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TECHNICAL AND ECONOMIC ASSESSMENT OF PROCESSES FOR THE PRODUCTION OF BUTANOL AND ACETONE—PHASE TWO: ANALYSIS OF RESEARCH ADVANCES

August 1984

Work Performed Under Contract No. AI01-81CS66001

Chem Systems, Inc.
Tarrytown, New York

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ENERGY CONVERSION
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TECHNOLOGIES PROGRAM

Technical and Economic Assessment of Processes for the Production of Butanol and Acetone — Phase Two: Analysis of Research Advances

(Chem Systems, Inc.)

August 1984

Sponsored by:
Energy Conversion and Utilization Technologies Division
Office of Energy Systems Research
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Through an Agreement with
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Prepared for:
Biocatalysis Research Project
Jet Propulsion Laboratory
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PREFACE

Phase Two: Analysis of Research Advances

This is the Final Report of a study effort performed for the United States Department of Energy/Energy Conversion Utilization Technologies Office (DOE/ECUT). The work was performed by Chem Systems Inc. under Contract No. 956277 with the Jet Propulsion Laboratory of the California Institute of Technologies, sponsored by The National Aeronautics and Space Administration under Contract NAS7-918.

ACKNOWLEDGEMENT

This report was performed as part of the Department of Energy's Energy Conversion and Utilization Technologies (ECUT) Program, under the direction of the Jet Propulsion Laboratory.

The authors gratefully acknowledge the cooperation of DOE and JPL staff members and, in particular, the guidance provided by Dr. J. Eberhardt of DOE and Dr. J. Moacanin of JPL.

ABSTRACT

The initial objective of this work was to develop a methodology for analyzing the impact of technological advances as a tool to help establish priorities for R&D options in the field of biocatalysis. As an example of a biocatalyzed process, butanol/acetone fermentation (ABE process) was selected as the specific topic of study. A base case model characterizing the technology and economics associated with the ABE process was developed in the previous first phase of study.

The project objectives were broadened in this second phase of work to provide parametric estimates of the economic and energy impacts of a variety of research advances in the hydrolysis, fermentation and purification sections of the process. The research advances analyzed in this study were based on a comprehensive literature review.

The six process options analyzed were:

- Continuous ABE fermentation
- Vacuum ABE fermentation
- Baelene solvent extraction
- HRI's Lignol process
- Improved prehydrolysis/dual enzyme hydrolysis
- Improved microorganism tolerance to butanol toxicity

Of the six options analyzed, only improved microorganism tolerance to butanol toxicity had a significant positive effect on energy efficiency and economics. This particular process option reduced the base case production cost (including 10% DCF return) by 20% and energy consumption by 16%.

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I. EXECUTIVE SUMMARY

The initial objective of this work was to develop a methodology for analyzing the impact of technological advances as a tool to help establish priorities for R&D options in the field of biocatalysis. As an example of a biocatalyzed process, butanol/acetone fermentation (ABE process) was selected as the specific topic of study. A base case model characterizing the technology and economics associated with the ABE process was developed in the previous first phase of study.(1)

The project objectives were broadened in this second phase of work to provide parametric estimates of the economic and energy impacts of a variety of research advances in the hydrolysis, fermentation and purification sections of the process. The research advances analyzed in this study were based on a comprehensive literature review, and the criteria employed in the selection process included information availability, technical feasibility, energy consumpton and economics. These advances were analyzed individually as well as in selected combinations in order to assess their overall impact relative to the base case. In addition, a hypothetical "best case", combining the best elements of each process improvement, was constructed for the overall production process.

The six process options analyzed were as follows:

- Continuous ABE fermentation
- Vacuum ABE fermentation
- Baelene solvent extraction

(1) "Technical and Economic Assessment of Processes for the Production of Butanol and Acetone", prepared by Chem Systems Inc. for Jet Propulsion Laboratory, sponsored through an agreement with NASA by Energy Conversion and Utilization Technologies Division, Office of Energy Systems Research, Dept. of Energy, September 1982 (JPL 9950-776).

- HRI's Lignol process.
- Improved prehydrolysis/dual enzyme hydrolysis
- Improved microorganism tolerance to butanol toxicity

Of the six options analyzed, four resulted in improved process economics. One of these, improved microorganism tolerance to butanol toxicity, had a significant positive effect. As indicated in Table I-1, this particular process option reduced the base case production cost (including 10% DCF return) from $2.60 per gallon to $2.09 per gallon. The others had marginally positive effects, with resulting production costs in the range of $2.49 to $2.60 per gallon. Insufficient data were available to adequately analyze the vacuum fermentation option, and this was not characterized quantitatively.

In the Phase I analysis, it was determined that the fermentation route to butanol and acetone resulted in energy consumption of about 5.4 trillion BTU for a 50 million gallon per year plant, a potential energy savings of nearly 40 percent relative to conventional methods of production. Each of the process options was analyzed herein to determine its energy consumption level relative to the base case. In the case of continuous fermentation, no additional energy savings were found (in fact, a small increase in energy requirements would result). The Lignol process offered savings of about 39 percent relative to the base case, due in large part to the energy credit accruing from by-product phenol and benzene. The Baelene process would result in reduced energy consumption of about 6 percent, and the dual enzyme system about 2 percent, relative to the base case. The savings which would accrue to increasing the butanol tolerance level total about 16 percent relative to the base case.

The factor which contributes most towards the poor economics of ABE fermentation compared to the conventional route in the toxic effect of butanol on microorganism activity, which limits product concentration

TABLE I-1

SUMMARY OF IMPACTS OF INDIVIDUAL PROCESS OPTIONS ON ABE PROCESS ECONOMICS

	Base Case	Continuous Fermentation	Lignol	Baelene	Dual Enzyme	Improved Butanol Tolerance
Investment, $MM						
Battery limits	92.8	88.5	135.6	102.4	86.1	75.6
Offsites	97.3	98.8	89.8	89.3	92.4	83.7
Total fixed investment	190.1	187.3	225.4	191.7	178.5	159.3
Cost of production, ¢/gal						
Raw materials	86.59	90.13	104.01	86.59	84.57	66.99
Utilities	43.37	43.79	36.88	39.09	43.39	35.08
Operating costs	15.16	13.70	20.29	16.31	14.36	13.10
Overhead expenses	17.37	15.91	21.76	18.16	16.50	15.10
By-product credit	(23.20)	(22.85)	(71.58)	(23.20)	(23.20)	(22.20)
Cash cost of production	139.28	140.67	111.37	136.95	135.61	108.07
Depreciation	56.58	55.16	72.20	58.82	52.92	46.98
Net cost of production	195.85	195.82	183.57	195.76	188.53	155.05
Selling price at 10% DCF	259.8	259.5	249.4	258.0	249.3	208.5

during fermentation. This analysis examined the sensitivity effect of higher solvent concentrations on process economics as well as on overall energy consumption. Three solvent base concentrations were studied: 1.0 percent (0.7 percent butanol), 2.1 percent (1.2 percent butanol), and 2.9 percent (1.7 percent butanol).

As the solvent concentration was increased, several relationships occur which contribute to decreasing the product cost: (1) battery limits capital cost is reduced due to the decreasing water content of the process streams, which results in reduced equipment volume requirements; (2) the corresponding reduction in fermenter volume results in reduced nutrient requirements; (3) the higher solvent concentration results in decreased steam requirements for purification.

Overall production costs (including 10 percent DCF return) were calculated to be $2.69 per gallon for the lowest case, $2.14 per gallon for the middle case, and $1.95 per gallon for the highest solvents case. Energy consumption levels were also impressive, exhibiting decreases of 16 percent and 33 percent for the latter two cases relative to the base case.

An assessment was also made of a "cumulative" case, incorporating the effects of continuous fermentation, Lignol and dual enzyme hydrolysis. This case yielded a net production cost of $2.38 per gallon, representing a reduction of about 22 cents per gallon relative to the base case. This is approximately equal to the sum of the individual reductions in production cost for each of the individual cases. A significant decrease in energy consumption of about 35 percent relative to the base case characterized the "cumulative" case.

The case which would offer the best economics for ABE fermentation would be a combinaton of all the potential research improvements incorporated into one design. This would include continuous fermentation, Lignol and dual enzyme hydrolysis processing options as well as improved microorganism tolerance to butanol toxicity, bringing the solvent concentration in the beer to 2.9 weight percent. This level of 2.9 percent solvents is arbitrarily chosen as the maximum realistic concentration obtainable in the near future ("best case").

The "best case" results in a production cost of $1.84 per gallon, which represents approximately a 20 percent price advantage over the conventional synthetic route. The associated energy requirement is calculated to be 68 percent lower than the base case and may be thought of as a "target" in terms of energy efficiency improvement.

Table I-2 summarizes required selling prices (at 10 percent DCF) and energy consumption levels for each of the options and combinations analyzed. From an economic standpoint, the impact of improved enzyme tolerance to butanol toxicity is clearly the most significant potential improvement. This particular option represents the bulk of the benefit indicated by the "best" case. Similarly, the impact of improved butanol tolerance is clearly manifested in a significantly reduced energy consumption level. This improvement, when coupled with the large reduction in net energy requirements accruing to the Lignol process (largely as a result of by-product credits resulting from production of phenol and benzene), yields a "best case" energy requirement of less than 35,000 Btu per gallon of product. This is about one-third of the energy required in the base case design.

The economic impact of alternative feedstock compositions for ABE production was examined, including a revised aspen composition, eucalyptus and corn stover. The revised aspen case produced mixed solvents for about $2.70 per gallon, which is slightly higher than the original aspen case because of the decreased sugar potential of the revised composition. The eucalyptus case resulted in a production cost of $2.82 per gallon, also because of the lower potential sugar content of eucalyptus and, to a lesser extent, higher capital-related costs. The

TABLE I-2

SUMMARY OF REQUIRED SELLING PRICES AND ENERGY CONSUMPTION LEVELS FOR OPTIONS ANALYZED

	Base Case	Continuous Fermentation	Lignol	Baelene	Dual Enzyme	Improved Butanol Tolerance	Cumulative	Best
Required selling price, ¢/gal	259.8	259.5	249.4	258.0	249.3	208.5	237.5	183.6
Energy consumption, MBtu/gal of product	107.4	110.4	65.4	100.4	105.6	89.8	69.4	34.6

corn stover case turned out to be quite promising, with production costs of $2.35 per gallon. This is by virtue of the fact that corn stover is field dried and contains only about 30 percent water (compared to 50 percent for the wood cases). However, corn stover may be an unsuitable year-round feedstock since it cannot be stored any length of time because of its sugar content.

Finally, the overall methodology developed for analyzing the ABE system was applied to the production of citric acid and furfural from wood-derived sugars, as an alternative example. A detailed process design was developed for this system and production economics were developed based on aspen feedstock. Various sensitivities were also explored as part of this case analysis.

II. INTRODUCTION

This report was prepared as part of the Department of Energy's Energy Conversion and Utilization Technologies (ECUT) Program, whose objective is to support long-term, high-risk applied research and development necessary to assure the availability of a future technology base that will enable a substantial increase in both the efficiency of energy conversion and utilization equipment and the increased use of non-critical fuels. It forms a segment of the Biocatalysis Research Project of the Energy Utilization Technology Sub-Program, which focuses on the engineering of biocatalyzed processes for producing chemicals.

The initial objective of the work was to develop a methodology for analyzing the impact of technological advances as a tool to help establish priorities for R&D options. As an example of a biocatalyzed process, butanol/acetone fermentation (ABE process) was selected as the specific topic of study. For ease of comparison with conventional production plants, a Gulf Coast location was hypothesized. Process economics were based on a size of 50 million gallons per year since this was deemed to be a reasonable size plant for producing alcohols for the fuel market. A final report on this first phase of study was published in September, 1982 as report number JPL 9950-776.(1)

The project objectives were broadened in this second phase of work to provide parametric estimates of the economic and energy impacts of a variety of research advances in the hydrolysis, fermentation and purification sections of the process. The research advances analyzed in this study were selected on the basis of a comprehensive literature review.(2) The criteria used in the selection process included information availability, technical feasibility, energy consumption and economics.

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(1) "Technical and Economic Assessment of Processes for the Production of Butanol and Acetone," prepared by Chem Systems Inc. for Jet Propulsion Laboratory, sponsored through an agreement with NASA by Energy Conversion and Utilization Technologies Division, Office of Energy Systems Research, U.S. Department of Energy, September, 1982 (JPL 9950-776).

(2) "Review of Literature Relevant to ABE Fermentation," prepared by Chem Systems Inc. for Jet Propulsion Laboratory, November 16, 1982.

It is intended that the generic nature of this work will enable the parametric analysis to be applicable not only to butanol/acetone production, but also to a wide range of chemicals which can be produced via similar processing routes. As an example, the methodology is applied to a specific case study as part of the analysis.

III. DETAILED REVIEW OF RESEARCH ADVANCES

A. Revised Base Case

The base case economics developed in Phase I have been revised slightly to reflect more realistic process conditions. Tables III-A-1 and III-A-2 are cost of production estimates for the lower yield base case process, Table III-A-1 using CSI utility costs and Table III-A-2 reflecting DOE utility costs. These data are summarized in Table III-A-3. The modifications from the Phase I work are relatively minor; however, they do reflect a more realistic case than the Phase I estimate. The yield on total sugar has been lowered from an optimistic 30.5 percent to a more realistic 27.54 percent. Utility prices used in the Phase I estimate were projections; utilities have now been updated to reflect historical prices for mid-1982 and are lower than the previous projections. Labor requirements have been increased from 46 to 60 men to reflect the greater complexity of operating a batch fermentation versus a continuous one.

The overall change is slight; using CSI utility numbers the revised low yield case gives a selling price at 10 percent DCF of 2.60 dollars per gallon compared to 2.58 dollars per gallon reported in the Phase I study. These revised figures are used as a basis for comparison throughout this study.

In terms of energy consumption, the figures developed in the Phase I analysis are still valid. The key benchmark is the total energy requirement for the lower yield case, which amounts to 5.37 trillion Btu (assuming a power plant heat rate of 10,000 Btu/kwh and a steam generation efficiency of 85 percent).

TABLE III-A-1

COST OF PRODUCTION ESTIMATE FOR ABE
PROCESS- LOW YIELD

CAPITAL SUMMARY

BASIS		CAPITAL COST	$MILLION
Location:	U.S. Gulf Coast	Battery Limits	92.8
	Mid-1982	Offsites	97.3
Capacity:	50.00 million gallons/yr		------
	22,680 metric tons/yr	Total Fixed Inv.	190.1
Str.Time:	8000 hours per year	Working Capital	16.2

PRODUCTION COST SUMMARY

RAW MATERIALS	UNITS PER GAL	PRICE, ¢/UNIT	ANNUAL COST, $M	CENTS PER GAL	DOLLARS/ MET TON
Aspen, lb	61.80965	1.0	30,906		
Sulfuric Acid, lb	.27790	4.3	598		
Calcium Hydroxide, lb	.19798	2.0	198		
Sodium Hydroxide, lb	.00880	26.0	114		
Corn, lb	.00850	4.5	19		
Ammonium Sulfate, lb	.44400	3.0	666		
Superphosphate(46), lb	1.91375	8.0	7,655		
Calcium Carbonate, lb	.88034	2.7	1,189		
Catalyst & Chemicals			1,950		

TOTAL RAW MATERIALS			43,295	86.59	1908.99
UTILITIES					
Power, kWH	1.78069	3.2	2,849		
Cooling Water, M Gal	.30312	5.8	879		
Process Water, M Gal	.03304	60.0	991		
Steam,50 psig, M Lb	.07363	391.0	14,395		
Steam,200 psig, M Lb	.01298	396.0	2,570		

TOTAL UTILITIES			21,684	43.37	956.10
OPERATING COSTS					
Labor, 60 Men @ $ 25,500		13 M/S	1,530		
Foremen, 13 Men @ $ 29,000		2 M/S	377		
Supervision, 3 Man @ $ 35,000		3 Man	105		
Maint., Material & Labor		6% of ISBL	5,568		

TOTAL OPERATING COST			7,580	15.16	334.22
OVERHEAD EXPENSES					
Direct Overhead	45% Lab. & Sup.		905		
Gen. Plant Overhead	65% Oper. Costs		4,927		
Insurance, Prop. Tax	1.5% Tot. Fix. Inv.		2,851		

TOTAL OVERHEAD EXPENSES			8,684	17.37	382.89
BY-PRODUCT CREDIT					
Carbon Dioxide, lb	8.12319	2.8	-11,373		
SCP, lb	.03051	15.0	-229		

TOTAL BY-PRODUCT CREDIT			-11,602	-23.20	-511.55
			==========	=======	==========
CASH COST OF PRODUCTION			69,642	139.28	3070.64
DEPRECIATION	20% ISBL + 10% OSBL		28,290	56.58	1247.36
			==========	=======	==========
NET COST OF PRODUCTION			97,932	195.85	4318.01
REQUIRED SALES PRICE AT 10% DCF				259.8	5727.2

TABLE III-A-2

COST OF PRODUCTION ESTIMATE FOR ABE PROCESS- LOW YIELD

CAPITAL SUMMARY

BASIS		CAPITAL COST	$MILLION
Location:	U.S. Gulf Coast	Battery Limits	92.8
	Mid-1982	Offsites	97.3
Capacity:	50.00 million gallons/yr		------
	22,680 metric tons/yr	Total Fixed Inv.	190.1
Str.Time:	8000 hours per year	Working Capital	15.7

PRODUCTION COST SUMMARY

RAW MATERIALS	UNITS PER GAL	PRICE, ¢/UNIT	ANNUAL COST, $M	CENTS PER GAL	DOLLARS/ MET TON
Aspen, lb	61.80965	1.0	30,906		
Sulfuric Acid, lb	.27790	4.3	598		
Calcium Hydroxide, lb	.19798	2.0	198		
Sodium Hydroxide, lb	.00880	26.0	114		
Corn, lb	.00850	4.5	19		
Ammonium Sulfate, lb	.44400	3.0	666		
Superphosphate(46), lb	1.91375	8.0	7,655		
Calcium Carbonate, lb	.88034	2.7	1,189		
Catalyst & Chemicals			1,950		

TOTAL RAW MATERIALS			43,295	86.59	1908.99
UTILITIES					
Power, kWH	1.78069	4.3	3,829		
Cooling Water, M Gal	.30312	5.8	879		
Process Water, M Gal	.03304	60.0	991		
Steam,50 psig, M Lb	.07363	295.0	10,860		
Steam,200 psig, M Lb	.01298	299.0	1,941		

TOTAL UTILITIES			18,500	37.00	815.70
OPERATING COSTS					
Labor, 60 Men @ $ 25,500		13 M/S	1,530		
Foremen, 13 Men @ $ 29,000		2 M/S	377		
Supervision, 3 Man @ $ 35,000		3 Man	105		
Maint., Material & Labor		6% of ISBL	5,568		

TOTAL OPERATING COST			7,580	15.16	334.22
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Direct Overhead	45% Lab. & Sup.		905		
Gen. Plant Overhead	65% Oper. Costs		4,927		
Insurance, Prop. Tax	1.5% Tot. Fix. Inv.		2,851		

TOTAL OVERHEAD EXPENSES			8,684	17.37	382.89
BY-PRODUCT CREDIT					
Carbon Dioxide, lb	3.12319	2.8	-11,373		
SCP, lb	.03051	15.0	-229		

TOTAL BY-PRODUCT CREDIT			-11,602	-23.20	-511.55
			==========	=======	==========
CASH COST OF PRODUCTION			66,457	132.91	2930.24
DEPRECIATION	20% ISBL + 10% OSBL		28,290	56.58	1247.36
			==========	=======	==========
NET COST OF PRODUCTION			94,747	189.48	4177.60
REQUIRED SALES PRICE AT 10% DCF				252.8	5573.8

TABLE III-A-3

SUMMARY OF PROCESS ECONOMICS AND ENERGY REQUIREMENTS FOR BASE CASE LOW YIELD ABE FERMENTATION

Basis: 50MM gal/year, U.S. Gulf Coast/Louisiana
Mid-1982

	CSI Utilities	DOE Utilities
Investment, $MM		
Battery limits	92.8	92.8
Offsites	97.3	97.3
Total fixed investment	190.1	190.1
Cost of production, ¢/gal		
Raw materials	86.59	86.59
Utilities	43.37	37.00
Operating costs	15.16	15.16
Overhead expenses	17.37	17.37
By-product credit	(23.20)	(23.20)
Cash cost of production	139.28	132.91
Depreciation	56.58	56.58
Net cost of production	195.85	189.48
Selling price at 10% DCF	259.8	252.8
Energy required, MBtu/gal of product	107.4	107.4

B. Vacuum Fermentation

Figure III-B-1 is a conceptual flowsheet of a vacuum fermentation system. Vacuum fermentation is a technique, which, when applied in ethanol fermentation, has several advantages over conventional continuous fermentation. Vacuum fermentation operates on the principle that the volatile components formed during fermentation can be immediately boiled off in greater concentrations in the vapor phase compared to the remaining liquid phase when a vacuum is applied. This results in a reduction in end product (in this case ethanol) inhibition in the fermentation liquor, resulting in much greater fermentation volumetric productivities when compared to conventional continuous fermentation. In addition, this enables the vacuum fermentation system to be able to efficiently ferment very high sugar concentrations (up to 33 percent)

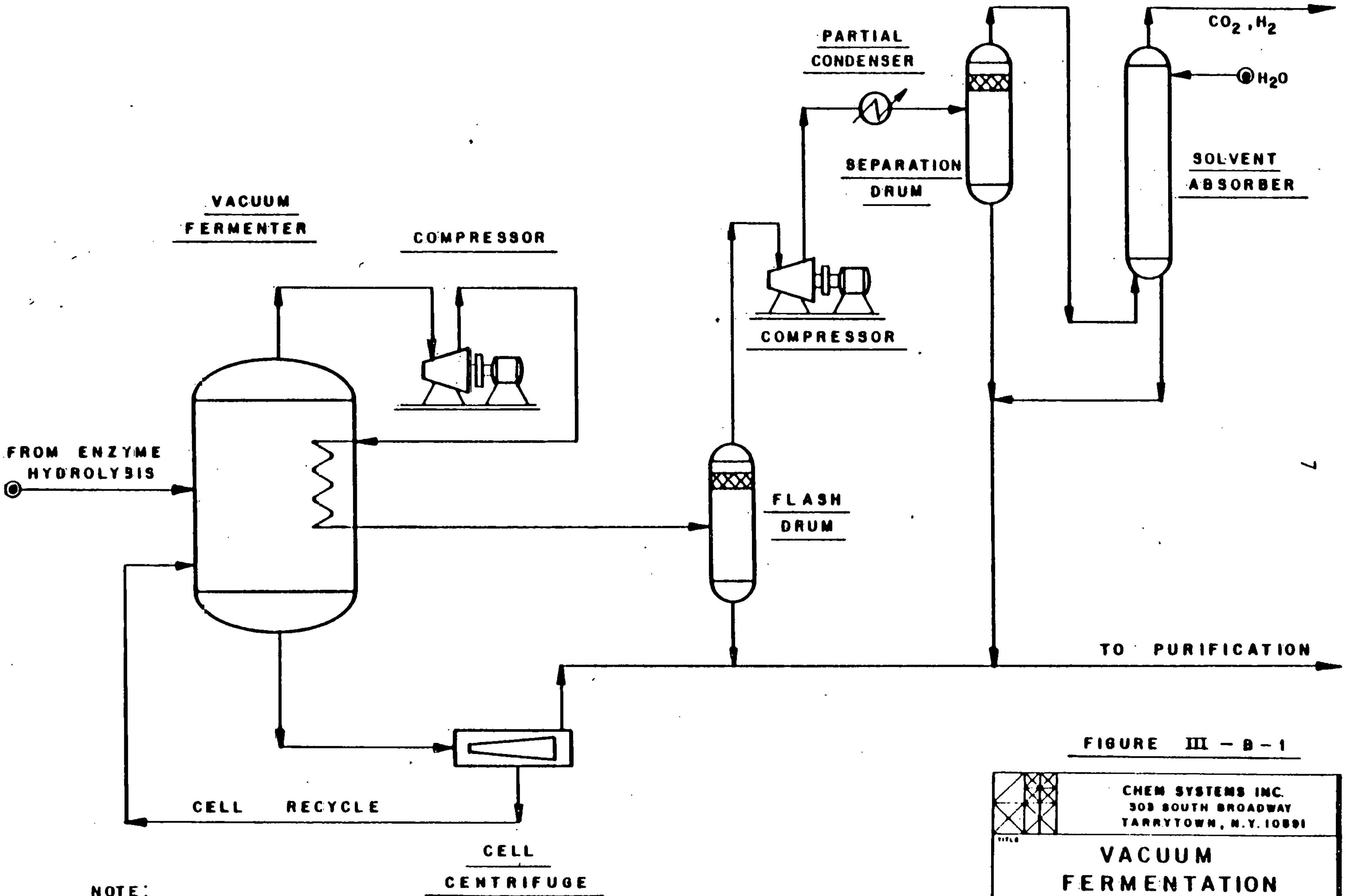

FIGURE III - B - 1

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VACUUM
FERMENTATION

NO 6003

NOTE:
ALL FLOWS IN LBS./HR.

without end product inhibition, thus resulting in significantly reduced distillation energy requirements due to the reduction in water in the system.

The ABE-water system forms two low boiling azeotropes on flashing. The first, ethanol-water, is not very important because the concentration of ethanol is so small in the liquor that it can be considered negligible. The second azeotrope, butanol-water, boils at a lower temperature than water, theoretically resulting in butanol enrichment in the vapor phase compared to the fermentation liquor. The available butanol-water azeotropic data are summarized in Table III-B-1 .

TABLE III-B-1

n-BUTANOL-WATER AZEOTROPIC DATA

Pressure mm Hg	Wt % H_2O	Wt % Butanol	BP, ^oC
100	49.8	50.2	48.0
270	46.6	53.7	70.0
755	42.8	57.2	92.4

Therefore at ABE fermentation conditions of 33^oC and about 38 mm Hg, the azeotropic composition is between 45-50 weight percent butanol. However, a search of the literature revealed no meaningful vapor pressure data for the butanol-water azeotrope, indicating that little or no research has or is being performed on ABE vacuum fermentation. This fact also prevents even a speculative analysis from being undertaken. However, it is known from preliminary experiments performed at JPL, that condensate from a 1 percent n-butanol solution at 33^oC produces aqueous n-butanol that is substantially enriched in butanol.

Although the exact benefit of ABE vacuum fermentation is unknown at present, the potential benefit may be significant. The enrichment of butanol in the vapor phase will enable fermentation of much higher sugar concentrations than are possible under conventional non-vacuum conditions. This will reduce the water content of the fermentation beer, thus reducing the energy requirements for purification. The extent of this reduction remains to be seen and depends on the exact vapor-liquid equilibrium concentrations occurring during the conditions of vacuum fermentation.

C. Continuous Fermentation

Design Basis

The conditions for continuous fermentation are based upon data obtained by Leung and Wang published in two recent papers [1,2].

Leung obtained maximum volumetric productivity of 2.5 g/l-hour at a fermentation residence time of approximately 5 hours. This represents over a 300 percent increase compared to the productivity obtained for a batch fermentation. Continuous fermentation process parameters are summarized in Table III-C-1.

TABLE III-C-1

CONTINUOUS FERMENTATION PARAMETERS

Temperature	37°C
Sugar concentration	5 weight percent
pH	5.0
Cell mass concentration	4.5-5 g/liter
Residence time	5 hours
Volumetric productivity	2.5 g/l-hr

(1) Leung, J.C.Y., and Wang, D.I.C., "Production of Acetone and Butanol by Clostridium Acetobutylicum in Continuous Culture Using Free Cells and Immobilized Cells," 2nd World Congress of Chemical Engineering and World Chemicals, Montreal, Oct., 1981.

(2) Leung, J.C.Y., and Wang, D.I.C., "Production of Acetone/Butanol by Cl. Acetobutylicum in Batch and Continuous Cultures, 72nd Annual AIChE Meeting, San Francisco, Nov., 1979.

The yield of solvents produced, sugar utilized and solvent product concentration are different than for the conventional batch base case. Product yield is 26.3 weight percent solvents produced based upon total sugar charged. Sugar utilization is approximately 85 percent. The end product slate is presented on a weight percent basis in Table III-C-2.

TABLE III-C-2

END PRODUCT DISTRIBUTION

	Weight Percent
Butanol	57.0
Acetone	35.0
Ethanol	8.0
	100.0

It has been assumed that the nutrient requirements for continuous fermentation are identical in concentration to those used in the base batch case. These are summarized in Table III-C-3.

TABLE III-C-3

FERMENTATION NUTRIENT REQUIREMENTS

Nutrient	Weight Percent
$(NH_4)_2SO_4$	0.1
Superphosphate	0.2
$CaCO_3$	0.2

Other trace chemicals such as $FeSO_4 \cdot 7H_2O$, and $MnSO_4 \cdot 3H_2O$ may have to be added; however, in the small quantities required they will have a negligible effect upon economics. C. Acetobutylicum culture growth and maintenance nutrient requirements are met with 5 weight percent corn mash medium as in the base case.

Process Design

A conceptualized flowsheet for the continuous fermentation section is presented in Figure III-C-1.

The continuous fermentation scheme is similar to the batch fermentation, excepting the fact that the fermenters are operated continuously. The C. Acetobutylicum inoculum is prepared in the same manner as in the batch case, except that the inoculum to the fermenters is fed continuously from the seed tanks. The first, second and third generation growth tanks can be operated batchwise as before. However, sufficient inoculum storage must be available in the seed tanks in order to operate the fermenters continuously. The fermenters are operated continuously in a concept similar to the continuous cascade system used in continuous ethanol fermentation. Three cascade trains containing two tanks each are utilized. Each train has a 50 percent capacity, with partial fermentation occurring in each tank until complete fermentation is realized in the last tank. In this way two trains are active at any one time, with the third train down for sterilization. Fermentation time is approximately 5 hours. The CO_2/H_2 fermenter off-gas is removed continuously to recovery as before. Although not considered here, the fermenter beer could be centrifuged to remove the C. Acetobutylicum cells to be recycled to the fermenters as is done with yeast cells in continuous ethanol fermentation. The rest of the continuous fermentation proceeds the same as the batch system.

Economics and Energy Requirements

Cost of production estimates for the incorporation of the continuous fermentation block into the base case are presented in Tables III-C-4 and III-C-5. Table III-C-4 shows the process economics using CSI-developed utility numbers, and Table III-C-5 utilizes DOE-derived utility numbers on a 1981 basis. Both cases are on a mid-1982 basis, for a plant located in Louisiana/U.S. Gulf Coast producing 50 million gallons per year of mixed solvents. The cost of production data are summarized and compared against the (revised) base case figures in Table III-C-6. An ISBL investment breakdown is provided in Table III-C-7.

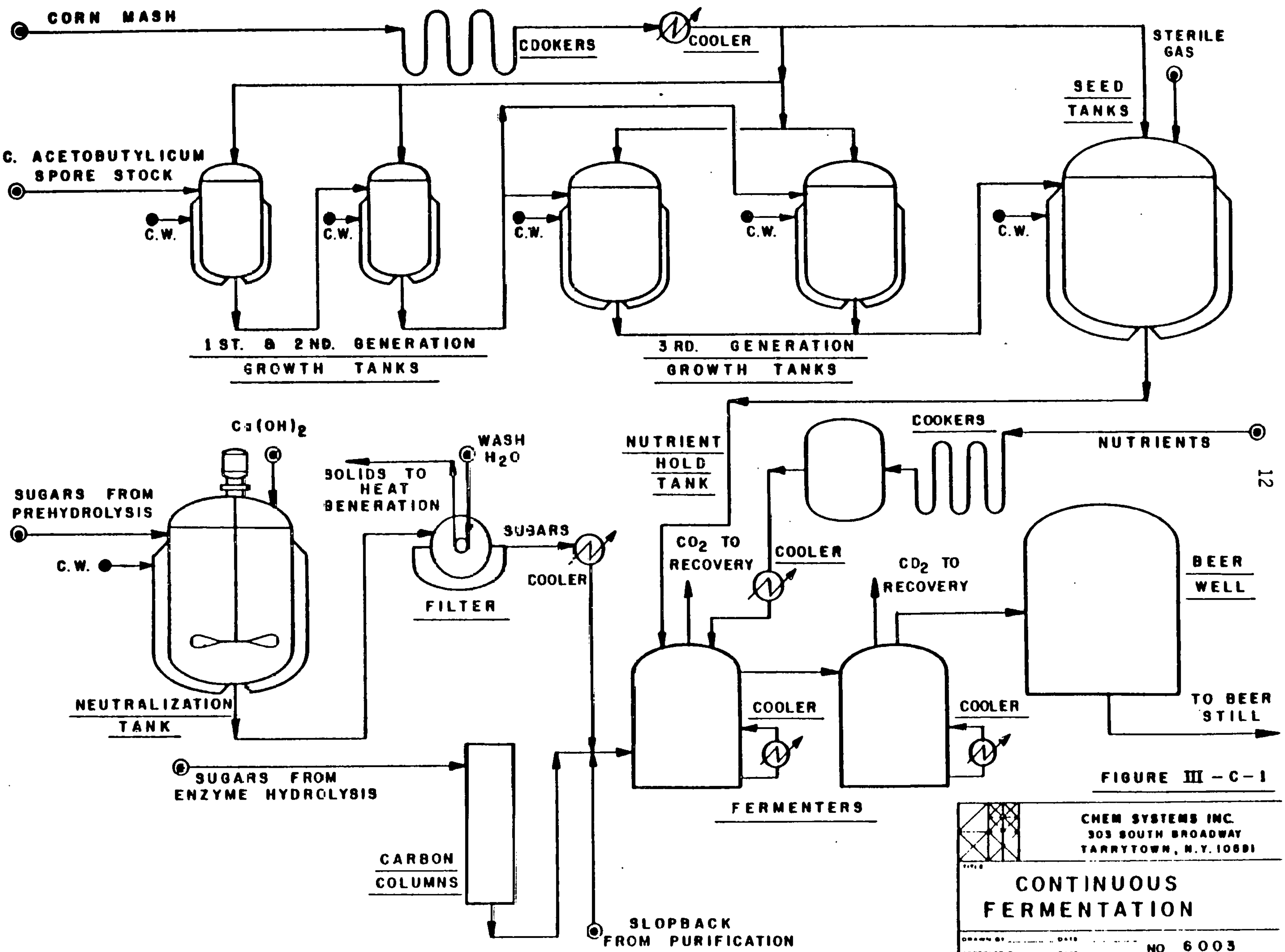

FIGURE III – C – 1

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CONTINUOUS FERMENTATION

NO 6003

TABLE III-C-4

COST OF PRODUCTION ESTIMATE FOR ABE PROCESS- CONTINUOUS FERMENT

CAPITAL SUMMARY

BASIS		CAPITAL COST	$MILLION
Location:	U.S. Gulf Coast	Battery Limits	88.5
	Mid-1982	Offsites	98.8
Capacity:	50.00 million gallons/yr		------
	22,680 metric tons/yr	Total Fixed Inv.	187.3
Str.Time:	8000 hours per year	Working Capital	15.9

PRODUCTION COST SUMMARY

RAW MATERIALS	UNITS PER GAL	PRICE, ¢/UNIT	ANNUAL COST, $M	CENTS PER GAL	DOLLARS/ MET TON
Aspen, lb	64.70349	1.0	32,353		
Sulfuric Acid, lb	.29058	4.3	625		
Calcium Hydroxide, lb	.18529	2.0	185		
Sodium Hydroxide, lb	.00930	26.0	121		
Corn, lb	.00892	4.5	20		
Ammonium Sulfate, lb	.45208	3.0	678		
Superphosphate(46), lb	1.96554	8.0	7,863		
Calcium Carbonate, lb	.90415	2.7	1,221		
Catalyst & Chemicals			2,000		

TOTAL RAW MATERIALS			45,066	90.13	1987.05
UTILITIES					
Power, kWH	1.75690	3.2	2,811		
Cooling Water, M Gal	.33955	5.8	985		
Process Water, M Gal	.03304	60.0	991		
Steam,50 psig, M Lb	.07702	391.0	15,057		
Steam,200 psig, M Lb	.01036	396.0	2,051		

TOTAL UTILITIES			21,896	43.79	965.43
OPERATING COSTS					
Labor, 46 Men @ $ 25,500		10 M/S	1,173		
Foremen, 9 Men @ $ 29,000		1 M/S	261		
Supervision, 3 Man @ $ 35,000		3 Man	105		
Maint., Material & Labor	6% of ISBL		5,310		

TOTAL OPERATING COST			6,849	13.70	301.99
OVERHEAD EXPENSES					
Direct Overhead	45% Lab. & Sup.		693		
Gen. Plant Overhead	65% Oper. Costs		4,452		
Insurance, Prop. Tax	1.5% Tot. Fix. Inv.		2,809		

TOTAL OVERHEAD EXPENSES			7,954	15.91	350.70
BY-PRODUCT CREDIT					
Carbon Dioxide, lb	8.00157	2.8	-11,203		
SCP, lb	.03000	15.0	-225		

TOTAL BY-PRODUCT CREDIT			-11,428	-22.85	-503.87
			==========	=======	==========
CASH COST OF PRODUCTION			70,337	140.67	3101.30
DEPRECIATION	20% ISBL + 10% OSBL		27,580	55.16	1216.05
			==========	=======	==========
NET COST OF PRODUCTION			97,917	195.82	4317.35
REQUIRED SALES PRICE AT 10% DCF				259.5	5721.9

TABLE III-C-5

COST OF PRODUCTION ESTIMATE FOR ABE PROCESS- CONTINUOUS FERMENT

CAPITAL SUMMARY

BASIS
Location: U.S. Gulf Coast
Mid-1982
Capacity: 50.00 million gallons/yr
22,680 metric tons/yr
Str.Time: 8000 hours per year

CAPITAL COST	$MILLION
Battery Limits	88.5
Offsites	98.8
Total Fixed Inv.	187.3
Working Capital	15.5

PRODUCTION COST SUMMARY

RAW MATERIALS	UNITS PER GAL	PRICE, c/UNIT	ANNUAL COST, $M	CENTS PER GAL	DOLLARS/ MET TON
Aspen, lb	64.70349	1.0	32,353		
Sulfuric Acid, lb	.29058	4.3	625		
Calcium Hydroxide, lb	.18529	2.0	185		
Sodium Hydroxide, lb	.00930	26.0	121		
Corn, lb	.00892	4.5	20		
Ammonium Sulfate, lb	.45208	3.0	678		
Superphosphate(46), lb	1.96554	8.0	7,863		
Calcium Carbonate, lb	.90415	2.7	1,221		
Catalyst & Chemicals			2,000		
TOTAL RAW MATERIALS			45,066	90.13	1987.05
UTILITIES					
Power, kWH	1.75690	4.3	3,778		
Cooling Water, M Gal	.33955	5.8	985		
Process Water, M Gal	.03304	60.0	991		
Steam,50 psig, M Lb	.07702	295.0	11,360		
Steam,200 psig, M Lb	.01036	299.0	1,549		
TOTAL UTILITIES			18,663	37.32	822.88
OPERATING COSTS					
Labor, 46 Men @ $ 25,500		10 M/S	1,173		
Foremen, 9 Men @ $ 29,000		1 M/S	261		
Supervision, 3 Man @ $ 35,000		3 Man	105		
Maint., material & Labor	6% of ISBL		5,310		
TOTAL OPERATING COST			6,849	13.70	301.99
OVERHEAD EXPENSES					
Direct Overhead	45% Lab. & Sup.		693		
Gen. Plant Overhead	65% Oper. Costs		4,452		
Insurance, Prop. Tax	1.5% Tot. Fix. Inv.		2,809		
TOTAL OVERHEAD EXPENSES			7,954	15.91	350.70
BY-PRODUCT CREDIT					
Carbon Dioxide, lb	8.00157	2.8	-11,203		
SCP, lb	.03000	15.0	-225		
TOTAL BY-PRODUCT CREDIT			-11,428	-22.85	-503.87
CASH COST OF PRODUCTION			67,104	134.20	2958.74
DEPRECIATION	20% ISBL + 10% OSBL		27,580	55.16	1216.06
NET COST OF PRODUCTION			94,684	189.36	4174.80
REQUIRED SALES PRICE AT 10% DCF				252.5	5566.2

TABLE III-C-6

SUMMARY OF PROCESS ECONOMICS AND ENERGY REQUIREMENTS FOR CONTINUOUS ABE FERMENTATION PROCESS

Basis: 50 MM gal/yr, U.S. Gulf Coast/Louisiana
Mid-1982 (1981 utilities for DOE)

	CSI Utilities		DOE Utilities	
	Continuous Fermentation	Base	Continuous Fermentation	Base
Investment, $MM				
Battery limits	88.5	92.8	88.5	92.8
Offsites	98.8	97.3	98.8	97.3
Total fixed investment	187.3	190.1	187.3	190.1
Cost of production, ¢/gal				
Raw materials	90.13	86.59	90.13	86.59
Utilities	43.79	43.37	35.39	37.00
Operating costs	13.70	15.16	13.70	15.16
Overhead expenses	15.91	17.37	15.91	17.37
By-product credit	(22.85)	(23.20)	(22.85)	(23.20)
Cash cost of production	140.67	139.28	132.27	132.91
Depreciation	55.16	56.58	55.16	56.58
Net cost of production	195.82	195.85	187.42	189.48
Selling price at 10% DCF	259.5	259.8	250.4	252.8
Energy required, MBtu/gal of product	110.4	107.4	110.4	107.4

TABLE III-C-7

ISBL INVESTMENT BREAKDOWN FOR ABE CONTINUOUS FERMENTATION

Section	Installed Cost, MM$
Raw materials handling	.5
Prehydrolysis	29.2
Enzyme production	3.3
Enzyme hydrolysis	9.4
Fermentation	6.3
Purification	6.9
Heat generation	4.6
CO_2 recovery	8.8
Engineering & contingencies	19.5
	88.5

As can be seen from the data in Table III-C-6, continuous fermentation offers nearly equal economics to the base batch case, with the continuous case showing a nominal advantage. This is due to the fact that although continuous fermentation offers ISBL capital advantage over batch due to the decreased fermentation time, this is offset by the reduced yield (26.3% versus 27.54%) reported for the continuous case. This results in increased raw material consumption as well as slightly increased ISBL and OSBL for the pretreatment and hydrolysis portion of the plant. If continuous fermentation were to show a yield equal to that of the lower yield batch case, it would offer a 4-5 cent per gallon advantage over the batch case.

