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# **Economic Feasibility Study of an Acid Hydrolysis-Based Ethanol Plant**

## **A Subcontract Report**

**Badger Engineers, Inc.**  
Cambridge, MA

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SERI Technical Monitor:  
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### EXECUTIVE SUMMARY AND CONCLUSIONS

This report presents the conceptual design and economic analysis of a process to convert lignocellulosic material, specifically mixed hardwood chips into fuel-grade ethanol. A key element of the technology is a novel plug flow reactor developed at Dartmouth College by Professors Grethlein and Converse in which dilute sulfuric acid is used in a continuous, short residence time, high temperature reactor to hydrolyse susceptible fractions in the feed. The products of the hydrolysis are fermentable sugars (hexose), pentose sugars, furfural, lignin residue, and a variety of minor organic constituents including hydroxymethyl-furfural (HMF); levulinic, acetic and formic acid; and terpenes and other extractives. The relative proportions of the products can be varied to some degree by adjusting reactor conditions. In the present study, reactor conditions have been specified which maximize the yield of fermentable sugars with furfural as the major process byproduct. The process also produces electric power in excess of its own requirements from the combustion of waste byproducts lignin and waste organics.

Two design capacities have been investigated, namely 25 million and 5 million gallons of ethanol per year. In addition, a second case representing an important design variation has been examined at each design capacity (Design Case II and Design Case IV). The resulting four cases are summarized in Table 1.1.

Table 1.1 Summary of Design Cases Studied

Design Case	Design* Capacity MM Gallon Ethanol/Yr	Number Hydrolysis Stages	Wood Feed Rate Dry Metric Tons/Hour	Byproducts Furfural MM Lb/Yr	Exported Electricity MW	Outside Utilities Required	Effluents
I Base Case	25	1	73.8	130.2	22	No	Landfill, Boiler Flue, Cooling Tower**
II Alternative Case	25	2	66	93.14	--***	No	Landfill, Boiler Flue, Cooling Tower
III Small Scale Plant	5	1	14.8	26	4.4	No	Landfill, Boiler Flue, Cooling Tower
IV Small Scale Plant	5	1	14.8	26	--	Electric Power 4.1 MW	Landfill, Boiler Flue, Cooling Tower

\*Nominal design capacity: actual design capacity for 25 million gallon is 26.2 and for 5 million gallon is 5.24.

\*\*Cooling tower refers to evaporation, drift, and blowdown.

\*\*\*Electrical needs generated onsite.

25 million gallons ethanol per year has been adopted as the base case since this is foreseen as the nominal capacity of typical large scale plants that would be built once the uncertainties due to feedstock availability, technology, and byproduct utilization have been satisfactorily answered and the advantages of economy of scale can be taken. The 5 million gallon scale was investigated because this is the probable size of the first commercial plant.

Hydrolysis reactor scale-up considerations in progressing from the existing  $\frac{1}{2}$ " diameter used at Dartmouth to the eventual 10" diameter for the large commercial scale plant have been addressed from the point of view of operating stability, heat transfer capacity, feed slurry handling, and yield of product.

Flexibility of location was an important consideration in the study since it is expected that this technology will be employed in more than one location. Thus, the base case plant has been designed to minimize environmental impact and to be essentially self-contained, generating all utilities onsite. Also, the capital cost estimate has been generated for a typical location in the Midwest.

An economic analysis has been carried out to determine, for each of the four design cases, the required ethanol selling price to yield a 15% discounted cash flow internal rate of return (IRR) after tax. Two parallel sets of calculations have been performed: one using economic assessment parameters requested by SERI and a second using parameters developed by Badger for this type of study. In addition, a calculation has been made on the SERI basis with a 25% IRR for the 5 million gallon plant (Design Case III) since a higher IRR than 15% may be required for the first commercial installation. The effect of the main economic parameters is shown in Figure 1.1.\*

The economic analysis is extended, for Design Cases I and III, to calculation of the ethanol price sensitivity to feedstock price and capital cost. The sensitivity of ethanol price to furfural selling price is also calculated for base case (Design Case I).

\*NOTE: An energy tax credit has been included in the Badger analysis of 10% of installed cost. This has the effect of reducing the ethanol selling price by 15¢ per gallon (\$1.95 to \$1.80) for the base case.

The technical risk involved in commercialization of the process based on the present level of development is analyzed and quantified, and possible process development strategies are assessed by a methodology developed by the Rand Corporation. Recommendations for future development are based on the results of the Rand Method. (Reference 30)

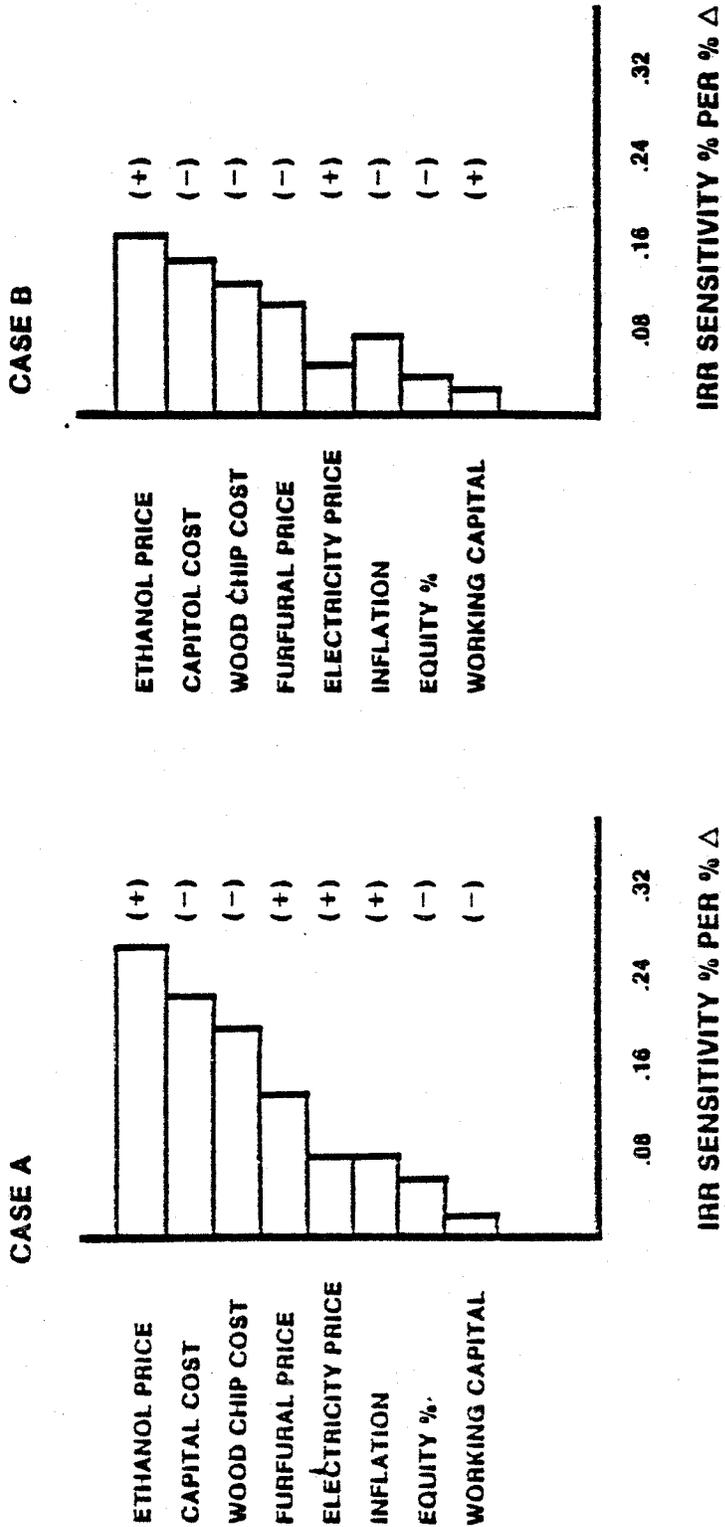
### Conclusions

The major conclusions reached in the study are listed below:

