium* bacteria convert carbohydrates to solvents. Two new developments are introduced: the progressing batch reactor for carbohydrate production by wood hydrolysis and a low-energy, liquid extraction process for product separation. Their impact on the traditional process is also discussed. The overall process comprises four major sections: feed handling, acid hydrolysis, fermentation, and solvents recovery.

The hardwood chips and sawdust used as feedstock are the source of lignocellulose most readily available on a commercial scale throughout the United States and Canada (1). Other cellulosic feedstocks such as bagasse, straw, and corn stover could also be used without significant process modification.

Preliminary production and capital costs are estimated for a base-case grassroots plant processing 200,000 dry metric ton of wood per year and producing 89 million lb of solvents. (The overall yield of the process is expected to be approximately 445 lb of solvents per dry metric ton of wood.) For the purposes of this study the plant is designed to be self-sufficient in utilities.

Established large markets now exist for the products. However, there is also a potential market for n-butanol and isopropanol as cosolvents for methanol in gasoline. Butanol can also be blended directly with diesel fuel if economically warranted (2,3).

A block flow diagram of the process is shown in Figure 1.

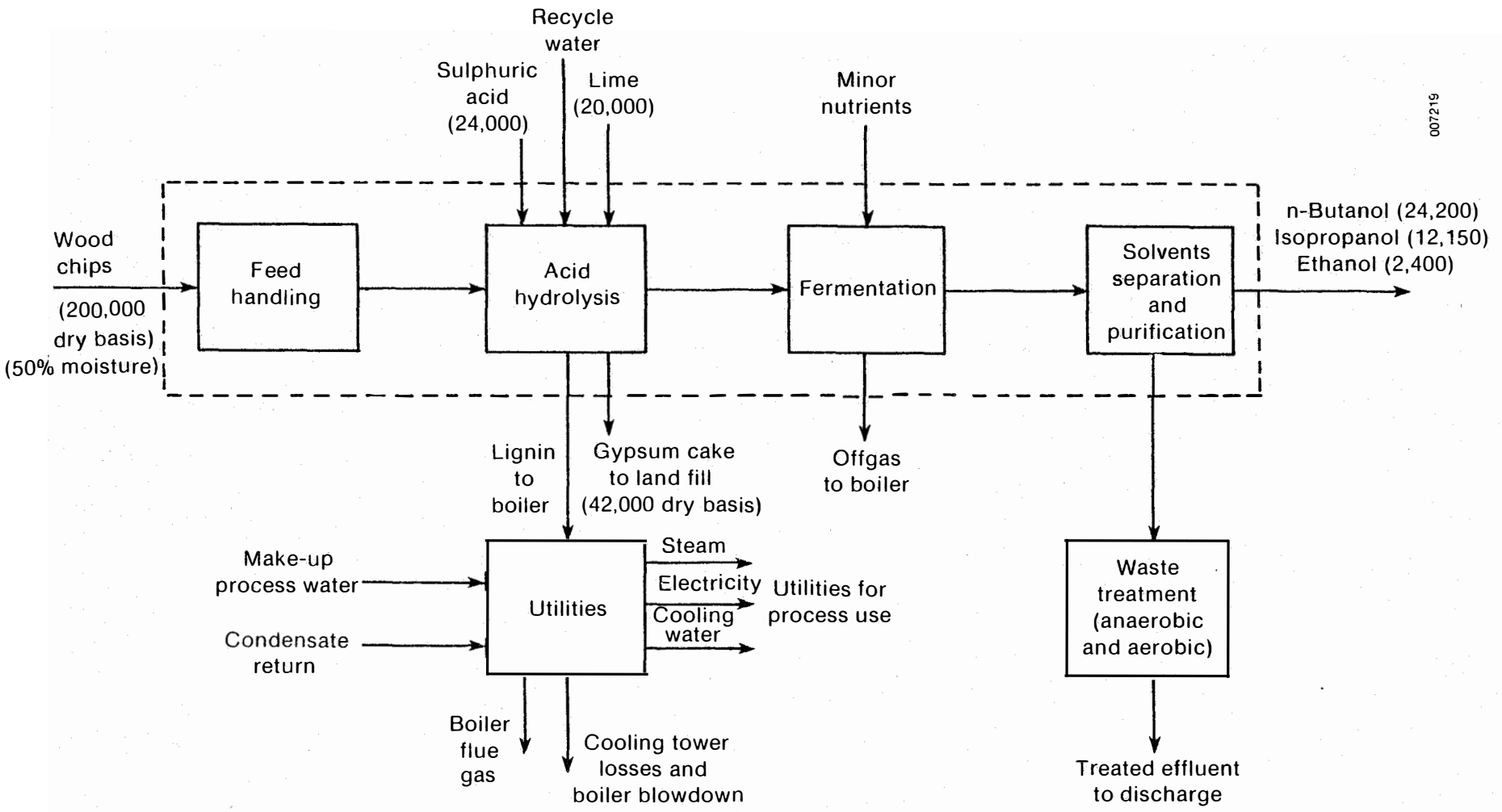
### Key Technology Developments

#### The Progressing Batch Reactor

The progressing batch reactor was developed at the Solar Energy Research Institute (SERI), Golden, Colo. The SERI system consists of a series of batch hydrolysis reactors operated so that acid solution contacts the wood chips in a continuous counter-current fashion (4).

Batch dilute acid hydrolysis to convert wood to fermentable sugars is proven technology (5,6,7,8,9), although the designs, details, and procedures necessary for building a reliable reactor are not all available in the literature. The existence of the Russian wood chip percolation industry, however, suggests that the design problems have been solved. The principal operational problems reported in the literature include fouling of the outlet pipes by degradation products and collapse, compaction, and plugging of the chips within the reactor as hydrolysis progresses (10).

Using the progressing batch method in this hydrolysis is expected to result in hexose sugar yields of 75% to 80% of theoretical and pentose sugar yields of greater than 90% of theoretical. The sugar concentration produced by the progressing batch reactor is 5% by weight total sugar at the above yield levels (4).



Note: Figures in parentheses are metric tons per year.