Based on the thermal equivalent of the electric power required (assumed to be 10,000 Btu/KWH) plus the enthalpy of the steam requirements (assuming an 85 percent generation efficiency), incorporation of continuous fermentation would result in an annual energy consumption of 5.52×10^{12} Btu. This represents a very slight increase (about 2 percent) over the base case in terms of energy consumption which, for all practical purposes, can be ignored.

D. Lignol Process

Design Basis

Although a good deal of lignin chemistry is speculative, the potential chemical products are impressive in number. Unfortunately, most processes currently being researched produce specialty, low volume chemicals and are difficult to justify on an economic basis.

One speculative route which may offer commercial promise involves hydrocracking and hydrodealkylation of lignin to a mixture of phenol, benzene and fuel oil. This process has been developed by Hydrocarbon Research Inc. and is reviewed herein as an alternative to utilizing the lignin for its fuel value as was done in the base case.

Lignin, as a by-product from pulp and paper manufacture, has been traditionally utilized for its fuel value in paper mills. The lignin, which comprises approximately 25 percent of the wood feed to a paper mill, is solubilized in the white liquor solution during pulping and separated from the cellulose fraction as black liquor. The black liquor is then concentrated and burnt as fuel; however, only 57 percent of the original lignin is a net fuel by-product because of furnace inefficiencies and utilization of some of the heat to concentrate the black liquor feed to the furnace. Alternatively, due to lignin's structure of aromatic rings linked with propyl groups and the presence of hydroxy and methoxy groups, lignin has the potential to be utilized as a feedstock for the production of several useful chemicals.

The structure of lignin contains only monoaromatics, which are almost exclusively derived from petroleum today. In addition, those aromatics have hydroxy and methoxy groups attached to them, which, in the case of petroleum-derived monoaromatics, have to be attached in a special process. Lignin, therefore, could potentially be used as a raw material for producing such chemicals as phenol, benzene, cresols and catechols.

Hydrocarbon Research has recently developed a process producing phenol, benzene and fuel oil from kraft lignin. Kraft lignin, which is dissolved in phenolic form in the black liquor, can be precipitated by acidification with carbon dioxide and filtered at 60-80°C to be used as feed to the HRI Lignol Process. The lignin feed is then hydrocracked in an ebullated bed reactor to hydrocarbon (fuel) gases and a liquid mixture of phenols, catechols and other hydrocarbons (fuel oil). The alkylphenols and liquid monoaromatics are then hydrodealkylated to yield phenol and benzene. Reported net yields are 20.2 percent to phenol, 14.4 percent to benzene and 10.9 percent to fuel oil, based upon original lignin. The hydrogen requirements for hydrodealkylation and hydrocracking are provided from makeup hydrogen. The plant fuel requirement is provided by some of the fuel oil produced.

HRI's Lignol Process has been slightly modified to handle a lignin feed from the ABE facility. The primary difference is that this lignin feed contains no ash or sulfur (H_2S); thus, the sulfur recovery section has been eliminated. The following assumptions and design parameters have been utilized to develop the preliminary design of the Lignol section:

- Overall process yields (weight percent based upon original lignin):

  20.2 to phenol
  14.1 to benzene
  13.1 to fuel gas
  29.1 to fuel oil

- Hydrogen is provided as makeup. The fuel oil and fuel gas produced as a by-product of the Lignol process are used as fuel for process steam requirements for both the Lignol process and other ABE plant sections.

- Net fuel oil to be credited to make steam other than that required for the Lignol process is 10.9 weight percent based upon original lignin. The remaining fuel oil (18.2%) meets the Lignol process steam requirements. All fuel gas produced (13.1%) goes to other process steam requirements.

- Product yields from lignin hydrocracking are presented in Tables III-D-1, III-D-2 and III-D-3.

TABLE III-D-1

GASEOUS HYDROCARBON COMPOSITION

	Wt % of Organic Lignin
CO	3.9
CO_2	1.8
CH_4	6.6
C_2H_6	1.9
C_3H_8	7.3
C_4H_8	1.0
C_4H_{10}	1.0
C_5H_{10}	0.7
C_5H_{12}	1.0
Total hydrocarbon gases	25.2

TABLE III-D-2

LIQUID HYDROCARBON COMPOSITION

	Wt % of Organic Lignin
H_2O	17.9
Hydrocarbons (C_6-300°F)	8.3
Hydrocarbons (300-465°F)	5.7
Phenols (300-465°F)	37.5
Catechols (465-500°F)	8.7
Heavies (500°F +)	2.4
Total liquid hydrocarbons + phenols	62.6
Total	105.7
H_2 Consumption	5.7

TABLE III-D-3

FRACTION (Wt %) COMPOSITION OF PHENOL FRACTION

Phenol	6.5
o-Cresol	3.6
m-p-Cresols	21.6
2, 4 Xylenol	7.0
p-Ethylphenol	33.2
o-n-Propylphenol	7.9
p-n-Propylphenol	20.1

Process Description

Figures III-D-1-a and III-D-1-b are flow diagrams representing the Lignol process block.

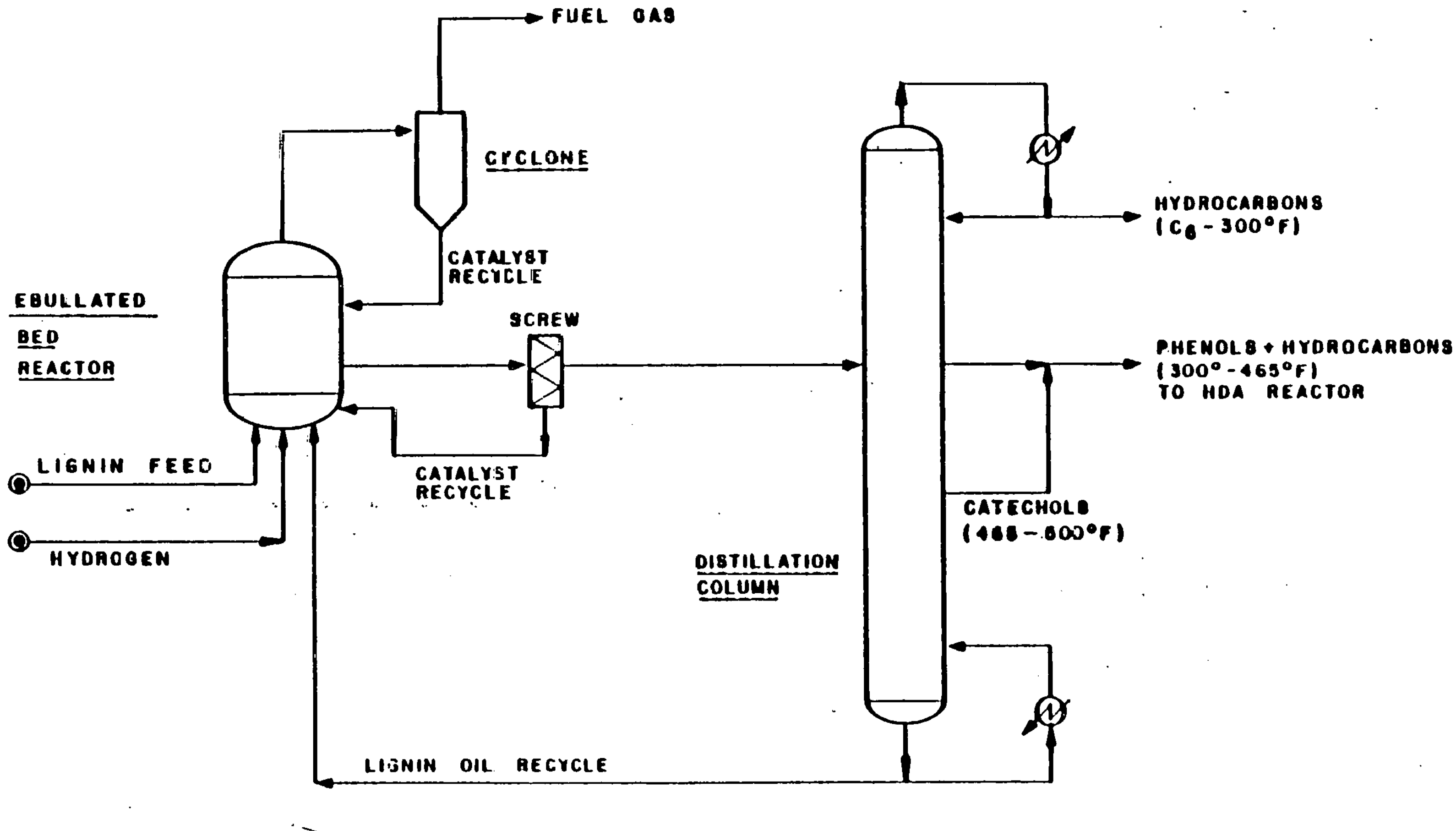


FIGURE III - D - 1a

CHEM SYSTEMS INC.
303 SOUTH BROADWAY
TARRYTOWN, N.Y. 10591

PHENOL, BENZENE PRODUCTION VIA LIGNOL PROCESS

NO 6003

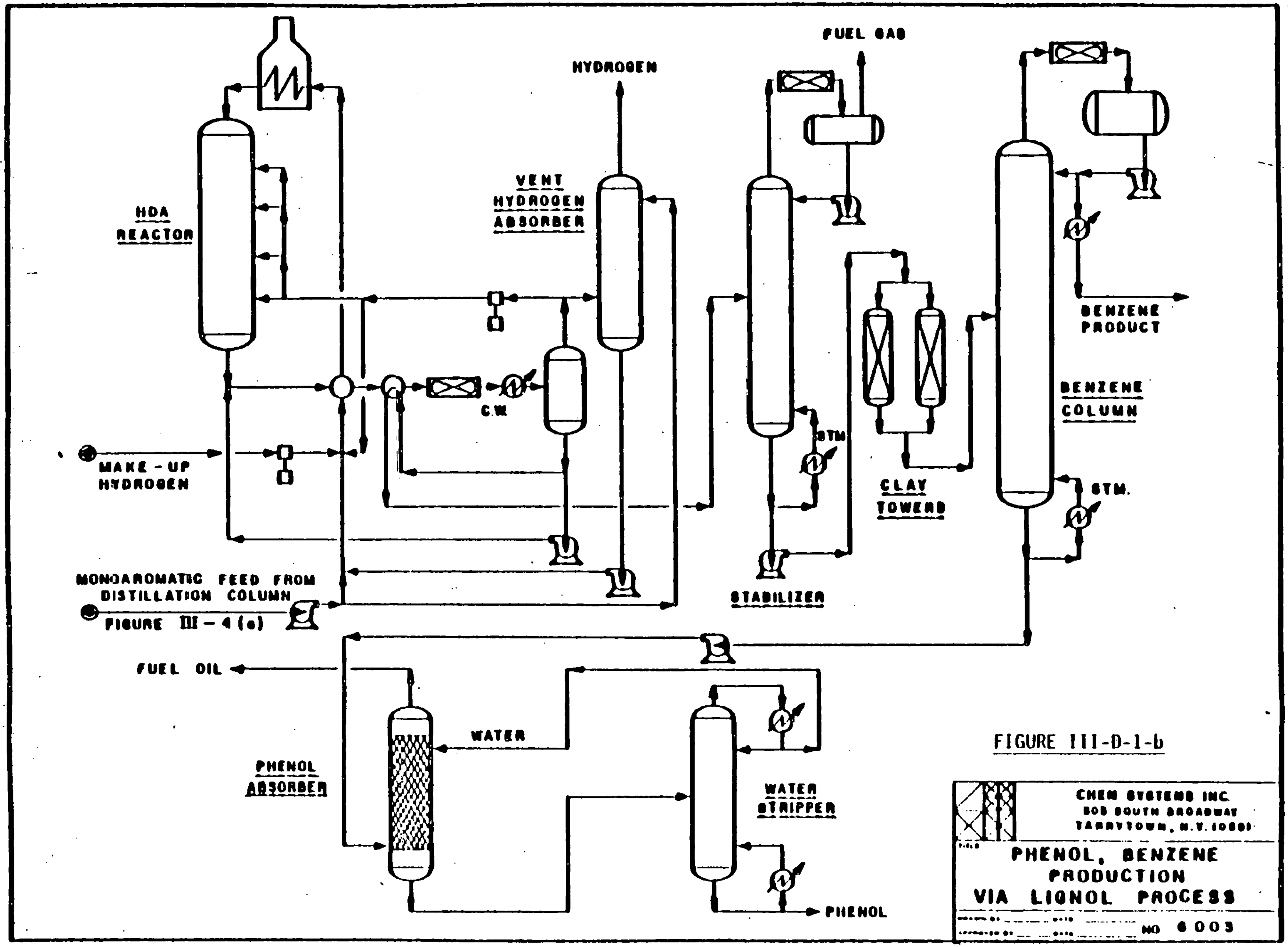


FIGURE III-D-1-b

The lignin-containing streams from enzyme hydrolysis, fermentation filter and the stillage filter are reacted with hydrogen and recycled lignin-derived tar in an ebullated bed reactor. The ebullated bed reactor concept enables heavy liquids and solids (fed at the bottom of the reactor) to be reacted in the presence of high pressure hydrogen and a catalyst. The upward velocity of the gas and liquid maintains catalyst movement with the reaction zone containing solids and liquid introduced with the feed, liquid and gaseous products, hydrogen and catalyst. Contact between catalyst and reactants is provided by the random motion of the constituents within the reactor.

The liquid products, namely, phenols, hydrocarbons, catechols and lignin-derived tar, are sent to a distillation column for fractionation. The lower boiling hydrocarbon fraction is taken overheads with successive higher boiling cuts taken off at intermediate stages. The heavy lignin tar is the column bottoms which is recycled to the hydrocracker. The overhead hydrocarbon fraction is utilized as fuel oil for plant needs and the two intermediate cuts, which contain the phenols, catechols and heavier hydrocarbons, are combined and sent to the hydrodealkylation reactor. The high pressure hydrodealkylation reaction produces phenol, benzene and fuel gas (mostly methane). The gaseous reactor products are condensed with the bulk of the unreacted hydrogen being recycled to the hydrodealkylation reactor from the flash drum. A portion of the gas is sent to a hydrogen absorber where a small portion of the feed is used to absorb benzene from the hydrogen stream and recycled back to the reactor. The pure hydrogen stream is then sent to the lignin hydrocracker as hydrogen makeup.

The liquid from the flash drum is preheated by exchange against reactor effluent and fed to a stabilizer. The stabilizer operates at 200-300 psig, which minimizes benzene losses and removes the light components as well as any water present in the feed. The overhead fuel gas is recovered and burned to produce steam in the heat generation section. The stabilizer bottoms is sent directly to the clay towers where trace impurities are removed. The effluent from the clay towers is sent to the benzene tower which produces high purity benzene as the overhead product. The benzene column bottoms containing phenol and a hydrocarbon fraction in the 300-465^{o}F boiling range is sent to a packed bed absorber. Here water is countercurrently contacted with the feed absorbing phenol. The phenol water solution constitutes the column bottoms and is sent to a water stripper where the water is removed overhead from the phenol product. The overhead hydrocarbon from the absorber is utilized as fuel oil for plant fuel requirements.

Economics and Energy Requirements

Tables III-D-4 and III-D-5 are cost of production estimates for the base case ABE low yield process with the Lignol process step substituted for the utilization of lignin for its fuel value as in the base case. Table III-D-4 uses CSI-derived utility costs and Table III-D-5 uses DOE-derived utility costs (1981 basis). These data are summarized and compared against the (revised) base case values in Table III-D-6. The facilities produce 50 million gallons per year mixed solvents at a facility located on the U.S. Gulf Coast/Louisiana in mid-1982. Table III-D-7 is an ISBL investment breakdown for the low yield with Lignol facility.

As can be seen from Table III-D-6, the Lignol case produces mixed solvents for 2.49 dollars per gallon compared to 2.60 dollars per gallon for the base case. This 11 cent per gallon advantage for the Lignol case is, of course, due to the higher by-product credit obtained from phenol and benzene, which more than offsets the increase in capital and utilities required for the Lignol processing step.

Based on the thermal equivalent of the electric power required (assumed to be 10,000 Btu/kwh) plus the enthalpy of the steam requirements (assuming an 85 percent generation efficiency) minus the energy credit accruing from the production of benzene and phenol as by-products, incorporation of the Lignol process would result in an annual energy consumption of 3.27 trillion Btu. This represents a decrease in energy consumption of about 39 percent relative to the base case.

TABLE III-D-4

COST OF PRODUCTION ESTIMATE FOR ABE PROCESS- LOW YIELD/LIGNOL

CAPITAL SUMMARY

BASIS		CAPITAL COST	$MILLION
Location:	U.S. Gulf Coast	Battery Limits	135.6
	Mid-1982	Offsites	89.8
Capacity:	50.00 million gallons/yr		------
	22,680 metric tons/yr	Total Fixed Inv.	225.4
Str.Time:	8000 hours per year	Working Capital	18.2

PRODUCTION COST SUMMARY

RAW MATERIALS	UNITS PER GAL	PRICE, ¢/UNIT	ANNUAL COST, $M	CENTS PER GAL	DOLLARS/ MET TON
Aspen, lb	61.00965	1.0	30,906		
Sulfuric Acid, lb	.27790	4.3	598		
Calcium Hydroxide, lb	.19798	2.0	198		
Sodium Hydroxide, lb	.00880	26.0	114		
Corn, lb	.00850	4.5	19		
Ammonium Sulfate, lb	.44400	3.0	666		
Superphosphate(46), lb	1.91375	8.0	7,655		
Calcium Carbonate, lb	.88034	2.7	1,189		
Hydrogen, lb	.28878	60.0	8,664		
Catalyst & Chemicals			2,000		

TOTAL RAW MATERIALS			52,009	104.01	2293.20
UTILITIES					
Power, kWH	2.33137	3.2	3,730		
Cooling Water, M Gal	.42232	5.8	1,225		
Process Water, M Gal	.03304	60.0	991		
Steam,50 psig, M Lb	.06390	391.0	12,492		

TOTAL UTILITIES			18,439	36.88	813.[illegible]
OPERATING COSTS					
Labor, 60 Men @ $ 25,500		13 M/S	1,530		
Foremen, 13 Men @ $ 29,000		2 M/S	377		
Supervision, 3 Man @ $ 35,000		3 Man	105		
Maint., Material & Labor		6% of ISBL	8,136		

TOTAL OPERATING COST			10,148	20.29	44[illegible]
OVERHEAD EXPENSES					
Direct Overhead		45% Lab. & Sup.	905		
Gen. Plant Overhead		65% Oper. Costs	6,596		
Insurance, Prop. Tax		1.5% Tot. Fix. Inv.	3,381		

TOTAL OVERHEAD EXPENSES			10,883	21.76	479.84
BY-PRODUCT CREDIT					
Carbon Dioxide, lb	8.12319	2.8	‾11,373		
SCP, lb	.03051	15.0	‾229		
Phenol, lb	1.02362	32.5	‾16,635		
Benzene, lb	.72972	20.7	‾7,553		

TOTAL BY-PRODUCT CREDIT			‾35,790	‾71.58	‾1578.06
			==========	=======	==========
CASH COST OF PRODUCTION			55,689	111.37	2455.45
DEPRECIATION	20% ISBL + 10% OSBL		36,100	72.20	1591.72
			==========	=======	==========
NET COST OF PRODUCTION			91,789	183.57	4047.17
REQUIRED SALES PRICE AT 10% DCF				249.4	5498.0

TABLE III-D-5

COST OF PRODUCTION ESTIMATE FOR ABE PROCESS- LOW YIELD/LIGNOL

CAPITAL SUMMARY

BASIS		CAPITAL COST	$MILLION
Location:	U.S. Gulf Coast	Battery Limits	135.6
	Mid-1982	Offsites	89.8
Capacity:	50.00 million gallons/yr		------
	22,680 metric tons/yr	Total Fixed Inv.	225.4
Str.Time:	8000 hours per year	Working Capital	17.9

PRODUCTION COST SUMMARY

RAW MATERIALS	UNITS PER GAL	PRICE, ¢/UNIT	ANNUAL COST, $M	CENTS PER GAL	DOLLARS/ MET TON
Aspen, lb	61.80965	1.0	30,906		
Sulfuric Acid, lb	.27790	4.3	598		
Calcium Hydroxide, lb	.19798	2.0	198		
Sodium Hydroxide, lb	.00880	26.0	114		
Corn, lb	.00850	4.5	19		
Ammonium Sulfate, lb	.44400	3.0	666		
Superphosphate(46), lb	1.91375	8.0	7,655		
Calcium Carbonate, lb	.88034	2.7	1,189		
Hydrogen, lb	.28878	60.0	8,664		
Catalyst & Chemicals			2,000		

TOTAL RAW MATERIALS			52,009	104.01	2293.20
UTILITIES					
Power, kWH	2.33137	4.3	5,013		
Cooling Water, M Gal	.42232	5.8	1,225		
Process Water, M Gal	.03304	60.0	991		
Steam,50 psig, M Lb	.06390	295.0	9,425		

TOTAL UTILITIES			16,654	33.31	734.31
OPERATING COSTS					
Labor, 60 Men @ $ 25,500	13 M/S		1,530		
Foremen, 13 Men @ $ 29,000	2 M/S		377		
Supervision, 3 Man @ $ 35,000	3 Man		105		
Maint., Material & Labor	6% of ISBL		8,136		

TOTAL OPERATING COST			10,148	20.29	447.4[illegible]
OVERHEAD EXPENSES					
Direct Overhead	45% Lab. & Sup.		905		
Gen. Plant Overhead	65% Oper. Costs		6,596		
Insurance, Prop. Tax	1.5% Tot. Fix. Inv.		3,381		

TOTAL OVERHEAD EXPENSES			10,883	21.76	479.84
BY-PRODUCT CREDIT					
Carbon Dioxide, lb	8.12319	2.8	-11,373		
SCP, lb	.03051	15.0	-229		
Phenol, lb	1.02362	32.5	-16,635		
Benzene, lb	.72972	20.7	-7,553		

TOTAL BY-PRODUCT CREDIT			-35,790	-71.58	-1578.04
			==========	=======	==========
CASH COST OF PRODUCTION			53,904	107.80	2376.75
DEPRECIATION	20% ISBL + 10% OSBL		36,100	72.20	1591.72
			==========	=======	==========
NET COST OF PRODUCTION			90,004	180.00	3968.47
REQUIRED SALES PRICE AT 10% DCF				245.5	5412.0

TABLE III-D-6

SUMMARY OF ABE FERMENTATION PROCESS ECONOMICS AND ENERGY REQUIREMENTS FOR LOW YIELD/LIGNOL PROCESS

Basis: 50MM gal/year, U.S. Gulf Coast/Louisiana, Mid-1982

	CSI Utilities		DOE Utilities	
	Lignol	Base	Lignol	Base
Investment, $MM				
Battery limits	135.6	92.8	135.6	92.8
Offsites	89.8	97.3	89.8	97.3
Total fixed investment	225.4	190.1	225.4	190.1
Cost of production, ¢/gal				
Raw materials	104.01	86.59	104.01	86.59
Utilities	36.88	43.37	33.31	37.00
Operating costs	20.29	15.16	20.29	15.16
Overhead expenses	21.76	17.37	21.76	17.37
By-product credit	(71.58)	(23.20)	(71.58)	(23.20)
Cash cost of production	111.37	139.28	107.80	132.91
Depreciation	72.20	56.58	72.20	56.58
Net cost of production	183.57	195.85	180.00	189.48
Selling price at 10% DCF	249.4	259.8	245.5	252.8
Energy required, MBtu/gal of product	65.4	107.4	65.4	107.4

TABLE III-D-7

ISBL INVESTMENT BREAKDOWN FOR LOW YIELD/LIGNOL PROCESS

	MM$
Raw materials handling	0.5
Prehydrolysis	28.3
Enzyme production	3.2
Enzyme hydrolysis	9.2
Fermentation	10.6
Purification	6.8
Heat generation	-
CO_2 recovery	9.1
Lignol	34.2
Overhead	17.7
Contingencies	16.0
Total	135.6

E. Baelene Solvent Extraction

Design Basis

Contacts with Baeol, Inc. have unfortunately not yielded the type of information required to perform a detailed technical and economic evaluation of their solvent extraction process. Therefore, a number of assumptions have been made, and these in conjunction with the information that was made available form the basis for the analysis.

The patent information(1) contains some extraction information on butanol and ethanol in various fluorocarbon solvents but contains no data for acetone. Subsequent conversations with Baeol indicated that new data had been obtained for an unknown solvent system which essentially extracted all of the butanol, acetone and ethanol. Unfortunately, the new data were only available via a secrecy agreement, which would not allow us to publish the results. Therefore based on the information that was made available in the patent, the following assumptions are made:

1. Baeol has indicated that, based on information received from experts previously working in the fermentation ABE industry, fermentation conditions can be set such that essentially all the acetone (along with a substantial quantity of water and to a lesser extent butanol and ethanol) can be stripped from the beer by the evolved CO_2 and H_2 gases in a manner similar to vacuum fermentation. In order to use the fermentation section base case design, however, it is assumed that no solvents are evaporated during fermentation, and that the feed to purification contains all the products formed during fermentation. If the acetone were evaporated during fermentation along with a significant quantity of water, then that stream would also have to be purified in some manner, e.g., by extraction or distillation. In either case, whether the acetone is or is not

(1) U.S. Patent 4,260,836, (April 7, 1981) to Sidney Levy.

evaporated, a similar amount of extraction has to take place, i.e., identical quantities of solvents must be recovered in some manner.

2. F-21, or monofluoro dichloromethane is chosen as the solvent. General ethanol and butanol extraction data are available from the patent, and it is assumed that acetone is extracted in a manner similar to butanol.

3. F-21 is 0.95 weight percent soluble in H_2O and water is 0.16 weight percent soluble in freon, both at 25^o and 1 atmosphere. Although these solubilities are comparatively low, it must be recalled that a very large quantity of water is involved (about 3 million pounds per hour), making the potential loss of freon meaningful. Following countercurrent extraction, a significant amount of F-21 (about 28,000 pounds per hour) is transferred to the dilute aqueous phase. Since F-21 is relatively expensive, it is economically unattractive to provide makeup F-21 to account for this loss. Therefore the F-21 must be recovered from the dilute aqueous stream. In lieu of any novel extraction process which might accomplish this separation, distillation of the dilute aqueous stream is the assumed method of recovery.

Process Description

A conceptualized flowsheet for the Baelene solvent extraction process is presented in Figure III-E-1.

A dilute mixed solvent stream containing approximately 1.43 weight percent total solvents from fermentation is contacted countercurrently with a solvent in a two stage mixer-settler extraction system. The solvent of choice here is monofluoro dichloromethane (F-21) which has the ability to extract ethanol, butanol and (it is assumed) acetone with very low miscibility with water. In addition, F-21 is extremely volatile,

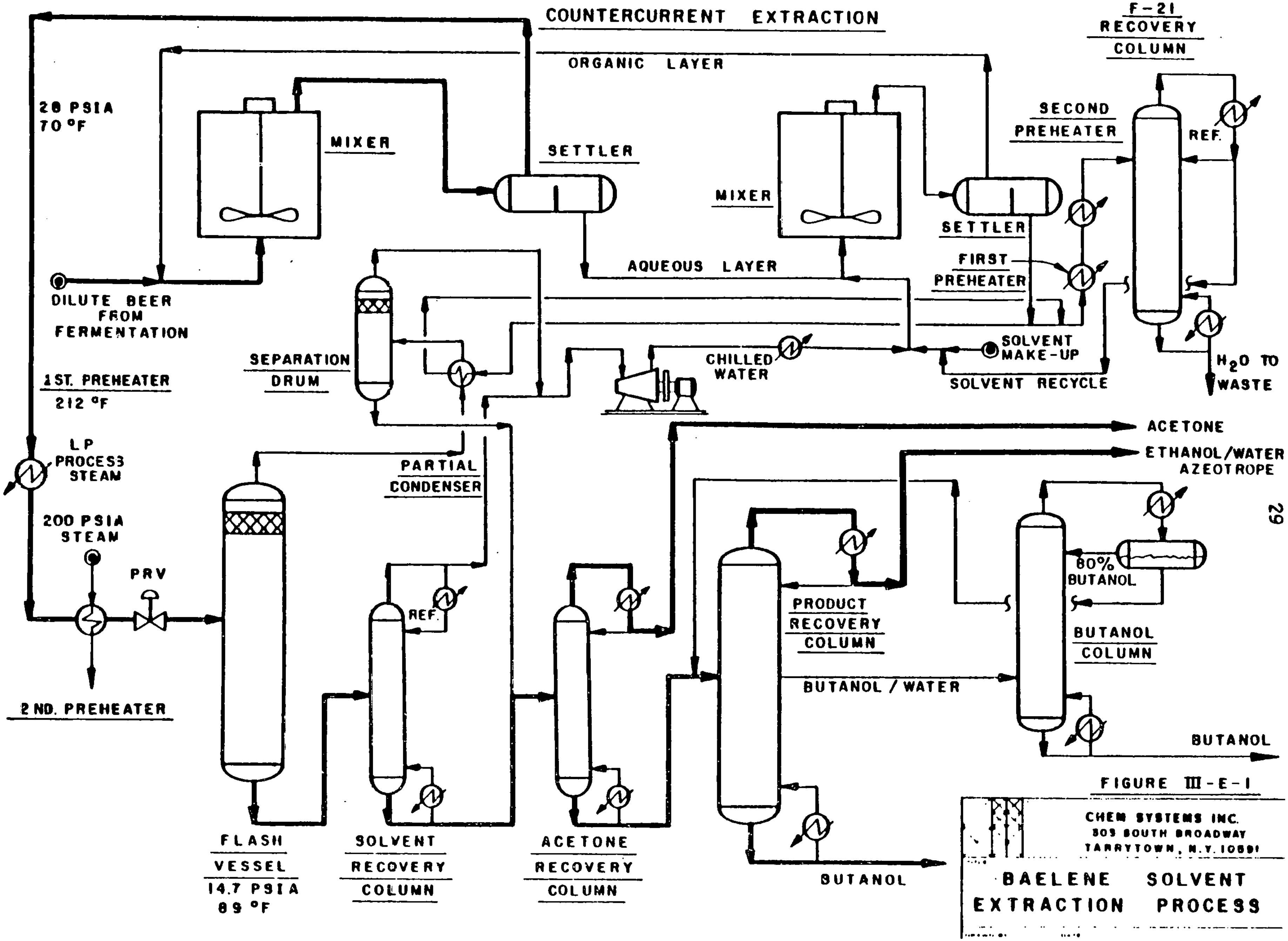


FIGURE III-E-1

BAELENE SOLVENT EXTRACTION PROCESS

having a boiling point of 48^{o}F at atmospheric pressure. In order for the extraction to take place in the liquid phase at 70^{o}F, the mixer-settler operation must be under 28 psia.

The mixer-settler operation takes place as follows. The dilute beer feed is contacted with the organic phase from the second stage, thoroughly mixed and decanted. The aqueous phase from the first stage becomes the feed to the second stage which is contacted with fresh recycle solvent. The organic phase from the first stage is the extract and contains nearly all the butanol, ethanol and acetone. The aqueous phase from the second settler which contains most of the water is sent to a solvent recovery column where the F-21 is recovered overhead. The feed to the solvent recovery column is at 70^{o}F and must be heated to the bubble point of the mixture, which is essentially all water. A first preheating step uses atmospheric steam generated in prehydrolysis to bring the mixture to 94^{o}F. Steam at 50 psig is used to bring the mixture from 94^{o}F to its bubble point. This sensible heat required to preheat the feed is the primary energy consumption due to the volume of water ($\sim$ 3 million pounds per hour) being heated. Additional energy is required to distill the F-21 in the recovery column; however, this is only a fraction of that required to preheat the feed. Refrigeration is required to condense the overhead F-21 vapors to be refluxed.

In this operation approximately 99 percent of the butanol and acetone and 85 percent of the ethanol are recovered in the extract. The extract contains approximately 8.9 percent butanol, 4.4 percent acetone and 8 percent ethanol with a trace of water. The remainder of the extract is the solvent, F-21.

The extract, which is at 28 psia and 70^{o}F, is sent through a series of preheating steps prior to flashing. The first preheating step is accomplished with atmospheric steam generated in the prehydrolysis section. The second preheating step, terminating in a feed temperature of 343^{o}F is accomplished with 200 psig steam. The extract stream is then letdown to atmospheric pressure where 60 percent of the feed is flashed at 90^{o}F.

A relatively pure F-21 vapor stream is obtained (97 percent F-21) containing only traces of butanol, ethanol, acetone and water, which are easily condensed in a partial condenser and separated in a drum.

The liquid stream from the flash vessel still contains about 68 percent F-21, 23 percent butanol and 7.5 percent acetone. This stream is sent to a solvent recovery column where the remaining solvent is taken overhead and combined with the vapor stream from the separation drum. This pure F-21 stream is compressed to 28 psia, condensed and recycled with fresh solvent makeup to the second stage mixer-settler.