- (1) The process design presented is operationally feasible from an engineering viewpoint. However, a high level of risk of capital cost growth and of performance shortfall would be expected without a process demonstration phase and a site-specific cost analysis. These two exercises should result in a sufficient level of confidence to justify commitment of funds by an industrial producer.
- (2) Recommendations for design and operation of the commercial-scale reactor have been made. These include measures to reduce the effects of tar pluggage, to ensure adequate heat transfer, and to facilitate pumping the high consistency wood slurry feed. These recommendations have been supported, where possible, by experimental work at Dartmouth College or Badger's Weymouth Laboratory.

Figure 1.1 Effect of Economic Parameters on IRR

FIG. 1.1  
 SENSITIVITY OF DCFIROR TO ECONOMIC PARAMETERS



- (3) Feedstock availability investigation reveals that sufficient quantities of waste wood from logging, lumber mill, and forest thinning operations would be available in most forested areas to supply a 25 million gallon plant. The scale of collection required is the same as that now being carried out by Burlington Electric Company for their McNeil wood-fired power plant. A conservative figure of \$42 per dry metric ton has been used in the economic assessment based on information from wood chip brokers, large scale chip users, forestry consultants, and published literature.
- (4) The market for fuel-grade ethanol is expected to grow to well over 1000 million gallons per year by 1990 primarily as a result of the increasing demand for gasoline blending components. Consequently, the market would experience little downward pressure on price as a result of the additional output from several 25 million gallon plants.

The question of the furfural market is not clear cut. Although furfural is a versatile chemical intermediate, its demand in the U.S. has been dropping in recent years as lower cost alternatives have eroded its traditional outlets. The introduction of the quantities of furfural produced by a 25 million gallon, acid-based, ethanol from wood plant would be expected to depress the price severely. As the price of furfural drops, however, its attractiveness as a chemical intermediate in the production of such commodity chemicals as adipic acid, maleic anhydride, ethylbenzene, and butanol increases.

Analysis of the potential market for furfural from published data (reference 10) shows that at 15¢ per pound a potential market of over 2 billion pounds per year and at 10¢ per pound a potential market of over 3 billion pounds per year exists. The above figures suggest that the combined ethanol and furfural markets have the potential to absorb the output of 15 to 20 large commercial scale plants.

The establishment of such major furfural markets will not occur overnight and could reasonably be expected to take 10 to 15 years to develop. However, future disruption of oil supplies with resultant increases in the prices of petrochemical feedstocks could accelerate this scenario.

- (5) A market for electric power is assumed to exist under the Public Utilities Regulatory Policies Act (PURPA). The 25 million gallon plant would produce a total of 44.5 megawatts which is below the present PURPA maximum limit of 80 megawatts. The price assumed in the study for this byproduct is 5¢ per kilowatt hour.
- (6) The commercial potential of the other organics produced has not been investigated in this study nor has credit been taken for them as products. However, these organics are concentrated in the process which facilitates recovery if warranted and technically feasible.

- (7) The selling price of fuel grade ethanol calculated by SERI and Badger economic assessment parameters is given in Table 1.2.

The figures suggest that the technology looks attractive in the long term. Economic assistance may be necessary to introduce the technology in the short term at small commercial scale. However, economic analysis of the small commercial plant suggests that this might be an attractive venture should a cheap local source of feedstock be available.

Table 1.2  
Selling Price of Denatured Ethanol for the Four Design Cases

Design Case	Ethanol Selling Price \$ Per Gallon		
	SERI Parameters		Badger Parameters
	15% IRR	25% IRR*	15% IRR
I 25 Million Gallon (Base)**	1.23		1.80
II 25 Million Gallon (Alternative)	1.65		2.23
III 5 Million Gallon***	1.42	2.15	2.83
IV 5 Million Gallon (Purchased Power)	1.63		2.85

\*For rationale behind using 25% DCFIROR for 5 million gallon per year cases see Page 9, Section 5.

\*\*Furfural Priced at ¢ 15/lb.

\*\*\*Furfural Priced at ¢ 55/lb.

### TECHNICAL APPROACH AND DESIGN BASIS

This report presents the conceptual design and economic analysis of a process to produce fuel-grade ethanol from mixed hardwood chips. The process technology is based on a novel plug flow acid hydrolysis reactor developed at Dartmouth College. This system uses a short residence time, high temperature, dilute acid reaction to hydrolyse the susceptible fractions in lignocellulose to hexose sugars, pentose sugars, furfural, and lignin residue. The proportion of each product is dependent on the reactor conditions used.

Four design cases are dealt with, covering two production capacities of 25 and 5 million gallons ethanol per year (Design Cases I and III); the 25 million gallon plant with a modified, two-stage wood hydrolysis reactor (Design Case II); and the 5 million gallon plant with purchased electric power (Design Case IV). The 25 million gallon scale has been used as the base case since this is foreseen as the typical large scale plant which will be built once uncertainties due to process technology, feedstock availability, and byproduct markets have been satisfactorily answered, allowing the advantages of economy of scale to be taken. The 5 million gallon plant has been designed and analyzed since this will be the most probable size of the first commercial plant.

A two stage hydrolysis reactor scheme was investigated since research and development work shows that higher yields of fermentable sugars can be obtained from such a scheme.

Finally, the cost-effectiveness of onsite electrical generation is determined for the 5 million gallon scale by investigating the economics of the plant using purchased rather than generated power.

The basic process has been split into the following units:

Unit 100	Feedstock Handling
Unit 200	Hydrolysis
Unit 300	Fermentation
Unit 400	Ethanol Purification
Unit 500	Furfural Recovery
Unit 600	Offsite Tankage
Unit 700	Waste Treatment
Unit 800	Utilities

A summary of the four design cases investigated is given in Table 2-1 on the following page.