Figure 1. Process Block Flow Diagram Showing Major Material Flows

This hydrolysis system seems particularly suited for butyl solvents fermentation for the following reasons:

- It produces high yields of sugars at a concentration of 5% by weight-ideal for the *Clostridium* fermentation.
- Both the hexose and the pentose sugars can be used by the *Clostridium* to produce solvents.

A pilot-scale reactor system to demonstrate the operational principle of the acid hydrolysis portion of the process has been built at the SERI laboratories.

### **Continuous Butyl Solvents Separation**

The butyl solvents fermentation process was a major source of n-butanol and acetone for many years (11,12,13,14,15). However, the high energy requirements for product separation and purification by batch distillation were a serious disadvantage. Steam consumption in the traditional batch distillation separation process was ~20 lb/lb of solvents.

A continuous separation process using primary steam stripping and liquid-liquid extraction was developed by Badger Engineers, Inc. Steam consumption using the liquid-liquid extraction process is calculated to be approximately 6 lb/lb of solvents or 30% of the requirements of the traditional batch distillation. Other utility requirements are also significantly reduced in this process.

### **Process Description**

The four main process sections are feed handling, acid hydrolysis, fermentation, and solvents recovery and purification. The process also includes power generation and wastewater treatment systems for stand-alone operation.

#### **Feed Handling**

The feed-handling section receives wood chips from trucks or rail cars. Debris (such as rocks and soil) is separated in a flume pond, and the chips are stored in a pile. Chips are fed to the process via a coarse screen and conveyor. A magnet separates tramp metal items.

#### **Acid Hydrolysis**

Figure 2 shows the equipment used to implement the progressing batch reactor process. The system consists of a series of batch reactors operated so that the acid contacts the wood chips in a continuous counter-current fashion. The first reaction vessel is isolated from the adjacent reactors by closing the valves on the inlet and outlet lines. The top hatch is opened, and wood chips are dumped or blown into the reactor. To obtain a higher sugar concentration in the outflow, the reactor is packed as densely as possible by compacting the chips with blasts of steam and using a mixture of large chips and sawdust in the feed. The sawdust partially fills the spaces between the large chips. The reactor is then sealed and purged of air by passing steam through the tank, which also preheats the chips.