The bottoms from the solvent recovery column is combined with the liquid condensate from the separation drum and sent to an acetone recovery column. Here the acetone is taken overhead as distillate and sent to product storage. The acetone recovery column bottoms, which contains most of the butanol and ethanol, is sent to the product recovery column. The product recovery column separates butanol and ethanol into separate components, the ethanol/H_2O azeotrope forming the overheads, the remaining water/butanol azeotrope is taken as an intermediate cut, and butanol forms the column bottoms.

The butanol/water azeotrope cut from the product recovery column is sent to a butanol column where pure butanol is recovered as the bottoms. The overheads is a 70 percent butanol/water stream which is decanted. The organic layer (80 percent butanol) is refluxed back to the column, while the aqueous layer (4 percent butanol) is recycled to the product recovery column.

Economics and Energy Requirements

Tables III-E-1 and III-E-2 represent cost of production estimates for the ABE low yield case with a Baelene solvent extraction scheme as described above using CSI and DOE utilities, respectively. These data are summarized and compared aganst the (revised) base case values in Table III-E-3. Both cases are for a 50 million gallon per year ABE facility located on the U.S. Gulf Coast in mid-1982.

TABLE III-E-1

COST OF PRODUCTION ESTIMATE FOR ABE PROCESS- LOW YIELD

CAPITAL SUMMARY

BASIS
Location: U.S. Gulf Coast
Mid-1982
Capacity: 50.00 million gallons/yr
22,680 metric tons/yr
Str.Time: 8000 hours per year

CAPITAL COST	$MILLION
Battery Limits	102.4
Offsites	89.3
Total Fixed Inv.	191.7
Working Capital	16.4

PRODUCTION COST SUMMARY

RAW MATERIALS	UNITS PER GAL	PRICE, ¢/UNIT	ANNUAL COST, $M	CENTS PER GAL	DOLLARS/ MET TON
Aspen, lb	61.80965	1.0	30,906		
Sulfuric Acid, lb	.27790	4.3	598		
Calcium Hydroxide, lb	.19798	2.0	198		
Sodium Hydroxide, lb	.00880	26.0	114		
Corn, lb	.00850	4.5	19		
Ammonium Sulfate, lb	.44400	3.0	666		
Superphosphate(46), lb	1.91375	8.0	7,655		
Calcium Carbonate, lb	.88034	2.7	1,189		
Catalyst & Chemicals			1,950		
TOTAL RAW MATERIALS			43,295	86.59	1908.99
UTILITIES					
Power, kWH	2.84385	3.2	4,550		
Cooling Water, M Gal	.17149	5.8	497		
Process Water, M Gal	.03304	60.0	991		
Steam,50 psig, M Lb	.04726	391.0	9,357		
Steam,200 psig, M Lb	.02096	396.0	4,150		
TOTAL UTILITIES			19,546	39.09	861.81
OPERATING COSTS					
Labor, 60 Men @ $ 25,500		13 M/S	1,530		
Foremen, 13 Men @ $ 29,000		2 M/S	377		
Supervision, 3 Man @ $ 35,000		3 Man	105		
Maint., Material & Labor		6% of ISBL	6,144		
TOTAL OPERATING COST			8,156	16.31	359.61
OVERHEAD EXPENSES					
Direct Overhead	45% Lab. & Sup.		905		
Gen. Plant Overhead	65% Oper. Costs		5,301		
Insurance, Prop. Tax	1.5% Tot. Fix. Inv.		2,875		
TOTAL OVERHEAD EXPENSES			9,082	18.16	400.46
BY-PRODUCT CREDIT					
Carbon Dioxide, lb	8.12317	2.8	¯11,373		
SCP, lb	.03051	15.0	¯229		
TOTAL BY-PRODUCT CREDIT			¯11,602	¯23.20	¯511.55
CASH COST OF PRODUCTION			68,478	136.95	3019.31
DEPRECIATION	20% ISBL + 10% OSBL		29,410	58.82	1296.75
NET COST OF PRODUCTION			97,888	195.76	4316.06
REQUIRED SALES PRICE AT 10% DCF				258.0	5607.5

TABLE III-E-2

COST OF PRODUCTION ESTIMATE FOR ABE PROCESS- LOW YIELD

CAPITAL SUMMARY

BASIS		CAPITAL COST	$MILLION
Location:	U.S. Gulf Coast	Battery Limits	102.4
	Mid-1982	Offsites	89.3
Capacity:	50.00 million gallons/yr		------
	22,680 metric tons/yr	Total Fixed Inv.	191.7
Str.Time:	8000 hours per year	Working Capital	16.2

PRODUCTION COST SUMMARY

RAW MATERIALS	UNITS PER GAL	PRICE, ¢/UNIT	ANNUAL COST, $M	CENTS PER GAL	DOLLARS/ MET TON
Aspen, lb	61.80965	1.0	30,906		
Sulfuric Acid, lb	.27790	4.3	598		
Calcium Hydroxide, lb	.19798	2.0	198		
Sodium Hydroxide, lb	.00880	26.0	114		
Corn, lb	.00850	4.5	19		
Ammonium Sulfate, lb	.44400	3.0	666		
Superphosphate(46), lb	1.91375	8.0	7,655		
Calcium Carbonate, lb	.88034	2.7	1,189		
Catalyst & Chemicals			1,950		

TOTAL RAW MATERIALS			43,295	86.59	1908.99
UTILITIES					
Power, kWH	2.84385	4.3	6,115		
Cooling Water, M Gal	.17149	5.8	497		
Process Water, M Gal	.03304	60.0	991		
Steam,50 psig, M Lb	.04786	295.0	7,059		
Steam,200 psig, M Lb	.02096	299.0	3,134		

TOTAL UTILITIES			17,796	35.59	784.66
OPERATING COSTS					
Labor, 60 Men @ $ 25,500		13 M/S	1,530		
Foremen, 13 Men @ $ 29,000		2 M/S	377		
Supervision, 3 Man @ $ 35,000		3 Man	105		
Maint., Material & Labor	6% of ISBL		6,144		

TOTAL OPERATING COST			8,156	16.31	359.61
OVERHEAD EXPENSES					
Direct Overhead	45% Lab. & Sup.		905		
Gen. Plant Overhead	65% Oper. Costs		5,301		
Insurance, Prop. Tax	1.5% Tot. Fix. Inv.		2,875		

TOTAL OVERHEAD EXPENSES			9,082	18.16	400.46
BY-PRODUCT CREDIT					
Carbon Dioxide, lb	8.12319	2.8	¯11,373		
SCP, lb	.03051	15.0	¯229		

TOTAL BY-PRODUCT CREDIT			¯11,602	¯23.20	¯511.55
			==========	=======	==========
CASH COST OF PRODUCTION			66,728	133.45	2942.17
DEPRECIATION	20% ISBL + 10% OSBL		29,410	58.82	1296.75
			==========	=======	==========
NET COST OF PRODUCTION			96,138	192.27	4238.91
REQUIRED SALES PRICE AT 10% DCF				254.1	5603.3

TABLE III-E-3

SUMMARY OF PROCESS ECONOMICS AND ENERGY REQUIREMENTS FOR ABE LOW YIELD BAELENE SOLVENT EXTRACTION CASE

Basis: 50 MM gal/yr US Gulf Coast
Mid-1982

	CSI Utilities		DOE Utilities	
	Baelene	Base	Baelene	Base
Investment, MM$				
Battery limits	102.4	92.8	102.4	92.8
Offsites	89.3	97.3	89.3	97.3
Total fixed investment	191.7	190.1	191.7	190.1
Cost of production, ¢/lb				
Raw materials	86.59	86.59	86.59	86.59
Utilities	39.09	43.37	35.59	37.00
Operating costs	16.31	15.16	16.31	15.16
Overhead expenses	18.16	17.37	18.16	17.37
By-product credit	(23.20)	(23.20)	(23.20)	(23.20)
Cash cost of production	136.95	139.28	133.45	132.91
Depreciation	58.82	56.58	58.82	56.58
Net cost of production	195.76	195.85	192.27	189.48
Selling price at 10% DCF	258.0	259.8	254.1	252.8
Energy requirements, MBtu/gal of product	100.4	107.4	100.4	107.4

In comparing these economics against revised base case economics, it can be seen that this solvent extraction scheme offers little advantage over conventional distillation. Less than two cents per gallon separates the two processes. This is due to the fact that although butanol, acetone and ethanol have been extracted with little energy consumption, the recovery of the F-21 solvent from the dilute aqueous waste stream is almost as costly in steam as the overall separation in conventional distillation. This is caused by the fact that the vast volume of water has to be heated to column temperature in both cases. Although there is a slight savings in steam consumption for the Baelene case, more power is used for the refrigeration system. This results in virtually identical economics for the two cases.

Based on the thermal equivalent of the electric power required (assumed to be 10,000 Btu/kwh) plus the enthalpy of the steam requirements (assuming an 85 percent generation efficiency), incorporation of the Baelene process would result in an annual energy consumption of about 5.02 trillion Btu. This represents a decrease of about 6 percent in energy consumption relative to the base case.

F. Dual Enzyme System

Design Basis

Investigations(1) of mild acid prehydrolysis as a pretreatment for enzyme hydrolysis have recently attempted to optimize the conditions of prehydrolysis and enzyme hydrolysis. The base case model analyzed previously chose prehydrolysis reactor conditions to be 0.5 percent sulfuric acid, residence time of 12 seconds and a temperature of 190^{o}C. It also assumed that the hydrolysis of amorphous cellulose and hemicellulose occurs almost instantaneously with a 95 mol percent conversion. This pretreatment resulted in an enzyme hydrolysis yield of 90 mol percent cellulose to glucose at a residence time of 24 hours, and an enzyme loading of RUT-C-30 produced enzymes of 12.5 IU/gm cellulose.

More recent data(2) have resulted in significantly better yields. Studies were conducted which compared enzyme hydrolysis yields at different prehydrolysis conditions for different enzyme loadings and enzyme systems. It was found that the most effective prehydrolysis conditions (the ones which made the cellulose most accessible to enzymatic attack) were at 203^{o}C, 0.275 percent H_2SO_4 and 7.9 second residence time. When only RUT-C-30 cellulase is used (RUT C-30 naturally produces primarily β-glucanase with little β-glucosidase (cellobiase), glucose yield is depressed because of the accumulation of cellobiase. When a small amount of β-glucosidase is added in the form of a NOVO cellobiase 250L solution, significantly higher glucose yields

(1) Grethlein, H.E., et al., "Annual Report on Acid Hydrolysis of Cellulosic Biomass," March 1980-March 31, 1981.
(2) Grethlein, H.E., et al., "Second Annual Report on Acid Hydrolysis of Cellulosic Biomass," March 1981-August 1982.

are realized. Quantitative yields can be obtained at about 12.5 hours enzyme hydrolysis residence time for enzyme loadings of .61 IU/ml (33.85 IU/gm solids) and .3 IU/ml (16.92 IU/gm solids) and 88 percent yield after 24 hours for an enzyme loading of .05 IU/ml (2.7 IU/gm solids). This relationship is illustrated in Figure III-F-1.

The prehydrolysis conditions which result in higher glucose enzyme hydrolysis yields apparently cause significant hemicellulose removal without extensive glucose formation or any glucose degradation. Hemicellulose removal is approximately 70-100 percent without significant furfural formation which leaves a porous cellulose structure that is readily accessible to enzymatic attack. In addition, the lignin is also altered such that approximately 50 percent is soluble in ethanol following prehydrolysis under these conditions.

Using these new data, modifications were made to the base case model changing the prehydrolysis, enzyme production and enzyme hydrolysis section design parameters. The design of all three sections remains essentially the same as for the base case model except a two enzyme system is incorporated into the enzyme production section. The introduction of β-glucosidase (cellobiase) into the enzyme hydrolyzers is accomplished via production of β-glucosidase via Aspergillus phoenicis QM 329. As mentioned previously, β-glucosidase is an enzyme which catalyzes the conversion of cellobiase to glucose and prevents its accumulation, thus increasing glucose yields compared to the single RUT-C-30 system. RUT-C-30 enzymes are still produced in the same manner as in the base case model.

The new design parameters for the prehydrolysis, enzyme production and enzyme hydrolysis sections are summarized in Table III-F-1.

Process Design

The process descriptions for the prehydrolysis, enzyme production and enzyme hydrolysis sections of the plant are essentially the same as for the base case except for some minor modifications. As mentioned in the

FIGURE III-F-1

HYDROLYSIS OF PRETREATED POPLAR WITH C30 CELLULASE AND NOVO CELLOBIASE AT 0.1 ml/100 ml

PRETREATMENT: 200°C, 0.50% ACID, 7.9 sec. (RUN 0924)

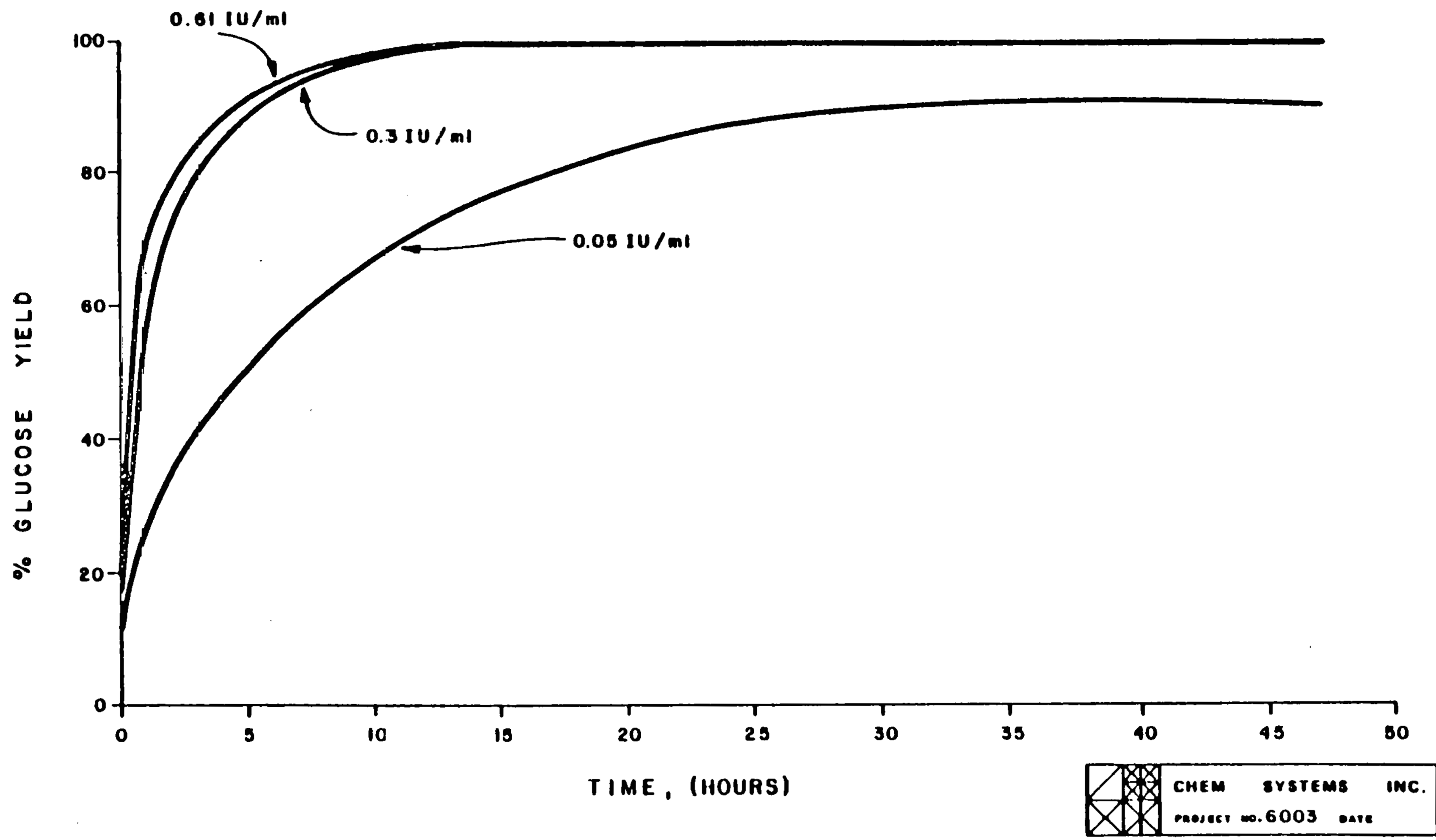


TABLE III-F-1

DESIGN PARAMETERS

Prehydrolysis

• Temperature	200°C
• Acid concentration	0.5 Wt%
• Residence Time	7.9 seconds

Enzyme Production

	RUT-C-30	QM 329
• Temperature	86°F	86°F
• pH	4.8	3.0
• Pressure	Atmospheric	Atmospheric
• Nutrients	(1)	(1)
• Oxygen	17,264 IU/mol O_2	17,264 IU/mol O_2
• Cell concentration	7 gms/liter	7 gms/liter
• Cell yield	0.26 gms mycellium/ gm cellulose	0.26 gms mycellium/ gm/cellulose
• Enzyme yield	1.63 gms enzyme/ gm cellulose based upon enzyme productivity of 114 IU/l/hr	7.17 gms enzyme/ gm cellulose based upon enzyme productivity of 500 IU/l/hr
• Cell recycle	.77 gm/gm cells	.77 gm/gm cells

(1) 1.0 percent cellulose
0.2 percent KH_2PO_4
0.03 percent $CaCl_2$
0.03 percent $MgSO_4 \cdot 7H_2O$
1.0 percent corn steep liquor

Enzyme Hydrolysis

• Temperature	122°F
• Pressure	Atmospheric
• pH	4.8
• Hydrolysis time	12.5 hours
• Hydrolysis conversion	100 mol percent conversion cellulose to glucose
• Terminal sugar concentration	5.82 percent
• Enzyme loading	17 IU/gm solids

design basis, the design parameters and performance characteristics of these sections have been changed to reflect recent advances in prehydrolysis pretreatment for enzyme hydrolysis. The prehydrolysis and enzyme hydrolysis sections are identical in design to the base case except for these changes in design and performance parameters. Figures III-F-2 and III-F-3 are flowsheets illustrating the prehydrolysis and enzyme hydrolysis sections. The reader is referred to the base case study for a detailed process description of these two sections(1). The enzyme production section design has been modified slightly to accommodate the two enzyme system discussed in the design basis.

A flowsheet representing the enzyme production section reflecting the modifications required to incorporate the two enzyme system is presented in Figure III-F-4. Essentially, the design is the same as for the base case except that two identical (in equipment) trains are required, one to produce RUT-C-30 enzymes as before, and another much smaller train to produce QM329 β-glucosidase. Although the equipment required for both trains is the same, the design parameters and performance characteristics are somewhat different, as discussed in the design basis. QM329 has a significantly higher productivity compared to RUT-C-30, and since only a small amount of β-glucosidase is required to significantly increase enzyme hydrolysis glucose yields, a much smaller production system is required.

Enzymes for enzyme hydrolysis of cellulose are produced from a mutation of T. Reesei fungus, RUT-C-30. β-glucosidase is produced by the QM329 strain of Aspergillus phoenicis. The RUT-C-30 is used as a seed to produce an enzyme mixture of endo-glucanase and β-glucosidase, and also to produce more cells (mycellium). QM 329 produces primarily β-glucosidase and more cells. Enzyme production for both systems takes

(1) "Technical and Economic Assessment of Processes for the Production of Butanol and Acetone," prepared by Chem Systems Inc. for Jet Propulsion Laboratory, sponsored through an agreement with NASA by Energy Conversion and Utilization Technology Divison, Office of Energy Systems Research, U.S. Department of Energy, September, 1982 (JPL 9950-776).

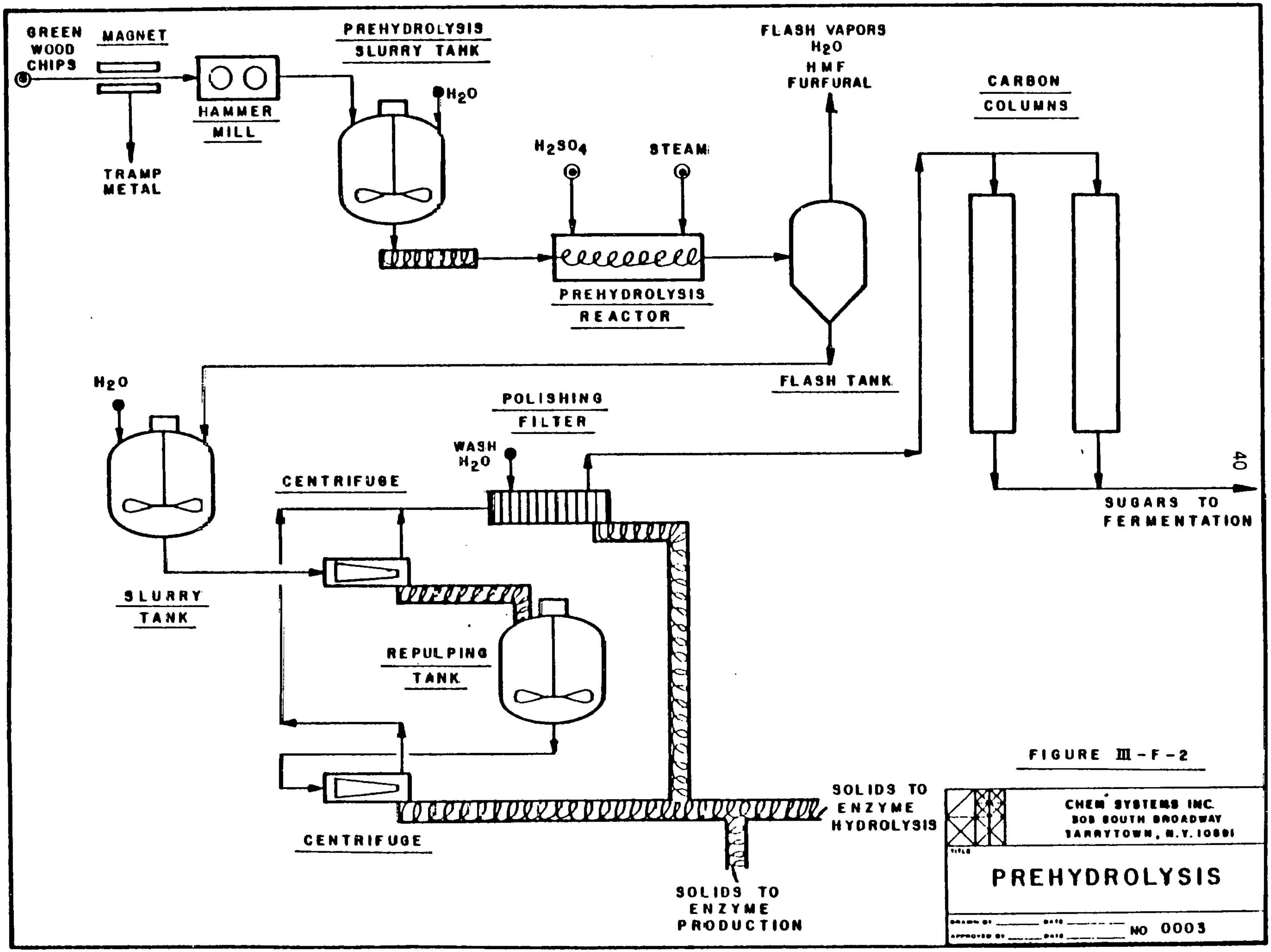


FIGURE III-F-2

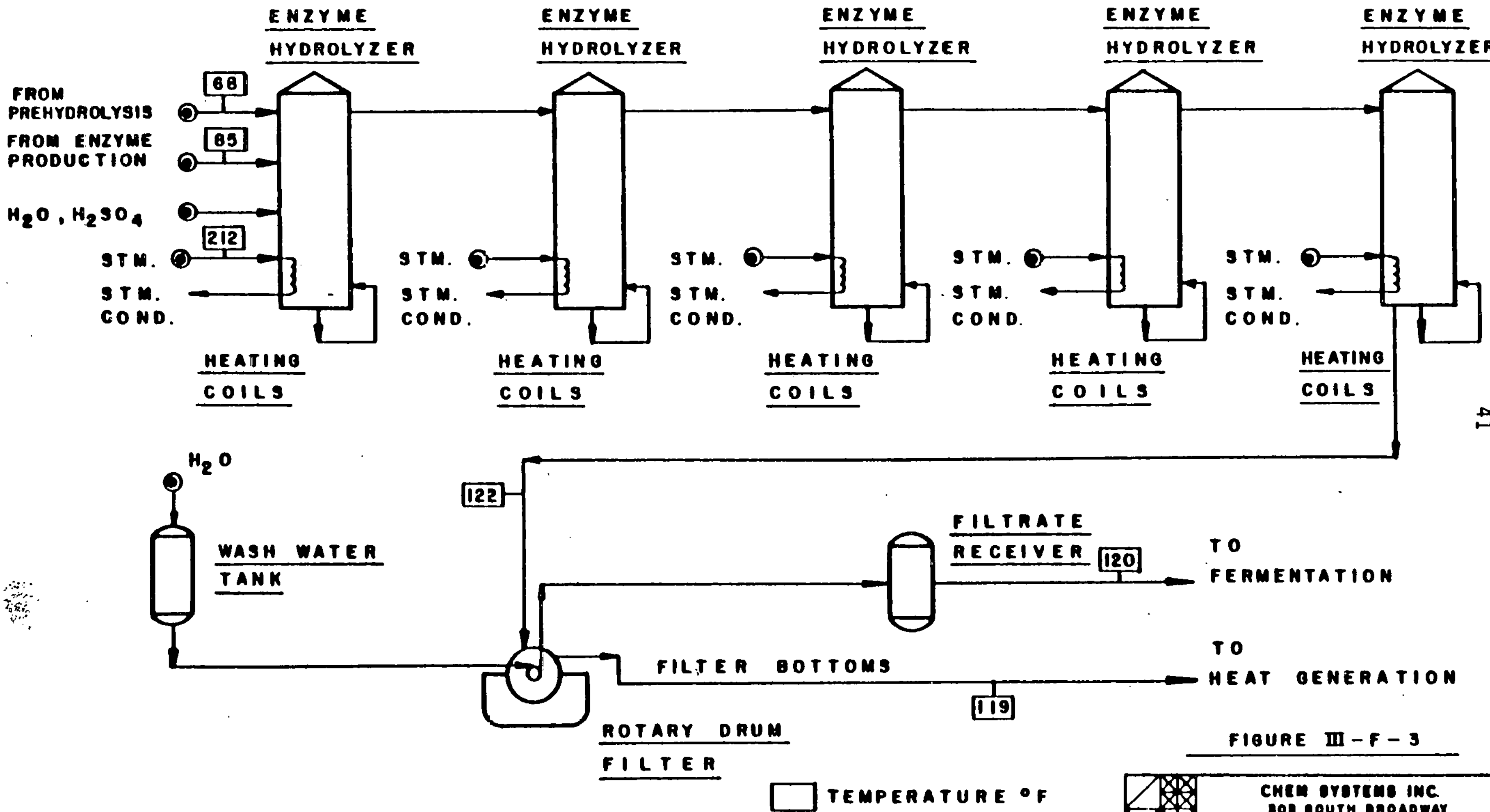


FIGURE III-F-3

CHEM SYSTEMS INC.
303 SOUTH BROADWAY
TARRYTOWN, N.Y. 10591

ENZYME HYDROLYSIS

NO. 0003

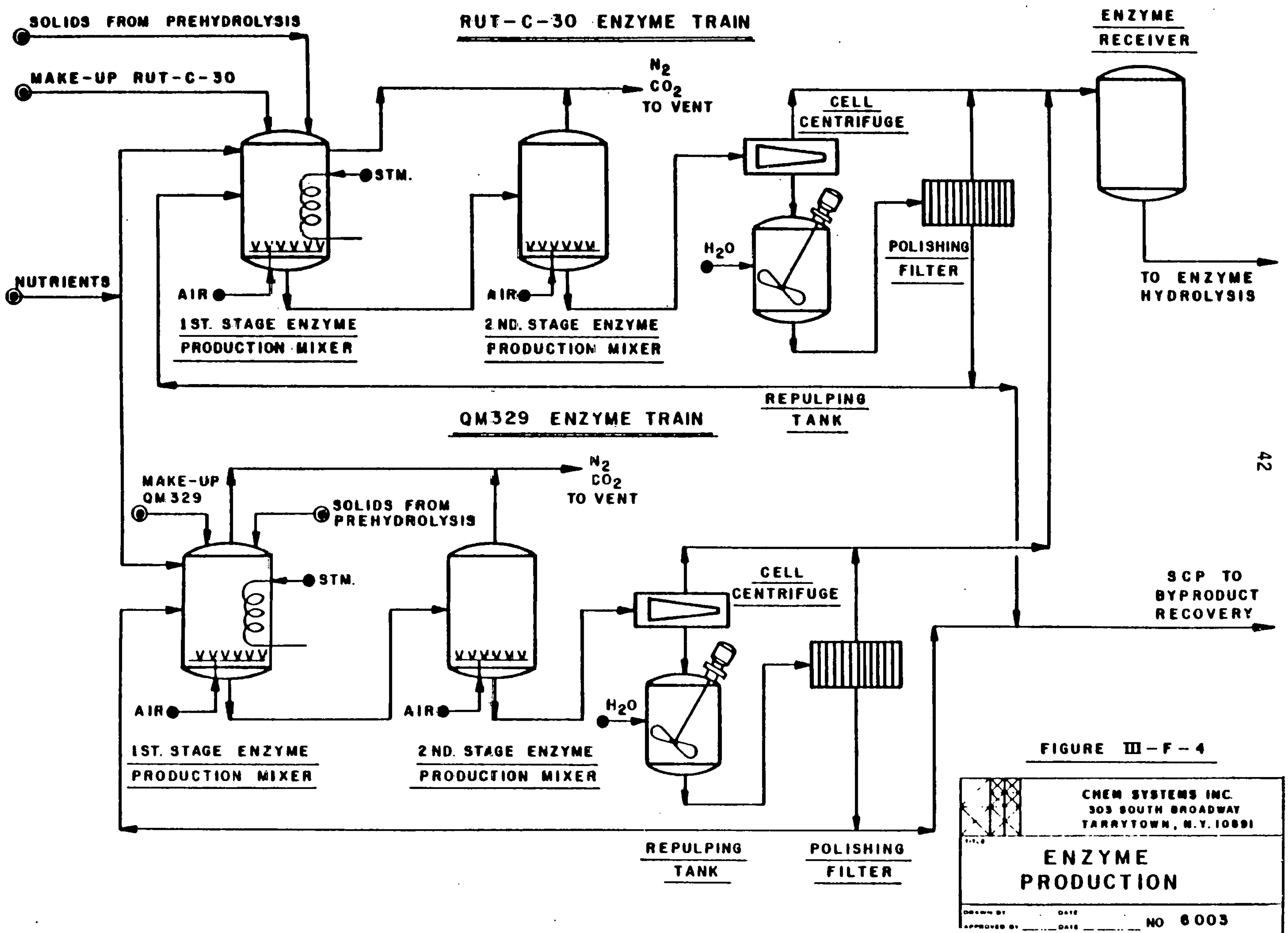


FIGURE III-F-4

place in a two-stage continuous fermentation system at 86°F and a pH of 4.8 for RUT-C-30 and 3.0 for QM329. Enzyme productivity is 114 IU/l/hr for RUT-C-30 and 500 IU/l/hr for QM329. Ligno-cellulose from the process is used as a carbon source, corn steep liquor is used as a nitrogen source and other inorganic salts are provided to complete the nutrient requirements. Air is sparged into the fermenters as an oxygen source. Recirculation pumps provide agitation in the vessels and temperature is maintained by steam heated coils.

Carbon dioxide and nitrogen are vented to the atmosphere. The product from the enzyme fermenters is sent to a cell centrifuge to remove most of the mycellium as the centrifuge bottoms. The bottoms are then repulped to eight percent (8 wt.%) solids and filtered and washed to recover the enzyme remaining on the original cake. The cake is split into two streams, one is recycled back to the enzyme fermenters to serve as the enzyme seed, the other is recovered as single cell protein by-product. The centrifuge overflow and filtrate are combined for both systems and sent to the enzyme receiver prior to entry into the enzyme hydrolyzers.

Economics and Energy Requirements

Tables III-F-1 and III-F-2 represent cost of production estimates for the ABE low yield case with the dual enzyme system, using CSI and DOE utilities respectively. These data are summarized and compared against the (revised) base case values in Table III-F-3. Both cases are for a 50 million gallon per year ABE facility located on the U.S. Gulf Coast in mid-1982.

In comparing these results to the revised base case economics, it can be seen that the dual enzyme system offers superior economics. The mixed solvents product can be produced for 249.3 cents per gallon using CSI utilities and 242.2 cents per gallon using DOE utilities. This is approximately a 10 cent per gallon price advantage for the dual enzyme system. This is primarily due to a small decrease in raw materials cost afforded by the enzyme hydrolysis yield increase and a substantial

TABLE III-F-1

COST OF PRODUCTION ESTIMATE FOR ABE PROCESS- LOW YIELD/DUAL ENZYM

CAPITAL_SUMMARY

BASIS		CAPITAL_COST	$MILLION
Location:	U.S. Gulf Coast	Battery Limits	86.1
	Mid-1982	Offsites	92.4
Capacity:	50.00 million gallons/yr		------
	22,680 metric tons/yr	Total Fixed Inv.	178.5
Str.Time:	8000 hours per year	Working Capital	15.5

PRODUCTION_COST_SUMMARY

RAW_MATERIALS	UNITS PER_GAL	PRICE, c/UNIT	ANNUAL COST,_$M	CENTS PER_GAL	DOLLARS/ MET_TON_
Aspen, lb	59.79676	1.0	29,900		
Sulfuric Acid, lb	.27790	4.3	598		
Calcium Hydroxide, lb	.19798	2.0	198		
Sodium Hydroxide, lb	.00880	26.0	114		
Corn, lb	.00850	4.5	19		
Ammonium Sulfate, lb	.44400	3.0	666		
Superphosphate(46), lb	1.91375	8.0	7,655		
Calcium Carbonate, lb	.88034	2.7	1,189		
Catalyst & Chemicals			1,950		

TOTAL RAW MATERIALS			42,289	84.57	1864.61
UTILITIES					
Power, kWH	1.76310	3.2	2,821		
Cooling Water, M Gal	.27023	5.8	784		
Process Water, M Gal	.02988	60.0	896		
Steam,50 psig, M Lb	.02405	391.0	4,702		
Steam,200 psig, M Lb	.06309	396.0	12,492		

TOTAL UTILITIES			21,695	43.39	956.57
OPERATING_COSTS					
Labor, 60 Men @ $ 25,500	13 M/S		1,530		
Foremen, 13 Men @ $ 29,000	2 M/S		377		
Supervision, 3 Man @ $ 35,000	3 Man		105		
Maint., Material & Labor	6% of ISBL		3,166		

TOTAL OPERATING COST			7,178	14.36	316.49
OVERHEAD_EXPENSES					
Direct Overhead	45% Lab. & Sup.		905		
Gen. Plant Overhead	65% Oper. Costs		4,666		
Insurance, Prop. Tax	1.5% Tot. Fix. Inv.		2,677		

TOTAL OVERHEAD EXPENSES			8,249	16.50	363.70
BY-PRODUCT_CREDIT					
Carbon Dioxide, lb	8.12319	2.8	-11,373		
SCP, lb	.03051	15.0	-229		

TOTAL BY-PRODUCT CREDIT			-11,602	-23.20	-511.55
			=========	=======	=========
CASH COST OF PRODUCTION			67,809	135.61	2989.61
DEPRECIATION	20% ISBL + 10% OSBL		26,460	52.92	1166.67
			=========	=======	=========
NET COST OF PRODUCTION			94,269	188.53	4156.49
REQUIRED SALES PRICE AT 10% DCF				249.3	5495.9

TABLE III-F-2

COST OF PRODUCTION ESTIMATE FOR ABE PROCESS- LOW YIELD/DUAL ENZYM

CAPITAL SUMMARY

BASIS		CAPITAL COST	$MILLION
Location:	U.S. Gulf Coast	Battery Limits	86.1
	Mid-1982	Offsites	92.4
Capacity:	50.00 million gallons/yr		------
	22,680 metric tons/yr	Total Fixed Inv.	178.5
Str.Time:	8000 hours per year	Working Capital	15.0

PRODUCTION COST SUMMARY

RAW MATERIALS	UNITS PER GAL	PRICE, ¢/UNIT	ANNUAL COST, $M	CENTS PER GAL	DOLLARS/ MET TON
Aspen, lb	59.79676	1.0	29,900		
Sulfuric Acid, lb	.27790	4.3	598		
Calcium Hydroxide, lb	.19798	2.0	198		
Sodium Hydroxide, lb	.00880	26.0	114		
Corn, lb	.00850	4.5	19		
Ammonium Sulfate, lb	.44400	3.0	666		
Superphosphate(46), lb	1.91375	8.0	7,655		
Calcium Carbonate, lb	.88034	2.7	1,189		
Catalyst & Chemicals			1,950		

TOTAL RAW MATERIALS			42,289	84.57	1864.61
UTILITIES					
Power, KWH	1.76310	4.3	3,791		
Cooling Water, M Gal	.27023	5.8	784		
Process Water, M Gal	.02988	60.0	896		
Steam,50 psig, M Lb	.02405	295.0	3,547		
Steam,200 psig, M Lb	.06309	299.0	9,432		

TOTAL UTILITIES			18,450	36.90	813.51
OPERATING COSTS					
Labor, 60 Men @ $ 25,500	13 M/S		1,530		
Foremen, 13 Men @ $ 29,000	2 M/S		377		
Supervision, 3 Man @ $ 35,000	3 Man		105		
Maint., Material & Labor	6% of ISBL		5,166		

TOTAL OPERATING COST			7,178	14.36	316.49
OVERHEAD EXPENSES					
Direct Overhead	45% Lab. & Sup.		905		
Gen. Plant Overhead	65% Oper. Costs		4,666		
Insurance, Prop. Tax	1.5% Tot. Fix. Inv.		2,677		

TOTAL OVERHEAD EXPENSES			8,249	16.50	363.70
BY-PRODUCT CREDIT					
Carbon Dioxide, lb	8.12319	2.8	-11,373		
SCP, lb	.03051	15.0	-229		

TOTAL BY-PRODUCT CREDIT			-11,602	-23.20	-511.55
			==========	=======	==========
CASH COST OF PRODUCTION			64,564	129.12	2846.76
DEPRECIATION	20% ISBL + 10% OSBL		26,460	52.92	1166.67
			==========	=======	==========
NET COST OF PRODUCTION			91,024	182.04	4013.43
REQUIRED SALES PRICE AT 10% DCF				242.2	5339.7

reduction in capital-related expenses due to the decreased volume of material being processed. Although the sugar concentration obtained following enzyme hydrolysis is higher (5.8 versus 5.4%) than for the base case, no overall savings in steam consumption is realized. This is because less steam is made from solid waste streams, since there is no unconverted cellulose at 100 percent yield.