Table 2.1 Summary of Design Cases

Design Case	Design* Capacity MM Gallon Ethanol/Yr	Number Hydrolysis Stages	Wood Feed Rate Dry Metric Tons/Hour	Byproducts Furfural MM Lb/Yr	Export Electricity MW	Outside Utilities Required	Effluents
I Base Case	25	1	73.8	130.2	22	No	Landfill, Boiler Flue, Cooling Tower**
II Alternative Case	25	2	66	93.14	--	No	Landfill, Boiler Flue, Cooling Tower
III Small Scale Plant	5	1	14.8	26	4.4	No	Landfill, Boiler Flue, Cooling Tower
IV Small Scale Plant	5	1	14.8	26	--	Electric Power 4.1 MM	Landfill, Boiler Flue, Cooling Tower

\*Nominal design capacity, actual design capacity for 25 million gallon is 26.2 and for 5 million gallon is 5.24.

\*\*Cooling tower refers to evaporation, drift and blowdown.

Heat and material balances for the base case process design as well as energy consumption by plant unit are provided at the end of this section for the base case plant.

In adopting 25 million gallons per year as the design basis for the large scale plant, a number of factors were considered. Most importantly, it was felt that the establishment of a wood supply infrastructure to support such an output is feasible in most of the forested areas of the country.

In point of fact, the wood supply requirement for a 25 million gallon per year plant (1772 dry metric tons per day) is similar in size and scope to existing commercial wood-gathering efforts employed at the wood-fired power plant operated by Burlington Electric Company of Burlington, Vermont.

Most commercial fermentation ethanol plants built today have capacities in the 50-60 mm gallon per year range. It was our first thought to select a capacity in that range for this study. However, the logistics of wood gathering for such a large plant limits the number of siting options available. Rather than let site-specific issues outweigh the general technical and economic conclusions of the study, it was decided to design a capacity where one could point to an existing wood-gathering system as a basis for costs. This maintains the intended focus of the study on the economic feasibility of the technology. It is intended in this way to make the study as generally applicable as possible since it is anticipated that plants will be operated in more than one location.

The process has been designed in this study to use mixed hardwood as a feed for two reasons. Firstly, this is the predominant waste wood in many areas and, secondly, much of the development at Dartmouth College has been carried out using mixed hardwood flour.

Softwood has been shown to be a suitable feedstock without process modification, however, the relative yields of sugars and furfural would be expected to change, resulting in more fermentable sugar and less furfural than with hardwoods.

Minor modifications to the process would allow use of lignocellulosics such as sugar cane bagasse, corn stover, etc., as well as cellulosic domestic garbage. (No sufficient furfural production would be expected with domestic garbage.)

The same reasoning that lead us to a conservative basis for plant capacity also lead us to the following additional general criteria of approach:

- (1) To minimize environmental impact such as water requirements, land area, access requirements and solid, liquid, and gaseous effluents discharge. The large plant (25 million gallon) has a water requirement of 1700 gallons per minute due to the fact that waste products are limited to landfill (gypsum, ash and tar solids), boiler flue gas, and cooling tower vapor and blowdown.
- (2) To maximize the flexibility of plant location. The plant has been designed with enough flexibility to allow it to operate in a number of likely locations. The capital and operating costs have been developed for a typical Midwest location. It is intended in this way to make the study as general and widely applicable as possible.
- (3) To specify equipment with the exception of the hydrolysis reactor itself, which is commercially available. Where there has been uncertainty as to the equipment application, representative samples of process material have been obtained, vendor discussions held, and equipment trials conducted where possible. In these cases, the equipment designs have been based on vendors' recommendations and specifications.

- (4) To use a scale of reactor not too different from that which will have been used in preceeding development. (Dartmouth experiments have been carried out with a  $\frac{1}{2}$ " diameter reactor. The planned scaled-up reactor at SERI will be around 1-2" in diameter and a small demonstration plant would have a reactor 3-4" 1.1 diameter. The diameters chosen for the commercial scales, i.e. 5 and 25 million gallon, have been tentatively set at 6" and 10" respectively.
- (5) To use a level of process integration which allows major energy savings without significantly affecting process flexibility.
- (6) "To control the water balance," i.e., to limit the quantity of water or steam added to the process for reasons such as dilution, washing, and heating. The energy required to remove and recycle water represents a major loss of process energy efficiency and, thus, lost revenues.
- (7) To define as closely as possible all extraneous organics produced. This enables determination of their effect on physical properties of mixtures, energy and material balances, and separation equipment necessary. The data and calculations may also serve as a guide for future studies into the economics of their separation and purification. Hydrolysate fermentability considerations also dictate that potentially toxic components be identified and their level minimized or the yeast culture adapted.

- (8) To attempt to design the plant to be essentially self-sufficient in its energy requirements by using wood hydrolysis residues to produce the required steam for the process. To maximize the energy cycle efficiency, consideration was given to the co-generation of electric power. Electricity is co-generated at both scales, and excess electricity is exported to the local utility under the provisions of the PURPA Act. The fourth design case of a 5 million gallon plant with purchased electricity is presented for comparison.
- (9) To design the plant so that a variety of lignocellulosic feeds, such as bagasse and agricultural residues, can be used with only minor variations to the equipment. The design of the basic process is based on mixed hardwood chips as a feed for two reasons: firstly, this is the predominant waste wood in many areas and, secondly, much of the development at Dartmouth College has been carried out using mixed hardwood. Softwood should be just as suitable a feed for the process; however, the relative yields of fermentable sugar and furfural would be expected to change, giving increased fermentable sugars and decreased furfural.

Basis for Design Cases

Design Case I (Base Case)

25 Million Gallons Ethanol per Year

The base case plant is designed to produce fuel-grade ethanol from hardwood chips using a single stage hydrolysis reaction. The nominal capacity of the plant was set at 25 million gallons fuel grade ethanol per year. However, the actual design capacity is, in fact, 26.2 million gallons per year. (The increased production rate is due to slightly better fermentation and recovery yields than anticipated in preliminary calculations.)

The ethanol is sold denatured by the addition of 5 volume percent gasoline. The denatured ethanol production rate is 27.5 million gallons per year.

The wood chip feedstock will be used at an average rate of 73.8 metric tons (dry matter) per hour (590400 metric tons per year). The average bone-dry composition of the feedstock for the purposes of the study was taken as in Table 2.2 (Reference 1).

Table 2.2 Composition of Wood Feedstock

<u>Component</u>	<u>% By Weight</u>	
Glucan	42.0	-
Mannan (Hexan Polymer)	3.22	
Galactan	0.98	
Xylan (Pentosan Polymer)	21.14	
Arabian (Pentosan Polymer)	2.85	
Lignin + Uronic Acid + Acetyl	25.34	0.6 Acetic 24.74 lignin
Extractives*	4.12	{ 2.06 α Pinene 2.06 Oleic acid
Ash	<u>0.35</u>	
	100%	(Dry Basis)

\*Extractives are characterized as alpha pinene and oleic acid.

Overall Process Yields

Ethanol Yield

The yield of fermentable sugars in the hydrolysis reaction assumed as the basis for design was 55% fermentable sugars on "potential hexose" in the hydrolysis reactor feed. Potential hexose in the feed is calculated as

$$\text{hexan polymer} \times 1.111$$

Potential hexose represents the maximum theoretical yield of fermentable sugars.