Reactors 2 and 3 are in the prehydrolysis stage where the hemicellulose fraction is hydrolyzed. The liquid (water and acid) is at 150°C. The amorphous five- and six-carbon

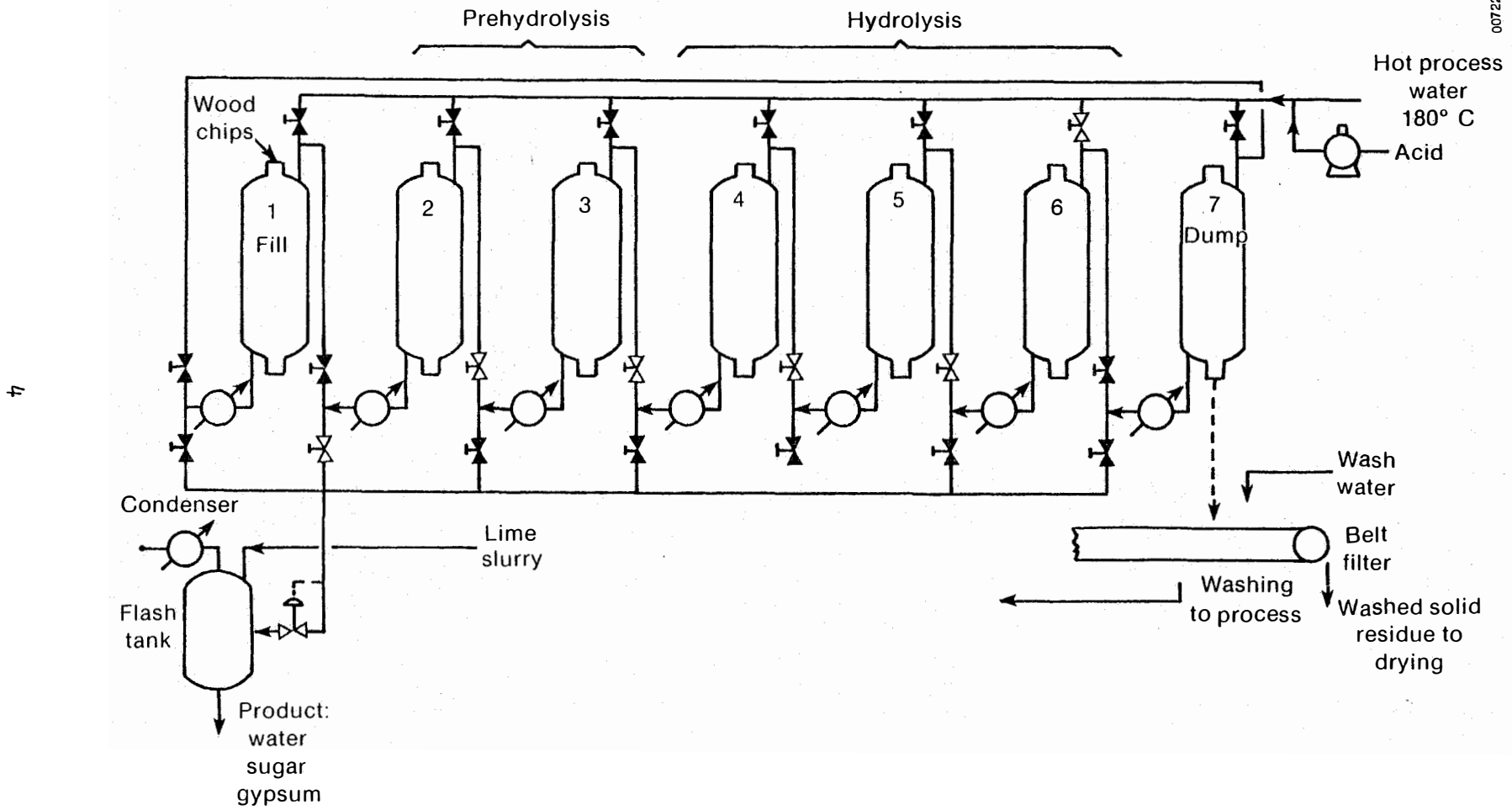


Figure 2. Schematic of Progressing Batch Reactor

polymers are readily hydrolyzed, but the product sugars undergo little degradation at this temperature. Thus, the hemicellulose sugar is recovered with high yield. The crystalline cellulose is hydrolyzed in reactors 4, 5, and 6 at 180°C. At these conditions, the crystalline cellulose is 92% hydrolyzed after 3 h.