TABLE III-F-3

SUMMARY OF PROCESS ECONOMICS AND ENERGY REQUIREMENTS FOR LOW YIELD/DUAL ENZYME ABE FERMENTATION

Basis: 50MM gal/year, U.S. Gulf Coast, Mid-1982

	CSI Utilities		DOE Utilities	
	Dual Enzyme	Base	Dual Enzyme	Base
Investment, MM$				
Battery limits	86.1	92.8	86.1	92.8
Offsites	92.4	97.3	92.4	97.3
Total fixed investment	178.5	190.1	178.5	190.1
Cost of production, ¢/gal				
Raw materials	84.57	86.59	84.57	86.59
Utilities	43.39	43.37	36.90	37.00
Operating costs	14.36	15.16	14.36	15.16
Overhead expenses	16.50	17.37	16.50	17.37
By-product credit	(23.20)	(23.20)	(23.20)	(23.20)
Cash cost of production	135.61	139.28	129.12	132.91
Depreciation	52.92	56.58	52.92	56.58
Net cost of production	188.53	195.85	182.04	189.48
Selling price at 10% DCF	249.3	259.8	242.2	252.8
Energy required, MBtu/gal of product	105.6	107.4	105.6	107.4

Based on the thermal equivalent of the electric power required (assumed to be 10,000 Btu/kwh) plus the enthalpy of the steam requirements (assuming an 85 percent generation efficiency), incorporation of mild acid prehydrolysis as a pretreatment for enzyme hydrolysis would result in an annual energy consumption of about 5.28 trillion Btu. This represents a saving of less than 2 percent of the energy consumed in the base case and, for all practical purposes, can be ignored.

G. Improved Butanol Tolerance of Cl. Acetobutylicum

Design Basis

The major economic limitation concerning ABE fermentation has been the inhibitory effect of the butanol product upon microorganism activity limiting product concentration. Cl. acetobutylicum, which is used to ferment cellulosic derived sugars to acetone, butanol and ethanol, is totally inhibited at a butanol concentration of about 1.1 weight percent in the fermentation beer. This translates to approximately 1.7-1.9 weight percent total solvents, depending upon the solvent ratio attained. Past attempts at solving the butanol toxicity problem have centered on mutation, adaptation and selection of tolerant strains of microorganisms. These attempts have yielded only marginally improved results. However, recent research at Colorado State University has defined the mechanism of butanol toxicity, and encouraging results have been obtained in mitigating the butanol toxicity problem.

The incorporation of aliphatic alcohols, such as butanol, into the cell membrane alters necessary interactions between membrane-bound protein and the fluid phospholipid environment of the membrane. The effect of the alcohols is to "tighten" the membrane or increase membrane fluidity. The incorporation of unsaturated fatty acids into the membranes serves to decrease membrane fluidity, thus hopefully offsetting the toxic effect of the alcohol. Alteration of the phospolipid environment by selective incorporation of specific unsaturated fatty acids, such as oleic and elaidic acid, increased the level of butanol tolerance during ABE fermentation by Cl. acetobutylicum.

Cl. acetobutylicum supplemented with elaidic acid obtained a butanol concentration of 1.2 weight percent and a total solvent concentration of 2.1 percent after 34.5 hours of fermentation time with a yield of 23 percent. This is approximately 10 percent better than that obtained when no fatty acid was added. Colorado State researchers have performed an extrapolation of the data, which indicates that 1.7 percent butanol and 2.9 percent total solvents are obtainable after 42 hours with a yield of

31.4 percent. Chem Systems has not reviewed the technical feasibility of this extrapolation, but has considered the economic impact of such a speculation in this analysis.

The base case fermentation conditions for the low yield case makes the following assumptions: Butanol concentration in the fermentation beer is approximately 1 percent, which translates to a total solvents concentration of 1.5 percent based upon a solvent ratio of 61.7:31.8:6.5, butanol:acetone:ethanol. Fermentation residence time is 48 hours after which 90 percent sugar conversion is attained with an overall yield on total sugar of 27.5 percent. The sugar concentration required in the fermenters to obtain these solvent concentrations is approximately 5 weight percent. Although the experimental results stated above indicate at 2.1 percent total solvents a yield of only 23 percent on total sugar at 81.4 percent conversion (28 percent yield on sugar consumed) at 34.5 hours, this yield is thought to be dependent primarily upon fermentation time. A 23 percent yield would be commercially unacceptable; therefore, it is assumed that by allowing the fermentation to continue to 42 hours, and based upon constant solvent formation rates, 100 percent sugar conversion can be realized. The only variable which controls butanol and solvent concentration at 100 percent sugar conversion in the fermentation beer is initial sugar concentration (and butanol tolerance of the microorganism). Setting the initial sugar concentration in the fermenters at approximately 6.4 weight percent yields a total solvents concentration of 2.1 percent. At a solvents ratio of 60.7:36.7:2.6, butanol:acetone:ethanol, the final butanol concentration is 1.2 percent. The yield based on total sugar is 31.4 percent at 100 percent consumption. This correlates well with the experimental case in which no fatty acid was added to the fermentation which was used as the control. Under those conditions 99.3 percent sugar conversion was attained after 39 hours with a 32.5 percent yield on total sugar. This yielded 1.9 percent total solvents and 1.1 percent butanol, the maximum obtainable before complete microorganism inhibition sets in.

As previously mentioned if approximately 8.7 weight percent sugars initially in fermentation were extrapolated to complete sugar conversion at 42 hours based upon previously attained solvent formation rates, 2.9 percent total solvents are obtained with 1.7 percent butanol in the fermentation beer. Again, this extrapolation is quite speculative since such results have not been demonstrated to date.

Economics and Energy Requirements

Tables III-G-1 and III-G-2 represent cost of production estimates for the increased butanol toxicity case at 2.1 percent total solvents, 100 percent sugar conversion, 42 hours fermentation time and a total yield of 31.4 percent. These data are summarized and compared against the (revised) base case values in Table III-G-3. The cases are for a 50 million gallon per year total solvents facility located on the U.S. Gulf Coast in mid-1983.

As can be seen from Table III-G-3, solvents can be produced for 209 and 203 cents per gallon, respectively, for CSI and DOE utilities, respectively. This represents a significant decrease in cost of production compared to the base case costs of 260 and 253 cents per gallon. The primary differences are the reduction in energy consumption and energy-related capital expenses for the increased solvent concentration case. The base case assumed 1.5 percent total solvents which in addition to having greater energy requirements for purification, significantly increases fermenter volume. This results in higher capital-related expenses and nutrient raw material requirements. These effects are further exacerbated by the yield assumptions (the base case yield is only 27.5 percent compared to 31.4 percent for the increased butanol tolerance case) which increases front end capital as well as higher wood raw material requirements.

If, as Colorado State has extrapolated, 2.9 percent total solvents were obtainable at 31.4 percent yield, even greater savings in energy consumption, nutrient raw materials, and capital-related expenses would be realized. Although the effect of continuing increases in solvent concentration in the fermentation beer will begin to level off at higher concentrations, the initial doubling or tripling of that concentration has a pronounced positive effect on lowering the cost of production.

TABLE III-G-1
COST OF PRODUCTION ESTIMATE FOR ABE PROCESS- IMPROVED BUTANOL TOL

CAPITAL SUMMARY

BASIS
Location: U.S. Gulf Coast
Mid-1982
Capacity: 50.00 million gallons/yr
22,680 metric tons/yr
Str.Time: 8000 hours per year

CAPITAL COST	$MILLION
Battery Limits	75.6
Offsites	83.7

Total Fixed Inv.	159.3
Working Capital	13.1

PRODUCTION COST SUMMARY

RAW MATERIALS	UNITS PER GAL	PRICE, ¢/UNIT	ANNUAL COST, $M	CENTS PER GAL	DOLLARS/ MET TON
Aspen, lb	54.57504	1.0	27,289		
Sulfuric Acid, lb	.24739	4.3	532		
Calcium Hydroxide, lb	.16189	2.0	162		
Sodium Hydroxide, lb	.00864	26.0	112		
Corn, lb	.00320	4.5	7		
Ammonium Sulfate, lb	.16665	3.0	250		
Superphosphate(46), lb	.72409	8.0	2,897		
Calcium Carbonate, lb	.33308	2.7	450		
Catalyst & Chemicals			1,800		

TOTAL RAW MATERIALS			33,499	66.99	1477.02
UTILITIES					
Power, kWH	1.57631	3.2	2,522		
Cooling Water, M Gal	.31868	5.8	924		
Process Water, M Gal	.01950	60.0	585		
Steam,50 psig, M Lb	.06910	391.0	13,509		

TOTAL UTILITIES			17,541	35.08	773.40
OPERATING COSTS					
Labor, 60 Men @ $ 25,500		13 M/S	1,530		
Foremen, 13 Men @ $ 29,000		2 M/S	377		
Supervision, 3 Man @ $ 35,000		3 Man	105		
Maint., Material & Labor	6% of ISBL		4,536		

TOTAL OPERATING COST			6,548	13.10	288.71
OVERHEAD EXPENSES					
Direct Overhead	45% Lab. & Sup.		905		
Gen. Plant Overhead	65% Oper. Costs		4,256		
Insurance, Prop. Tax	1.5% Tot. Fix. Inv.		2,389		

TOTAL OVERHEAD EXPENSES			7,551	15.10	332.94
BY-PRODUCT CREDIT					
Carbon Dioxide, lb	7.76488	2.8	-10,871		
SCP, lb	.03050	15.0	-229		

TOTAL BY-PRODUCT CREDIT			-11,100	-22.20	-489.43
			==========	=======	==========
CASH COST OF PRODUCTION			54,038	108.07	2382.64
DEPRECIATION	20% ISBL + 10% OSBL		23,490	46.98	1035.72
			==========	=======	==========
NET COST OF PRODUCTION			77,528	155.05	3418.36
REQUIRED SALES PRICE AT 10% DCF				208.5	4597.8

TABLE III-B-2

COST OF PRODUCTION ESTIMATE FOR ABE PROCESS- IMPROVED BUTANOL TOL

CAPITAL SUMMARY

BASIS		CAPITAL COST	$MILLION
Location:	U.S. Gulf Coast	Battery Limits	75.6
	Mid-1982	Offsites	83.7
Capacity:	50.00 million gallons/yr		------
	22,680 metric tons/yr	Total Fixed Inv.	159.3
Str.Time:	8000 hours per year	Working Capital	12.8

PRODUCTION COST SUMMARY

RAW MATERIALS	UNITS PER GAL	PRICE, ¢/UNIT	ANNUAL COST, $M	CENTS PER GAL	DOLLARS/ MET TON
Aspen, lb	54.57504	1.0	27,289		
Sulfuric Acid, lb	.24739	4.3	532		
Calcium Hydroxide, lb	.16189	2.0	162		
Sodium Hydroxide, lb	.00864	26.0	112		
Corn, lb	.00320	4.5	7		
Ammonium Sulfate, lb	.16665	3.0	250		
Superphosphate(46), lb	.72409	8.0	2,897		
Calcium Carbonate, lb	.33308	2.7	450		
Catalyst & Chemicals			1,800		

TOTAL RAW MATERIALS			33,499	66.99	1477.02
UTILITIES					
Power, kWH	1.57631	4.3	3,389		
Cooling Water, M Gal	.31868	5.8	924		
Process Water, M Gal	.01950	60.0	585		
Steam,50 psig, M Lb	.06910	295.0	10,192		

TOTAL UTILITIES			15,091	30.18	665.38
OPERATING COSTS					
Labor, 60 Men @ $ 25,500		13 M/S	1,530		
Foremen, 13 Men @ $ 29,000		2 M/S	377		
Supervision, 3 Man @ $ 35,000		3 Man	105		
Maint., Material & Labor	6% of ISBL		4,536		

TOTAL OPERATING COST			6,548	13.10	288.71
OVERHEAD EXPENSES					
Direct Overhead	45% Lab. & Sup.		905		
Gen. Plant Overhead	65% Oper. Costs		4,256		
Insurance, Prop. Tax	1.5% Tot. Fix. Inv.		2,389		

TOTAL OVERHEAD EXPENSES			7,551	15.10	332.94
BY-PRODUCT CREDIT					
Carbon Dioxide, lb	7.76488	2.8	-10,871		
SCP, lb	.03050	15.0	-229		

TOTAL BY-PRODUCT CREDIT			-11,100	-22.20	-489.43
			==========	=======	==========
CASH COST OF PRODUCTION			51,588	103.17	2274.63
DEPRECIATION	20% ISBL + 10% OSBL		23,490	46.98	1035.72
			==========	=======	==========
NET COST OF PRODUCTION			75,078	150.15	3310.35
REQUIRED SALES PRICE AT 10% DCF				203.2	4479.8

TABLE III-G-3

SUMMARY OF ABE PROCESS ECONOMICS AND ENERGY REQUIREMENTS-INCREASED BUTANOL TOLERANCE

Basis: 50 MM gal/yr, U.S. Gulf Coast, Mid-1982

	CSI Utilities		DOE Utilities	
	Increased Tolerance	Base	Increased Tolerance	Base
Investment, $MM				
Battery limits	75.6	92.8	75.6	92.8
Offsites	83.7	97.3	83.7	97.3
Total fixed investment	159.3	190.1	159.3	190.1
Cost of production, ¢/gal				
Raw materials	66.99	86.59	66.99	86.59
Utilities	35.08	43.37	30.18	37.00
Operating costs	13.10	15.16	13.10	15.16
Overhead expenses	15.10	17.37	15.10	17.37
By-product credit	(22.20)	(23.20)	(22.20)	(23.20)
Cash cost of production	108.07	139.28	103.17	132.91
Depreciation	46.98	56.58	46.98	56.58
Net cost of production	155.05	195.85	150.15	189.48
Selling price at 10% DCF	208.5	259.8	203.2	252.8
Energy required, MBtu/gal of product	89.8	107.4	89.8	107.4

Based on the thermal equivalent of the electric power required (assumed to be 10,000 Btu/kwh) plus the enthalpy of the steam requirements (assuming an 85 percent generation efficiency), incorporation of improved butanol tolerance would result in an annual energy consumption of about 4.49 trillion Btu. This represents a saving of about 16 percent of the energy required in the base case.

IV. TECHNOLOGY ASSESSMENT

A. Sensitivity of Cost of Production (COP) to Solvent Concentration in Beer

Design Basis

The factor which contributes most towards the poor economics of ABE fermentation is the toxic effect of butanol on microorganism activity, which limits product concentration during fermentation. However, this raises the issue of what the effect on process economics would be if higher solvent concentrations could be obtained, without sacrificing fermentation yield. The key question is to determine the solvent concentration at which ABE fermentation becomes competitive with synthetic routes. The encouraging results obtained at Colorado State University in understanding the mechanism of butanol toxicity and in increasing the solvent concentration have made this not merely a speculative exercise, but a realistically obtainable goal.

Experimental work at Colorado State has already obtained butanol concentrations of 1.2 weight percent (and a total solvent concentration of 2.1 percent) after 34.5 hours of fermentation. These results were obtained using Cl. acetobutylicum supplemented with elaidic acid. The results are about 10 percent better than the maximum obtained previously with Cl. acetobutylicum, i.e., 1.1 percent butanol and 1.9 percent total solvents. However, the fatty acid-supplemented fermentation obtained only a 23 percent yield after 34.5 hours, whereas the unsupplemented fermentation obtained 32.5 percent yield after 39 hours. This indicates a decrease in microorganism activity in the presence of the higher butanol concentration, however, the encouraging result is that microorganism activity was not completely inhibited. Extrapolation of the experimental data indicates that, if constant solvent formation rates are assumed, 1.7 percent butanol and 2.9 percent total solvents would be obtained after 42 hours fermentation with a yield of 31.4 percent.

Three points are available to determine the relationship between solvent beer concentration and cost of production (COP) for the base case process. Since it is desirable to analyze the effect of concentration on COP while maintaining relatively high fermentation yields, a yield for all three points of 31.4 percent is assumed. This correlates well with the unsupplemented fermentation results obtained by Colorado State as well as with the maximum yield base case analyzed previously (approximately 32.0 percent yield) for cellulosic derived sugars. The three solvent beer concentrations are 1.0 percent (0.7 percent butanol) which corresponds to the maximum yield case, 2.1 percent (1.2 percent butanol) and 2.9 percent (1.7 percent butanol). The economics for the maximum yield case have been adjusted to reflect a fermentation time of 42 hours, instead of 72 hours, to allow for a meaningful comparison of the results. The sugar concentrations initially required in fermentation to obtain the various solvent concentrations are summarized in Table IV-A-1.

TABLE IV-A-1

INITIAL FERMENTATION SUGAR CONCENTRATION REQUIRED FOR VARIOUS SOLVENT CONCENTRATIONS

Solvent Concentration (wt %)	Sugar Concentration (wt %)
1.0	3.0
2.1	6.4
2.9	8.7

In order to obtain the higher sugar concentrations in fermentation, correspondingly higher sugar concentrations must be obtained during enzyme hydrolysis by controlling water additions at various points in the front end of the process. Primarily this is done by increasing enzyme hydrolysis solids concentration, decreasing water addition during repulping and by controlling the dilution effect of the slopback recycle. Therefore the higher solvent concentration cases have higher enzyme hydrolysis solids concentration, lower repulping water addition and lower percentage of slopback recycle. These parameters affect the enzyme hydrolysis yield and front end capital requirements as well as nutrient raw material requirements. The higher the solvent concentration (and therefore, sugar concentration) the lower the capital (due to decreased volume throughput) and the lower nutrient raw material requirements (which is somewhat offset by the higher slopback obtainable for the dilute case).

The most important design parameters for the three solvent concentration cases are summarized in Table IV-A-2.

TABLE IV-A-2

SUMMARY OF ABE DESIGN PARAMETERS

Solvents, wt %	1.0	2.1	2.9
Butanol, wt %	0.7	1.2	1.7
Repulping solids, wt %	8.0	8.33	10.7
Enzyme hydrolysis solids, wt %	8.0	12.0	12.0
Fermentation yield, %	31.4	31.4	31.4
Fermentation time, hours	42	42	42
Slopback, %	50	9.7	9.4
Solvent ratio, wt %	72:25:3	60.7:36.7:2.6	60.7:36.7:2.6
Sugar conversion, %	97*	100	100

*Ideally, it would be desirable to compare each case at the same sugar conversion level. Unfortunately, the analysis is limited by the data available. However, the difference between 97 percent and 100 percent is thought to be insignificant.

Economics and Energy Requirements

Tables IV-A-3 through IV-A-8 represent cost of production estimates for ABE fermentation at the three solvent concentrations and design parameters discussed previously: 1.0, 2.1, and 2.9 weight percent. Both CSI and DOE utility costs are used for each case. These data are summarized in Table IV-A-9. All cases are for a 50 million gallon per year total solvents facility located on the U.S. Gulf Coast in mid-1982.

TABLE IV-A-3

COST OF PRODUCTION ESTIMATE FOR ABE PROCESS- MAX YIELD

CAPITAL SUMMARY

BASIS		CAPITAL COST	$MILLION
Location:	U.S. Gulf Coast	Battery Limits	93.0
	Mid-1982	Offsites	109.9
Capacity:	50.00 million gallons/yr		------
	22,680 metric tons/yr	Total Fixed Inv.	202.9
Str.Time:	8000 hours per year	Working Capital	16.6

PRODUCTION COST SUMMARY

RAW MATERIALS	UNITS PER GAL	PRICE, ¢/UNIT	ANNUAL COST, $M	CENTS PER GAL	DOLLARS/ MET TON
Aspen, lb	54.58140	1.0	27,292		
Sulfuric Acid, lb	.24510	4.3	527		
Calcium Hydroxide, lb	.17767	2.0	178		
Sodium Hydroxide, lb	.00784	26.0	102		
Corn, lb	.01238	4.5	28		
Ammonium Sulfate, lb	.35826	3.0	537		
Superphosphate(46), lb	1.55767	8.0	6,231		
Calcium Carbonate, lb	.71632	2.7	967		
Catalyst & Chemicals			1,900		

TOTAL RAW MATERIALS			37,762	75.52	1665.02
UTILITIES					
Power, kWH	2.05031	3.2	3,281		
Cooling Water, M Gal	.32312	5.8	937		
Process Water, M Gal	.03021	60.0	906		
Steam,50 psig, M Lb	.10624	375.0	19,922		
Steam,200 psig, M Lb	.01141	381.0	2,174		

TOTAL UTILITIES			27,220	54.44	1200.20
OPERATING COSTS					
Labor, 60 men @ $ 25,500		13 M/S	1,530		
Foremen, 13 Men @ $ 29,000		2 M/S	377		
Supervision, 3 Man @ $ 35.000		3 Man	105		
Maint., Material & Labor		6% of ISBL	5,580		

TOTAL OPERATING COST			7,592	15.18	334.75
OVERHEAD EXPENSES					
Direct Overhead	45% Lab. & Sup.		905		
Gen. Plant Overhead	65% Oper. Costs		4,935		
Insurance, Prop. Tax	1.5% Tot. Fix. Inv.		3,043		

TOTAL OVERHEAD EXPENSES			8,884	17.77	391.70
BY-PRODUCT CREDIT					
Carbon Dioxide, lb	7.76805	2.8	-10,876		
SCP, lb	.03461	15.0	-260		

TOTAL BY-PRODUCT CREDIT			-11,135	-22.27	-490.98
			==========	=======	==========
CASH COST OF PRODUCTION			70,323	140.64	3100.68
DEPRECIATION	20% ISBL + 10% OSBL		29,590	59.18	1304.68
			==========	=======	==========
NET COST OF PRODUCTION			99,913	199.82	4405.36
REQUIRED SALES PRICE AT 10% DCF				268.6	5922.5

TABLE IV-A-4

COST OF PRODUCTION ESTIMATE FOR ABE PROCESS- MAX YIELD

CAPITAL SUMMARY

BASIS		CAPITAL COST	$MILLION
Location:	U.S. Gulf Coast	Battery Limits	93.0
	Mid-1982	Offsites	109.9
Capacity:	50.00 million gallons/yr		------
	22,680 metric tons/yr	Total Fixed Inv.	202.9
Str.Time:	8000 hours per year	Working Capital	16.1

PRODUCTION COST SUMMARY

RAW MATERIALS	UNITS PER GAL	PRICE, ¢/UNIT	ANNUAL COST, $M	CENTS PER GAL	DOLLARS/ MET TON
Aspen, lb	54.58140	1.0	27,292		
Sulfuric Acid, lb	.24510	4.3	527		
Calcium Hydroxide, lb	.17767	2.0	178		
Sodium Hydroxide, lb	.00784	26.0	102		
Corn, lb	.01238	4.5	28		
Ammonium Sulfate, lb	.35826	3.0	537		
Superphosphate(46), lb	1.55767	8.0	6,231		
Calcium Carbonate, lb	.71652	2.7	967		
Catalyst & Chemicals			1,900		

TOTAL RAW MATERIALS			37,762	75.52	1665.02
UTILITIES					
Power, kWH	2.05081	4.3	4,409		
Cooling Water, M Gal	.32312	5.8	937		
Process Water, M Gal	.03021	60.0	906		
Steam,50 psig, M Lb	.10624	295.0	15,672		
Steam,200 psig, M Lb	.01141	299.0	1,706		

TOTAL UTILITIES			23,631	47.26	1041.92
OPERATING COSTS					
Labor, 60 Men @ $ 25,500		13 M/S	1,530		
Foremen, 13 Men @ $ 29,000		2 M/S	377		
Supervision, 3 Man @ $ 35,000		3 Man	105		
Maint., Material & Labor		6% of ISBL	5,580		

TOTAL OPERATING COST			7,592	15.18	334.75
OVERHEAD EXPENSES					
Direct Overhead	45% Lab. & Sup.		905		
Gen. Plant Overhead	65% Oper. Costs		4,935		
Insurance, Prop. Tax	1.5% Tot. Fix. Inv.		3,043		

TOTAL OVERHEAD EXPENSES			8,884	17.77	391.70
BY-PRODUCT CREDIT					
Carbon Dioxide, lb	7.76805	2.8	-10,876		
SCP, lb	.03461	15.0	-260		

TOTAL BY-PRODUCT CREDIT			-11,135	-22.27	-490.98
			==========	=======	==========
CASH COST OF PRODUCTION			66,733	133.46	2942.40
DEPRECIATION	20% ISBL + 10% OSBL		29,590	59.18	1304.68
			==========	=======	==========
NET COST OF PRODUCTION			96,323	192.64	4247.08
REQUIRED SALES PRICE AT 10% DCF				260.8	5749.6

TABLE IV-A-5

COST OF PRODUCTION ESTIMATE FOR ABE
PROCESS- IMPROVE BUT TOL-2.1

CAPITAL SUMMARY

BASIS		CAPITAL COST	$MILLION
Location:	U.S. Gulf Coast	Battery Limits	75.6
	Mid-1982	Offsites	93.7
Capacity:	50.00 million gallons/yr		------
	22,680 metric tons/yr	Total Fixed Inv.	159.3
Str.Time:	8000 hours per year	Working Capital	13.4

PRODUCTION COST SUMMARY

RAW MATERIALS	UNITS PER GAL	PRICE, ¢/UNIT	ANNUAL COST, $M	CENTS PER GAL	DOLLARS/ MET TON
Aspen, lb	54.57504	1.0	27,289		
Sulfuric Acid, lb	.24739	4.3	532		
Calcium Hydroxide, lb	.16189	2.0	162		
Sodium Hydroxide, lb	.00864	26.0	112		
Corn, lb	.01237	4.5	28		
Ammonium Sulfate, lb	.28664	3.0	430		
Superphosphate(46), lb	1.24623	8.0	4,985		
Calcium Carbonate, lb	.57327	2.7	774		
Catalyst & Chemicals			1,800		

TOTAL RAW MATERIALS			36,112	72.22	1592.26
UTILITIES					
Power, kWH	1.57631	3.2	2,522		
Cooling Water, M Gal	.31868	5.8	924		
Process Water, M Gal	.01950	60.0	585		
Steam,50 psig, M Lb	.06910	391.0	13,509		

TOTAL UTILITIES			17,541	35.08	773.40
OPERATING COSTS					
Labor, 60 Men @ $ 25,500		13 M/S	1,530		
Foremen, 13 Men @ $ 29,000		2 M/S	377		
Supervision, 3 Man @ $ 35,000		3 Man	105		
Maint., Material & Labor	6% of ISBL		4,536		

TOTAL OPERATING COST			6,548	13.10	288.71
OVERHEAD EXPENSES					
Direct Overhead	45% Lab. & Sup.		905		
Gen. Plant Overhead	65% Oper. Costs		4,256		
Insurance, Prop. Tax	1.5% Tot. Fix. Inv.		2,389		

TOTAL OVERHEAD EXPENSES			7,551	15.10	332.94
BY-PRODUCT CREDIT					
Carbon Dioxide, lb	7.91237	2.8	-11,078		
SCP, lb	.03050	15.0	-229		

TOTAL BY-PRODUCT CREDIT			-11,307	-22.61	-498.53
			==========	=======	==========
CASH COST OF PRODUCTION			56,445	112.88	2488.78
DEPRECIATION	20% ISBL + 10% OSBL		23,490	46.98	1035.72
			==========	=======	==========
NET COST OF PRODUCTION			79,935	159.86	3524.50
REQUIRED SALES PRICE AT 10% DCF				213.8	4713.0

TABLE IV-A-6

COST OF PRODUCTION ESTIMATE FOR ABE
PROCESS- IMPROVE BUT TOL-2.1

CAPITAL SUMMARY

BASIS		CAPITAL COST	$MILLION
Location:	U.S. Gulf Coast	Battery Limits	75.6
	Mid-1982	Offsites	83.7
Capacity:	50.00 million gallons/yr		------
	22,680 metric tons/yr	Total Fixed Inv.	159.3
Str.Time:	8000 hours per year	Working Capital	13.0

PRODUCTION COST SUMMARY

RAW MATERIALS	UNITS PER GAL	PRICE, c/UNIT	ANNUAL COST, $M	CENTS PER GAL	DOLLARS/ MET TON
Aspen, lb	54.57504	1.0	27,289		
Sulfuric Acid, lb	.24739	4.3	532		
Calcium Hydroxide, lb	.16189	2.0	162		
Sodium Hydroxide, lb	.00864	26.0	112		
Corn, lb	.01237	4.5	28		
Ammonium Sulfate, lb	.28664	3.0	430		
Superphosphate(46), lb	1.24623	8.0	4,985		
Calcium Carbonate, lb	.57327	2.7	774		
Catalyst & Chemicals			1,800		

TOTAL RAW MATERIALS			36,112	72.22	1592.26
UTILITIES					
Power, kWH	1.57631	4.3	3,389		
Cooling Water, M Gal	.31868	5.8	924		
Process Water, M Gal	.01950	60.0	585		
Steam,50 psig, M Lb	.06910	295.0	10,192		

TOTAL UTILITIES			15,091	30.18	665.38
OPERATING COSTS					
Labor, 60 Men @ $ 25,500		13 M/S	1,530		
Foremen, 13 Men @ $ 29,000		2 M/S	377		
Supervision, 3 Man @ $ 35,000		3 Man	105		
Maint., Material & Labor		6% of ISBL	4,536		

TOTAL OPERATING COST			6,548	13.10	288.71
OVERHEAD EXPENSES					
Direct Overhead	45% Lab. & Sup.		905		
Gen. Plant Overhead	65% Oper. Costs		4,256		
Insurance, Prop. Tax	1.5% Tot. Fix. Inv.		2,389		

TOTAL OVERHEAD EXPENSES			7,551	15.10	332.94
BY-PRODUCT CREDIT					
Carbon Dioxide, lb	7.91237	2.8	-11,078		
SCP, lb	.03050	15.0	-229		

TOTAL BY-PRODUCT CREDIT			-11,307	-22.61	-498.53
			==========	=======	==========
CASH COST OF PRODUCTION			53,995	107.98	2380.76
DEPRECIATION	20% ISBL + 10% OSBL		23,490	46.98	1035.72
			==========	=======	==========
NET COST OF PRODUCTION			77,485	154.96	3416.48
REQUIRED SALES PRICE AT 10% DCF				208.4	4595.0

TABLE IV-A-7

COST OF PRODUCTION ESTIMATE FOR ABE
PROCESS- IMPROVE BUT TOL-2.9

CAPITAL SUMMARY

BASIS		CAPITAL COST	$MILLION
Location:	U.S. Gulf Coast	Battery Limits	71.5
	Mid-1982	Offsites	74.8
Capacity:	50.00 million gallons/yr		------
	22,680 metric tons/yr	Total Fixed Inv.	146.3
Str.Time:	8000 hours per year	Working Capital	12.3

PRODUCTION COST SUMMARY

RAW MATERIALS	UNITS PER GAL	PRICE, ¢/UNIT	ANNUAL COST, $M	CENTS PER GAL	DOLLARS/ MET TON
Aspen, lb	55.07302	1.0	27,538		
Sulfuric Acid, lb	.24739	4.3	532		
Calcium Hydroxide, lb	.16095	2.0	161		
Sodium Hydroxide, lb	.01488	26.0	193		
Corn, lb	.01237	4.5	28		
Ammonium Sulfate, lb	.20636	3.0	310		
Superphosphate(46), lb	.89715	8.0	3,589		
Calcium Carbonate, lb	.41268	2.7	557		
Catalyst & Chemicals			1,800		

TOTAL RAW MATERIALS			34,708	69.41	1530.33
UTILITIES					
Power, KWH	1.30193	3.2	2,083		
Cooling Water, M Gal	.30938	5.8	897		
Process Water, M Gal	.01456	60.0	437		
Steam,50 psig, M Lb	.05455	391.0	10,665		

TOTAL UTILITIES			14,082	28.16	620.89
OPERATING COSTS					
Labor, 60 Men @ $ 25,500		13 M/S	1,530		
Foremen, 13 Men @ $ 29,000		2 M/S	377		
Supervision, 3 Man @ $ 35,000		3 Man	105		
Maint., Material & Labor		6% of ISBL	4,290		

TOTAL OPERATING COST			6,302	12.60	277.87
OVERHEAD EXPENSES					
Direct Overhead	45% Lab. & Sup.		905		
Gen. Plant Overhead	65% Oper. Costs		4,096		
Insurance, Prop. Tax	1.5% Tot. Fix. Inv.		2,194		

TOTAL OVERHEAD EXPENSES			7,196	14.39	317.29
BY-PRODUCT CREDIT					
Carbon Dioxide, lb	7.91237	2.8	-11,078		
SCP, lb	.03050	15.0	-229		

TOTAL BY-PRODUCT CREDIT			-11,307	-22.61	-498.53
			==========	=======	==========
CASH COST OF PRODUCTION			50,981	101.96	2247.85
DEPRECIATION	20% ISBL + 10% OSBL		21,780	43.56	960.32
			==========	=======	==========
NET COST OF PRODUCTION			72,761	145.51	3208.18
REQUIRED SALES PRICE AT 10% DCF				194.5	4288.3

TABLE IV-A-8

COST OF PRODUCTION ESTIMATE FOR ABE
PROCESS- IMPROVE BUT TOL-2.9

CAPITAL SUMMARY

BASIS		CAPITAL COST	$MILLION
Location:	U.S. Gulf Coast	Battery Limits	71.5
	Mid-1982	Offsites	74.8
Capacity:	50.00 million gallons/yr		------
	22,680 metric tons/yr	Total Fixed Inv.	146.3
Str.Time:	8000 hours per year	Working Capital	12.1

PRODUCTION COST SUMMARY

RAW MATERIALS	UNITS PER GAL	PRICE, c/UNIT	ANNUAL COST, $M	CENTS PER GAL	DOLLARS/ MET TON
Aspen, lb	55.07302	1.0	27,538		
Sulfuric Acid, lb	.24739	4.3	532		
Calcium Hydroxide, lb	.16095	2.0	161		
Sodium Hydroxide, lb	.01488	26.0	193		
Corn, lb	.01237	4.5	28		
Ammonium Sulfate, lb	.20636	3.0	310		
Superphosphate(46), lb	.89715	8.0	3,589		
Calcium Carbonate, lb	.41268	2.7	557		
Catalyst & Chemicals			1,800		

TOTAL RAW MATERIALS			34,708	69.41	1530.33
UTILITIES					
Power, kWH	1.30193	4.3	2,799		
Cooling Water, M Gal	.30938	5.8	897		
Process Water, M Gal	.01456	60.0	437		
Steam,50 psig, M Lb	.05455	295.0	8,046		

TOTAL UTILITIES			12,179	24.36	537.02
OPERATING COSTS					
Labor, 60 Men @ $ 25,500		13 M/S	1,530		
Foremen, 13 Men @ $ 29,000		2 M/S	377		
Supervision, 3 Man @ $ 35,000		3 Man	105		
Maint., Material & Labor		6% of ISBL	4,290		

TOTAL OPERATING COST			6,302	12.60	277.87
OVERHEAD EXPENSES					
Direct Overhead	45% Lab. & Sup.		905		
Gen. Plant Overhead	65% Oper. Costs		4,096		
Insurance, Prop. Tax	1.5% Tot. Fix. Inv.		2,194		

TOTAL OVERHEAD EXPENSES			7,196	14.39	317.29
BY-PRODUCT CREDIT					
Carbon Dioxide, lb	7.91237	2.8	-11,078		
SCP, lb	.03050	15.0	-229		

TOTAL BY-PRODUCT CREDIT			-11,307	-22.61	-498.53
			=========	=======	=========
CASH COST OF PRODUCTION			49,079	98.15	2163.98
DEPRECIATION	20% ISBL + 10% OSBL		21,780	43.56	960.32
			=========	=======	=========
NET COST OF PRODUCTION			70,859	141.71	3124.30
REQUIRED SALES PRICE AT 10% DCF				190.4	4196.7

TABLE IV-A-9

SUMMARY OF ABE FERMENTATION PROCESS ECONOMICS AND ENERGY REQUIREMENTS

Basis: U.S. Gulf Coast, mid-1982, 50 million gallons per year.