The overall recovery of sugars from the hydrolysis unit was calculated as 98.56%.

The fermentation yield of ethanol from sugars was 95% of theoretical. (Theoretical fermentation yield is 51.1%.) The ethanol recovery yield was calculated as 99.5%.

The above yield assumption and feedstock composition corresponds to an overall process yield of 44.3 gallons of undenatured fuel-grade ethanol per dry metric ton of wood.

This is calculated thus:

$$\frac{\text{pounds}}{\text{metric ton}} \times \text{fraction hexose} \times \text{hydrolysis yield} \times \frac{\text{sugar}}{\text{recovery yield}} \times \text{fermentation yield} \times \frac{\text{ethanol}}{\text{recovery yield}} \div \frac{\text{pounds ethanol}}{\text{gallon}}$$

or:

$$(2205) \times 0.462 \times 1.11 \times 0.55 \times 0.9856 \times 0.95 \times 0.511 \times 0.995 \div 6.61 = 44.32$$

Since gasoline is added at 5% by volume as a denatureant the yield of denatured ethanol is:

$$44.32 \times 1.05 = 46.5 \text{ Gallons/Metric Ton}$$

The calculated product fuel grade ethanol composition is:

<u>Component</u>	<u>% By Weight</u>
Ethanol	99.5
Acetaldehyde	0.412
Acetone	0.028
MEK	0.021
Butyraldehyde	0.034
n Propanol	<u>0.005</u>
	100%

Furfural Yield

A major byproduct of the process is chemical grade furfural which is produced at the rate of 130.2 million pounds per year as commercial grade furfural. The basis for this furfural production rate was an assumed furfural yield in hydrolysis of 60% of theoretical. Hydrolysis furfural yield is therefore calculated for Design Case I as:

$$\frac{\text{pounds}}{\text{metric ton}} \times \frac{\text{fraction pentosan}}{\text{polymer}} \times \frac{\text{molec weight pentose}}{\text{molec weight polymer}} \times \frac{\text{molec weight furfural}}{\text{molec weight pentose}} \times \text{fraction yield}$$

or:

$$2205 \times 0.2399 \times 150/132 \times 96/150 \times 0.6 = 231 \text{ pounds/metric ton}$$

At the reactor conditions chosen in Design Case I, complete conversion of the pentose sugars to furfural and degradation products takes place.

Recovery of furfural from the hydrolysis unit is 96%.  
Recovery of furfural from the furfural recovery unit is 99%.  
Furfural production is therefore:

$$231 \times 0.96 \times 0.99 = 219.5 \text{ pounds/metric ton of wood}$$

Furfural is sold as commercial grade containing 0.5% water.

#### Electricity

A second major plant byproduct is excess electrical power which is exported at a rate of 22 MW.

The plant is designed to generate all of its own utilities.

The plant stream factor is 8,000 hours per year.

The plant is designed to produce no process liquid effluents under normal operation. Environmental emissions are limited to gypsum, boiler ash, and waste treatment sludges for landfill, boiler combustion gases, and cooling tower water vapor and blowdown.

Design Case II (Alternative Case)

25 Million Gallons per Year - Two Stage Hydrolysis Reaction

The alternative case plant is designed to produce fuel grade ethanol from hardwood chips using a two-stage hydrolysis reaction. The ethanol capacity and product analysis is the same as for the base case (27.5 MM Gal Denatured Ethanol). The wood chip feed requirement is 66 metric tons (dry matter) per hour with the same average chemical composition as for the base case. The alternative case design produces byproduct furfural at the rate of 93.14 million pounds per year as commercial grade furfural. The alternative case plant does not export electricity although, like the base case, it generates all of its own utilities. As with the base case, the only effluents are ash, landfill sludges, boiler stack gases, and cooling tower water vapor and blowdown.

The design of the alternative case process is based on a yield of fermentable sugars of 50% of potential hexose converted in the first stage hydrolysis and 55% of the remaining potential hexose in the second stage. This gives a hydrolysis yield of 66.5% fermentable sugars and an overall plant yield of 49.6 U.S. gallons undenatured ethanol per dry metric ton.

At the conditions chosen for Design Case II, only 43% of the potential pentose sugars are converted to furfural.

Design Case III (Reduced Capacity Plant)

5 Million Gallons Ethanol per Year - Onsite Power Generation

A reduced capacity plant has also been designed to produce 5 million gallons, nominal, (5.24 million gallons, design) per year. This is a scaled-down version of the large base case plant except that the reduced scale does not require parallel process units in the hydrolysis section.

The ethanol is sold denatured by addition of 5 volume percent gasoline. The denatured ethanol production rate is 5.5 MM gal/yr.

This plant uses 14.76 metric tons of wood chips per hour (118100 metric tons per year).

The production rate of byproducts is: Furfural 26 million pounds per year and electricity 4.4 MW.

As with Design Case I, at the reactor condition chosen, complete conversion of the pentose sugars to furfural and other degradation products takes place.

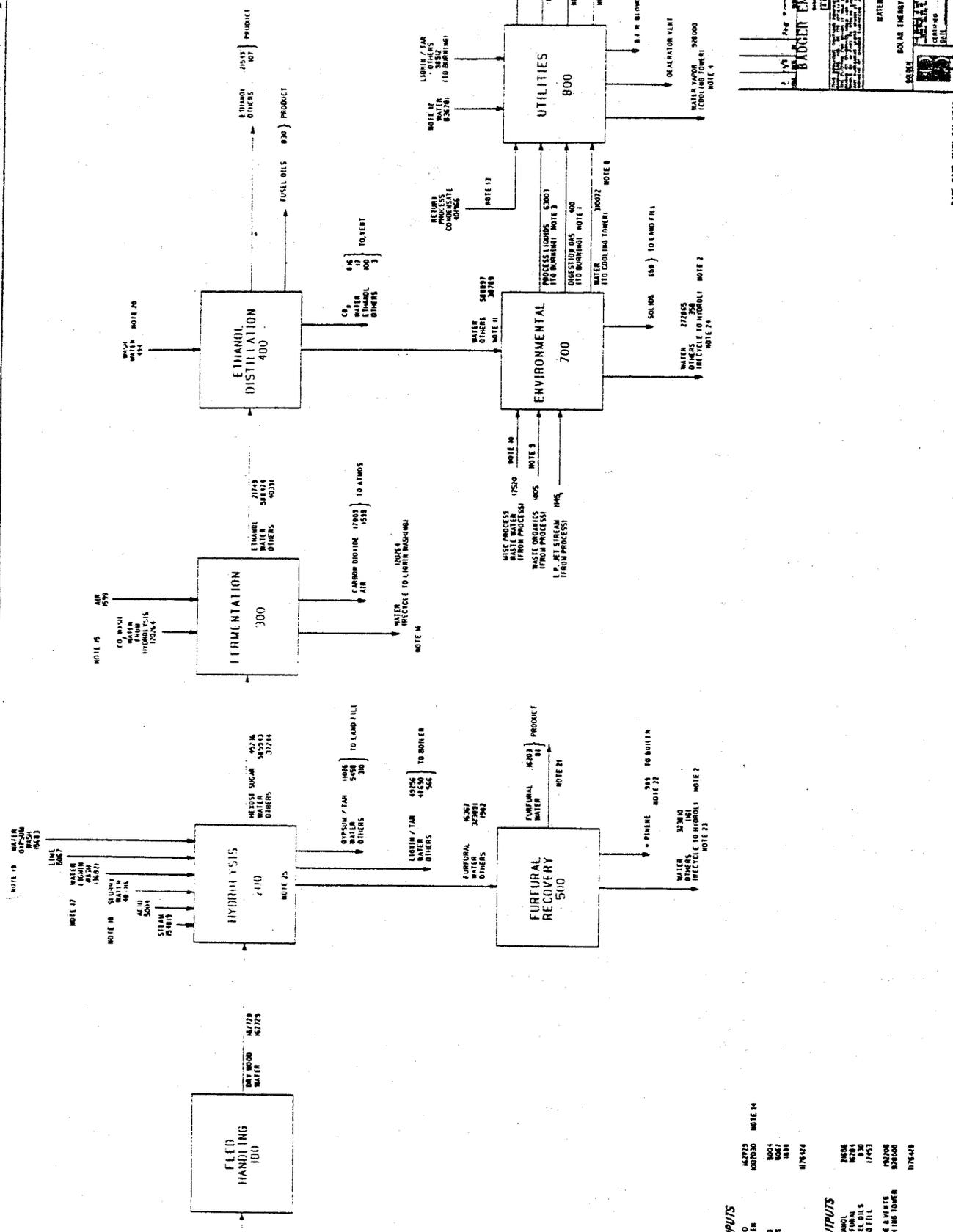
Design Case IV (Reduced Capacity Plant)