The liquid flow from reactor to reactor is controlled by simply closing the valves leading to the hot water and flash lines and opening the valve connecting the reactors. The temperature of each reactor stage is controlled with a feed heater/cooler. In the configuration shown, reactor 7 is being emptied. The lignin and unreacted crystalline cellulose are blown out by the pressure of the steam in the reactor. Residual sugars are washed out from the solid residue on a belt filter.

After about 30 min the flow configuration is changed. Fresh liquid introduced into reactor 5 flows through reactors 5, 4, and 3 at 180°C (hydrolysis) and through reactors 2 and 1 at 150°C (prehydrolysis). Reactor 6 is emptied and reactor 7 is filled. After 30 min the flow is changed again, and feed is introduced into reactor 4 and taken from reactor 7.

Lime slurry is introduced into the flash vessel to neutralize the sulfuric acid and adjust pH to 7 before fermentation. Gypsum is separated and washed in a series of hydrocyclones and dewatered in a rotary filter. Gypsum cake is disposed of by landfill. Lignin residue is washed to recover residual sugars, dried, and fed to a boiler for process energy.

### Fermentation

A sugar solution containing hexose and pentose sugars at a total concentration of 5% by weight is sent to fermentation. Additional nutrients, if required, are sterilized and added to the fermentation medium separately. Fermentation is carried out by strains of *Clostridium butylicum* to produce butyl solvents in the approximate ratio of 6 parts n-butanol: 3 parts isopropanol: 1 part ethanol [some butyl solvents fermentations produce acetone rather than isopropanol (14)]. The fermentation system consists of 24 stainless-steel-lined mild steel tanks of 200,000 gallons each and is operated in the traditional batchwise manner. The operating cycle, which lasts for approximately 48 h, consists of filling, broth deaeration, and inoculation; fermentation; emptying; and cleaning and sterilization.

Nitrogen or fermenter offgas can be used to deaerate the broth before inoculation and to prevent vacuum after sterilization. During fermentation the pH is controlled above 5.5 by adding lime. Vent gas from the fermenters containing hydrogen and carbon dioxide is scrubbed to remove products and fed to the boiler.

### Solvents Recovery and Purification

Beer from the fermentation containing approximately 1.1% butanol, 0.5% isopropanol, 0.1% ethanol, 0.2% suspended solids, and minor quantities of organic acids at pH 6.0 is fed continuously to the beer stripper. This produces an overhead stream containing about 50% by weight total solvents. Beer column overhead vapors are deodorized by passing them through a copper shavings tank; they are then condensed and sent to the liquid extraction system.

Solvents are extracted from the beer column overhead condensate by multistage counter-current contact with liquid extractant. The extract, consisting of extractant and mixed alcohols, is flashed to atmospheric pressure to recover the extractant, which is

condensed and recycled. Heat for the flash column reboiler is provided by condensing the beer column overheads. Mixed alcohols from the flash column are separated by distillation. A flow diagram of the solvents separation and purification process is given in Figure 3. Calculations indicate that this processing scheme gives almost complete recovery of butanol and isopropanol. There is, however, a significant loss of ethanol in the beer column bottoms. The recovery of ethanol in this system is 60%.

### Utilities and Offsites

The process is completely self-sufficient in utilities with systems for the generation of steam and condensate, electric power, boiler feed water, cooling water, plant and instrument air, nitrogen, tempered water, and brine refrigeration. The fuel requirements for the plant come from offgas from the fermenters and anaerobic digestion as well as lignin residue from hydrolysis. Raw water requirements must be imported and are principally used for make-up to the recirculated cooling water system. Additional units in the offsites include product storage facilities, flume, plant buildings (control room/lab and change house/maintenance shops), plant fencing, and internal roads.