	Fermenter Effluent Product Concentration		
	1.0 Percent Solvents	2.1 Percent Solvents	2.9 Percent Solvents
Investment, MM$			
Battery limits	93.0	75.6	71.5
Offsites	109.9	83.7	74.8
Total fixed investment	202.9	159.3	146.3
Cost of production, ¢/gal			
Raw materials	75.52	72.22	69.41
Utilities	54.44	35.08	28.16
Operating costs	15.18	13.10	12.60
Overhead expenses	17.77	15.10	14.39
By-product credit	(22.27)	(22.61)	(22.61)
Cash cost of production	140.64	112.88	101.96
Depreciation	59.18	46.98	43.56
Net cost of production	199.82	159.86	145.51
Selling price at 10% DCF	268.6	213.8	194.5
Energy required, MBtu/gal of product	145.6	89.8	71.6

As can be seen from Table IV-A-9, COP is reduced from 268.6 cents per gallon at 1.0 percent solvents, to 213.8 cents per gallon at 2.1 percent solvent, to 194.5 at 2.9 percent. These values represent the sales price required at 10 percent DCF, using CSI utility values. As solvent concentration is increased there are several relationships that occur which contribute to the decreasing cost of production:

- Inside Battery Limits (ISBL) capital is reduced due to the decreasing water content of the process streams, which results in reduced equipment volume requirements.

- The corresponding reduction in fermenter volume results in reduced nutrient requirements for the higher solvent concentration cases, although this is somewhat offset by the larger slopback percentage (thus recycling nutrients) for the 1.0 percent case.

- The decreased steam requirement during purification due to the higher solvent concentrations results in reduced utility cost and Outside Battery Limits (OSBL) capital (i.e., smaller coal-fired steam system required).

- The overall decreased capital results in an across-the-board reduction in capital-related expenses (operating costs, overheads, depreciation and DCF return).

Aspen wood raw material requirements are essentially the same for all cases due to the assumed constant enzyme hydrolysis and fermentation yield, however, there are some slight variations due to sugar recovery differences for different repulping solids concentrations.

Figure IV-A-1 represents the relationship between total solvent concentration and sales price at 10 percent DCF for ABE fermentation given the design basis outlined previously. The graph indicates that the greatest savings per unit of increase in solvent concentration are afforded in the low concentration range, between 0.5 and around 2.5 percent total solvents. At greater than 2.5 percent the effect of increasing solvents concentration begins to level off, although it does continue to decrease. For example, at 3.5 percent, COP is approximately 186.0 cents per gallon, only 8.5 cents per gallon lower than at 2.9 percent. However, if the results of this analysis are compared to the COP of synthetic butanol and acetone, it is seen that ABE fermentation becomes competitive at around 2.1 percent solvents, and at 2.9 percent solvents offers a significant advantage over the synthetic route.

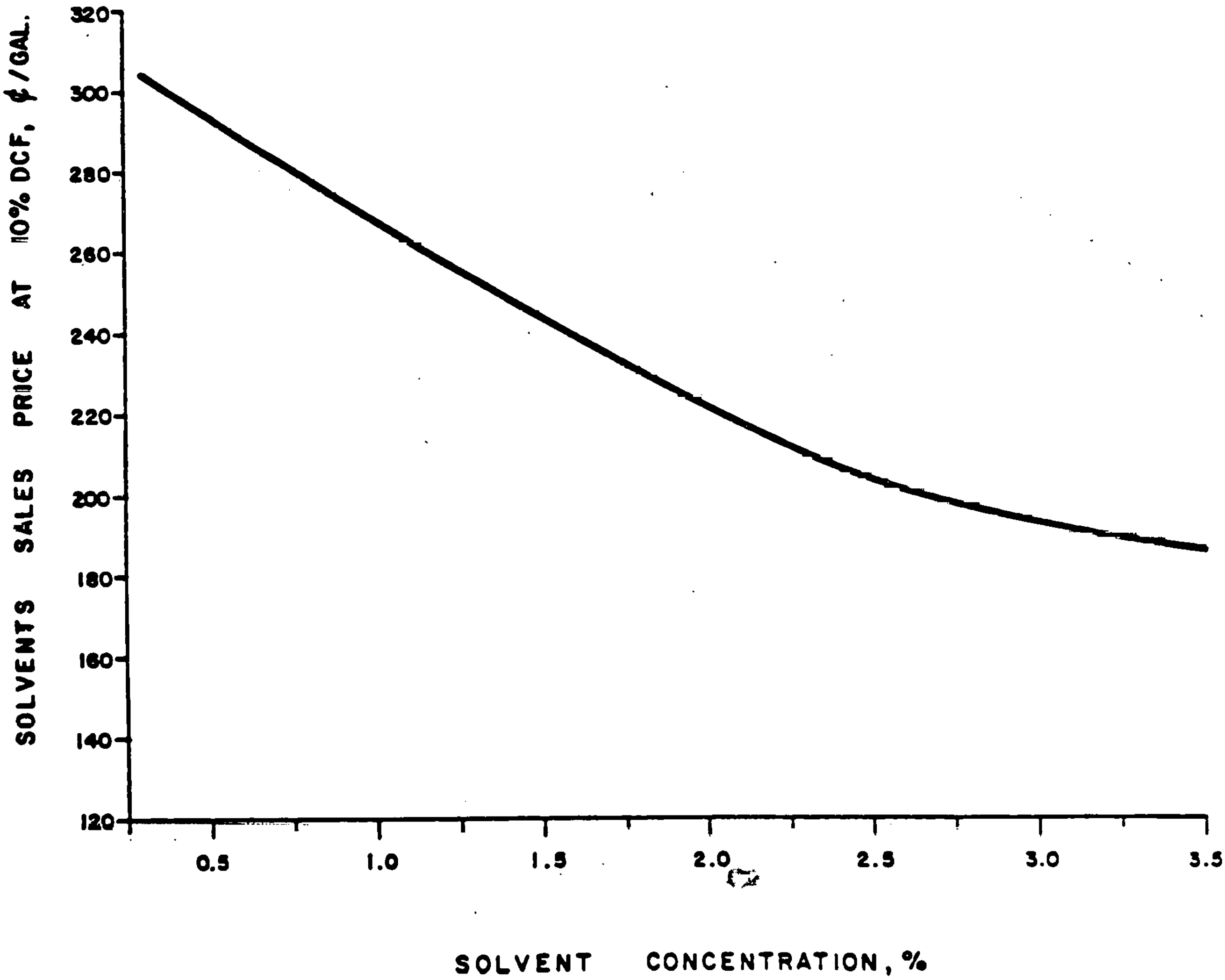
Based on the thermal equivalent of the electric power required (assumed to be 10,000 Btu/kwh) plus the enthalpy of the steam requirements (assuming an 85 percent generation efficiency), improving the butanol tolerance to the levels examined herein would result in annual energy consumption levels as follows:

	Fermenter Effluent Product Concentration		
	1.0 Percent	2.1 Percent	2.9 Percent
Annual energy consumption (Btu x 10^{12})	7.28	4.49	3.58

FIGURE IV-A-1

ABE FERMENTATION PROCESS ECONOMICS VERSUS SOLVENT CONCENTRATION IN FERMENTATION BEER

BASIS: 50 MM GALS. / YR.
LOW YIELD CASE
U.S. GULF COAST
MID. 1982



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Recalling that the base case (i.e., lower yield case) exhibited an annual energy consumption level of 5.37 trillion Btu, the speculative 2.9 percent total solvents case exhibits an impressive saving of one third of the total energy required in the base case.

B. Cumulative Effect of Most Promising Research Options

Design Basis

Six areas of potential process improvements were selected and analyzed in Section III relative to their effect on ABE process economics. This analysis was based on the most recent research data available with the objective being the selection of the most promising research areas which could have the greatest impact on ABE process economics. The options analyzed were as follows:

- Continuous ABE fermentation
- Vacuum ABE fermentation
- Baelene solvent extraction
- HRI's Lignol Process
- Improved prehydrolysis/dual enzyme hydrolysis
- Improved microorganism tolerance to butanol toxicity

Of the six options analyzed, four resulted in improved process economics. One of these, improved microorganism tolerance to butanol toxicity, had a significant positive effect and is analyzed separately. The other three, continuous fermentation, Lignol and prehydrolysis/dual enzyme hydrolysis, each had only a marginally positive effect on economics. However, a new ABE fermentation facility could incorporate all three of these process improvements to take advantage of the cumulative effect of each.

An analysis of the base case low yield case incorporating continuous fermentation, Lignol and dual enzyme hydrolysis has been performed. This case has an initial sugar concentration in fermentation of 5.2 percent, which results in a total solvents concentration of 1.4 percent. Slopback recycle is about 8 percent.

The design parameters for this case are summarized in Tables IV-B-1 and IV-B-2 and are identical to the individual cases analyzed previously in Section III.

Economics and Energy Requirements

Tables IV-B-3 and IV-B-4 represent cost of production estimates for the "cumulative" case, incorporating continuous fermentation, Lignol and dual enzyme process options at 1.4 percent total solvents in the fermentation beer, using CSI and DOE utilities, respectively. These data are summarized and compared against (revised) base case values in Table IV-B-5. This case is for a facility producing 50 million gallons per year of total solvents at a plant located on the U.S. Gulf Coast in mid-1982.

TABLE IV-B-5

SUMMARY OF ABE PROCESS ECONOMICS AND ENERGY REQUIREMENTS, CONTINUOUS FERMENTATION, LIGNOL, DUAL ENZYME

Basis: U.S. Gulf Coast, mid-1982, 50 million gallons per year.

	CSI Utilities		DOE Utilities	
	Cumulative	Base	Cumulative	Base
Investment, MM$				
Battery limits	118.4	92.8	118.4	92.8
Offsites	88.3	97.3	88.3	97.3
Total fixed investment	206.7	190.1	206.7	190.1
Cost of production, ¢/gal				
Raw materials	104.68	86.59	104.68	86.59
Utilities	38.27	43.37	34.18	37.00
Operating costs	18.23	15.16	18.23	15.16
Overhead expenses	19.86	17.37	19.86	17.37
By-product credit	(71.02)	(23.20)	(71.02)	(23.20)
Cash cost of production	110.01	139.28	105.92	132.91
Depreciation	65.02	56.58	65.02	56.58
Net cost of production	175.03	195.85	170.94	189.48
Selling price at 10% DCF	237.5	259.8	233.0	252.8
Energy required, MBtu/gal of product	69.4	107.4	69.4	107.4

TABLE IV-8-1

DESIGN PARAMETERS

Prehydrolysis

• Temperature	200°C
• Acid concentration	0.5 Wt%
• Residence time	7.9 seconds

Enzyme production

	RUT-C-30	QM 329
Temperature	86°F	86°F
pH	4.8	3.0
Pressure	Atmospheric	Atmospheric
Nutrients	(1)	(1)
Oxygen	17,264 IU/mol O_2	17,264 IU/mol O_2
Cell concentration	7 gms/liter	7 gms/liter
Cell yield	0.26 gms mycelium/ gm cellulose	0.26 gms mycelium/ gm cellulose
Enzyme yield	1.53 gms enzyme/ gm cellulose based upon enzyme productivity of 114 IU/l/hr	7.17 gms enzyme/ gm cellulose based upon enzyme productivity of 114 IU/l/hr
Cell recycle	0.77 gm/gm cells	0.77 gm/gm cells

(1) 1.0 percent cellulose
0.2 percent KH_2PO_4
0.03 percent $CaCl_2$
0.03 percent $MgSO_4 \cdot 7\ H_2O$
1.0 percent corn steep liquor

Enzyme Hydrolysis

• Temperature	122°F
• Pressure	Atmospheric
• pH	4.8
• Hydrolysis time	12.5 hours
• Hydrolysis conversion	100 mol percent conversion cellulose to glucose
• Terminal sugar concentration	5.82 percent
• Enzyme loading	17 IU/gm solids

Continuous fermentation

Sugar utilization	95 percent
Solvent yield	26.3 percent
Temperature	37°C
Sugar concentration	52 Wt%
pH	5.0
Cell mass concentration	4.5-5 g/liter
Residence time	5 hours
Volumetric productivity	2.5 g/l-hr

End Product Distribution

	Wt Percent
Butanol	57.0
Acetone	35.0
Ethanol	8.0
	100.0

TABLE IV-B-2

LIGNOL PROCESS DESIGN PARAMETERS

- Overall process yields: (weight percent based upon original lignin)

  20.2 to phenol
  14.1 to benzene
  13.1 to fuel gas
  29.1 to fuel oil

Gaseous Hydrocarbon Composition

	Organic Lignin (Wt %)
CO	3.9
CO_2	1.8
CH_4	6.6
C_2H_6	1.9
C_3H_8	7.3
C_4H_8	1.0
C_4H_{10}	1.0
C_5H_{10}	0.7
C_5H_{12}	1.0
Total hydrocarbon gases	25.2

Liquid Hydrocarbon Composition

	Organic Lignin (Wt %)
H_2O	17.9
Hydrocarbons (C_6-300°F)	8.3
Hydrocarbons (300-465°F)	5.7
Phenols (300-465°F)	36.5
Catechols (465-500°F)	8.7
Heavies (500°F +)	2.4
Total liquid hydrocarbons + phenols	62.6
Total	105.7
H_2 consumption	5.7

Fraction (Wt %) Composition of Phenol Fraction

Phenol	6.5
o-Cresol	3.6
m-p-Cresols	21.6
2,4 Xylenol	7.0
p-Ethylphenol	33.2
o-n-Propylphenol	7.9
p-n-Propylphenol	20.1

TABLE IV-B-3

COST OF PRODUCTION ESTIMATE FOR ABE PROCESS- CUMULATIVE

CAPITAL SUMMARY

BASIS		CAPITAL COST	$MILLION
Location:	U.S. Gulf Coast	Battery Limits	118.4
	Mid-1982	Offsites	88.3
Capacity:	50.00 million gallons/yr		------
	22,680 metric tons/yr	Total Fixed Inv.	206.7
Str.Time:	8000 hours per year	Working Capital	16.7

PRODUCTION COST SUMMARY

RAW MATERIALS	UNITS PER GAL	PRICE, ¢/UNIT	ANNUAL COST, $M	CENTS PER GAL	DOLLARS/ MET TON
Aspen, lb	61.60486	1.0	30,804		
Sulfuric Acid, lb	.27669	4.3	595		
Calcium Hydroxide, lb	.17499	2.0	175		
Sodium Hydroxide, lb	.00880	26.0	114		
Corn, lb	.01237	4.5	28		
Ammonium Sulfate, lb	.45208	3.0	678		
Superphosphate(46), lb	1.96554	8.0	7,863		
Calcium Carbonate, lb	.90415	2.7	1,221		
Hydrogen, lb	.28878	60.0	8,664		
Catalyst & Chemicals			2,200		

TOTAL RAW MATERIALS			52,342	104.68	2307.85
UTILITIES					
Power, kWH	2.25648	4.3	4,852		
Cooling Water, M Gal	.39068	5.8	1,133		
Process Water, M Gal	.03360	60.0	1,008		
Steam,50 psig, M Lb	.06845	295.0	10,096		

TOTAL UTILITIES			17,089	34.18	753.4[illegible]
OPERATING COSTS					
Labor, 60 Men @ $ 25,500		13 M/S	1,530		
Foremen, 13 Men @ $ 29,000		2 M/S	377		
Supervision, 3 Man @ $ 35,000		3 Man	105		
Maint., Material & Labor		6% of ISBL	7,104		

TOTAL OPERATING COST			9,116	18.23	401.[illegible]
OVERHEAD EXPENSES					
Direct Overhead	45% Lab. & Sup.		905		
Gen. Plant Overhead	65% Oper. Costs		5,925		
Insurance, Prop. Tax	1.5% Tot. Fix. Inv.		3,100		

TOTAL OVERHEAD EXPENSES			9,931	19.86	437.8[illegible]
BY-PRODUCT CREDIT					
Carbon Dioxide, lb	8.00157	2.8	-11,203		
SCP, lb	.03000	15.0	-225		
Phenol, lb	1.01930	32.5	-16,565		
Benzene, lb	.72666	20.7	-7,521		

TOTAL BY-PRODUCT CREDIT			-35,514	-71.02	-1565.87
			==========	=======	==========
CASH COST OF PRODUCTION			52,964	105.92	2335.30
DEPRECIATION	20% ISBL + 10% OSBL		32,510	65.02	1433.43
			==========	=======	==========
NET COST OF PRODUCTION			85,474	170.94	3768.73
REQUIRED SALES PRICE AT 10% DCF				**233.0**	**5137.6**

TABLE IV-B-4

COST OF PRODUCTION ESTIMATE FOR ABE PROCESS- CUMULATIVE

CAPITAL SUMMARY

BASIS		CAPITAL COST	$MILLION
Location:	U.S. Gulf Coast	Battery Limits	118.4
	Mid-1982	Offsites	88.3
Capacity:	50.00 million gallons/yr		------
	22,680 metric tons/yr	Total Fixed Inv.	206.7
Str.Time:	8000 hours per year	Working Capital	17.0

PRODUCTION COST SUMMARY

RAW MATERIALS	UNITS PER GAL	PRICE, c/UNIT	ANNUAL COST, $M	CENTS PER GAL	DOLLARS/ MET TON
Aspen, lb	61.60486	1.0	30,804		
Sulfuric Acid, lb	.27669	4.3	595		
Calcium Hydroxide, lb	.17499	2.0	175		
Sodium Hydroxide, lb	.00880	26.0	114		
Corn, lb	.01237	4.5	28		
Ammonium Sulfate, lb	.45208	3.0	678		
Superphosphate(46), lb	1.96554	8.0	7,863		
Calcium Carbonate, lb	.90415	2.7	1,221		
Hydrogen, lb	.28878	60.0	8,664		
Catalyst & Chemicals			2,200		

TOTAL RAW MATERIALS			52,342	104.68	2307.85
UTILITIES					
Power, kWH	2.25648	3.2	3,611		
Cooling Water, M Gal	.39068	5.8	1,133		
Process Water, M Gal	.03360	60.0	1,008		
Steam,50 psig, M Lb	.06845	391.0	13,382		

TOTAL UTILITIES			19,134	38.27	843.64
OPERATING COSTS					
Labor, 60 Men @ $ 25,500		13 M/S	1,530		
Foremen, 13 Men @ $ 29,000		2 M/S	377		
Supervision, 3 Man @ $ 35,000		3 Man	105		
Maint., Material & Labor		6% of ISBL	7,104		

TOTAL OPERATING COST			9,116	18.23	401.[illegible]
OVERHEAD EXPENSES					
Direct Overhead	45% Lab. & Sup.		905		
Gen. Plant Overhead	65% Oper. Costs		5,925		
Insurance, Prop. Tax	1.5% Tot. Fix. Inv.		3,100		

TOTAL OVERHEAD EXPENSES			9,931	19.86	437.89
BY-PRODUCT CREDIT					
Carbon Dioxide, lb	8.00157	2.8	¯11,203		
SCP, lb	.03000	15.0	¯225		
Phenol, lb	1.01930	32.5	¯16,565		
Benzene, lb	.72666	20.7	¯7,521		

TOTAL BY-PRODUCT CREDIT			¯35,514	¯71.02	¯1565.87
			==========	=======	==========
CASH COST OF PRODUCTION			55,009	110.01	2425.45
DEPRECIATION	20% ISBL + 10% OSBL		32,510	65.02	1433.43
			==========	=======	==========
NET COST OF PRODUCTION			87,519	175.03	3858.88
REQUIRED SALES PRICE AT 10% DCF				237.5	5236.1

As can be seen from Table IV-8-5, mixed solvents can be produced for 237.5 and 233.0 cents per gallon at 10 percent DCF, respectively, for the cumulative case. This is a reduction of approximately 22.0 cents per gallon from the base case low yield COP of 258.5 cents per gallon, and is approximately equal to the sum of the reductions in COP for each of the individual cases analyzed previously.

Based on the thermal equivalent of the electric power required (assumed to be 10,000 Btu/kwh) plus the enthalpy of the steam requirements (assuming an 85 percent generation efficiency), minus the energy credit accruing from the production of benzene and phenol as by-products of the Lignol process, incorporation of the "cumulative" case would result in an annual energy consumption of 3.47 trillion Btu. This represents a significant decrease in energy consumption of 35 percent relative to the base case.

C. "Best Case" - Cumulative Research Options Plus 2.9 Percent Solvents

Design Basis

The case which would offer the best economics for ABE fermentation would be a combination of all the potential research improvements incorporated into one design. This would include continuous fermentation, Lignol and dual enzyme hydrolysis processing options as well as improved microorganism tolerance to butanol toxicity, bringing the solvent concentration in the beer to 2.9 weight percent. This level of 2.9 percent solvents is arbitrarily chosen as the maximum realistic concentration obtainable in the near future. All the design parameters from the previous analysis of these process options are identical, except that it is assumed that continuous fermentation can achieve a yield of 31.4 percent (instead of 26.3 percent) at 100 percent sugar conversion in the 5 hours previously used for continuous fermentation. Initial sugar concentration is 8.7 percent in fermentation and 9.0 percent slopback recycle is used.

Economics and Energy Requirements

Tables IV-C-1 and IV-C-2 represent a cost of production (COP) estimate for the "best case" ABE fermentation using CSI and DOE utility costs, respectively. These data are summarized and compared against (revised) base case values in Table IV-C-3. The analysis is for a facility producing 50 million gallons per year of mixed solvents at a U.S. Gulf Coast location in mid-1982.

TABLE IV-C-3

SUMMARY OF ABE FERMENTATION PROCESS ECONOMICS AND ENERGY REQUIREMENTS, BEST CASE

Basis: U.S. Gulf Coast, mid-1982, 50 million gallons per year.

	CSI Utilities		DOE Utilities	
	"Best"	Base	"Best"	Base
Investment, MM$				
Battery limits	99.8	92.8	99.8	92.8
Offsites	65.0	97.3	65.0	97.3
Total fixed investment	164.8	190.1	164.8	190.1
Cost of production, ¢/gal				
Raw materials	83.16	86.59	83.16	86.59
Utilities	29.91	43.37	28.28	37.00
Operating costs	16.00	15.16	16.00	15.16
Overhead expenses	17.15	17.37	17.15	17.37
By-product credit	(64.04)	(23.20)	(64.04)	(23.20)
Cash cost of production	82.18	139.28	80.55	132.91
Depreciation	52.92	56.58	52.92	56.58
Net cost of production	135.10	195.85	133.47	189.48
Selling price at 10% DCF	183.6	259.8	181.8	252.8
Energy required, MBtu/gal of product	34.6	107.4	34.6	107.4

As can be seen from Table IV-C-3, the best case improvements afford an 11 cent per gallon savings over the 2.9 percent solvent case without the additional improvements previously analyzed. More importantly, at 183.6 cents per gallon, the "best case" ABE fermentation process offers approximately a 20 percent price advantage over the conventional synthetic routes.

TABLE IV-C-1

COST OF PRODUCTION ESTIMATE FOR ABE PROCESS- BEST

CAPITAL SUMMARY

BASIS		CAPITAL COST	$MILLION
Location:	U.S. Gulf Coast	Battery Limits	99.8
	Mid-1982	Offsites	65.0
Capacity:	50.00 million gallons/yr		-------
	22,680 metric tons/yr	Total Fixed Inv.	164.8
Str.Time:	8000 hours per year	Working Capital	13.9

PRODUCTION COST SUMMARY

RAW MATERIALS	UNITS PER GAL	PRICE, ¢/UNIT	ANNUAL COST, $M	CENTS PER GAL	DOLLARS/ MET TON
Aspen, lb	52.67715	1.0	26,340		
Sulfuric Acid, lb	.23659	4.3	509		
Calcium Hydroxide, lb	.15439	2.0	154		
Sodium Hydroxide, lb	.01302	26.0	169		
Corn, lb	.01237	4.5	23		
Ammonium Sulfate, lb	.22243	3.0	334		
Superphosphate(46), lb	.96708	8.0	3,869		
Calcium Carbonate, lb	.44486	2.7	601		
Hydrogen, lb	.24589	60.0	7,377		
Catalyst & Chemicals			2,200		

TOTAL RAW MATERIALS			41,580	83.16	1833.34
UTILITIES					
Power, kWH	1.74801	3.2	2,797		
Cooling Water, M Gal	.35158	5.8	1,020		
Process Water, M Gal	.12999	60.0	3,900		
Steam,50 psig, M Lb	.03703	391.0	7,239		

TOTAL UTILITIES			14,956	29.91	659.44
OPERATING COSTS					
Labor, 60 Men @ $ 25,500		13 M/S	1,530		
Foremen, 13 Men @ $ 29,000		2 M/S	377		
Supervision, 3 Man @ $ 35,000		3 Man	105		
Maint., Material & Labor		6% of ISBL	5,988		

TOTAL OPERATING COST			8,000	16.00	352.74
OVERHEAD EXPENSES					
Direct Overhead	45% Lab. & Sup.		905		
Gen. Plant Overhead	65% Oper. Costs		5,200		
Insurance, Prop. Tax	1.5% Tot. Fix. Inv.		2,472		

TOTAL OVERHEAD EXPENSES			8,577	17.15	378.19
BY-PRODUCT CREDIT					
Carbon Dioxide, lb	8.00157	2.8	-11,203		
SCP, lb	.03000	15.0	-225		
Phenol, lb	.87145	32.5	-14,162		
Benzene, lb	.62127	20.7	-6,430		

TOTAL BY-PRODUCT CREDIT			-32,020	-64.04	-1411.83
			==========	======	========
CASH COST OF PRODUCTION			41,093	82.18	1311.88
DEPRECIATION	20% ISBL + 10% OSBL		26,460	52.92	1166.67
			==========	======	========
NET COST OF PRODUCTION			67,553	135.10	2978.56
REQUIRED SALES PRICE AT 10% DCF				183.6	4047.1

TABLE IV-C-2

COST OF PRODUCTION ESTIMATE FOR ABE PROCESS- BEST

CAPITAL SUMMARY

BASIS		CAPITAL COST	$MILLION
Location:	U.S. Gulf Coast	Battery Limits	99.8
	Mid-1982	Offsites	65.0
Capacity:	50.00 million gallons/yr		------
	22,680 metric tons/yr	Total Fixed Inv.	164.8
Str.Time:	8000 hours per year	Working Capital	13.7

PRODUCTION COST SUMMARY

RAW MATERIALS	UNITS PER GAL	PRICE, ¢/UNIT	ANNUAL COST, $M	CENTS PER GAL	DOLLARS/ MET TON
Aspen, lb	52.67715	1.0	26,340		
Sulfuric Acid, lb	.23659	4.3	509		
Calcium Hydroxide, lb	.15439	2.0	154		
Sodium Hydroxide, lb	.01302	26.0	169		
Corn, lb	.01237	4.5	28		
Ammonium Sulfate, lb	.22243	3.0	334		
Superphosphate(46), lb	.96708	8.0	3,869		
Calcium Carbonate, lb	.44486	2.7	601		
Hydrogen, lb	.24589	60.0	7,377		
Catalyst & Chemicals			2,200		

TOTAL RAW MATERIALS			41,580	83.16	1833.34
UTILITIES					
Power, KWH	1.74801	4.3	3,758		
Cooling Water, M Gal	.35158	5.8	1,020		
Process Water, M Gal	.12999	60.0	3,900		
Steam,50 psig, M Lb	.03703	295.0	5,462		

TOTAL UTILITIES			14,140	28.28	623.46
OPERATING COSTS					
Labor, 60 Men @ $ 25,500		13 M/S	1,530		
Foremen, 13 Men @ $ 29,000		2 M/S	377		
Supervision, 3 Man @ $ 35,000		3 Man	105		
Maint., Material & Labor	6% of ISBL		5,988		

TOTAL OPERATING COST			8,000	16.00	352.74
OVERHEAD EXPENSES					
Direct Overhead	45% Lab. & Sup.		905		
Gen. Plant Overhead	65% Oper. Costs		5,200		
Insurance, Prop. Tax	1.5% Tot. Fix. Inv.		2,472		

TOTAL OVERHEAD EXPENSES			8,577	17.15	378.19
BY-PRODUCT CREDIT					
Carbon Dioxide, lb	8.00157	2.8	-11,203		
SCP, lb	.03000	15.0	-225		
Phenol, lb	.87143	32.5	-14,162		
Benzene, lb	.62127	20.7	-6,430		

TOTAL BY-PRODUCT CREDIT			-32,020	-64.04	-1411.83
			=========	=======	=========
CASH COST OF PRODUCTION			40,277	80.55	1775.90
DEPRECIATION	20% ISBL + 10% OSBL		26,460	52.92	1166.67
			=========	=======	=========
NET COST OF PRODUCTION			66,737	133.47	2942.58
REQUIRED SALES PRICE AT 10% DCF				[illegible]	[illegible]

Based on the thermal equivalent of the electric power required (assumed to be 10,000 Btu/kwh) plus the enthalpy of the steam requirements (assuming an 85 percent generation efficiency), minus the energy credits accruing as a result of the production of phenol and benzene as by-products of the Lignol process, incorporation of the speculative "best" case would result in an annual energy consumption of 1.73 trillion Btu. This represents an impressive saving of about 68 percent in energy consumption relative to the base case and may be thought of as a "target" in terms of improving energy efficiency.

V. SENSITIVITY OF ABE FERMENTATION PROCESS ECONOMICS TO VARIATION IN FEEDSTOCK COST

Several cellulosic sources were considered as potential feedstocks for the ABE fermentation process. The factors which determine feedstock desirability are price, availability, high potential sugars, and potential for rapid, high yield growth. Aspen wood (Populus tremuloides) was chosen for the base case because of its high potential sugar content, abundance in North America and potential for rapid, high yield reforestation (especially some of the new hybrid poplars). Other woods considered potentially good cellulosic sources are black cottonwood, eastern cottonwood and eucalyptus, all hardwoods. Softwoods were not considered because, in an enzyme hydrolysis process, softwoods are relatively impervious to pretreatment. This results in low enzyme hydrolysis yields compared to hardwoods. Since the cottonwoods are also members of the poplar genus, they were considered too similar in characteristics to aspen. Eucalyptus, being a species that grows in abundance in the southern United States and having been shown to be very inexpensive to grow and harvest,[1] was chosen as a potential feedstock. Agricultural wastes are another potential source of cellulosics, and corn stover has been selected as a representative of this group.

In each case examined herein, identical feedstock prices were assumed. This enabled the analysis to consider strictly compositional effects associated with the different feedstocks.

It should be noted that reported wood compositions are of very uncertain accuracy. Compositions of individual wood species can vary by as much as 5-10 percent from tree to tree. As a result, reported wood compositions vary from source to source, and the best approach is to take an average value within the reported range.

(1) Mitre Technical Report No. 7347, Silvicultural Biomass Farms, Volume II, "The Biomass Potential of Short-Rotation Farms," May 1977.

The aspen wood composition which was used to generate the base cases has been revised to reflect a more detailed breakdown of its hemicellulose content, and is included as part of the feedstock variation analysis.

A. Revised Aspen (Populus tremuloides)

Design Basis

The original aspen feedstock composition showed the hemicellulose component to be composed of xylan and hexosan. The xylan was thought to be composed entirely of D-xylose, a five carbon sugar, and hexosan, a mixture of D-glucose, D-mannose and D-galactan, all six carbon sugars. In reality, the xylan and hexosan fractions of the hemicellulose contain additional components. Xylan, in addition to D-xylose, contains glucuronic acid and acetyl groups attached intermittently to the D-xylose groups. The hemicellulose hexoses (D-glucose, galactose and mannose) also contain attached acetyl groups. Table V-A-1 summarizes the original aspen composition used and the revised aspen composition which reflects these additional hemicellulose components. The revised aspen composition, in effect, has decreased the percentage of potential xylose, and potential hemicellulose-derived hexose, while adding the acetyl and glucuronic acid groups. Quantitatively this translates to an overall decrease in total potential sugars from 39.7 percent to 37.7 percent.

The cost of production analysis for the feedstock variation cases are performed on the base case low yield process with 5.0 percent sugars in fermentation, 27.54 percent yield and 1.4 percent solvents in the beer.

The revised aspen composition case does not use any recycle slopback. This is due to the total potential sugar being reduced, such that any recycle would dilute the sugar stream below 5.0 percent.

Economics

Tables V-A-2 and V-A-3 represent cost of production (COP) estimates for the revised aspen feedstock case using CSI and DOE utility costs, respectively. These data are summarized and compared against the (revised) base case values in Table V-A-4. This analysis is for a facility producing 50 million gallons per year of mixed solvents on the U.S. Gulf Coast in mid-1982.

TABLE V-A-1

SUMMARY OF POPULUS TREMULOIDES COMPOSITIONS
(Weight Percent, Green Wood)

	Original	Revised
Water	50.0	50.0
Crystalline cellulose	20.7	22.0
Amorphous cellulose	3.7	3.9
Lignin	8.3	7.9
Ash	0.1	0.1
Extractives	1.9	1.5
Hemicellulose C_5*	10.9	7.9
Hemicellulose C_6**	4.4	3.5
D-glucuronic acid	-	1.6
o-acetyl	-	1.6
Total	100.0	100.0

*Hemicellulose C_5 sugars are: D-xylose and D-arabinose.
**Hemicellulose C_6 sugars are: D-glucose, D-mannose, and D-galactose.

TABLE V-A-4

SUMMARY OF ABE PROCESS ECONOMICS
REVISED ASPEN COMPOSITION

Basis: U.S. Gulf Coast, mid-1982, 50 million gallons per year.

	CSI Utilities		DOE Utilities	
	Revised	Base	Revised	Base
Investment, MM$				
Battery limits	98.7	92.8	98.7	92.8
Offsites	98.4	97.3	98.4	97.3
Total fixed investment	197.1	190.1	197.1	190.1
Cost of production, ¢/gal				
Raw materials	93.44	86.59	93.44	86.59
Utilities	41.62	43.37	36.91	37.00
Operating costs	15.87	15.16	15.87	15.16
Overhead expenses	18.04	17.37	18.04	17.37
By-product credit	(23.20)	(23.20)	(23.20)	(23.20)
Cash cost of production	145.75	139.28	141.05	132.91
Depreciation	59.16	56.58	59.16	56.58
Net cost of production	204.91	195.85	200.21	189.48
Selling price at 10% DCF	270.7	259.8	265.5	252.8

TABLE V-A-2

COST OF PRODUCTION ESTIMATE FOR ABE
PROCESS - LOW YIELD/REV POPLAR

CAPITAL SUMMARY

BASIS		CAPITAL COST	$MILLION
Location:	U.S. Gulf Coast	Battery Limits	98.7
	Mid-1982	Offsites	98.4
Capacity:	50.00 million gallons/yr		------
	22,680 metric tons/yr	Total Fixed Inv.	197.1
Str.Time:	8000 hours per year	Working Capital	16.8

PRODUCTION COST SUMMARY

RAW MATERIALS	UNITS PER GAL	PRICE, ¢/UNIT	ANNUAL COST, $M	CENTS PER GAL	DOLLARS/ MET TON
Aspen, lb	66.57139	1.0	33,287		
Sulfuric Acid, lb	.28158	4.3	605		
Calcium Hydroxide, lb	.17609	2.0	176		
Sodium Hydroxide, lb	.00900	26.0	117		
Corn, lb	.01237	4.5	28		
Ammonium Sulfate, lb	.48907	3.0	734		
Superphosphate(46), lb	2.12553	8.0	8,503		
Calcium Carbonate, lb	.97772	2.7	1,320		
Catalyst & Chemicals			1,950		

TOTAL RAW MATERIALS			46,720	93.44	2059.98
UTILITIES					
Power, kWH	1.86240	3.2	2,980		
Cooling Water, M Gal	.31068	5.8	901		
Process Water, M Gal	.03670	60.0	1,101		
Steam,50 psig, M Lb	.08441	375.0	15,827		

TOTAL UTILITIES			20,809	41.62	917.51
OPERATING COSTS					
Labor, 60 Men @ $ 25,500		13 M/S	1,530		
Foremen, 13 Men @ $ 29,000		2 M/S	377		
Supervision, 3 Man @ $ 35,000		3 Man	105		
Maint., Material & Labor	6% of ISBL		5,922		

TOTAL OPERATING COST			7,934	15.87	349.83
OVERHEAD EXPENSES					
Direct Overhead	45% Lab. & Sup.		905		
Gen. Plant Overhead	65% Oper. Costs		5,157		
Insurance, Prop. Tax	1.5% Tot. Fix. Inv.		2,956		

TOTAL OVERHEAD EXPENSES			9,019	18.04	397.67
BY-PRODUCT CREDIT					
Carbon Dioxide, lb	8.12319	2.8	-11,373		
SCP, lb	.03051	15.0	-229		

TOTAL BY-PRODUCT CREDIT			-11,602	-23.20	-511.55
			==========	=======	==========
CASH COST OF PRODUCTION			72,880	145.75	3213.43
DEPRECIATION	20% ISBL + 10% OSBL		29,580	59.16	1304.24
			==========	=======	==========
NET COST OF PRODUCTION			102,460	204.91	4517.67
REQUIRED SALES PRICE AT 10% DCF				270.7	5967.7

TABLE V-A-3

COST OF PRODUCTION ESTIMATE FOR ABE
PROCESS - LOW YIELD/REV POPLAR

CAPITAL SUMMARY

BASIS		CAPITAL COST	$MILLION
Location:	U.S. Gulf Coast	Battery Limits	98.7
	Mid-1982	Offsites	98.4
Capacity:	50.00 million gallons/yr		------
	22,680 metric tons/yr	Total Fixed Inv.	197.1
Str.Time:	8000 hours per year	Working Capital	16.5

PRODUCTION COST SUMMARY

RAW MATERIALS	UNITS PER GAL	PRICE, c/UNIT	ANNUAL COST, $M	CENTS PER GAL	DOLLARS/ MET TON
Aspen, lb	66.57139	1.0	33,287		
Sulfuric Acid, lb	.28158	4.3	605		
Calcium Hydroxide, lb	.17609	2.0	176		
Sodium Hydroxide, lb	.00900	26.0	117		
Corn, lb	.01237	4.5	28		
Ammonium Sulfate, lb	.48907	3.0	734		
Superphosphate(46), lb	2.12553	8.0	8,503		
Calcium Carbonate, lb	.97772	2.7	1,320		
Catalyst & Chemicals			1,950		

TOTAL RAW MATERIALS			46,720	93.44	2059.98
UTILITIES					
Power, kWH	1.86240	4.3	4,004		
Cooling Water, M Gal	.31068	5.8	901		
Process Water, M Gal	.03670	60.0	1,101		
Steam,50 psig, M Lb	.08441	295.0	12,450		

TOTAL UTILITIES			18,457	36.91	813.80
OPERATING COSTS					
Labor, 60 Men @ $ 25,500		13 M/S	1,530		
Foremen, 13 Men @ $ 29,000		2 M/S	377		
Supervision, 3 Man @ $ 35,000		3 Man	105		
Maint., Material & Labor	6% of ISBL		5,922		

TOTAL OPERATING COST			7,934	15.87	349.83
OVERHEAD EXPENSES					
Direct Overhead	45% Lab. & Sup.		905		
Gen. Plant Overhead	65% Oper. Costs		5,157		
Insurance, Prop. Tax	1.5% Tot. Fix. Inv.		2,956		

TOTAL OVERHEAD EXPENSES			9,019	18.04	397.67
BY-PRODUCT CREDIT					
Carbon Dioxide, lb	8.12319	2.8	11,373		
SCP, lb	.03051	15.0	-229		

TOTAL BY-PRODUCT CREDIT			-11,602	-23.20	-511.55
			==========	=======	==========
CASH COST OF PRODUCTION			70,528	141.05	3109.72
DEPRECIATION	20% ISBL + 10% OSBL		29,580	59.16	1304.24
			==========	=======	==========
NET COST OF PRODUCTION			100,108	200.21	4413.97
REQUIRED SALES PRICE AT 10% DCF				265.5	5854.5

As can be seen from Table V-A-4, the revised aspen composition case produces mixed solvents for 270.7 and 265.5 cents per gallon, respectively. This is about 12 cents per gallon higher than the original aspen composition case and is due primarily to the increased raw materials cost associated with the decreased sugar potential of the revised composition.