5 Million Gallons per Year with Purchased Electric Power

A reduced capacity plant has also been designed to produce 5 million gallons ethanol per year, nominal, (5.24 million gallons, actual). This is the same plant as Design Case III except that facilities for power generation have been omitted and electric power is purchased. This allows investigation of the cost-effectiveness of onsite power generation. The plant uses 14.76 metric tons of wood chips per hour (118100 metric tons per year). Demand rate of electricity is 4.1 MW.

Completing this section are the material and energy balances for the base case 25 million gallon per year plant (Design Case I). On the following page is the overall material balance. This is followed by a drawing presenting the material balances for each unit. The overall energy balance is then presented followed, in turn, by a unit by unit summary of energy consumptions.

Heat and material balances have been presented for Design Case I only. The differences between that case and the others have been described elsewhere in this section.



PROCESS INPUTS

WOOD	46213	NOTE 14
WATER	400230	
ACID	5067	
ETHANOL	5067	
FUEL OILS	1834	
LAND FILL	17652	
STEAM	21819	

PROCESS OUTPUTS

ETHANOL	2149
FUEL OILS	1834
LAND FILL	17652
STEAM	21819
COOLING WATER	37800
OTHERS	10358

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DATE CASE DRAWN 04/76

PROJECT NO. E0461-106-24

NOTES FOR MATERIAL BALANCE: (Flow Sheet 106-24)

1. Boiler flue accounts only for fuels mass and does not include combustion air.
2. "Others" in recycle to MT-223 (Section 200) is counted as water.
3. Process liquids from stillage evaporation.
4. Cooling tower includes evaporation drift and blowdown.
5. Steam to process is sum of streams 801, 802, and 803 on flowsheet 106-18.
6. Fresh process water to process from ring main (37880 lb/hr) comprises gypsum washwater (15683 lb/hr) lignin wash (16557 lb/h), water make-up to MT-223, (5186 lb/h), and fusel oil wash (494 lb/h).
7. Boiler feedwater blowdown (15000 lb/h) goes to unit 700.
8. Water to cooling tower (310072 lb/hr) is from anaerobic/aerobic treatment. This is composed of steam 702 (290598 lb/h) plus BFW blowdown (15000 lb/h) plus molecular sieve runoff (1954 lb/hr) plus flume washings (2520 lb/h). This is approximately 33% of total water make-up to cooling tower.
9. Waste organics include alpha pinene (985 lb/h) and organics in flume washings (20 lb/h).

10. Miscellaneous process water is: Boiler feed water blowdown (15000 lb/h) plus flume washings (2520 lb/h).
11. Water from unit 400 is beer column bottoms (586943 lb/h) plus molec sieve runoff (1954 lb/h).
12. Total water make-up to process (836781 lb/h) is boiler make-up (180964 lb/hr) plus cooling tower make-up (617928 lb/h) plus water to Lignin washing (16557 lb/h) plus water to gypsum washing (15683 lb/h) plus water to fused oil washing (454 lb/h) plus MT-223 make-up water (5186 lb/h).
13. Return process condensate is streams 806 + 802 + 805 on flowsheet 106 - 18.
14. Total water to process (1002030 lb/h) is: Total make-up water (836781 lb/h) plus wood moisture (162729 lb/h) plus flume washings (2520 lb/h).
15. Wash water for CO<sub>2</sub> is from MT 223 (120264 lb/h).
16. Washings from CO<sub>2</sub> scrubber (120264 lb/h) are sent to lignin washing.
17. Lignin wash water (136821 lb/h) is CO<sub>2</sub> washings from unit 300 (120264 lb/h) plus fresh process water from ring main (16557 lb/h).
18. Slurry water to hydrolysis (483116 lb/h) is from MT 223.
19. Gypsum wash water (15683 lb/h) is fresh process water from ring main.

20. Fusel oil wash water.
21. Stream 506 from furfural recovery to product storage.
22. Stream 507 from furfural recovery to MT-702 (Drawing 106-14).
23. Furfural azeotrope column bottoms (324971 lb/h) to MT-223.
24. Condensate from evaporator (273223 lb/h) recycled to MT-223.
25. MT-223 (Hydrolysis process water tank) mass balance is:  
  
Inputs: 273223 lb/hr from unit 700 + 324971 lb/h from unit 500 + 5186 lb/hr fresh make-up from ring main.  
  
Outputs: 120264 lb/h to CO<sub>2</sub> scrubber + 483116 lb/hr hydrolysis slurry water.

Overall Material Balance  
(Design Case I)

<u>Inputs</u>			<u>Outputs</u>	
<u>Item</u>	<u>Flow</u> <u>Lb/Hr</u>		<u>Item</u>	<u>Flow</u> <u>Lb/Hr</u>
Wood	162,729	P	Ethanol	21,656
Water	1,002,030	R	Furfural	16,284
Acid	5,004	O	Fusel Oils	830
Lime	5,067	C	Landfill*	17,453
Air	<u>1,599</u>	E	Flue/Vents**	192,206
Total	1,176,429	S	Cooling Tower***	<u>928,000</u>
		S	Total	1,176,429

\*This item includes gypsum, boiler ash, waste treatment sludges, and tars.

\*\*This item does not include combustion air.

\*\*\*This item includes entrainment allowance and blowdown.