### Waste Treatment

Stillage bottoms from the beer column contain approximately 20,000 ppm COD. It is first cooled by heat exchange with incoming beer, filtered to remove suspended solids, and then treated by anaerobic digestion. A substantial quantity of the digester effluent is recycled to the hydrolysis unit. The net blowdown is further treated aerobically, producing an effluent suitable for discharge. The methane-rich gas produced in the anaerobic digester is sent to the boiler. The stillage solids are also sent to the boiler and burned along with lignin. The gypsum produced from neutralization of the hydrolysate is sent to landfill disposal.

### Process Economics

#### Capital Cost

The capital cost of the base-case plant processing 200,000 dry metric ton/yr of wood and producing 89 million lb/yr of butyl solvents has been estimated to be \$71 million. This figure is a preliminary estimate based partly on estimates carried out elsewhere and adjusted for capacity and inflation (1,13). The installed plant cost including both onsite and offsite units is based on an estimate for the U.S. Gulf Coast. Additional components of capital cost have also been estimated and include engineering cost, land cost, spare parts, start-up fees, construction allowance for a typical southeastern U.S. location, and estimating contingency. A breakdown of the capital cost by plant section and additional factors is given in Table 1. The most expensive process section is the fermentation. This section alone accounts for almost 50% of the onsite capital cost.

#### Revenues and Production Cost

The annual revenues from such a plant were estimated at \$28.8 million. This was based on present market prices for butanol (36¢/lb), isopropanol (31¢/lb), and ethanol (25¢/lb) (15). (This figure excludes potential revenues from any furfural produced in the hydrolysis section. The potential revenue from sale of separated stillage solids for animal feed is likewise not considered.) Production costs and revenues are shown in Table 2.