B. Eucalyptus

Design Basis

The genus Eucalyptus contains over 500 identifiable species which grow in some 50 countries including the United States. Eucalyptus are broad-leaved evergreen trees which are highly adaptable to climate and site. Despite this, in general, this genus does not do well in areas where rapid drops in temperature occur. Therefore, Eucalyptus species grown for energy use would do best in Northern California (*E. globulus*) and the Southeastern United States. Table V-B-1 represents a composition for a typical Eucalyptus species, namely *E. globulus*.(1)

TABLE V-B-1

COMPOSITION OF E. GLOBULUS
(Green Wood, Wt % Total Wood)

	Percent
Water	50.0
Crystalline cellulose	20.6
Amorphous cellulose	3.7
Lignin	12.2
Ash	0.1
Extractives	2.5
Hemicellulose C_5	6.7
Hemicellulose C_6	1.7
O-glucuronic acid	1.3
o-acetyl	1.2
Total	100.0

(1) Dillner B., et al. "The Breeding of E. globulus on the Basis of Wood Density, Chemical Composition and Growth Rate," Symposium on the Production and Industrial Utilization of Eucalyptus, Lisbon, Portugal, 1970.

The Eucalyptus wood contains 32.7 percent total potential sugars, significantly lower than the aspen wood. The Eucalyptus case is performed on the same basis as the aspen cases (low yield); however, slopback recycle is 2.3 percent.

Economics

Tables V-B-2 and V-B-3 represent cost of production (COP) estimates for the low yield case using Eucalyptus wood feedstock with CSI and DOE utilities, respectively. These data are summarized and compared against the revised aspen composition case (Table V-A-4) in Table V-B-4. This case is for a facility producing 50 million gallons per year of mixed solvents on the U.S. Gulf Coast in mid-1982.

TABLE V-B-4

SUMMARY OF ABE PROCESS ECONOMICS USING EUCALYPTUS WOOD

Basis: U.S. Gulf Coast, mid-1982, 50 million gallons per year.

	CSI Utilities		DOE Utilities	
	Euca-lyptus	Revised Aspen	Euca-lyptus	Revised Aspen
Investment, MM$				
Battery limits	106.0	98.7	106.0	98.7
Offsites	100.3	98.4	100.3	98.4
Total fixed investment	206.3	197.1	206.3	197.1
Cost of production, ¢/gal				
Raw materials	103.23	93.44	103.23	93.44
Utilities	35.46	41.62	32.28	36.91
Operating costs	16.74	15.87	16.74	15.87
Overhead expenses	18.88	18.04	18.88	18.04
By-product credit	(23.30)	(23.20)	(23.20)	(23.20)
Cash cost of production	151.11	145.75	147.93	141.05
Depreciation	62.46	59.16	62.46	59.16
Net cost of production	213.57	204.91	210.38	200.21
Selling price at 10% DCF	281.6	270.7	278.1	265.5

TABLE V-B-2

COST OF PRODUCTION ESTIMATE FOR ABE
PROCESS- LOW YIELD/EUCALYPTUS

CAPITAL SUMMARY

BASIS		CAPITAL COST	$MILLION
Location:	U.S. Gulf Coast	Battery Limits	106.0
	Mid-1982	Offsites	100.3
Capacity:	50.00 million gallons/yr		------
	22,680 metric tons/yr	Total Fixed Inv.	206.3
Str.Time:	8000 hours per year	Working Capital	17.4

PRODUCTION COST SUMMARY

RAW MATERIALS	UNITS PER GAL	PRICE, ¢/UNIT	ANNUAL COST, $M	CENTS PER GAL	DOLLARS/ MET TON
Eucalyptus Wood, lb	76.41436	1.0	38,209		
Sulfuric Acid, lb	.32108	4.3	690		
Calcium Hydroxide, lb	.19939	2.0	199		
Sodium Hydroxide, lb	.01750	26.0	228		
Corn, lb	.01237	4.5	28		
Ammonium Sulfate, lb	.47787	3.0	717		
Superphosphate(46), lb	2.07659	8.0	8,307		
Calcium Carbonate, lb	.95515	2.7	1,290		
Catalyst & Chemicals			1,950		

TOTAL RAW MATERIALS			51,618	103.23	2275.92
UTILITIES					
Power, kWH	1.93969	3.2	3,104		
Cooling Water, M Gal	.36058	5.8	1,046		
Process Water, M Gal	.03712	60.0	1,114		
Steam,50 psig, M Lb	.06650	375.0	12,469		

TOTAL UTILITIES			17,732	35.46	781.83
OPERATING COSTS					
Labor, 60 Men @ $ 25,500		13 M/S	1,530		
Foremen. 13 Men @ $ 29,000		2 M/S	377		
Supervision, 3 Man @ $ 35,000		3 Man	105		
Maint., Material & Labor		6% of ISBL	6,360		

TOTAL OPERATING COST			8,372	16.74	369.14
OVERHEAD EXPENSES					
Direct Overhead	45% Lab. & Sup.		905		
Gen. Plant Overhead	65% Oper. Costs		5,442		
Insurance, Prop. Tax	1.5% Tot. Fix. Inv.		3,094		

TOTAL OVERHEAD EXPENSES			9,442	18.88	416.30
BY-PRODUCT CREDIT					
Carbon Dioxide, lb	8.12319	2.8	-11,373		
SCP, lb	.03051	15.0	-229		

TOTAL BY-PRODUCT CREDIT			-11,602	-23.20	-511.55
			==========	=======	==========
CASH COST OF PRODUCTION			75,561	151.11	3331.64
DEPRECIATION	20% ISBL + 10% OSBL		31,230	62.46	1376.99
			==========	=======	==========
NET COST OF PRODUCTION			106,791	213.57	4708.63
REQUIRED SALES PRICE AT 10% DCF				281.6	6208.0

TABLE V-B-3

COST OF PRODUCTION ESTIMATE FOR ABE
PROCESS - LOW YIELD/EUCALYPTUS

CAPITAL SUMMARY

BASIS		CAPITAL COST	$MILLION
Location:	U.S. Gulf Coast	Battery Limits	106.0
	Mid-1982	Offsites	100.3
Capacity:	50.00 million gallons/yr		------
	22,680 metric tons/yr	Total Fixed Inv.	206.3
Str.Time:	8000 hours per year	Working Capital	17.2

PRODUCTION COST SUMMARY

RAW MATERIALS	UNITS PER GAL	PRICE, ¢/UNIT	ANNUAL COST, $M	CENTS PER GAL	DOLLARS/ MET TON
Eucalyptus Wood, lb	76.41436	1.0	38,209		
Sulfuric Acid, lb	.32108	4.3	690		
Calcium Hydroxide, lb	.19939	2.0	199		
Sodium Hydroxide, lb	.01750	26.0	228		
Corn, lb	.01237	4.5	28		
Ammonium Sulfate, lb	.47787	3.0	717		
Superphosphate(46), lb	2.07659	8.0	8,307		
Calcium Carbonate, lb	.95515	2.7	1,290		
Catalyst & Chemicals			1,950		

TOTAL RAW MATERIALS			51,618	103.23	2275.92
UTILITIES					
Power, kWH	1.93969	4.3	4,171		
Cooling Water, M Gal	.36058	5.8	1,046		
Process Water, M Gal	.03712	60.0	1,114		
Steam,50 psig, M Lb	.06650	295.0	9,809		

TOTAL UTILITIES			16,139	32.28	711.59
OPERATING COSTS					
Labor, 60 Men @ $ 25,500		13 M/S	1,530		
Foremen, 13 Men @ $ 29,000		2 M/S	377		
Supervision, 3 Man @ $ 35,000		3 Man	105		
Maint., Material & Labor	6% of ISBL		6,360		

TOTAL OPERATING COST			8,372	16.74	369.14
OVERHEAD EXPENSES					
Direct Overhead	45% Lab. & Sup.		905		
Gen. Plant Overhead	65% Oper. Costs		5,442		
Insurance, Prop. Tax	1.5% Tot. Fix. Inv.		3,094		

TOTAL OVERHEAD EXPENSES			9,442	18.88	416.30
BY-PRODUCT CREDIT					
Carbon Dioxide, lb	8.12319	2.8	-11,373		
SCP, lb	.03051	15.0	-229		

TOTAL BY-PRODUCT CREDIT			-11,602	-23.20	-511.55
			==========	=======	==========
CASH COST OF PRODUCTION			73,968	147.93	3261.40
DEPRECIATION	20% ISBL + 10% OSBL		31,230	62.46	1376.99
			==========	=======	==========
NET COST OF PRODUCTION			105,198	210.38	4638.39
REQUIRED SALES PRICE AT 10% DCF				278.1	6131.3

As can be seen from Table V-B-4, mixed solvents can be produced from Eucalyptus via ABE fermentation for 281.6 and 278.1 cents per gallon, respectively. These values represent the sales price at 10 percent DCF. This is about 11 cents per gallon higher than the cost of production using aspen wood (revised case) as the feedstock. This is due primarily to the increase in raw materials cost for Eucalyptus associated with its lower potential sugar content and, to a lesser extent, the higher capital-related expenses associated with the greater volumetric throughput required. These disadvantages are somewhat offset by the fact that Eucalyptus contains more lignin than aspen, resulting in more by-product steam being produced, thus decreasing utility costs.

C. Corn Stover

Design Basis

Corn stover is an agricultural waste which includes all parts of the corn plant excluding the corn kernels. Corn stover is composed of the following corn plant components by weight, as shown in Table V-C-1.

TABLE V-C-1

COMPOSITION OF CORN STOVER PLANT

	Corn Stover (Wt %)
Cob	21.3
Leaf	11.7
Husk	13.0
Stalk	53.9

Following the harvesting of corn, part of the remaining corn stover is plowed under to be used as fertilizer. The rest is collected as waste, usually in a field dried state which reduces the water content from about 50 percent to 30 percent. This is assumed to be the state of the corn stover feedstock to be used in the ABE facility.

A typical corn stover composition for field dried material is presented in Table V-C-2. For comparison, green corn stover (50 percent water) composition is also given.

TABLE V-C-2

CORN STOVER COMPOSITION
(Weight Percent)

	Field Dried	Green
Water	30.7	50.0
Crystalline cellulose	22.3	16.1
Amorphous cellulose	3.9	2.8
Lignin	7.0	5.1
Ash	3.9	2.8
Extractives	1.1	0.8
Hemicellulose C_5	16.1	11.6
Hemicellulose C_6	6.5	4.7
Sucrose	5.3	3.8
Soluble protein	0.4	0.3
Insoluble protein	2.6	1.9

Table V-C-2 indicates that corn stover is quite different in composition from hardwoods. Corn stover contains several components not present in significant quantities in woods. These are soluble and insoluble proteins as well as sucrose. Sucrose, or natural plant sugar, is a disaccharide ($C_{12}H_{24}O_{12}$) which hydrolyzes very easily to d-glucose and d-fructose. Therefore, corn stover contains an additional potential sugar component compared to wood, which brings its total potential sugar content (on a 50 percent water basis) to 39 percent.

Because of the fact that field dried corn stover has much less water than green wood, the sugar concentrations obtainable with corn stover in the base case design are very high. For example, the sugar concentration after enzyme hydrolysis for the wood cases is only about 5.0 percent, but for the corn stover case it is 6.4 percent. Therefore, to maintain consistency during fermentation, at 5.0 percent sugar the corn stover case must be diluted. This is accomplished by using a 25.3 percent slopback recycle, which also recycles nutrients.

Economics

Table V-C-3 and Table V-C-4 represent cost of production (COP) estimates for ABE fermentation with corn stover as the feedstock using CSI and DOE utilities, respectively. These data are summarized and compared against the revised aspen composition case (Table V-A-4) in Table V-C-5. This analysis is for a facility producing 50 million gallons per year of mixed solvents on the U.S. Gulf Coast in mid-1982.

TABLE V-C-5

SUMMARY OF ABE PROCESS ECONOMICS
CORN STOVER FEEDSTOCK

Basis: U.S. Gulf Coast, mid-1982, 50 million gallons per year.

	CSI Utilities		DOE Utilities	
	Corn Stover	Revised Aspen	Corn Stover	Revised Aspen
Investment, MM$				
Battery limits	88.8	98.7	88.8	98.7
Offsites	97.3	98.4	97.3	98.4
Total fixed investment	186.1	197.1	186.1	197.1
Cost of production, ¢/gal				
Raw materials	65.84	93.44	65.84	93.44
Utilities	44.86	41.62	38.96	36.91
Operating costs	14.68	15.87	14.68	15.87
Overhead expenses	16.93	18.04	16.93	18.04
By-product credit	(23.20)	(23.20)	(23.20)	(23.20)
Cash cost of production	119.11	145.75	113.20	141.05
Depreciation	54.98	59.16	54.98	59.16
Net cost of production	174.08	204.91	168.18	200.21
Selling price at 10% DCF	235.7	270.7	229.2	265.5

The corn stover based ABE fermentation process produces mixed solvents for 235.7 and 229.2 cents per gallon. These values represent the sales price at a 10 percent DCF rate of return. This is about 40 cents per gallon better than the revised aspen case, a significant decrease. This is by virtue of the fact that the corn stover is field dried and contains only 30.7 percent water compared to 50 percent for the wood cases. This

TABLE V-C-3

COST OF PRODUCTION ESTIMATE FOR ABE
PROCESS - LOW YIELD/CORN STOVE

CAPITAL SUMMARY

BASIS		CAPITAL COST	$MILLION
Location:	U.S. Gulf Coast	Battery Limits	88.8
	Mid-1982	Offsites	97.3
Capacity:	50.00 million gallons/yr		------
	22,680 metric tons/yr	Total Fixed Inv.	186.1
Str.Time:	8000 hours per year	Working Capital	14.5

PRODUCTION COST SUMMARY

RAW MATERIALS	UNITS PER GAL	PRICE, ¢/UNIT	ANNUAL COST, $M	CENTS PER GAL	DOLLARS/ MET TON
Corn Stover, lb	44.51609	1.0	22,259		
Sulfuric Acid, lb	.24869	4.3	535		
Calcium Hydroxide, lb	.15969	2.0	160		
Sodium Hydroxide, lb	.00790	26.0	103		
Corn, lb	.01237	4.5	28		
Ammonium Sulfate, lb	.36538	3.0	548		
Superphosphate(46), lb	1.58771	8.0	6,351		
Calcium Carbonate, lb	.73026	2.7	986		
Catalyst & Chemicals			1,950		

TOTAL RAW MATERIALS			32,919	65.84	1451.48
UTILITIES					
Power, kWH	1.63091	4.3	3,507		
Cooling Water, M Gal	.30163	5.8	875		
Process Water, M Gal	.03030	60.0	909		
Steam,50 psig, M Lb	.09619	295.0	14,189		

TOTAL UTILITIES			19,480	38.96	858.92
OPERATING COSTS					
Labor, 60 Men @ $ 25,500		13 M/S	1,530		
Foremen, 13 Men @ $ 29,000		2 M/S	377		
Supervision, 3 Man @ $ 35,000		3 Man	105		
Maint., Material & Labor	6% of ISBL		5,328		

TOTAL OPERATING COST			7,340	14.68	323.64
OVERHEAD EXPENSES					
Direct Overhead	45% Lab. & Sup.		905		
Gen. Plant Overhead	65% Oper. Costs		4,771		
Insurance, Prop. Tax	1.5% Tot. Fix. Inv.		2,791		

TOTAL OVERHEAD EXPENSES			8,468	16.93	373.37
BY-PRODUCT CREDIT					
Carbon Dioxide, lb	8.12319	2.8	-11,373		
SCP, lb	.03051	15.0	-229		

TOTAL BY-PRODUCT CREDIT			-11,602	-23.20	-511.55
			==========	=======	==========
CASH COST OF PRODUCTION			56,605	113.20	2495.85
DEPRECIATION	20% ISBL + 10% OSBL		27,490	54.98	1212.09
			==========	=======	==========
NET COST OF PRODUCTION			84,095	168.18	3707.94
REQUIRED SALES PRICE AT 10% DCF				229.2	5054.3

TABLE V-C-4

COST OF PRODUCTION ESTIMATE FOR ABE PROCESS - LOW YIELD/CORN STOVE

CAPITAL SUMMARY

BASIS		CAPITAL COST	$MILLION
Location:	U.S. Gulf Coast	Battery Limits	88.8
	Mid-1982	Offsites	97.3
Capacity:	50.00 million gallons/yr		------
	22,680 metric tons/yr	Total Fixed Inv.	186.1
Str.Time:	8000 hours per year	Working Capital	14.9

PRODUCTION COST SUMMARY

RAW MATERIALS	UNITS PER GAL	PRICE, ¢/UNIT	ANNUAL COST, $M	CENTS PER GAL	DOLLARS/ MET TON
Corn Stover, lb	44.51609	1.0	22,259		
Sulfuric Acid, lb	.24869	4.3	535		
Calcium Hydroxide, lb	.15969	2.0	160		
Sodium Hydroxide, lb	.00790	26.0	103		
Corn, lb	.01237	4.5	28		
Ammonium Sulfate, lb	.36538	3.0	548		
Superphosphate(46), lb	1.58771	8.0	6,351		
Calcium Carbonate, lb	.73026	2.7	986		
Catalyst & Chemicals			1,950		

TOTAL RAW MATERIALS			32,919	65.84	1451.48
UTILITIES					
Power, kWH	1.63891	3.2	2,610		
Cooling Water, M Gal	.30168	5.8	875		
Process Water, M Gal	.03030	60.0	909		
Steam,50 psig, M Lb	.09619	375.0	18,037		

TOTAL UTILITIES			22,431	44.86	989.03
OPERATING COSTS					
Labor, 60 Men @ $ 25,500		13 M/S	1,530		
Foremen, 13 Men @ $ 29,000		2 M/S	377		
Supervision, 3 Man @ $ 35,000		3 Man	105		
Maint., Material & Labor	6% of ISBL		5,328		

TOTAL OPERATING COST			7,340	14.68	323.64
OVERHEAD EXPENSES					
Direct Overhead	45% Lab. & Sup.		905		
Gen. Plant Overhead	65% Oper. Costs		4,771		
Insurance, Prop. Tax	1.5% Tot. Fix. Inv.		2,791		

TOTAL OVERHEAD EXPENSES			8,468	16.93	373.37
BY-PRODUCT CREDIT					
Carbon Dioxide, lb	8.12319	2.8	-11,373		
SCP, lb	.03051	15.0	-229		

TOTAL BY-PRODUCT CREDIT			-11,602	-23.20	-511.55
			==========	=======	==========
CASH COST OF PRODUCTION			59,556	119.11	2625.96
DEPRECIATION	20% ISBL + 10% OSBL		27,490	54.98	1212.09
			==========	=======	==========
NET COST OF PRODUCTION			87,046	174.08	3838.05
REQUIRED SALES PRICE AT 10% DCF				235.7	5196.5

results in a significant decrease in raw material expenses associated with the decreased raw material required. To a lesser extent there is a decrease in capital related expenses due to the smaller front end equipment requirements associated with the smaller feed. However, it should be noted that corn stover may be an unsuitable year round feedstock since it cannot be stored any length of time because of its sugar content. As a result, a truly realistic assessment would have to modify the hypothetical plant design in one of two ways: either the plant would have to be designed several times larger to produce the same annual output in a shorter time period, or it would have to be designed to allow for use of a different feedstock during seasons when corn stover is unavailable. Incorporation of either of these two alternatives would undoubtedly result in capital costs sufficiently high as to make the plant non-competitive.

D. Aspen Feedstock Price Sensitivity

Design Basis

The price of aspen feedstock (as well as cellulose feedstocks) which has been used for all cases heretofore is 1.0 cents per pound (or $20 per ton) wet. Subsequent information has indicated that aspen wood is considered to be of very low quality and value by most of the paper and pulp industry, and its estimated cost is $40-50 per cord, delivered. A cord contains approximately 2.6-3.0 tons of wet wood. At 2.6 tons per cord, the range of wood cost is $15.4 per ton (0.77 cents per pound) at $40 per cord to $19.2 per ton (0.96 cents per pound) at $50 per cord. Recognizing that the 2.6 tons per cord is a conservative number, the real price range of aspen wood is probably about 0.5-1.0 cents per pound.

The aspen feedstock price sensitivity analysis was conducted by selecting three wood prices and running cost of production estimates on the low yield base case at those prices. The three prices were 0.5 cents per pound, 1.0 cents per pound, and 1.5 cents per pound for green wood. The original aspen wood composition was used.

Economics

Tables V-D-1 through V-D-6 represent cost of production (COP) estimates for ABE fermentation at three different aspen wood prices using CSI and DOE utilities, respectively. These data are summarized in Table V-D-7. This analysis is for a facility producing 50 million gallons per year mixed solvents on the U.S. Gulf Coast in mid-1982.

TABLE V-D-7

SUMMARY OF ABE FERMENTATION PROCESS ECONOMICS
ASPEN WOOD PRICE SENSITIVITY

Bases: 50 million gallons per year, U.S. Gulf Coast, mid-1982.

	CSI(1) Util.	DOE(1) Util.	CSI(2) Util.	DOE(2) Util.	CSI(3) Util.	DOE(3) Util.
Investment, MM$						
Battery limits	92.8	92.8	92.8	92.8	92.8	92.8
Offsites	97.6	97.6	97.6	97.6	97.6	97.6
Total fixed investment	190.4	190.4	190.4	190.4	190.4	190.4
Cost of production, ¢/gal						
Raw materials	55.70	55.70	86.60	86.60	117.51	117.51
Utilities	41.99	35.04	41.99	35.04	41.99	35.04
Operating costs	15.16	15.16	15.16	15.16	15.16	15.16
Overhead expenses	17.38	17.38	17.38	17.38	17.38	17.38
By-product credit	(23.20)	(23.20)	(23.20)	(23.20)	(23.20)	(23.20)
Cash cost of production	107.02	100.07	137.93	130.98	168.83	161.88
Depreciation	56.64	56.64	56.64	56.64	56.64	56.64
Net cost of production	163.66	156.71	194.57	187.61	225.47	218.52
Selling price at 10% DCF	225.20	217.4	258.5	250.9	292.0	284.4

(1) Aspen wood at 0.5¢/lb.
(2) Aspen wood at 1.0¢/lb (base case).
(3) Aspen wood at 1.5¢/lb.

As can be seen from Table V-D-7, the range of COPs at 10 percent DCF is 225.0 cents per gallon at 0.5 cents per pound to 292.0 cents per gallon at 1.5 cents per pound. An increase in feedstock price of 1.0 cents per pound increases the cost of production 67 cents per gallon or approximately 30 percent. Figure V-D-1 illustrates the relationship between aspen wood feedstock and sales price at 10 percent DCF.

TABLE V-D-1

COST OF PRODUCTION ESTIMATE FOR ABE PROCESS- LOW YIELD

CAPITAL SUMMARY

BASIS		CAPITAL COST	$MILLION
Location:	U.S. Gulf Coast	Battery Limits	92.8
	Mid-1982	Offsites	97.6
Capacity:	50.00 million gallons/yr		------
	22,680 metric tons/yr	Total Fixed Inv.	190.4
Str.Time:	8000 hours per year	Working Capital	14.5

PRODUCTION COST SUMMARY

RAW MATERIALS	UNITS PER GAL	PRICE, ¢/UNIT	ANNUAL COST, $M	CENTS PER GAL	DOLLARS/ MET TON
Aspen, lb	61.80965	.5	15,453		
Sulfuric Acid, lb	.27790	4.3	598		
Calcium Hydroxide, lb	.19798	2.0	198		
Sodium Hydroxide, lb	.00880	26.0	114		
Corn, lb	.01237	4.5	28		
Ammonium Sulfate, lb	.44400	3.0	666		
Superphosphate(46), lb	1.91375	8.0	7,655		
Calcium Carbonate, lb	.88034	2.7	1,189		
Catalyst & Chemicals			1,950		

TOTAL RAW MATERIALS			27,851	55.70	1228.00
UTILITIES					
Power, kWH	1.78069	3.2	2,849		
Cooling Water, M Gal	.30312	5.8	879		
Process Water, M Gal	.03304	60.0	991		
Steam,50 psig, M Lb	.07363	375.0	13,806		
Steam,200 psig, M Lb	.01298	381.0	2,473		

TOTAL UTILITIES			20,998	41.99	925.84
OPERATING COSTS					
Labor, 60 Men @ $ 25,500	13 M/S		1,530		
Foremen, 13 Men @ $ 29,000	2 M/S		377		
Supervision, 3 Man @ $ 35,000	3 Man		105		
Maint., Material & Labor	6% of ISBL		5,568		

TOTAL OPERATING COST			7,580	15.16	334.22
OVERHEAD EXPENSES					
Direct Overhead	45% Lab. & Sup.		905		
Gen. Plant Overhead	65% Oper. Costs		4.927		
Insurance, Prop. Tax	1.5% Tot. Fix. Inv.		2.856		

TOTAL OVERHEAD EXPENSES			8.688	17.38	383.09
BY-PRODUCT CREDIT					
Carbon Dioxide, lb	8.12319	2.8	¯11,373		
SCP, lb	.03051	15.0	¯229		

TOTAL BY-PRODUCT CREDIT			¯11,602	¯23.20	¯511.55
			==========	=======	==========
CASH COST OF PRODUCTION			53,515	107.02	2359.60
DEPRECIATION	20% ISBL + 10% OSBL		28,320	56.64	1248.69
			==========	=======	==========
NET COST OF PRODUCTION			81,835	163.66	3608.28
REQUIRED SALES PRICE AT 10% DCF				225.0	4959.9

TABLE V-D-2

COST OF PRODUCTION ESTIMATE FOR ABE PROCESS- LOW YIELD

CAPITAL SUMMARY

BASIS		CAPITAL COST	$MILLION
Location:	U.S. Gulf Coast	Battery Limits	92.8
	Mid-1982	Offsites	97.6
Capacity:	50.00 million gallons/yr		------
	22,680 metric tons/yr	Total Fixed Inv.	190.4
Str.Time:	8000 hours per year	Working Capital	16.1

PRODUCTION COST SUMMARY

RAW MATERIALS	UNITS PER GAL	PRICE, ¢/UNIT	ANNUAL COST, $M	CENTS PER GAL	DOLLARS/MET TON
Aspen, lb	61.80965	1.0	30,906		
Sulfuric Acid, lb	.27790	4.3	598		
Calcium Hydroxide, lb	.19798	2.0	198		
Sodium Hydroxide, lb	.00880	26.0	114		
Corn, lb	.01237	4.5	28		
Ammonium Sulfate, lb	.44400	3.0	666		
Superphosphate(46), lb	1.91375	8.0	7,655		
Calcium Carbonate, lb	.88034	2.7	1,189		
Catalyst & Chemicals			1,950		

TOTAL RAW MATERIALS			43,304	86.60	1909.37
UTILITIES					
Power, kWH	1.78069	3.2	2,849		
Cooling Water, M Gal	.30312	5.8	879		
Process Water, M Gal	.03304	60.0	991		
Steam,50 psig, M Lb	.07363	375.0	13,806		
Steam,200 psig, M Lb	.01298	381.0	2,473		

TOTAL UTILITIES			20,998	41.99	925.84
OPERATING COSTS					
Labor, 60 Men @ $ 25,500		13 M/S	1,530		
Foremen, 13 Men @ $ 29,000		2 M/S	377		
Supervision, 3 Man @ $ 35,000		3 Man	105		
Maint., Material & Labor	6% of ISBL		5,568		

TOTAL OPERATING COST			7,580	15.16	334.22
OVERHEAD EXPENSES					
Direct Overhead	45% Lab. & Sup.		905		
Gen. Plant Overhead	65% Oper. Costs		4,927		
Insurance, Prop. Tax	1.5% Tot. Fix. Inv.		2,856		

TOTAL OVERHEAD EXPENSES			8,688	17.38	383.09
BY-PRODUCT CREDIT					
Carbon Dioxide, lb	8.12319	2.8	-11,373		
SCP, lb	.03051	15.0	-229		

TOTAL BY-PRODUCT CREDIT			-11,602	-23.20	-511.55
			==========	=======	==========
CASH COST OF PRODUCTION			68,969	137.93	3040.96
DEPRECIATION	20% ISBL + 10% OSBL		28,320	56.64	1248.69
			==========	=======	==========
NET COST OF PRODUCTION			97,289	194.57	4289.65
REQUIRED SALES PRICE AT 10% DCF				258.5	5698.6

TABLE V-D-3

COST OF PRODUCTION ESTIMATE FOR ABE PROCESS- LOW YIELD

CAPITAL SUMMARY

BASIS		CAPITAL COST	$MILLION
Location:	U.S. Gulf Coast	Battery Limits	92.8
	Mid-1982	Offsites	97.6
Capacity:	50.00 million gallons/yr		------
	22,680 metric tons/yr	Total Fixed Inv.	190.4
Str.Time:	8000 hours per year	Working Capital	17.6

PRODUCTION COST SUMMARY

RAW MATERIALS	UNITS PER GAL	PRICE, c/UNIT	ANNUAL COST, $M	CENTS PER GAL	DOLLARS/ MET TON
Aspen, lb	61.80965	1.5	46,360		
Sulfuric Acid, lb	.27790	4.3	598		
Calcium Hydroxide, lb	.19798	2.0	198		
Sodium Hydroxide, lb	.00880	26.0	114		
Corn, lb	.01237	4.5	28		
Ammonium Sulfate, lb	.44400	3.0	666		
Superphosphate(46), lb	1.91375	8.0	7,655		
Calcium Carbonate, lb	.88034	2.7	1,189		
Catalyst & Chemicals			1,950		

TOTAL RAW MATERIALS			58,757	117.51	2590.73
UTILITIES					
Power, kWH	1.78069	3.2	2,849		
Cooling Water, M Gal	.30312	5.8	879		
Process Water, M Gal	.03304	60.0	991		
Steam,50 psig, M Lb	.07363	375.0	13,806		
Steam,200 psig, M Lb	.01298	381.0	2,473		

TOTAL UTILITIES			20,998	41.99	925.84
OPERATING COSTS					
Labor, 60 Men @ $ 25,500	13 M/S		1,530		
Foremen, 13 Men @ $ 29,000	2 M/S		377		
Supervision, 3 Man @ $ 35,000	3 Man		105		
Maint., Material & Labor	6% of ISBL		5,568		

TOTAL OPERATING COST			7,580	15.16	334.22
OVERHEAD EXPENSES					
Direct Overhead	45% Lab. & Sup.		905		
Gen. Plant Overhead	65% Oper. Costs		4,927		
Insurance, Prop. Tax	1.5% Tot. Fix. Inv.		2,856		

TOTAL OVERHEAD EXPENSES			8,688	17.38	383.09
BY-PRODUCT CREDIT					
Carbon Dioxide, lb	8.12319	2.8	-11,373		
SCP, lb	.03051	15.0	-229		

TOTAL BY-PRODUCT CREDIT			-11,602	-23.20	-511.55
			==========	=======	==========
CASH COST OF PRODUCTION			84,422	168.83	3722.33
DEPRECIATION	20% ISBL + 10% OSBL		28,320	56.64	1248.69
			==========	=======	==========
NET COST OF PRODUCTION			112,742	225.47	4971.01
REQUIRED SALES PRICE AT 10% DCF				292.0	6437.3

TABLE V-D-4

COST OF PRODUCTION ESTIMATE FOR ABE PROCESS- LOW YIELD

CAPITAL SUMMARY

BASIS
Location: U.S. Gulf Coast
Mid-1982
Capacity: 50.00 million gallons/yr
22,680 metric tons/yr
Str.Time: 8000 hours per year

CAPITAL COST	$MILLION
Battery Limits	92.8
Offsites	97.6

Total Fixed Inv.	190.4
Working Capital	14.0

PRODUCTION COST SUMMARY

RAW MATERIALS	UNITS PER GAL	PRICE, ¢/UNIT	ANNUAL COST, $M	CENTS PER GAL	DOLLARS/ MET TON
Aspen, lb	61.80965	.5	15,453		
Sulfuric Acid, lb	.27790	4.3	598		
Calcium Hydroxide, lb	.19798	2.0	198		
Sodium Hydroxide, lb	.00880	26.0	114		
Corn, lb	.01237	4.5	28		
Ammonium Sulfate, lb	.44400	3.0	666		
Superphosphate(46), lb	1.91375	8.0	7,655		
Calcium Carbonate, lb	.88034	2.7	1,189		
Catalyst & Chemicals			1,950		

TOTAL RAW MATERIALS			27,851	55.70	1228.00
UTILITIES					
Power, kWH	1.78069	3.2	2,849		
Cooling Water, M Gal	.30312	5.8	879		
Process Water, M Gal	.03304	60.0	991		
Steam,50 psig, M Lb	.07363	295.0	10,860		
Steam,200 psig, M Lb	.01298	299.0	1,941		

TOTAL UTILITIES			17,521	35.04	772.51
OPERATING COSTS					
Labor, 60 Men @ $ 25,500		13 M/S	1,530		
Foremen, 13 Men @ $ 29,000		2 M/S	377		
Supervision, 3 Man @ $ 35,000		3 Man	105		
Maint., Material & Labor		6% of ISBL	5,568		

TOTAL OPERATING COST			7,580	15.16	334.22
OVERHEAD EXPENSES					
Direct Overhead	45% Lab. & Sup.		905		
Gen. Plant Overhead	65% Oper. Costs		4,927		
Insurance, Prop. Tax	1.5% Tot. Fix. Inv.		2,856		

TOTAL OVERHEAD EXPENSES			8,688	17.38	383.09
BY-PRODUCT CREDIT					
Carbon Dioxide, lb	8.12319	2.8	-11,373		
SCP, lb	.03051	15.0	-229		

TOTAL BY-PRODUCT CREDIT			-11,602	-23.20	-511.55
			==========	=======	==========
CASH COST OF PRODUCTION			50,038	100.07	2206.27
DEPRECIATION	20% ISBL + 10% OSBL		28,320	56.64	1248.69
			==========	=======	==========
NET COST OF PRODUCTION			78,358	156.71	3454.96
REQUIRED SALES PRICE AT 10% DCF				217.4	4792.4

TABLE V-D-5

COST OF PRODUCTION ESTIMATE FOR ABE PROCESS- LOW YIELD

CAPITAL SUMMARY

BASIS		CAPITAL COST	$MILLION
Location:	U.S. Gulf Coast	Battery Limits	92.8
	Mid-1982	Offsites	97.6
Capacity:	50.00 million gallons/yr		------
	22,680 metric tons/yr	Total Fixed Inv.	190.4
Str.Time:	8000 hours per year	Working Capital	15.6

PRODUCTION COST SUMMARY

RAW MATERIALS	UNITS PER GAL	PRICE, ¢/UNIT	ANNUAL COST, $M	CENTS PER GAL	DOLLARS/ MET TON
Aspen, lb	61.80955	1.0	30,906		
Sulfuric Acid, lb	.27790	4.3	598		
Calcium Hydroxide, lb	.19798	2.0	198		
Sodium Hydroxide, lb	.00380	26.0	114		
Corn, lb	.01237	4.5	28		
Ammonium Sulfate, lb	.44400	3.0	666		
Superphosphate(46 %), lb	1.91375	8.0	7,655		
Calcium Carbonate, lb	.88034	2.7	1,189		
Catalyst & Chemicals			1,950		

TOTAL RAW MATERIALS			43,304	86.60	1909.37
UTILITIES					
Power, kWH	1.78069	3.2	2,849		
Cooling Water, M Gal	.30312	5.8	879		
Process Water, M Gal	.03304	60.0	991		
Steam,50 psig, M Lb	.07363	295.0	10,860		
Steam,200 psig, M Lb	.01298	299.0	1,941		