Overall Energy Balance

(Design Case I)

<u>Inputs</u>			<u>Outputs</u>	
<u>Item</u>	<u>Heat Value</u> MM Btu/Hr		<u>Item</u>	<u>Heat Value</u> MM Btu/Hr
Wood	1,425	P	Ethanol	252
Process Water	48	R	Furfural	165
Misc. Chemicals	<u>55</u>	O	Fusel Oils	8
Total	1,528	C	Electricity	75
		E	Flue and Vents	179
		S	Cooling Tower	834
		S	Landfill	<u>15</u>
			Total	1,528

\*Heat value is given as lower heating value

$$\text{Thermal Efficiency} = \frac{\text{Heating Value of Useful Products}}{\text{Total Heat Inputs}} = \frac{500}{1,528} = 32.7\%$$

Energy Consumption by Unit

(Design Case I)

<u>Unit</u>	<u>Electrical Energy</u> <u>MM Btu per Hour</u>	<u>Steam Energy</u> <u>MM Btu per Hour</u>
100 Feedstock Handling	0.5	2.0
200 Hydrolysis*	34.0	257.0
300 Fermentation	2.0	5.0
400 Ethanol Purification	9.5	42.0
500 Furfural Recovery	0.3	75.0
600 Offsite Tankage	0.8	-
700 Waste Treatment	0.6	182.0
800 Utilities	17.0	11.0
Miscellaneous (Lighting, Winterization, etc.)	4.0	11.0
		<u>585.0</u>

\*Includes size reduction of wood particles.

BOILER - TURBOGENERATOR HEAT BALANCE

Boiler Fuels

Item	Mat. Bal. Stream Number	Feedrate lb/Hr	Heat of Combustion Btu/lb	Heat Content MM Btu/Hr
Dry Lignin/Tar	222	49800	12000	598
Evaporator Liquids	704	63003	6600*	416
Fusel Oils	406	830	11000	9
Digester Gas	-	-	-	<u>4</u>
			Total	1027

\*Contains 39% Water

Steam Raised

Boiler Efficiency 85%

Feedwater Temperature 240°F

Steam Produced: 715000 lb/Hr at 1100 PSIA, 874°F

Process Steam Required: 192890 lb/Hr at 1100 PSIA  
20500 lb/Hr at 150 PSIG  
344600 lb/Hr at 50 PSIG

Electricity Generated at Turbine Efficiency of 78%

	Actual		Actual
	Water	Steam	Power
	Rate	Rate	Generated
Exit Steam Level	lb/MW Hr	lb/Hr	MW
Condensing Steam	8500	157010	18.5
150 PSIG Steam	20792	20500	0.9
50 PSIG Steam	15250	344600	<u>22.6</u>
<u>Total Electricity Generated</u>			42.0
Process Electricity Required (MW)			20
Export Electricity (MW)			22

## PROCESS CONSIDERATIONS

### Introduction

This section details the rationale used in establishing the sequence of unit operations for each process unit. In each instance the base case is discussed and followed, where applicable, by a discussion of the differences between the base case and the alternative cases. Scale-up considerations for the hydrolysis reactor itself are also discussed.

The source materials upon which the design is based are varied and include the following:

- (1) Experimental work carried out at Dartmouth College, Badger's Weymouth Laboratory, JGC (Japan), and equipment vendors' laboratories.
- (2) Consultation with Professors Grethlein and Converse, developers of the hydrolysis reactor at Dartmouth College.
- (3) Consultation with Badger engineering design specialists.
- (4) Literature survey.
- (5) Discussions with engineers at SERI.
- (6) Documents obtained from SERI engineers.
- (7) Badger in-house data.
- (8) Consultation with forestry experts of the Burlington Electric Company, Burlington, Vermont.

Unit 100 - Feedstock Handling and Storage

The feedstock handling and storage area consists of rail and truck offloading facilities, a chip flume, chip storage, and front loaders and conveyors for feeding the wood chips to the process. It is anticipated that most of the feed-stock will be delivered by railcar in order to minimize local trucking nuisance. It is, planned, however, to receive 5 to 10% of the feedstock directly by truck.

A flume pond has been included in the design both to wash the chips and to allow separation of large, heavy, tramp material such as soil, rocks, and metal items.

The chip pile has been sized to allow two weeks chip storage. The use of rail delivery also permits extra storage at remote rail loading heads. Delivery to these intermediate collection points will be by truck.

The chip pile will be managed by two front-end loaders, one for pile management and one for feeding the process. The loader used for pile management acts as a spare for the process feed loader. This has worked very well at both the Moran and McNeil wood burning power plants of the Burlington Electric Company of Burlington, Vermont. Management of the chip pile will be on a first in, first fed handling policy. This plus a twice a year turnover of the chip pile has proven effective in reducing microbial degradation and minimizing the risk of spontaneous combustion at both Burlington Electric plants.

Why not put flume after chip pile . . .

## Unit 200 - Hydrolysis

### Hydrolysis Reaction

The hydrolysis reaction is the key step in the process and represents the major deviation from existing process technology; it thus represents the major uncertainty in the process. (For further discussion and quantification of the technical risks see Section 6.0.) It has therefore been discussed below in considerable detail.

Other unit operations comprising the hydrolysis unit are also described and discussed in the following pages.

The hydrolysis reactor is a short residence time, high temperature plug flow reactor based on that developed at Dartmouth College (Reference 1). The reactor is designed to hydrolyse lignocellulosic materials fed as aqueous acidic slurries. The slurry feed is heated by high pressure steam injection as it enters the reactor. The major hydrolysis products are normally hexose sugars, furfural, hydroxymethyl furfural (HMF), tars, and lignin residue with a variety of minor organic constituents. Yield figures used in the study have been based directly on earlier figures obtained during operation of the Dartmouth reactor (Reference 1 and 2). Yield figures used are 55% hexose sugars (on potential hexose) and 60% of theoretical for furfural. Results obtained from later experiments performed under subcontract during the present study suggest that the above yield assumptions are conservative.

Other extraneous organics produced by wood hydrolysis are extractives and tall oils characterized by terpenes (alpha-pinene) and oleic acid, respectively; acetic acid from acetyl; and formic acid and levulinic acid from partial HMF breakdown.

A program to test a scaled-up version of the Dartmouth reactor is underway at the SERI research facility in Golden, Colorado. Data from this is not expected however until sometime in mid-1985.

### Reactor Scale-up Considerations

#### (1) Reactor Dimensions

The commercial-scale reactor diameter has been tentatively set at 10" for the 25 million gallon plant and 6" for the 5 million gallon plant. This is based on the premise that preceding development stages will be carried out on reactors of increasing diameter. The Dartmouth work has been conducted on a reactor of 4" diameter. The interim stages of development are anticipated to be, firstly, the planned reactor at the SERI laboratories and, secondly, an integrated demonstration process. The reactor sizes in these two intermediate developments should be 1" to 2" for the SERI work and 2" to 3" for the demonstration plant.

A tentative reactor specification is given in figure 3.1. This shows the approximate dimensions required for the reactor at various stages of development.