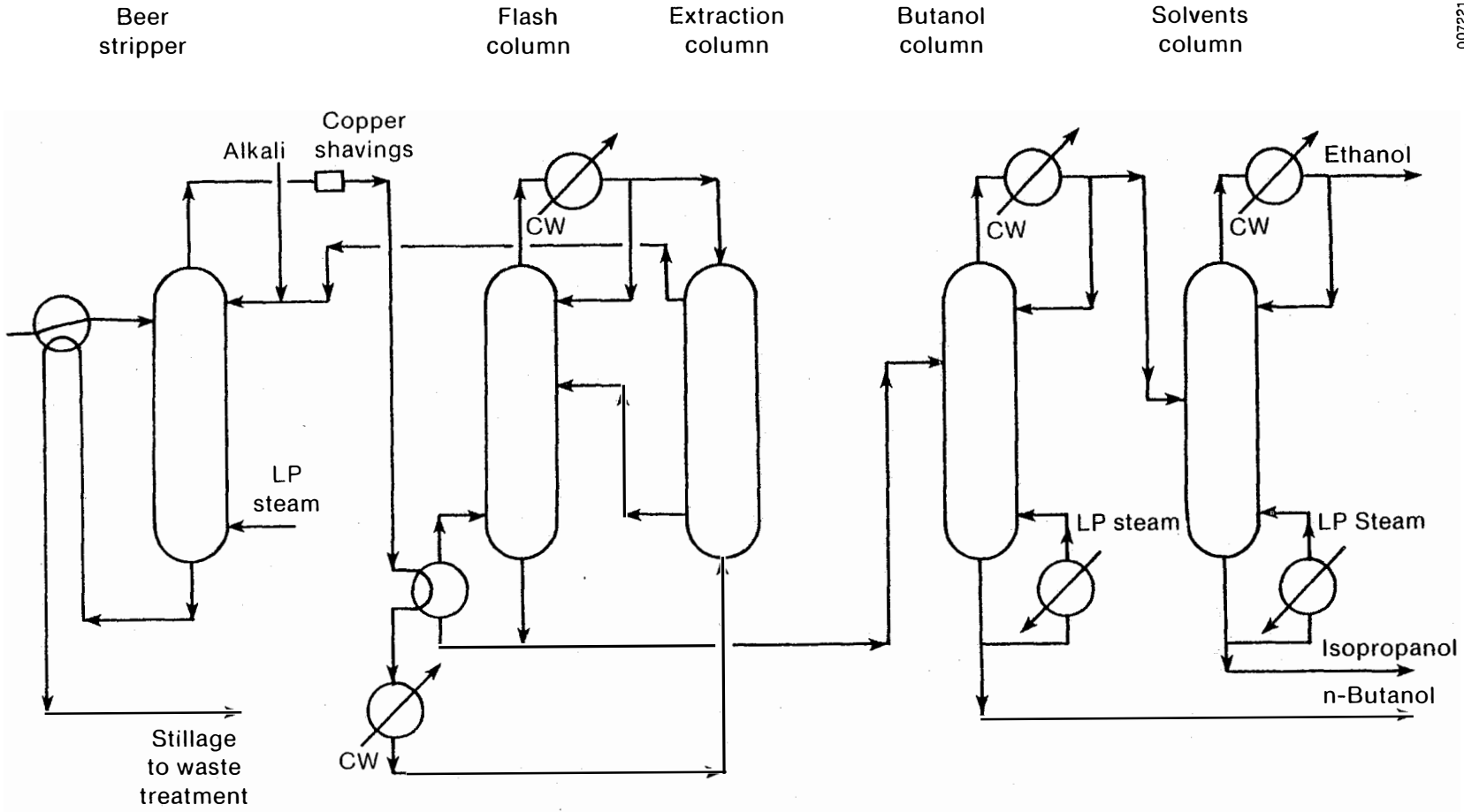


Figure 3. Solvents Separation and Purification System

**Table 1. Capital Cost Analysis**

Unit	Estimated Cost (10 <sup>6</sup> \$)
Feed handling	0.6
Hydrolysis	8.5
Fermentation	12.2
Recovery and purification	4.3
Tankage	5.0
Environmental	6.7
Utilities	26.0
 Total (U.S. Gulf Coast)	 63.3
Engineering overhead plus fee	3.6
Spare parts	1.2
Start-up fees	0.6
Land cost	1.0
Allowance for typical southeastern location	(-)2.7
Estimating contingency	3.7
 Total installed capital cost	 70.7

### Economic Analysis

The after-tax discounted cash flow internal rate of return (DCFIROR) for the base-case process (200,000 dry metric ton/yr feedstock) was calculated to be 8%. The basis of this calculation is given in Table 3. The effect of scale of operation on the DCFIROR was calculated for a range of feedstock capacities from 100,000 to 600,000 dry metric ton/yr. The return appears to be very sensitive to plant capacity. The calculated rate of return varies from near zero at a throughput of 100,000 dry metric ton/yr to more than 20% at 600,000 dry metric ton/yr. The project generates a 16% rate of return at a feedstock capacity of 400,000 dry metric ton/yr. The effect of plant capacity on the DCFIROR is shown in Figure 4a.

Because of the preliminary nature of the process design, considerable uncertainty is associated with the capital cost estimate. To assess the effect of such uncertainty, the DCFIROR was calculated for a range of capital costs around that actually estimated. The estimated capital cost is \$71 million and the range chosen is \$50 million to \$100 million.

The range represents +40% to -30% around the actual estimate. The results of the calculations suggest that the DCFIROR is sensitive to errors in the estimate: At the upper end of the range (\$100 million) the return is almost zero and at the lower end of the range (\$50 million) the return is nearly 14%. The results of this calculation are shown in Figure 4b. Further conclusions from the economic analysis are that the DCFIROR is sensitive to the debt/equity ratio and to the price of feedstock.