TOTAL UTILITIES			17,521	35.04	772.51
OPERATING COSTS					
Labor, 60 Men @ $ 25,500		13 M/S	1,530		
Foremen, 13 men @ $ 29,000		2 M/S	377		
Supervision, 3 Man @ $ 35,000		3 Man	105		
Maint., Material & Labor	6% of ISBL		5,568		

TOTAL OPERATING COST			7,580	15.16	334.22
OVERHEAD EXPENSES					
Direct Overhead	45% Lab. & Sup.		905		
Gen. Plant Overhead	65% Oper. Costs		4,927		
Insurance, Prop. Tax	1.5% Tot. Fix. Inv.		2,856		

TOTAL OVERHEAD EXPENSES			8,688	17.38	383.09
BY-PRODUCT CREDIT					
Carbon Dioxide, lb	8.12319	2.8	-11,373		
SCP, lb	.03051	15.0	-229		

TOTAL BY-PRODUCT CREDIT			-11,602	-23.20	-511.55
			==========	=======	==========
CASH COST OF PRODUCTION			65,491	130.98	2887.64
DEPRECIATION	20% ISBL + 10% OSBL		28,320	56.64	1248.69
			==========	=======	==========
NET COST OF PRODUCTION			93,811	187.61	4136.32
REQUIRED SALES PRICE AT 10% DCF				250.9	5531.1

TABLE V-D-6

COST OF PRODUCTION ESTIMATE FOR ABE PROCESS- LOW YIELD

CAPITAL SUMMARY

BASIS		CAPITAL COST	$MILLION
Location:	U.S. Gulf Coast	Battery Limits	92.8
	mid-1982	Offsites	97.6
Capacity:	50.00 million gallons/yr		------
	22,680 metric tons/yr	Total Fixed Inv.	190.4
Str.Time:	8000 hours per year	Working Capital	17.1

PRODUCTION COST SUMMARY

RAW MATERIALS	UNITS PER GAL	PRICE, ¢/UNIT	ANNUAL COST, $M	CENTS PER GAL	DOLLARS/ MET TON
Aspen, lb	61.80965	1.5	46,360		
Sulfuric Acid, lb	.27790	4.3	598		
Calcium Hydroxide, lb	.19798	2.0	198		
Sodium Hydroxide, lb	.00880	26.0	114		
Corn, lb	.01237	4.5	28		
Ammonium Sulfate, lb	.44400	3.0	666		
Superphosphate(46), lb	1.91375	8.0	7,655		
Calcium Carbonate, lb	.88034	2.7	1,139		
Catalyst & Chemicals			1,950		

TOTAL RAW MATERIALS			58,757	117.51	2590.73
UTILITIES					
Power, kWH	1.78069	3.2	2,849		
Cooling Water, M Gal	.30312	5.8	879		
Process Water, M Gal	.03304	60.0	991		
Steam,50 psig, M Lb	.07363	295.0	10,860		
Steam,200 psig, M Lb	.01298	299.0	1,941		

TOTAL UTILITIES			17,521	35.04	772.51
OPERATING COSTS					
Labor, 60 Men @ $ 25,500		13 M/S	1,530		
Foremen, 13 Men @ $ 29,000		2 M/S	377		
Supervision, 3 Man @ $ 35,000		3 Man	105		
Maint., Material & Labor	6% of ISBL		5,568		

TOTAL OPERATING COST			7,580	15.16	334.22
OVERHEAD EXPENSES					
Direct Overhead	45% Lab. & Sup.		905		
Gen. Plant Overhead	65% Oper. Costs		4,927		
Insurance, Prop. Tax	1.5% Tot. Fix. Inv.		2,856		

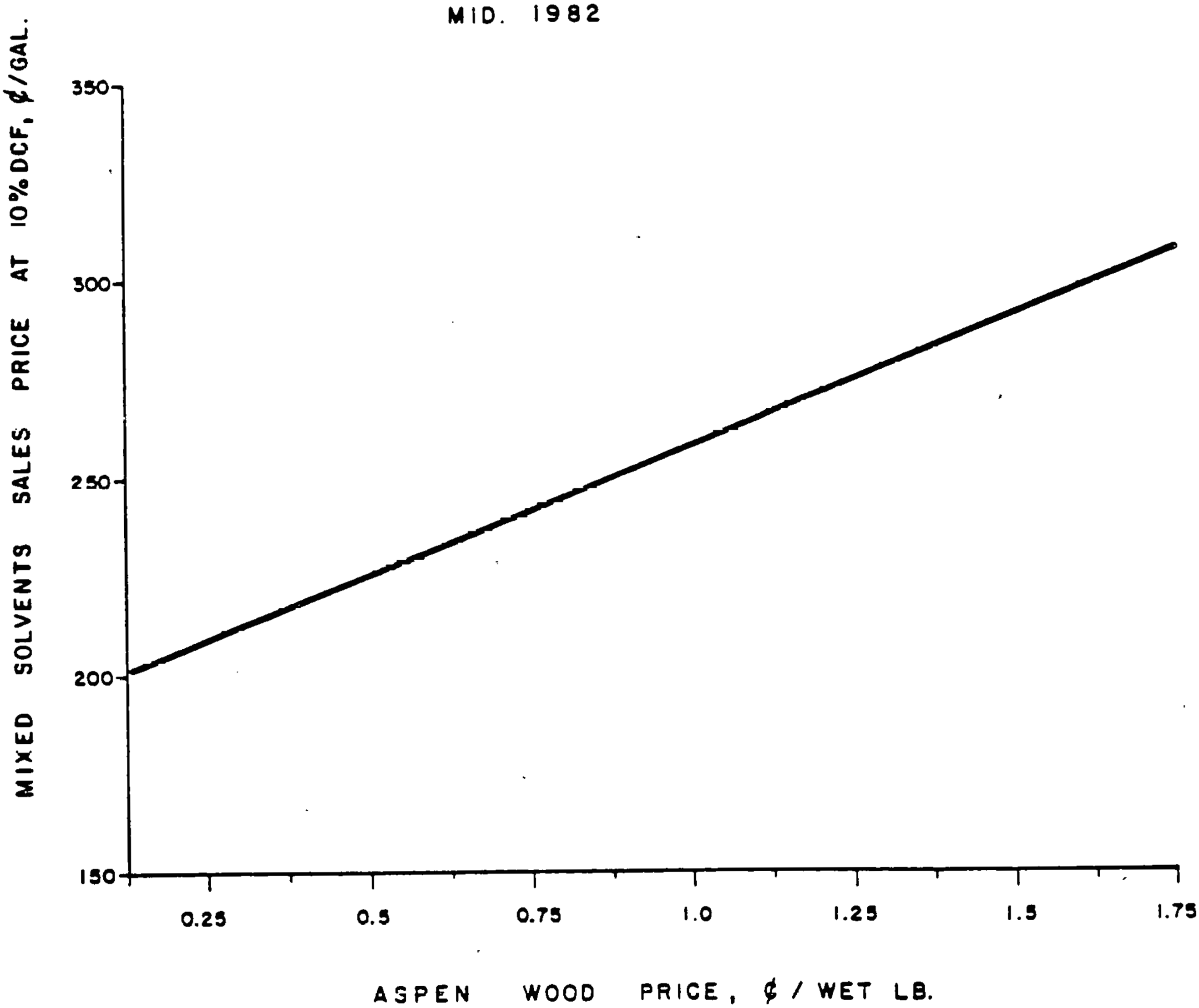
TOTAL OVERHEAD EXPENSES			8,688	17.38	383.09
BY-PRODUCT CREDIT					
Carbon Dioxide, lb	8.12319	2.8	-11,373		
SCP, lb	.03051	15.0	-229		

TOTAL BY-PRODUCT CREDIT			-11,602	-23.20	-511.55
			==========	=======	==========
CASH COST OF PRODUCTION			80,944	161.88	3569.00
DEPRECIATION	20% ISBL + 10% OSBL		28,320	56.64	1248.69
			==========	=======	==========
NET COST OF PRODUCTION			109,264	218.52	4817.69
REQUIRED SALES PRICE AT 10% DCF				284.4	6269.8

FIGURE V-D-1

SENSITIVITY OF ABE COST OF PRODUCTION TO ASPEN WOOD COST

BASIS: 50 MM GALS. / YR.
LOW YIELD CASE
U.S. GULF COAST
MID. 1982



CHEM SYSTEMS INC.
PROJECT NO. 6003 DATE

E. Revised Conventional Synthetic Routes to Butanol and Acetone[1]

Economics

Tables V-E-1 through V-E-4 represent cost of production (COP) estimates for propylene-based synthetic routes to butanol and acetone using CSI and DOE utility costs, respectively. These data are summarized in Table V-E-5. All cases produce 50 million gallons per year of products at a facility located on the U.S. Gulf Coast in mid-1982.

TABLE V-E-5

SUMMARY OF PROPYLENE-BASED SYNTHETIC ROUTES TO BUTANOL AND ACETONE

Basis: 50 million gallons per year, U.S. Gulf Coast, mid-1982.

	Butanol		Acetone	
	CSI Utilities	DOE Utilities	CSI Utilities	DOE Utilities
Investment, MM$				
Battery limits	61.2	61.2	41.2	41.2
Offsites	35.1	35.1	44.5	44.5
Total fixed investment	96.3	96.3	85.7	85.7
Cost of production, ¢/gal				
Raw materials	150.24	150.24	122.82	122.82
Utilities	5.13	3.86	20.07	16.81
Operating costs	9.45	9.45	7.05	7.05
Overhead expenses	9.98	9.98	8.10	8.10
By-product credit	(21.24)	(21.24)	(2.18)	(2.18)
Cash cost of production	153.56	152.29	155.86	152.60
Depreciation	31.50	31.50	25.38	25.38
Net cost of production	185.05	183.79	181.24	177.98
Selling price at 10% DCF	221.0	219.6	217.9	214.3

(1) Note that these revisions were made in order to incorporate more up-to-date feedstock and utility values consistent with the analysis that was performed for the biomass-based cases.

TABLE V-E-1

COST OF PRODUCTION ESTIMATE FOR BUTANOL
PROCESS- PROPYLENE CARBONYLAT

CAPITAL SUMMARY

BASIS		CAPITAL COST	$MILLION
Location:	U.S. Gulf Coast	Battery Limits	61.2
	Mid-1982	Offsites	35.1
Capacity:	50.00 million gallons/yr		------
	22,680 metric tons/yr	Total Fixed Inv.	96.3
Str.Time:	8000 hours per year	Working Capital	12.1

PRODUCTION COST SUMMARY

RAW MATERIALS	UNITS PER GAL	PRICE, ¢/UNIT	ANNUAL COST, $M	CENTS PER GAL	DOLLARS/ MET TON
Propylene (94.5%), Lb	5.33651	21.5	57,371		
Synthesis Gas, MSCF	.08525	240.0	10,230		
Hydrogen, MSCF	.03792	323.0	6,124		
Catalyst & Chemicals			1,400		

TOTAL RAW MATERIALS			75,125	150.24	3312.40
UTILITIES					
Power, kWH	.65406	3.2	1,047		
Cooling Water, M Gal	.05613	5.8	163		
Inert Gas, MSCF	.00135	1.0	1		
Fuel, MM BTU	-.02772	236.0	-3,271		
Steam,600 psig, M Lb	.02161	428.0	4,625		

TOTAL UTILITIES			2,564	5.13	113.03
OPERATING COSTS					
Labor, 32 men @ $ 25,500		6 M/S	816		
Foremen, 7 Men @ $ 29,000		1 M/S	203		
Supervision, 1 man @ $ 35,000		1 Man	35		
Maint., Material & Labor		6% of ISBL	3,672		

TOTAL OPERATING COST			4,726	9.45	208.38
OVERHEAD EXPENSES					
Direct Overhead	45% Lab. & Sup.		474		
Gen. Plant Overhead	65% Oper. Costs		3,072		
Insurance, Prop. Tax	1.5% Tot. Fix. Inv.		1,444		

TOTAL OVERHEAD EXPENSES			4,991	9.98	220.05
BY-PRODUCT CREDIT					
Isobutanol, Lb	.74542	28.5	-10,623		

TOTAL BY-PRODUCT CREDIT			-10,623	-21.24	-468.38
			==========	=======	==========
CASH COST OF PRODUCTION			76,782	153.56	3385.48
DEPRECIATION	20% ISBL + 10% OSBL		15,750	31.50	694.45
			==========	=======	==========
NET COST OF PRODUCTION			92,532	185.05	4079.93
REQUIRED SALES PRICE AT 10% DCF				221.0	4871.8

TABLE V-E-2

COST OF PRODUCTION ESTIMATE FOR ACETONE
PROCESS- IPA DEHYDROGENATION

CAPITAL SUMMARY

BASIS		CAPITAL COST	$MILLION
Location:	U.S. Gulf Coast	Battery Limits	41.2
	Mid-1982	Offsites	44.5
Capacity:	50.00 million gallons/yr		------
	22,680 metric tons/yr	Total Fixed Inv.	85.7
Str.Time:	8000 hours per year	Working Capital	11.6

PRODUCTION COST SUMMARY

RAW MATERIALS	UNITS PER GAL	PRICE, ¢/UNIT	ANNUAL COST, $M	CENTS PER GAL	DOLLARS/ MET TON
Propylene (95%), Lb	5.59520	21.5	60,152		
Catalyst & Chemicals			1,260		

TOTAL RAW MATERIALS			61,412	122.82	2707.76
UTILITIES					
Power, kWH	.99506	3.2	1,592		
Cooling Water, M Gal	.29826	5.8	365		
Process Water, M Gal	.00060	60.0	18		
Steam, 200 psig, M Lb	.03968	381.0	7,559		

TOTAL UTILITIES			10,034	20.07	442.43
OPERATING COSTS					
Labor, 32 men @ $ 25,500		6 M/S	816		
Foremen, 7 Men @ $ 29,000		1 M/S	203		
Supervision, 1 Man @ $ 35,000		1 Man	35		
Maint., Material & Labor	6% of ISBL		2,472		

TOTAL OPERATING COST			3,526	7.05	155.47
OVERHEAD EXPENSES					
Direct Overhead	45% Lab. & Sup.		474		
Gen. Plant Overhead	65% Oper. Costs		2,292		
Insurance, Prop. Tax	1.5% Tot. Fix. Inv.		1,285		

TOTAL OVERHEAD EXPENSES			4,052	8.10	178.65
BY-PRODUCT CREDIT					
Propane, Lb	.26595	8.2	-1,090		

TOTAL BY-PRODUCT CREDIT			-1,090	-2.18	-48.08
			==========	=======	==========
CASH COST OF PRODUCTION			77,933	155.86	3436.23
DEPRECIATION	20% ISBL + 10% OSBL		12,690	25.38	559.53
			==========	=======	==========
NET COST OF PRODUCTION			90,623	181.24	3995.75
REQUIRED SALES PRICE AT 10% DCF				217.9	4804.1

TABLE V-E-3

COST OF PRODUCTION ESTIMATE FOR BUTANOL
PROCESS- PROPYLENE CARBONYLAT

CAPITAL SUMMARY

BASIS		CAPITAL COST	$MILLION
Location:	U.S. Gulf Coast	Battery Limits	61.2
	Mid-1982	Offsites	35.1
Capacity:	50.00 million gallons/yr		------
	22,680 metric tons/yr	Total Fixed Inv.	96.3
Str.Time:	8000 hours per year	Working Capital	12.0

PRODUCTION COST SUMMARY

RAW MATERIALS	UNITS PER GAL	PRICE, ¢/UNIT	ANNUAL COST, $M	CENTS PER GAL	DOLLARS/ MET TON
Propylene (94.5%), Lb	5.33651	21.5	57,371		
Synthesis Gas, MSCF	.08525	240.0	10,230		
Hydrogen, MSCF	.03792	323.0	6,124		
Catalyst & Chemicals			1,400		

TOTAL RAW MATERIALS			75,125	150.24	3312.40
UTILITIES					
Power, kWH	.65406	4.3	1,406		
Cooling Water, M Gal	.05613	5.8	163		
Inert Gas, MSCF	.00135	1.0	1		
Fuel, MM BTU	-.02772	236.0	-3,271		
Steam,600 psig, M Lb	.02161	336.0	3,630		

TOTAL UTILITIES			1,929	3.86	35.06
OPERATING COSTS					
Labor, 32 Men @ $ 25,500		6 M/S	316		
Foremen, 7 Men @ $ 29,000		1 M/S	203		
Supervision, 1 Man @ $ 35,000		1 Man	35		
Maint., Material & Labor		6% of ISBL	3,672		

TOTAL OPERATING COST			4,726	9.45	208.38
OVERHEAD EXPENSES					
Direct Overhead	45% Lab. & Sup.		474		
Gen. Plant Overhead	65% Oper. Costs		3,072		
Insurance, Prop. Tax	1.5% Tot. Fix. Inv.		1,444		

TOTAL OVERHEAD EXPENSES			4,991	9.98	220.05
BY-PRODUCT CREDIT					
Isobutanol, Lb	.74542	28.5	-10,623		

TOTAL BY-PRODUCT CREDIT			-10,623	-21.24	-468.38
			==========	=======	==========
CASH COST OF PRODUCTION			76,148	152.29	3357.51
DEPRECIATION	20% ISBL + 10% OSBL		15,750	31.50	694.45
			==========	=======	==========
NET COST OF PRODUCTION			91,898	183.79	4051.96
REQUIRED SALES PRICE AT 10% DCF				219.6	4841.2

TABLE V-E-4

COST OF PRODUCTION ESTIMATE FOR ACETONE
PROCESS- IPA DEHYDROGENATION

CAPITAL SUMMARY

BASIS		CAPITAL COST	$MILLION
Location:	U.S. Gulf Coast	Battery Limits	41.2
	Mid-1982	Offsites	44.5
Capacity:	50.00 million gallons/yr		------
	22,680 metric tons/yr	Total Fixed Inv.	85.7
Str. Time:	8000 hours per year	Working Capital	11.3

PRODUCTION COST SUMMARY

RAW MATERIALS	UNITS PER GAL	PRICE, ¢/UNIT	ANNUAL COST, $M	CENTS PER GAL	DOLLARS/MET TON
Propylene (95%), Lb	5.59520	21.5	60,152		
Catalyst & Chemicals			1,260		

TOTAL RAW MATERIALS			61,412	122.82	2707.76
UTILITIES					
Power, kWH	.99506	3.2	1,592		
Cooling Water, M Gal	.29826	5.8	865		
Process Water, M Gal	.00060	60.0	18		
Steam, 200 psig, M Lb	.03968	299.0	5,932		

TOTAL UTILITIES			8,407	16.81	370.70
OPERATING COSTS					
Labor, 32 Men @ $ 25,500		6 M/S	816		
Foremen, 7 Men @ $ 29,000		1 M/S	203		
Supervision, 1 Man @ $ 35,000		1 Man	35		
Maint., Material & Labor	6% of ISBL		2,472		

TOTAL OPERATING COST			3,526	7.05	155.47
OVERHEAD EXPENSES					
Direct Overhead	45% Lab. & Sup.		474		
Gen. Plant Overhead	65% Oper. Costs		2,292		
Insurance, Prop. Tax	1.5% Tot. Fix. Inv.		1,285		

TOTAL OVERHEAD EXPENSES			4,052	8.10	178.65
BY-PRODUCT CREDIT					
Propane, Lb	.26595	8.2	-1,090		

TOTAL BY-PRODUCT CREDIT			-1,090	-2.18	-48.08
			==========	=======	==========
CASH COST OF PRODUCTION			76,306	152.60	3364.49
DEPRECIATION	20% ISBL + 10% OSBL		12,690	25.38	559.53
			==========	=======	==========
NET COST OF PRODUCTION			38,996	177.98	3924.02
REQUIRED SALES PRICE AT 10% DCF				214.3	4725.8

The major revision to the COP analysis of the synthetic routes is a change in propylene raw material price. The propylene (chemical grade) price of 25 cents per pound reported previously was the forecasted price used at the time the Phase I report was issued. However, the actual historical price for the mid-1982 time frame is 21.5 cents per pound, and this price is reflected in the revised COP analysis. In addition, minor changes in utility costs to reflect historical costs were made and outside battery limits (OSBL) capital cost was recalculated on a more realistic basis.

As can be seen from Table V-E-5, butanol can be produced for 221.0 cents per gallon and acetone 218.0 cents per gallon using CSI utility costs. These values represent the sales price at 10 percent DCF, and represent improvements relative to the original basis of 235.1 and 232.4 cents per gallon, respectively, for butanol and acetone.

VI. SAMPLE CASE STUDY: PRODUCTION OF CITRIC ACID AND FURFURAL FROM WOOD-DERIVED SUGARS

A. Introduction

The base case and subsequent parametric analysis have heretofore studied various scenarios for acetone, butanol, and ethanol (ABE) production via fermentation of wood (or biomass) derived sugars. However, there are potentially many other chemicals of commercial importance which could be produced by a biological route. The objective of this case study is to demonstrate the adaptability of the methodology utilized in generating the base case studies towards producing other biologically derived chemicals.

There are many chemicals which can be produced by biological routes. Biological routes include enzyme catalyzed reactions and reactions which occur as a result of cell metabolism in specific media, such as fermentation. Since the methodology used to analyze the ABE base cases produces five and six carbon sugars as precursors from biomass sources, the other chemicals considered are limited to those which can be produced from these sugars. Some potential candidates are listed in Table VI-A-1.

TABLE VI-A-1

REPRESENTING LIST OF POTENTIAL SUGAR-BASED CHEMICALS

Acetic acid	Ethanol
Lactic acid	Propionic acid
Citric acid	Riboflavin and other vitamins
Gluconic acid	Penicillan
Fumaric acid	Streptomycin and other antibiotics
Itaconic acid	Glycerol
2,3-Butanediol	HMF
Furfural	Formic acid
Levulinic acid	

Ethanol is not a viable candidate for study herein since so much work is being done on ethanol from biomass by other workers. The vitamins and antibiotics may also be eliminated since they are produced by the pharmaceutical industry in very small volumes. Glycerol is eliminated since even under the most favorable circumstances some ethanol is made during fermentation. Itaconic acid, levulinic acid, gluconic acid, 2,3-butanediol and HMF have little commercial importance and were not considered. The remaining chemicals were evaluated based on the perceived potential for improving conventional technology economics and the current demand for the product in relation to a reasonable size plant, which would be required to take advantage of economy of scale. Table VI-A-2 lists these chemicals and summarizes current U.S. capacity and demand (if known) and current U.S. list prices and market prices (if known).

TABLE VI-A-2

U.S. PRODUCTION CAPACITY, DEMAND AND PRICES FOR SELECTED CHEMICALS

	U.S. Capacity, MM lb/yr	Demand, MM lb/yr	List Price, ¢/lb
Acetic acid	4,610	2,778	25.0
Citric acid	350		81(2)-119.0(3)
Furfural	200(1)		66.0
Propionic acid	240		33.0
Fumaric acid	86		53(4)-69.0(5)
Formic acid	85		51.5(6)
Lactic acid	18	??	92(7)-94.0(8)

(1) Quaker Oats has recently closed down some of its capacity and is believed to be importing furfural from the People's Republic of China.
(2) Anhydrous.
(3) Hydrous.
(4) Technical grade.
(5) Food grade.
(6) 95 percent.
(7) Technical grade, 88 percent.
(8) Food grade, 88 percent.

It should be noted that furfural and formic acid are not produced biologically, but are produced via acid catalyzed decomposition of xylose and HMF, respectively. They would only be produced as coproducts in conjunction with a biologically produced primary product.

Although acetic acid is by far the most important potential product from a commercial point of view, the biological route will probably not be at all competitive with the conventional route in the foreseeable future. Acetic acid is conventionally made by methanol carbonylation, and at the current depressed methanol price of 45 cents per gallon (6.8 cents per pound), acetic acid can be produced for about 14 cents per pound at 10 percent DCF. These economics, coupled with the fact that current capacity is considerably larger than current demand, result in an actual market price for acetic acid that is significantly below its 25.0 cent per pound list price. Since the biological route must first make ethanol, and the best biological cases produce ethanol (without purification) for about 20 cents per pound, it is very unlikely that the biological acetic acid case will be competitive unless there is a significant rise in methanol prices.

Citric acid, however, has a current list price which ranges from 81-119.0 cents per pound, which leaves a greater margin for producing it from wood-derived sugars at a lower cost of production. Although one company uses corn as feedstock, citric acid is conventionally made via fermentation of a relatively expensive raw material, molasses (3.8 cents per pound). By contrast, wood costs about 1.0 cents per pound. The wood-based route would also have the additional advantage of producing furfural as a by-product. Furfural currently lists for 66.0 cents per pound, and although the traditional demand for furfural at this price has slackened, new uses for furfural could arise from favorable production economics. Low-cost furfural could open up a variety of possible uses for furfural as a chemical building block. These potential applications include expanded production of THF, 1,4 butanediol and adipic acid as well as the traditional uses of furfural, i.e., furfural alcohol, furfuraldehyde and furan resins.

B. Design Basis

Enzyme Hydrolysis

The design basis for the front end of the citric acid/furfural fermentation facility is basically the same as for the ABE base cases. The front end includes the following plant sections:

- Raw materials handling
- Prehydrolysis
- Enzyme hydrolysis
- Enzyme production

The plant capacity is 100 million pounds per year of citric acid. The only change in the design basis for the front end sugar producing sections is the concentration of sugar resulting from enzyme hydrolysis. The optimum glucose concentration for citric acid fermentation is 15 weight percent. Because of the limited solids concentration possible in enzyme hydrolysis, however, this level is not obtainable without concentration. In order to get as close as possible to 15 percent without concentration, a high solids concentration, i.e., 15 percent, is used in enzyme hydrolysis. This gives approximately 7.3 percent glucose and 10.1 percent total sugars (glucose and xylose). Only the glucose is fermented to citric acid, with the xylose subsequently decomposed to make furfural. The 7.3 percent glucose stream is concentrated to 15 percent by multiple effect evaporation.

Fermentation

The concentration of the sugar streams from prehydrolysis and enzyme hydrolysis is accomplished with a quadruple effect evaporator and a steam gain of 3.36 pounds H_2O evaporated per pound of steam. Note that this analysis assumes that the C_5 sugar stream from prehydrolysis passes through the fermentation operation unchanged. This assumption would need to be confirmed in actual practice.

Citric acid is produced from hexose units through the usual metabolic systems leading into the Krebs cycle via pyruvic acid. One mole of hexose can yield one mole of citric acid or of itaconic acid. Consequently, it is assumed that the itaconic acid biosynthesis passes through the same metabolic system as does citric acid biosynthesis. Both compounds accumulate when the Krebs cycle is blocked or inhibited. A diagram describing the Krebs cycle is illustrated in Figure VI-B-1.

In non-inhibited cultures, either citric or itaconic acid may serve as a good carbon source for growth of the respective organism. Enzymatic inhibition may be due to the low pH of the fermentation and by specific inhibitors such as copper ions, or by one of a series of organic compounds which exert effects similar to that of copper ion.

Citric acid is produced commercially by the aerobic fermentation of sugar solutions. Existing commercial operations use either sugar beet molasses or dextrose (corn sugar) as the sugar source. The fermentation process can be summarized by the following equation:

Microorganism + Substrate ⟶ More Microorganisms + Metabolic Products

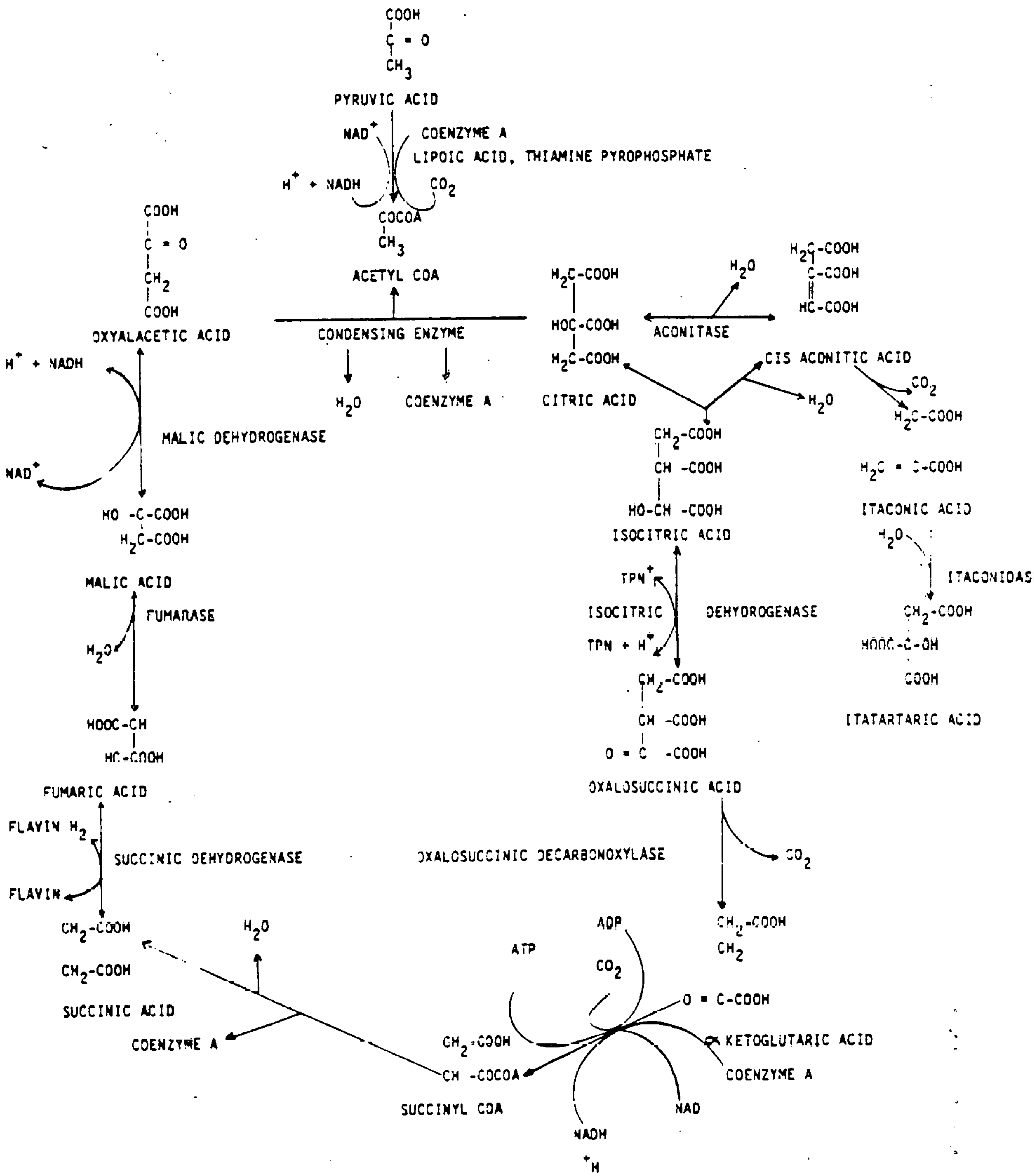
Basically, the citric acid synthesis involves the inoculation of the sugar solution with a special strain of the microorganism, Aspergillus niger, which under carefully controlled conditions converts the sugar to citric acid.

The production of citric acid via submerged fermentation involves four basic steps:

- Preparation of a "seed mash"
- Inoculation of the molasses solution with the seed mash
- Batch fermentation
- Citric acid recovery

FIGURE VI-8-1

KREBS TERMINAL RESPIRATORY CYCLE AND DERIVATIVES



Researchers have expended considerable effort over the years to determine the factors that provide optimum conversion of the contained sugar to citric acid. High sugar conversion is extremely sensitive to the concentration of the metallic ions Cu^{+2}, Zn^{+2}, Fe^{+3}, and Mn^{+2} in the molasses solution. Sufficient amounts of nutrients must also be added to the solution to enable the microorganism to reach a suitable level of growth. Nutrients commonly used include ammonium nitrate, magnesium sulfate heptahydrate, and potassium dihydrogen phosphate. Producers undoubtedly rely heavily upon processing experience in determining the combination of metallic ions and nutrients necessary for high citric acid yield.

The nutrient media used are summarized in Table VI-B-1.

TABLE VI-B-1

COMPOSITION OF FERMENTATION SOLUTION

	Weight Percent
15 percent glucose solution	99.1
Dry inoculant	0.5
NH_4NO_3	0.2
KH_2PO_4	0.1
$MgSO_4$ $7H_2O$	0.1

The design parameters used for citric acid fermentation are summarized in Table VI-B-2.

TABLE VI-B-2

CITRIC ACID FERMENTATION DESIGN PARAMETERS

• Microorganism	Aspergillus Niger
• Feed	15 weight percent glucose
• Temperature	28-33°C
• pH	3.5
• Pressure	10-15 psig
• Fermentation time	4-7 days
• Yield	70 weight percent on glucose; 100 percent glucose conversion; remainder goes to cell maintenance and CO_2.

Furfural Production

The sugar stream emanating from citric acid fermentation contains approximately 3.8 weight percent xylose. The contained xylose can be converted to furfural in an acid catalyzed thermal decomposition reaction:

$$C_5H_{10}O_5 \xrightarrow{H_2SO_4} C_4H_3OCHO + 3H_2O$$
$$C_4H_3OCHO \longrightarrow \text{Tars}$$

The furfural formed undergoes further decomposition to tars, although at a somewhat lower rate than xylose decomposition. Therefore, in order to optimize the formation of furfural, residence time must be kept sufficiently low as to minimize the furfural decomposition reaction. Studies at Dartmouth have resulted in the development of a kinetic model describing hemicellulose kinetics. Part of this model describes the kinetics of xylose decomposition as a function of xylose concentration, acid concentration, reaction temperature and residence time. From this kinetic expression, furfural yields can be calculated as a function of the above variables.

The furfural production reactor is a liquid phase plug flow type reactor similar to the prehydrolysis reactor except no solids are present. In order to simulate plug flow in a liquid phase reactor, a series of plates or baffles are situated inside the reactor to minimize backmixing. The unoptimized reactor conditions for furfural production are summarized in Table VI-B-3.

Furfural Recovery

Furfural recovery is accomplished by conventional azeotropic distillation using a three column system. One of the columns, the lights column, removes any light components (such as methanol) formed during hemicellulose hydrolysis. This column is generally very small. The design parameters for the other two columns (the azeotropic and dehydration columns) are summarized in Table VI-B-4.

TABLE VI-B-3

FURFURAL PRODUCTION DESIGN BASIS

• Xylose concentration	2.76 weight percent
• Temperature	243°C
• Acid concentration	1.84 weight percent
• Residence time	21.6 seconds
• Xylose conversion	99.2 percent
• Furfural yield	66.9 mole percent; remainder goes to tars
• Flash	90.3 percent furfural flashes 20.8 percent H_2O flashes

TABLE VI-B-4

FURFURAL RECOVERY DESIGN BASIS

Azeotrope column	
Feed	Saturated vapor at dew point, 33 percent furfural
Overhead	Water/furfural azeotrope 65/35 percent BP - 98°C
Decantation	
Organic layer	84 percent furfural/16 percent water
Aqueous layer	18 percent furfural/82 percent water
Reflux ratio	2.4
Actual trays	18
Dehydration column	
Feed	84 percent furfural/16 percent water
Overhead	35 percent furfural/65 percent water
Reflux ratio	0.34
Actual trays	10
Overall steam consumption	3.75 pounds steam/pound furfural

Process Description

The following plant sections have the same process description as the base cases:

- Raw materials handling
- Prehydrolysis
- Enzyme production
- Enzyme hydrolysis
- CO_2 recovery
- Heat generation

Process schemes for these plant sections are identical to the base cases and the reader is referred to earlier sections of this report for the process flow diagrams. The remainder of the plant is divided up into the following plant sections:

- Citric acid fermentation
- Furfural production
- Furfural recovery

Citric acid fermentation, which also includes citric acid recovery, is illustrated in Figure VI-B-2. The solubles from prehydrolysis (xylose and glucose) are combined with the solubles from enzyme hydrolysis (glucose) and sent to a pH adjustment tank. Here pH adjusted stream is filtered to remove the $CaSO_4$ formed during neutralization, with the filter cake being thoroughly washed to recover most of the solubles. The filtrate is then concentrated to 15 weight percent glucose in a multiple effect evaporator with four effects. The concentrated sugar stream is ready for fermentation, with a small side stream of sugar being diverted for seed preparation.

The seed mash is prepared by inoculating the dilute solution in the seed tank with a special strain of Aspergillus niger. Potassium dihydrogen phosphate and magnesium sulfate are also added. The acidity of the mixture is then adjusted between pH=5-6 by adding HCl. The mixture is incubated at 25-30°C for one to three days while air is continuously sparged through the solution.