Figure 3.1 Hydrolysis Reactor Specifications

FIGURE 3.1

HYDROLYSIS REACTOR SPECIFICATION

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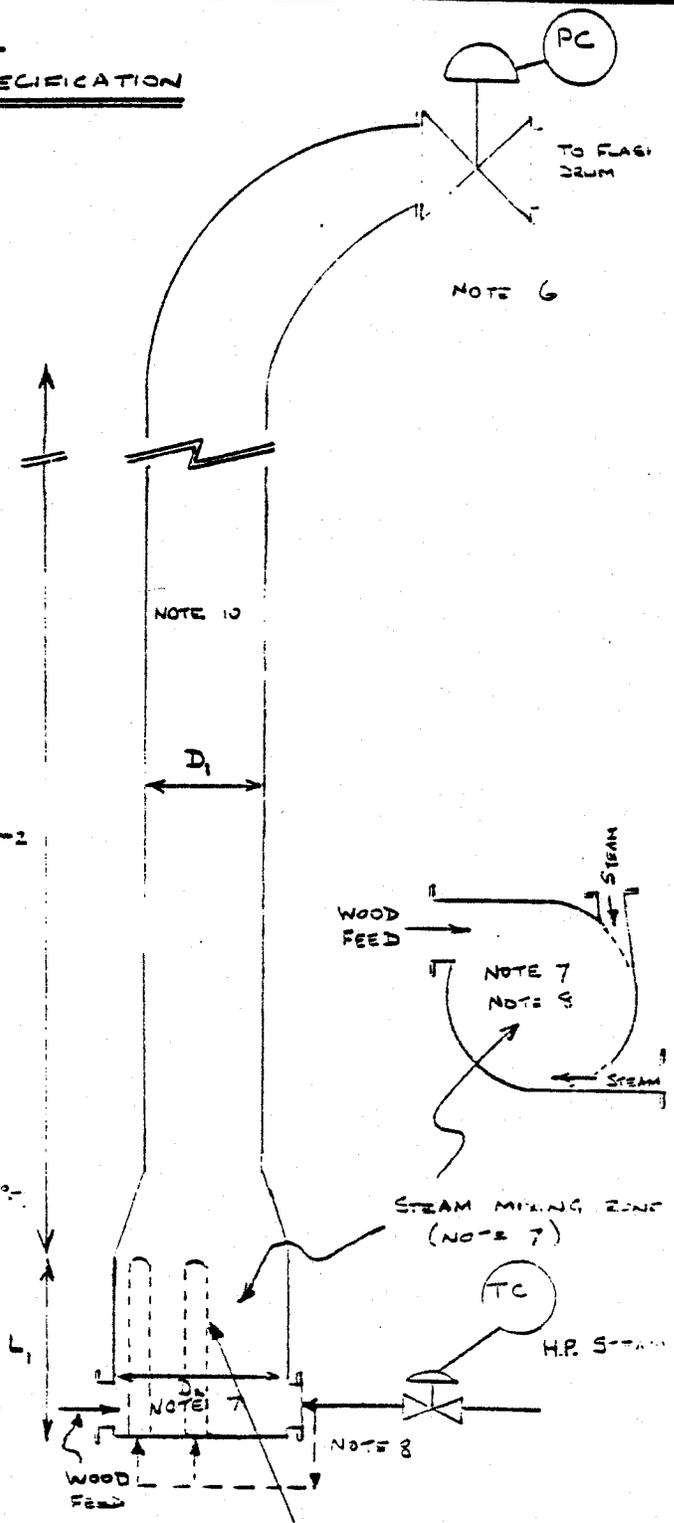
DEVELOPMENT STAGE

REACTOR DIMENSIONS (INCHES)

NOTE	D <sub>1</sub>	D <sub>2</sub>	L <sub>1</sub>	L <sub>2</sub>
DARTMOUTH	0.5	0.5	7.5	SEE NOTE 5
S.E.R.I.	1	2	10	NOTE 5
1 MM GAL	2-3	7	30	NOTE 5
5 MM GAL	6	12	43	NOTE 5
25 MM GAL	10	15	48	NOTE 5

NOTES:

- DARTMOUTH BENCH-SCALE REACTOR IN OPERATION
- S.E.R.I. REACTOR UNDER CONSTRUCTION
- 25 MM GAL PLANT HAS TWO REACTORS
- MATERIALS OF CONSTRUCTION: ZIRCONIUM-LINED MILD STEEL OR TEFLON-LINED MILD STEEL FOR TEMP. < 500°F.
- REACTOR LENGTH FROM STEAM MIXER TO OUTLET VALVE DEPENDS ON DESIRED RESIDENCE TIME.
- DARTMOUTH USES EXIT DRIFICE. LARGER REACTORS SHOULD USE PRESSURE CONTROL VALVE IN OUTLET.
- TANGENTIAL STEAM AND SLURRY INLET (SEE DIAGRAM INSET)
- MULTIPLE STEAM INJECTION PORTS WITH 0.04" - 0.08" PERFORATIONS (SEE INSET)
- DESIGN PRESSURE: 1200 PSIG / FULL VAC.; DESIGN TEMPERATURE 750°F (ZIRCONIUM LINED)
- REACTOR TUBE SHOULD BE CLEAN AND FREE FROM ANY OBSTRUCTIONS WHICH COULD CAUSE BLOCKAGE.



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The required residence time determines the reactor length. In the commercial scale reactor a residence time of 5.5 seconds based on reaction kinetics demonstrated at Dartmouth has been assumed. This equates to a reactor length of 27.5 feet with a reactor velocity of 5 feet per second. If a greater residence time is shown to be necessary this can be accommodated by increase of the reactor length.

(2) Tar Deposition and Blockage

In hydrolysis trials of extended duration (2 hours) carried out at Dartmouth during the present study blockage due to tar laydown was found to be a major problem. The exact cause of tar deposition is unknown, however, it is believed to be the product of furfural degradation and interactions between furfural, xylose, HMF glucose and dissolved lignin. The large scale reactor has been designed to reduce this problem by increasing the reactor velocity to 5 feet per second. Results from the extended duration trials at Dartmouth suggest that increased velocity reduces tar laydown in the reactor (Reference 8). Furthermore, exit orifice blocking experienced by Dartmouth at small scale with the 1 mm orifice should much less marked on the large scale where the reactor exit control valve is much larger. In addition, the large scale process has been designed with two parallel reactors and an installed spare to maintain continuous operation should it prove necessary to shut off one of the reactors on a regular basis to clear out excessive tar deposition. It may also be possible to construct the reactor outlet lines with teflon lined mild steel for temperatures up to 500°F. If this were allowable, teflon would reduce adhesion of tarry solids.

Excessive unreacted cellulosic solids in the reactor effluent may also contribute to blockage by adhering to any tar present on surfaces.