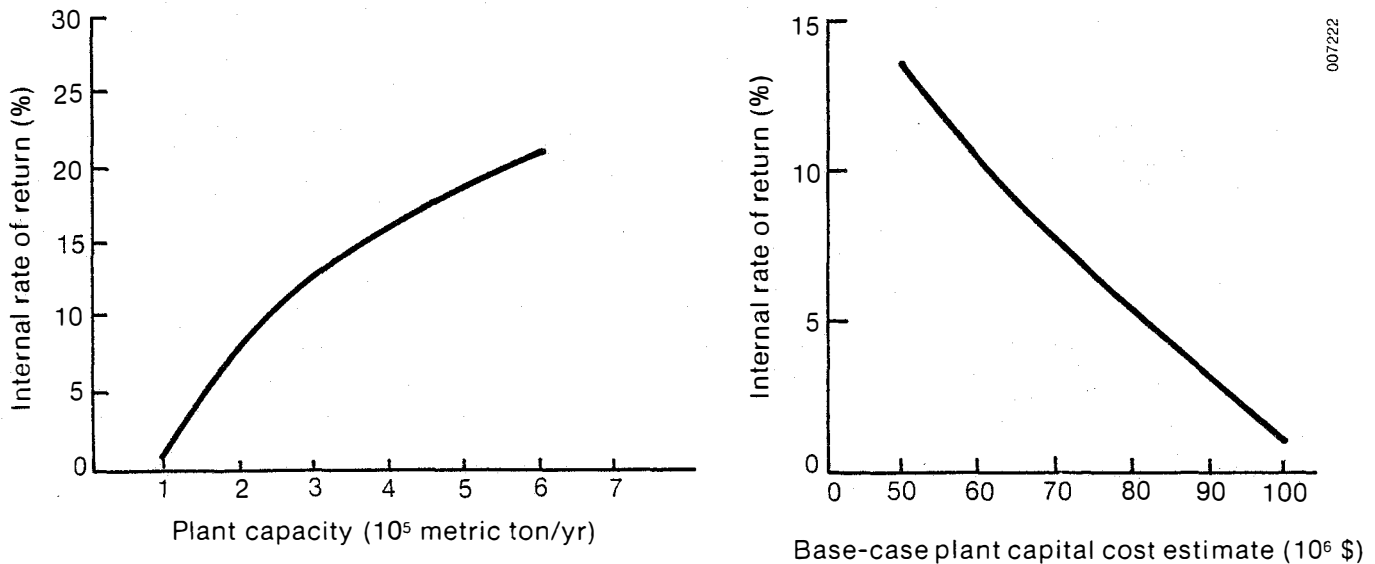
Table 2. Base-Case Revenue and Production Costs

<u>Revenues</u>				
<u>Item</u>	<u>Units</u>	<u>Units/yr</u>	<u>\$/lb</u>	<u>\$/yr</u>
n-Butanol	10 <sup>6</sup> lb	53.4	0.36	19,220,000
i-Propanol	10 <sup>6</sup> lb	26.7	0.31	8,280,000
Ethanol	10 <sup>6</sup> lb	5.34	0.25	1,340,000
Total revenue				28,840,000
<u>Costs</u>				
Raw Materials				
<u>Item</u>	<u>Units</u>	<u>Usage/yr</u>	<u>\$/Unit</u>	<u>\$/yr</u>
Wood chips	metric ton	200,000	\$40	8,000,000
Sulfuric acid	metric ton	24,000	\$65	1,560,000
Lime	metric ton	20,000	\$50	1,000,000
Others				250,000
Total raw materials				10,810,000
Labor				<u>\$/yr</u>
Operators	40 @ \$25,000/yr			1,000,000
Supervisors	4 @ \$30,000/yr			120,000
Total labor				1,120,000
Maintenance (@ 4% capital)				2,800,000
Overheads				
Direct @ 50% labor				560,000
General @ 50% maintenance & labor				1,960,000
Total overheads				2,520,000
Taxes and Insurance (@ 2% of investment)				1,400,000
Landfill Disposal (@ 3.5%/metric ton)				150,000
Total operating cost				17,680,000
Depreciation (@ 10% per year)				7,000,000
Total production cost				24,680,000

**Table 3. Parameters for Economic Analysis**

Feed capacity (base case)	200,000 dry metric ton/yr
Operating factor	8000 h/yr
Inflation rate	0
Base year for dollars	1986
Year of cost information	1986
Construction start date	3rd Q 1986
Investment drawdown	monthly
First commercial operation	2nd Q 1988
Operating lifetime	15 yr
Tax depreciation schedule	ACRS*
Taxes and insurance	2% of total investment
Income tax rate	45%
Investment tax credit	10%
Energy tax credit	0
Multiple or single venture company	Multiple
Ratio of debt to capitalization	0
Escalation rate on capital	0
Escalation rate on operations	0
Operating capacity schedule	50% first 6 months 100% thereafter
Salvage value	10% of total investment

ACRS - Accelerated Cost Recovery System allowed by law over 5 years. Yearly percent write-offs are: 15, 22, 21, 21, 21 (Reference 20).