The 15 percent sugar solution in the fermentor is inoculated with a portion of the seed mash. The charge of seed mash is in the range of 2-8

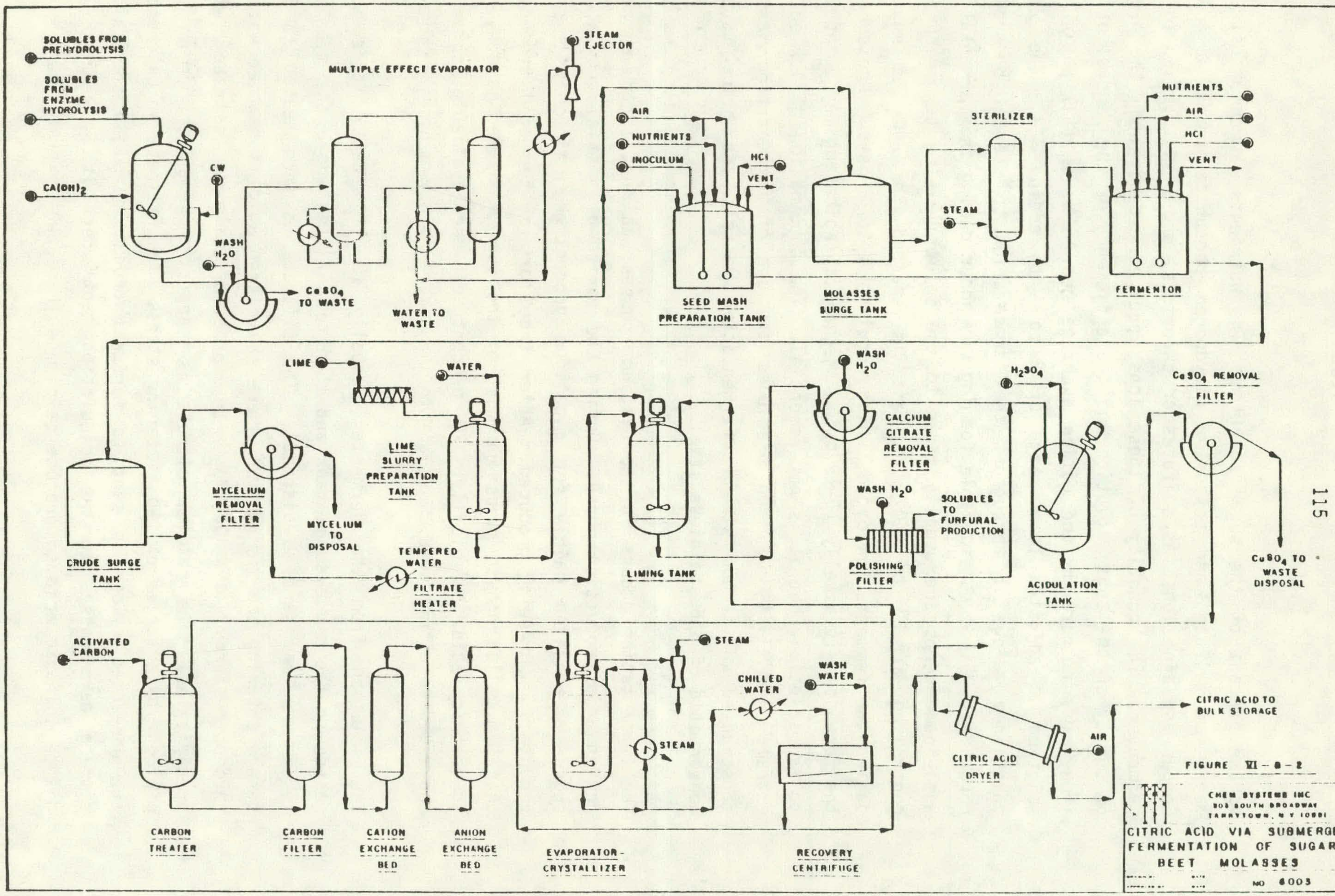


FIGURE VI-8-2

CITRIC ACID VIA SUBMERGED FERMENTATION OF SUGAR BEET MOLASSES volume percent of the sugar solution. The nutrients, ammonium nitrate, magnesium sulfate, and potassium dihydrogen phosphate are also added. Copper and zinc ions, as sulfates or chlorides, are also supplied to the fermenter. The quantity of these ions required depends upon the degree of purity of the sugar solution and the efficiency of the removal of the iron and manganese in the cation exchange bed. The initial pH of the solution is adjusted to a range of 2 to 4 by adding HCl. The pH is maintained below 3.5 throughout the fermentation to prevent by-product formation. Since considerable foaming may occur during the fermentation, anti-foam agents are usually added to the solution. The temperature is controlled at 27 to 33^{o}C and air is sparged through the mixture during the fermentation cycle, which requires 4 to 7 days. Pressure in the fermenter is usually 10-15 psig. Mechanical agitating may be applied to the solution, but it is not essential. The combination of low pH and extreme sensitivity to iron precludes the use of fermentation vessels constructed of carbon steel. Fermentation vessels are hence usually constructed of 316 stainless steel.

The fermentation occurs in two distinct phases; an initial growth phase and an acid production phase. During the period of initial growth, the sugar is utilized mainly for mycellium production with only a small amount of acid being produced. After the start of the acid production phase, mycellia growth stops and most of the sugar consumed is converted to citric acid. Typically, 70 percent of the available sugar is converted to citric acid.

The solution from the fermenter, usually 10-12 percent citric acid, passes to the crude surge tank and then to a filter where the mycellium is removed and sent to disposal. The filtrate is then heated to 60^{o}C before entering the liming tank where the citric acid is reacted with an aqueous slurry of $Ca(OH)_2$ to produce calcium citrate which precipitates. The calcium citrate is separated from the liming tank effluent by a series of two filtering steps. The filtrate which contains the unreacted xylose is sent to furfural production while the calcium citrate cake passes to the acidulation tank where it is reacted with dilute sulfuric acid to produce calcium sulfate, which precipitates, and citric acid. The calcium sulfate is removed by filtration and the filtrate, a solution of citric acid, is sent to the decolorizing tank. Here, the solution is treated with activated carbon to remove color bodies. After filtration to remove the carbon, the solution passes through the cation and anion exchange resin bed for demineralization. The demineralized solution is then concentrated in the evaporator-crystallizer. The hot slurry enters the centrifuge where the crystals are washed and separated. A portion of the mother liquor from the centrifuge is recycled to the evaporator-crystallizer while the remainder is returned to the liming tank. The citric acid crystals are dried and then sent to bulk storage. Operation of the evaporator-crystallizer, centrifuge and dryer above 40-45^{o}C results in anhydrous citric acid. When the concentration and recovery operation is conducted at 25-30^{o}C, citric acid monohydrate is produced.

Furfural Production and Recovery

Flow diagrams illustraing furfural production and recovery are presented in Figures VI-B-3 and VI-B-4.

The filtrate from the polishing filter, which contains 3.8 weight percent xylose, is sent to a dilution tank. Here water is added to dilute the xylose solution to about 2.75 weight percent.

The dilute xylose solution then enters the furfural plug flow reactor (PFR) where acid is first mixed in to bring its concentration to 1.84 weight percent, relative to water. The reactants are then heated to 243^{o}C nearly instantaneously (approximately 0.5 seconds) and the decomposition reactions take place. Reactor residence time is about 21-22 seconds. The reactor is zirconium clad to prevent corrosion in the acidic environment, and sieve plates are spaced intermittently down the reactor barrel to prevent back mixing. Under these conditions 99.2 percent of the xylose is converted with a furfural yield of about 67 mole percent. A rapid action ball valve quenches the reactor in an operation that is nearly continuous. The pressure is letdown to atmospheric across the valve, flashing most of the furfural (90 percent) and some of the

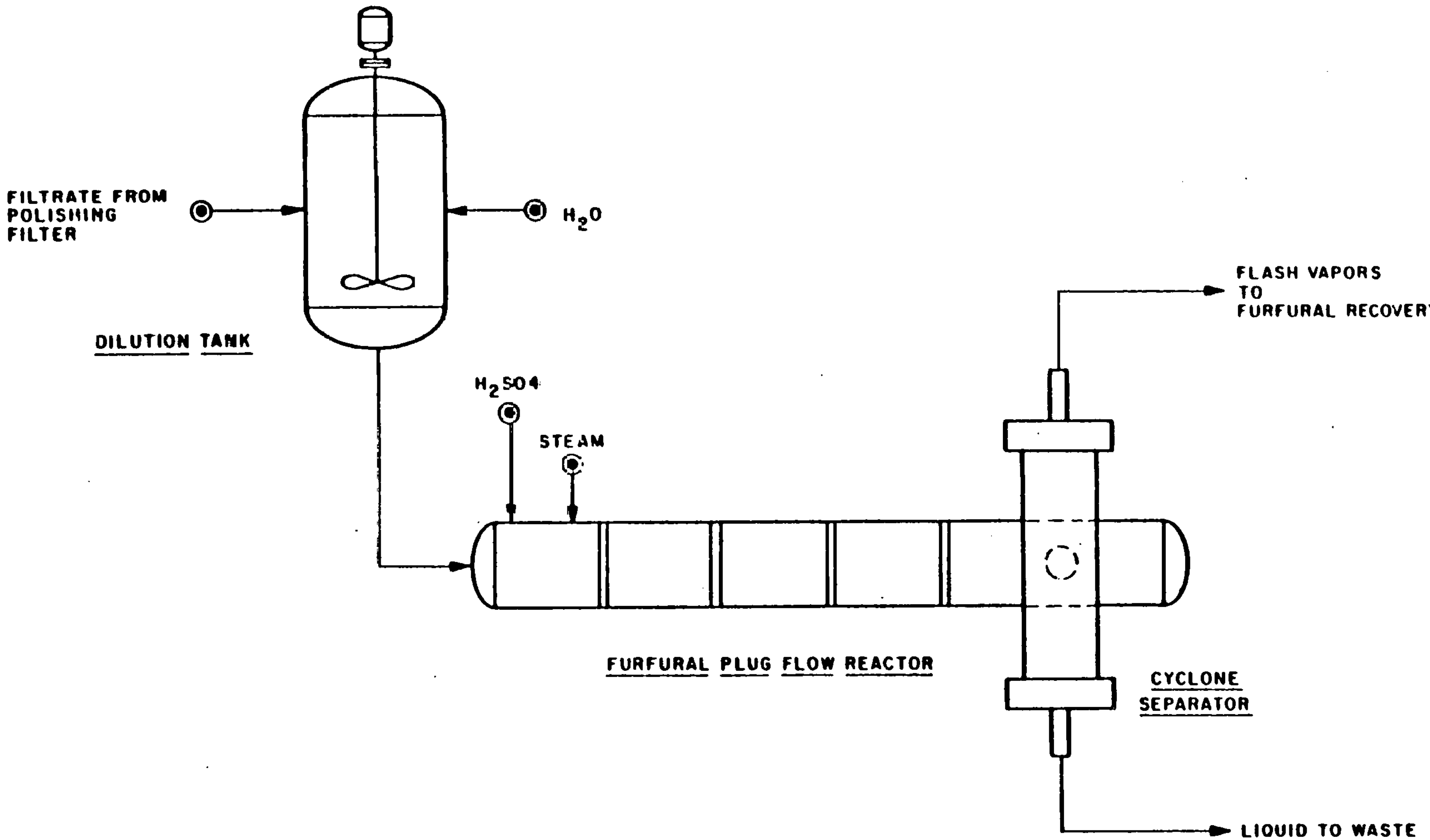


FIGURE VI-B-3

CHEM SYSTEMS INC.
303 SOUTH BROADWAY
TARRYTOWN, N.Y. 10591

FURFURAL PRODUCTION

NO 6003

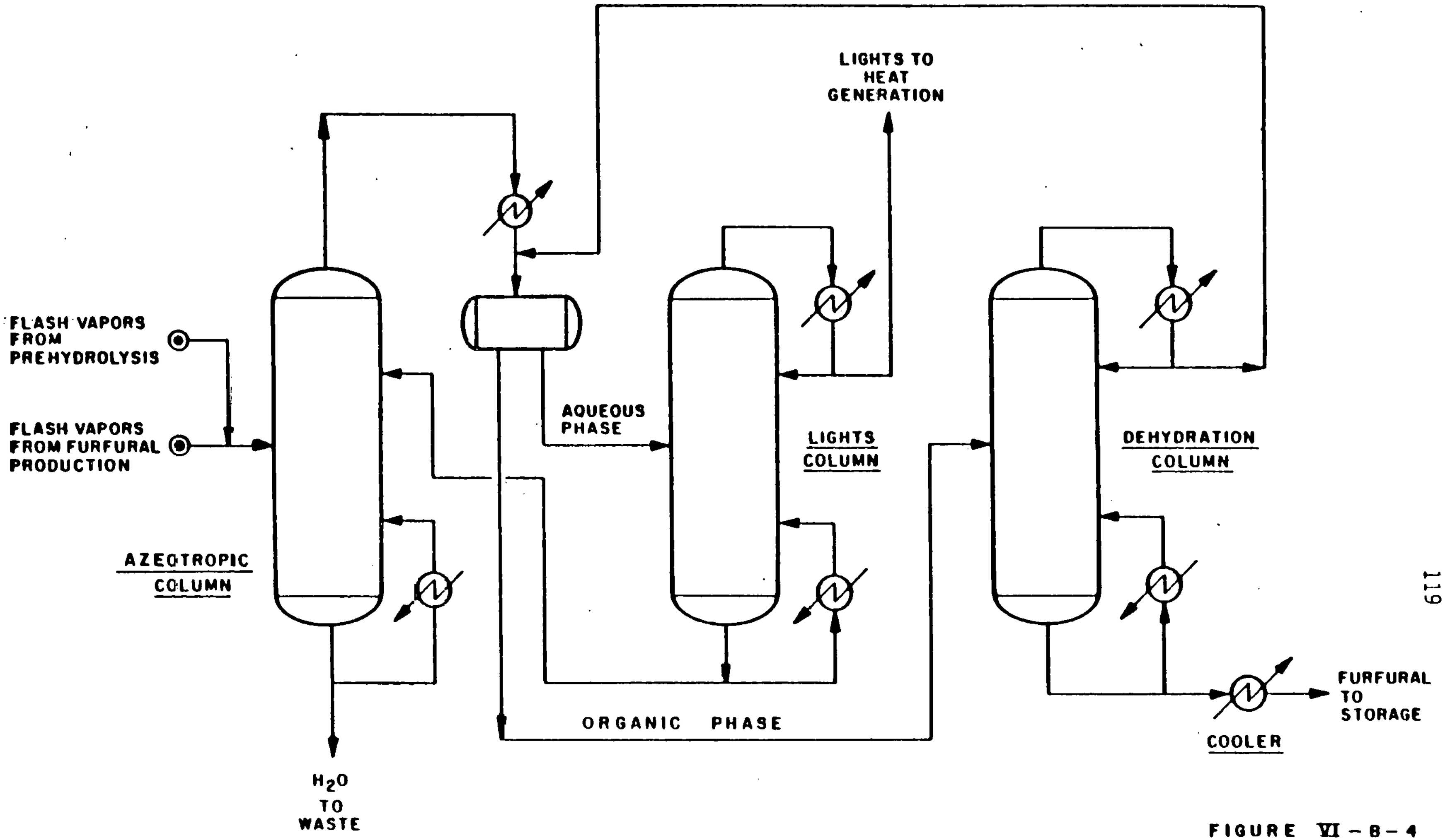


FIGURE VI - B - 4

CHEM SYSTEMS INC.
303 SOUTH BROADWAY
TARRYTOWN, N.Y. 10591

TITLE

FURFURAL RECOVERY

DRAWN BY DATE
APPROVED BY DATE NO 6003

water (21 percent). The liquid and vapor components are separated in a cylcone on the low pressure side of the valve. The flash vapors, containing about 4.8 percent furfural, are combined with the flash vapors from prehydrolysis and sent to furfural recovery.

The furfural/water vapor stream, a saturated vapor at its dew point, (210^{o}C) enters the azeotrope distillation column where the furfural/water azeotrope is taken overhead. The overhead vapor, composed of 35 weight percent furfural and 65 weight percent water, is condensed at 208^{o}F and sent to a decanter where it separates into two layers. The lower organic layer, composed of 84 percent furfural and 16 percent water is sent to a dehydration column. The upper layer, composed of 18 percent furfural and 82 percent water, (and if we considered it, some methanol) is sent to a lights column, where low boilers, such as methanol, are removed in the overhead. the column bottoms, which is essentially the aqueous layer from the decanter (18 percent furfural, 82 percent water) is returned to the azeotropic column as reflux. The reflux ratio is 2.4. The azeotropic column bottoms, which contains all the water and less than 0.2 percent furfural, (as well as other heavies such as HMF and acetic acid) is sent to waste treatment.

The furfural rich organic layer from the decanter enters a dehydration column where the remainder of the water is removed. The furfural product stream, which forms the column bottoms, is cooled and filtered to remove any residual solids prior to storage.

The overheads from the dehydration column, which is the water/furfural azeotrope, is condensed and recycled to the azeotropic column decanter.

C. Economics

Tables VI-C-1 and VI-C-2 represent cost of production estimates for the wood based citric acid fermentation facility, using both CSI and DOE utility costs, respectively. Tables VI-C-3 and VI-C-4 represent cost of production estimates for the conventional molasses based citric acid fermentation.

TABLE VI-C-1

COST OF PRODUCTION ESTIMATE FOR CITRIC ACID
PROCESS- FERMENTATION

CAPITAL SUMMARY

BASIS		CAPITAL COST	$MILLION
Location:	U.S. Gulf Coast	Battery Limits	90.2
	Mid-1982	Offsites	53.3
Capacity:	100.00 million pounds/yr		------
	45,360 metric tons/yr	Total Fixed Inv.	143.5
Str.Time:	8000 hours per year	Working Capital	10.3

PRODUCTION COST SUMMARY

RAW MATERIALS	UNITS PER LB	PRICE, ¢/UNIT	ANNUAL COST, $M	CENTS PER LB	DOLLARS/ MET TON
Aspen, Lb	5.11300	1.0	5,113		
H2SO4, Lb	1.22970	4.3	5,288		
Ca(OH)2, Lb	.04130	2.0	83		
Na(OH), Lb	.00130	26.0	34		
Amm. Nitr. (33.5%), Lb	.03000	7.5	225		
KH2PO4, Lb	.01500	50.0	750		
MgSO4 7H2O, Lb	.01200	11.5	138		
Lime, Lb	.57500	1.5	863		
Catalyst & Chemicals			500		

TOTAL RAW MATERIALS			12,993	12.99	286.44
UTILITIES					
Power, kWH	.48670	3.2	1,557		
Cooling Water, M Gal	.09030	5.8	524		
Process Water, M Gal	.00278	60.0	167		
Steam,55 psig, M Lb	.00086	375.0	323		
Steam,200 psig, M Lb	.01890	381.0	7,201		

TOTAL UTILITIES			9,771	9.77	215.42
OPERATING COSTS					
Labor, 60 Men @ $ 25,500	13 M/S		1,530		
Foremen, 13 Men @ $ 29,000	2 M/S		377		
Supervision, 3 Man @ $ 35,000	3 Man		105		
Maint., Material & Labor	6% of ISBL		5,412		

TOTAL OPERATING COST			7,424	7.42	163.67
OVERHEAD EXPENSES					
Direct Overhead	45% Lab. & Sup.		905		
Gen. Plant Overhead	65% Oper. Costs		4,826		
Insurance, Prop. Tax	1.5% Tot. Fix. Inv.		2,152		

TOTAL OVERHEAD EXPENSES			7,883	7.88	173.80
BY-PRODUCT CREDIT					
CO2, Lb	.42830	2.8	-1,199		
Furfural, Lb	.25750	20.0	-5,150		
SCP, Lb	.00286	15.0	-43		

TOTAL BY-PRODUCT CREDIT			-6,392	-6.39	-140.92
			==========	=======	==========
CASH COST OF PRODUCTION			31,679	31.68	698.40
DEPRECIATION	20% ISBL + 10% OSBL		23,370	23.37	515.22
			==========	=======	==========
NET COST OF PRODUCTION			55,049	55.05	1213.62
REQUIRED SALES PRICE AT 10% DCF				75.2	1658.0

TABLE VI-C-2

COST OF PRODUCTION ESTIMATE FOR CITRIC ACID
PROCESS- FERMENTATION

CAPITAL SUMMARY

BASIS		CAPITAL COST	$MILLION
Location:	U.S. Gulf Coast	Battery Limits	90.2
	Mid-1982	Offsites	53.3
Capacity:	100.00 million pounds/yr		------
	45,360 metric tons/yr	Total Fixed Inv.	143.5
Str.Time:	8000 hours per year	Working Capital	10.1

PRODUCTION COST SUMMARY

RAW MATERIALS	UNITS PER LB	PRICE, ¢/UNIT	ANNUAL COST, $M	CENTS PER LB	DOLLARS/ MET TON
Aspen, Lb	5.11300	1.0	5,113		
H2SO4, Lb	1.22970	4.3	5,288		
Ca(OH)2, Lb	.04130	2.0	83		
Na(OH), Lb	.00130	26.0	34		
Amm. Nitr. 33.5%, Lb	.03000	7.5	225		
KH2PO4, Lb	.01500	50.0	750		
MgSO4 7H2O, Lb	.01200	11.5	130		
Lime, Lb	.57500	1.5	863		
Catalyst & Chemicals			500		

TOTAL RAW MATERIALS			12,993	12.99	286.44
UTILITIES					
Power, kWH	.48670	4.3	2,093		
Cooling Water, M Gal	.09030	5.8	524		
Process Water, M Gal	.00278	60.0	167		
Steam,55 psig, M Lb	.00086	295.0	254		
Steam,200 psig, M Lb	.01890	299.0	5,651		

TOTAL UTILITIES			8,688	8.69	191.54
OPERATING COSTS					
Labor, 60 Men @ $ 25,500		13 M/S	1,530		
Foremen, 13 Men @ $ 29,000		2 M/S	377		
Supervision, 3 Man @ $ 35,000		3 Man	105		
Maint., Material & Labor	6% of ISBL		5,412		

TOTAL OPERATING COST			7,424	7.42	163.67
OVERHEAD EXPENSES					
Direct Overhead	45% Lab. & Sup.		905		
Gen. Plant Overhead	65% Oper. Costs		4,826		
Insurance, Prop. Tax	1.5% Tot. Fix. Inv.		2,152		

TOTAL OVERHEAD EXPENSES			7,883	7.88	173.80
BY-PRODUCT CREDIT					
CO2, Lb	.42830	2.8	-1,199		
Furfural, Lb	.25750	20.0	-5,150		
SCP, Lb	.00286	15.0	-43		

TOTAL BY-PRODUCT CREDIT			-6,392	-6.39	-140.92
			==========	=======	==========
CASH COST OF PRODUCTION			30,596	30.60	674.52
DEPRECIATION	20% ISBL + 10% OSBL		23,370	23.37	515.22
			==========	=======	==========
NET COST OF PRODUCTION			53,966	53.97	1189.74
REQUIRED SALES PRICE AT 10% DCF				74.0	1631.9

TABLE VI-C-3

COST OF PRODUCTION ESTIMATE FOR CITRIC ACID
PROCESS- MOLASSES FERMENT.

CAPITAL SUMMARY

BASIS		CAPITAL COST	$MILLIC
Location:	U.S. Gulf Coast	Battery Limits	67.
	Mid-1982	Offsites	27.
Capacity:	100.00 million pounds/yr		-----
	45,360 metric tons/yr	Total Fixed Inv.	95.
Str.Time:	8000 hours per year	Working Capital	8.

PRODUCTION COST SUMMARY

RAW MATERIALS	UNITS PER LB	PRICE, c/UNIT	ANNUAL COST, $M	CENTS PER LB	DOLLARS MET TON
Beet Molasses, Lb	3.75000	3.8	14,250		
Lime, Lb	.57500	1.5	863		
Sulfuric Acid, Lb	1.00000	4.3	4,300		
Ammonium Nitrate, Lb	.03000	7.5	225		
MgSO4, Lb	.01200	11.5	138		
KH2PO4, Lb	.01500	50.0	750		
Catalyst & Chemicals			740		

TOTAL RAW MATERIALS			21,265	21.27	468.82
UTILITIES					
Power, kWH	.40000	3.2	1,280		
Cooling Water, M Gal	.06540	5.8	379		
Process Water, M Gal	.00410	60.0	246		
Steam,200 psig, M Lb	.00750	381.0	2,857		

TOTAL UTILITIES			4,763	4.76	105.00
OPERATING COSTS					
Labor, 46 Men @ $ 25,500		10 M/S	1,173		
Foremen, 13 Men @ $ 29,000		2 M/S	377		
Supervision, 0 Man @ $	0	0 Man	0		
Maint., Material & Labor	6% of ISBL		4,026		

TOTAL OPERATING COST			5,576	5.58	122.93
OVERHEAD EXPENSES					
Direct Overhead	45% Lab. & Sup.		698		
Gen. Plant Overhead	65% Oper. Costs		3,624		
Insurance, Prop. Tax	1.5% Tot. Fix. Inv.		1,425		

TOTAL OVERHEAD EXPENSES			5,747	5.75	126.70
BY-PRODUCT CREDIT					

TOTAL BY-PRODUCT CREDIT			0	.00	.00
			==========	=======	==========
CASH COST OF PRODUCTION			37,351	37.35	823.44
DEPRECIATION	20% ISBL + 10% OSBL		16,210	16.21	357.37
			==========	=======	==========
NET COST OF PRODUCTION			53,561	53.56	1180.81
REQUIRED SALES PRICE AT 10% DCF				67.5	1487.2

TABLE VI-C-4

COST OF PRODUCTION ESTIMATE FOR CITRIC ACID
PROCESS- MOLASSES FERMENT.

CAPITAL SUMMARY

BASIS		CAPITAL COST	$MILLION
Location:	U.S. Gulf Coast	Battery Limits	67.1
	Mid-1982	Offsites	27.9
Capacity:	100.00 million pounds/yr		------
	45,360 metric tons/yr	Total Fixed Inv.	95.0
Str.Time:	8000 hours per year	Working Capital	8.5

PRODUCTION COST SUMMARY

RAW MATERIALS	UNITS PER LB	PRICE, c/UNIT	ANNUAL COST, $M	CENTS PER LB	DOLLARS/ MET TON
Beet Molasses, Lb	3.75000	3.8	14,250		
Lime, Lb	.57500	1.5	863		
Sulfuric Acid, Lb	1.00000	4.3	4,300		
Ammonium Nitrate, Lb	.03000	7.5	225		
KH2PO4, Lb	.01500	50.0	750		
MgSO4, Lb	.01200	11.5	138		
Catalyst & Chemicals			740		

TOTAL RAW MATERIALS			21,265	21.27	468.82
UTILITIES					
Power, kWH	.40000	4.3	1,720		
Cooling Water, M Gal	.06540	5.8	379		
Process Water, M Gal	.00410	60.0	246		
Steam,200 psig, M Lb	.00750	299.0	2,242		

TOTAL UTILITIES			4,588	4.59	101.14
OPERATING COSTS					
Labor, 46 Men @ $ 25,500		10 M/S	1,173		
Foremen, 13 Men @ $ 29,000		2 M/S	377		
Supervision, 0 Man @ $	0	0 Man	0		
Maint., Material & Labor	6% of ISBL		4,026		

TOTAL OPERATING COST			5,576	5.58	122.93
OVERHEAD EXPENSES					
Direct Overhead	45% Lab. & Sup.		698		
Gen. Plant Overhead	65% Oper. Costs		3,624		
Insurance, Prop. Tax	1.5% Tot. Fix. Inv.		1,425		

TOTAL OVERHEAD EXPENSES			5,747	5.75	126.70
BY-PRODUCT CREDIT					

TOTAL BY-PRODUCT CREDIT			0	.00	.00
			==========	=======	==========
CASH COST OF PRODUCTION			37,176	37.18	819.59
DEPRECIATION	20% ISBL + 10% OSBL		16,210	16.21	357.37
			==========	=======	==========
NET COST OF PRODUCTION			53,386	53.39	1176.95
REQUIRED SALES PRICE AT 10% DCF				67.3	1483.1

using CSI and DOE utility costs, respectively. These data are summarized in Table VI-C-5. All the facilities produce 100 million pounds per year of citric acid at a plant located on the U.S. Gulf Coast in mid-1982.

TABLE VI-C-5

SUMMARY OF CITRIC ACID PRODUCTION ECONOMICS

Basis: 100 MM lb/yr
U.S. Gulf Coast, Mid-1982

	Wood Based Fermentation		Molasses Based Fermentation	
	CSI	DOE	CSI	DOE
Investment, MM $				
Battery limits	90.2	90.2	67.1	67.1
Offsites	53.3	53.3	27.9	27.9
Total fixed investment	143.5	143.5	95.0	95.0
Cost of production, ¢/lb				
Raw materials	12.99(1)	12.99(1)	21.27(2)	21.27(2)
Utilities	9.77	8.69	4.76	4.59
Operating costs	7.42	7.42	5.58	5.58
Overhead expenses	7.88	7.88	5.75	5.75
By-product credit	(6.39)(3)	(6.39)(3)	-	-
Cash cost of production	31.67	30.59	37.36	37.19
Depreciation	23.37	23.37	16.21	16.21
Net cost of production	55.04	53.96	53.57	53.40
Selling price at 10% DCF	75.2	74.0	67.5	67.3

(1) Aspen wood at 1.0¢/lb, H_2SO_4 at 4.3¢/lb, lime at 1.5¢/lb.
(2) Molasses at 3.8¢/lb, H_2SO_4 at 4.3¢/lb, lime at 1.5¢/lb
(3) Furfural at 20¢/lb, CO_2 at 2.8¢/lb.

Table VI-C-5 indicates that the conventional molasses fermentation technology produces citric acid for 67.5 cents per pound compared to 75.2 cents per pound for the wood based fermentation. These values are the sales price at 10 percent DCF. The conventional technology offers approximately an 11 percent price advantage compared to the wood-based fermentation. This is due primarily to the large advantage in capital-related expenses of the conventional technology, inasmuch as there is no front end sugar producing facility as in the wood-based plant. The capital-related expenses advantage more than offsets the 8 cent per pound raw material advantage of the wood-based plant. Note, however, that molasses prices are very depressed at the present time. A return to historical pricing levels for molasses will increase the raw material cost advantage accruing to the wood-based route.

Economics for the wood-based case are based upon a nominal furfural by-product credit of 20 cents per pound, well below its list price of 66 cents per pound. A sensitivity analysis is provided to see at what furfural by-product value the wood technology becomes competitive with the conventional route. Figure VI-C-1 illustrates this relationship. As can be seen from the figure, the economics of the two routes to citric acid become equal at a furfural by-product credit of about 47.5 cents per pound, which is not an unreasonable number.

Figure VI-C-2 illustrates the sensitivity of citric acid's selling price to aspen wood feedstock price at 10 percent DCF and a furfural by-product value of 20 cents per pound. This figure indicates that the economics are not very sensitive to wood price, and parity is not reached even if the wood has zero cost.

FIGURE VI-C-1

RELATIONSHIP BETWEEN FURFURAL BYPRODUCT VALUE
AND
CITRIC ACID SALES PRICE

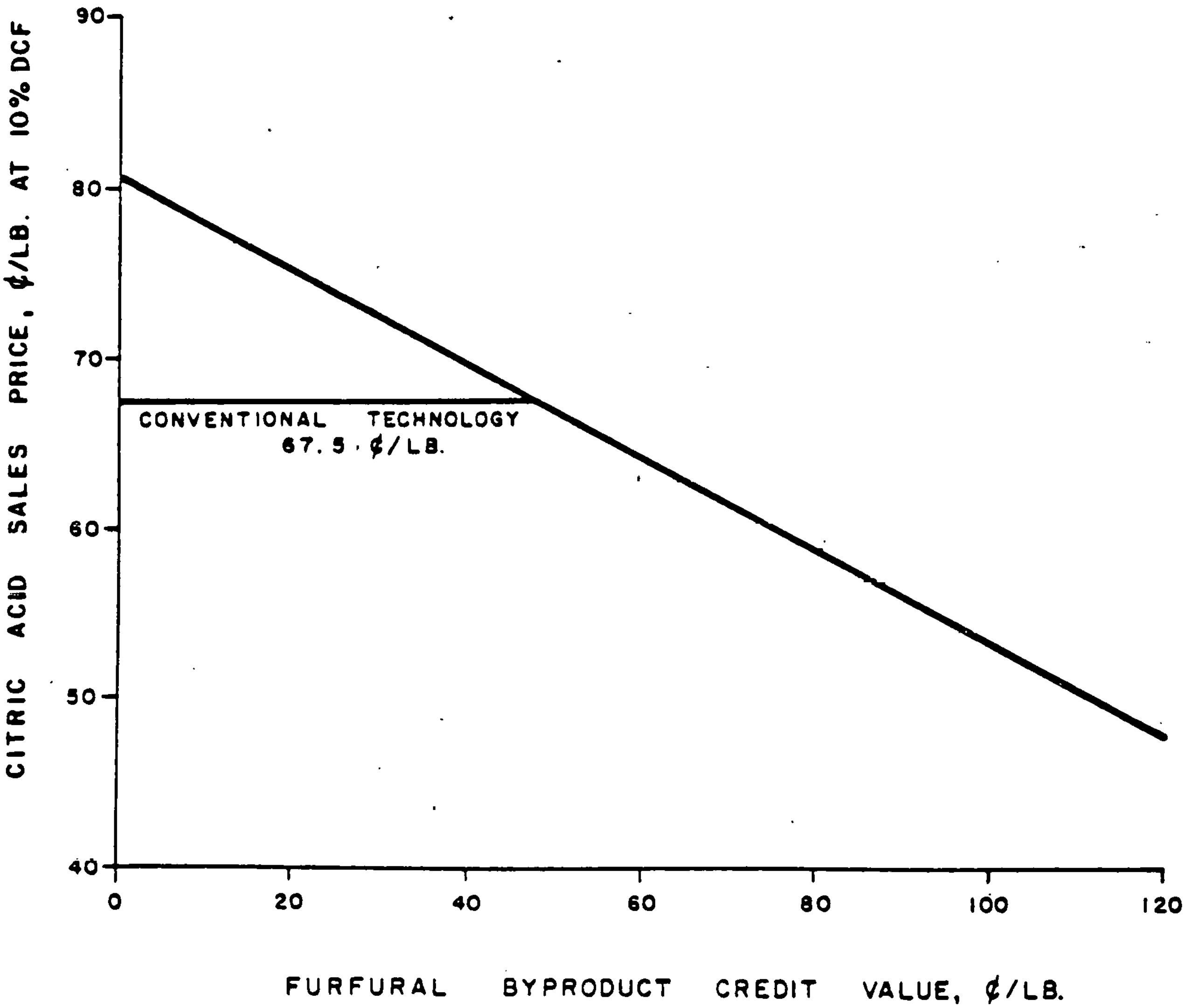


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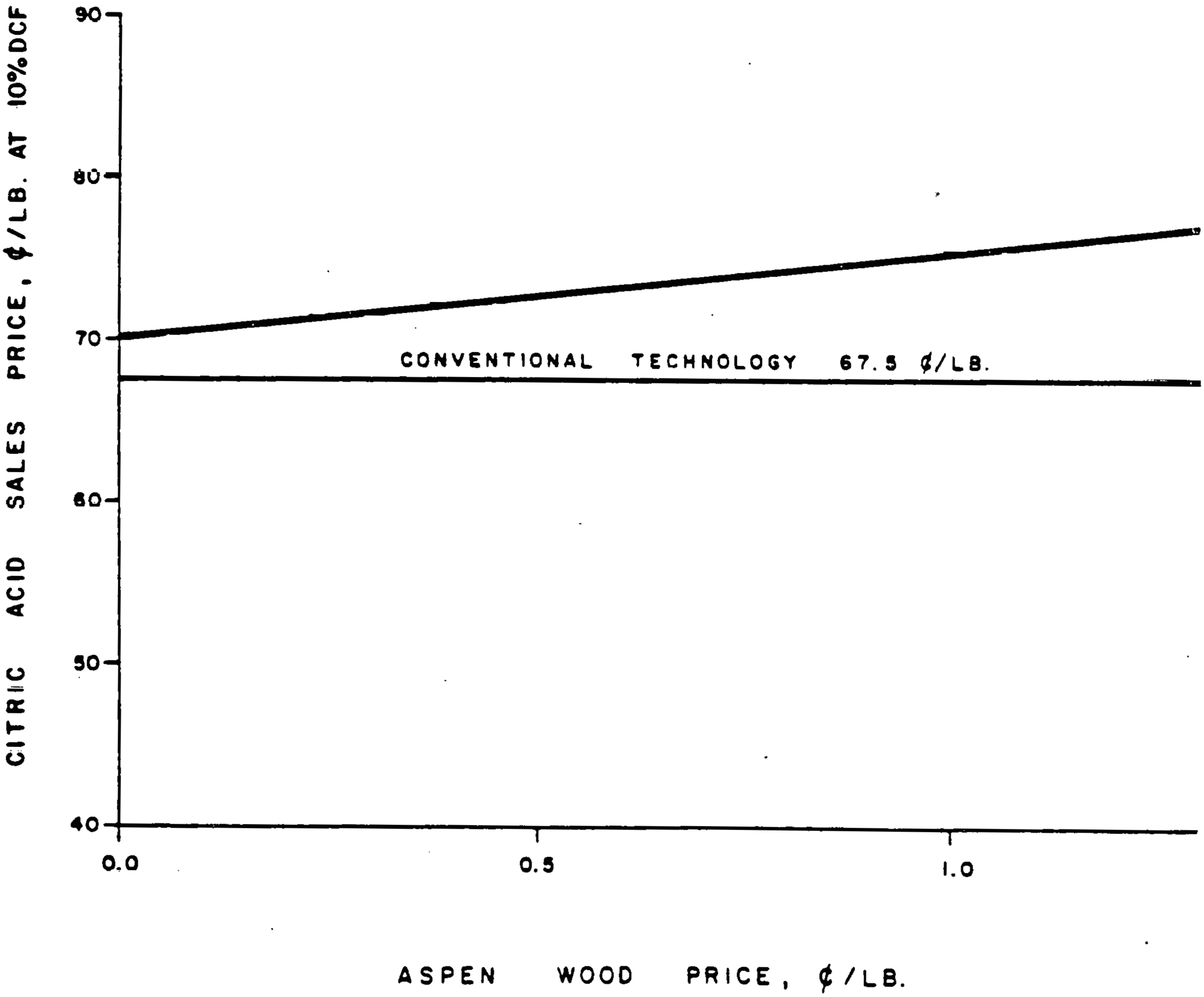
FIGURE VI-C-2

SENSITIVITY OF CITRIC ACID SALES PRICE TO ASPEN WOOD FEEDSTOCK



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