Note: Base-case plant capacity:  $2 \times 10^5$  metric ton/yr

**Figure 4. Effect on Process Economics of (a) Plant Capacity and (b) Estimated Base-Case Plant Capital Cost**

## Conclusion

In this paper we presented a process that produces butyl solvents, butanol, isopropanol, and ethanol from wood or other lignocellulosic feedstock. Two new elements of technology were introduced: (1) the processing batch reactor developed at SERI that produces high yields of fermentable sugars (hexoses and pentoses) at the appropriate concentration for the butyl solvents fermentation; and (2) a novel method of separating products using liquid-liquid extraction, which reduces the separation energy required to about 30% of energy required in the conventional batch method.

Economic analysis suggests that the project is attractive at a feedstock capacity of 400,000 dry MTA or larger (178 million lb/yr solvents). There are, however, uncertainties associated with the project because of the relatively early stage of development of the key elements of the process technology and the sensitivity of the DCFIROR to estimated capital cost. These uncertainties will be reduced by pilot work being conducted at SERI. Further development of the process is recommended based on the encouraging conclusions of the present study. A further conclusion from this analysis is that the process economics would benefit greatly from reduced capital cost of the fermentation section. This could perhaps be accomplished by developing a continuous fermentation process. Such fermentation technology has been demonstrated on laboratory scale, but as far as is known, has not been developed to pilot scale (16,17,18,19).

## References

1. Badger Engineers Inc., *Economic Feasibility of an Acid Hydrolysis Based Ethanol Plant*, SERI/DOE Subcontract No. ZX-3-030-96-2, Solar Energy Research Institute, Golden, CO: September 1984.
2. Manolakis, E., "Butanol from Waste Agricultural Products: A Business Plan," R.F. Webb Corporation, Ontario, Canada; for Agriculture Canada, AERD Program File No. 34SZ.01916-3-EL28.
3. Schoutens, G.H., et al., *Process Biochemistry*, February 1986, p. 30.
4. Wright, J.D., et al., ASC Annual Meeting, DOE Subcontract No. DE-AC02-83CH 10093, September 1985.
5. Wenzl, H.F.J., *The Chemical Technology of Wood*, NY: Academic Press, 1970, pp. 202-210.
6. Harris, E.E., and E. Beglinger, *Ind. Eng. Chem.*, Vol. 38, September 1945, pp. 890-895.
7. Gilbert, N., et al., *Ind. Eng. Chem.*, Vol. 44, July 1952, pp. 1712-1720.
8. Mendelsohn, H.R., et al., *Chem. Eng.*, 15 June 1981, pp. 61-65.
9. Burton, R.J., "The New Zealand Wood Hydrolysis Process," *Proc. of Ethanol from Biomass Conf.*, Winnipeg, Canada, 1982.
10. Korolov, I.I., et al., "Percolation Hydrolysis Using Alternating Liquid Flows," *Hydrolytic and Wood Chemical Industry*, No. 2, (In Russian) 1961.
11. Prescott, S.C., and C.G. Dunn, *Industrial Microbiology*, 3rd Ed., NY: McGraw Hill, 1959.
12. Spivey, M.J., *Process Biochemistry*, Vol. 13, No. 11, November 1978.
13. Cowley, R.S., et al., 5th International Alcohol Fuels Conference, Auckland, New Zealand, 13-18 May 1982.
14. Besch, S.C., *Ind. Eng. Chem.*, Vol. 44, No. 7, 1952.
15. Fouad, M., et al., *Acta. Biol. Acad. Sci. Hungarian*, Vol. 27 (2-3), 1976, p. 107.
16. *Chemical Marketing Reporter*, March 23, 1986.
17. Wang, D.I.C., et al., "Production of Acetone/Butanol by *Clostridium acetobutylicum* in Batch and Continuous Cultures," 72nd Annual AIChE Meeting, San Francisco, CA, November 1979.
18. Kolot, F.B., *Process Biochemistry*, Vol. 19, No. 1, February 1984.
19. Bullock, J.D., et al., eds., "The Acetone-Butanol Fermentation and Related Topics 1980-1983," published by Science and Technology Letters (*Biotechnology Letters*) 1983.

20. Marlatt, J.A., et al., *Biotechnology Progress*, Vol. 2, No. 1, 1986, p. 23.
21. U.S. Master Tax Guide - Sections 1165D; 1178A.