(3) Mixing and Heat Transfer

In scaling up the hydrolysis reactor care must be taken to address the poor mixing and heat transfer problems encountered by the Dartmouth researchers. During the experiments at Dartmouth, the reactor proved difficult to control and this can be attributed in part to the presence of steam vapor. Consequently, a reactor operating pressure has been chosen which corresponds to an equilibrium temperature 36°F higher than the chosen reactor temperature. This will give substantial temperature difference for condensation. Previous work on vapor condensation (Reference 12) indicates that this temperature difference causes complete collapse of an injected steam bubble within 0.5 seconds. This should result in improved control over reactor residence time and temperature due to elimination of steam vapor from the reactor and increased heat transfer rate. The larger diameter reactor will also afford a longer path of travel for injected steam bubbles before they reach the center of the reactor tube. This effectively gives steam a longer time for condensation. The above measures should also eliminate the possibility of steam-hammer sometimes experienced in this type of equipment.

Control should also be improved by the higher reactor velocity. The chosen velocity of 5 feet per second will give more uniform mixing in the reactor than obtained at Dartmouth where the velocity was around 0.25-0.7 feet per second. In addition, a tangential steam inlet as used at Dartmouth in recent work will be used for the large scale reactor to improve mixing. (See Figure 3.1). Consideration should also be given to use of a "lance" type axial injectors for further improvement to steam mixing.

The volume of the mixing chamber in the hydrolysis reactor has been calculated on the basis of published heat transfer measurements on steam injection mixers. These indicated a typical heat transfer coefficient of  $0.4 \times 10^6$  BTU per hour °F cubic foot suggesting a mixing chamber volume of 7 cubic feet (15" diameter x 4 feet long) for the base case design (Reference 27). This is supported by a second heat transfer calculation based on the length and number of heat transfer units for steam desuperheating and condensing, giving a required mixing chamber length of 3.1 feet for a diameter of 15" (Reference 31). A major scale-up issue will center around the reactor mixing tee heat transfer capacity.

(4) Reactor Operating Conditions

Choice of the reactor operating conditions residence time, temperature and acidity are based on operating data and reaction kinetics demonstrated experimentally at Dartmouth College. These kinetics show a trend towards higher fermentable sugar yield at higher temperature and shorter residence time. Higher temperatures also give increased conversion of xylose to furfural. This is shown in Figures 3.2 and 3.3 which are based on experimental data from the Dartmouth College (Reference 1).

The reactor operating conditions selected as a design basis are as follow:

Temperature	509°F = 265°C
Pressure	985 psig
Acidity	0.6% Sulphuric Acid
Residence Time	5.5 Seconds (Excluding Mixing Tee and Exit Piping)

A high degree of operating flexibility has been incorporated into the design.\* However, reactor optimisation is beyond the scope of this study since it depends on a number of considerations which can only be determined during actual operation. These include not merely maximising hexose yield but also:\*

- o Reactor Stability to Blockage by tar and polymer deposition.
- o Reactor Stability to unreacted cellulosic solids in the effluent which may cause blockage.\*
- o Furfural or other byproduct yields.
- o Wood feed particle size.
- o Available steam pressure.
- o Acid and lime costs.
- o Formation of difficult-to-separate extraneous organics which may be inhibitory to the yeast fermentation.

\*Reactor conditions can easily be varied to minimize unreacted effluent cellulose. For example at 509°F, 1% acid and 5.5 records residence time, Dartmouth kinetics (Reference 1, pg 140) predict only 3.7% of the feed cellulose remains unconverted. Unreacted cellulose has been taken to be zero in the process material balance. Furthermore, under these conditions the yield of fermentable sugars obtained by Dartmouth is over 55% (figure 3.2). (Reactor optimisation during future development may, in fact, show that the 0.6% acid concentration used in the present design needs to be increased.)

Figure 3.2

Hydrolysis of Hardwood to Glucose as a Function of Reaction Time and Temperature (Experimental Data)

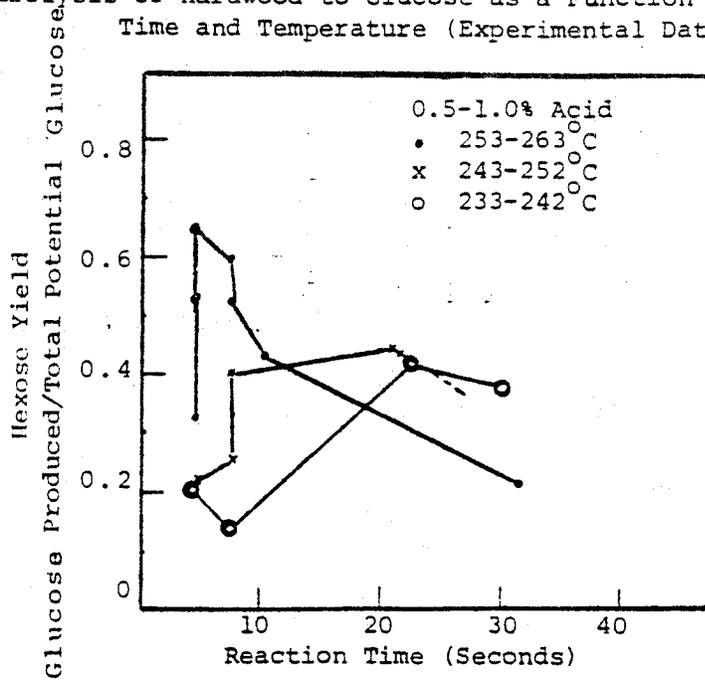


Figure 3.3

Furfural Yield from Hardwood as a Function of Temperature (Experimental Data)

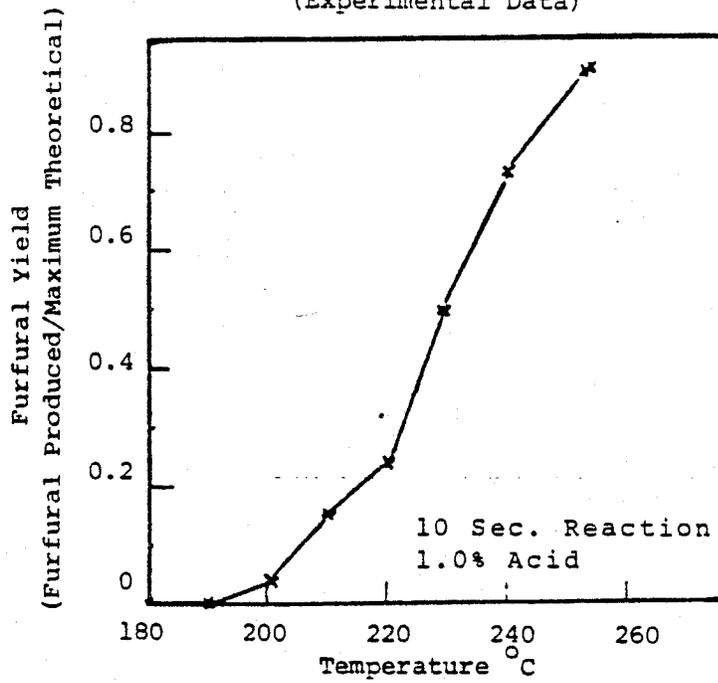


Figure 3.2  
Hydrolysis of Hardwood to Glucose as a Function of Reaction Time and Temperature (Experimental Data)

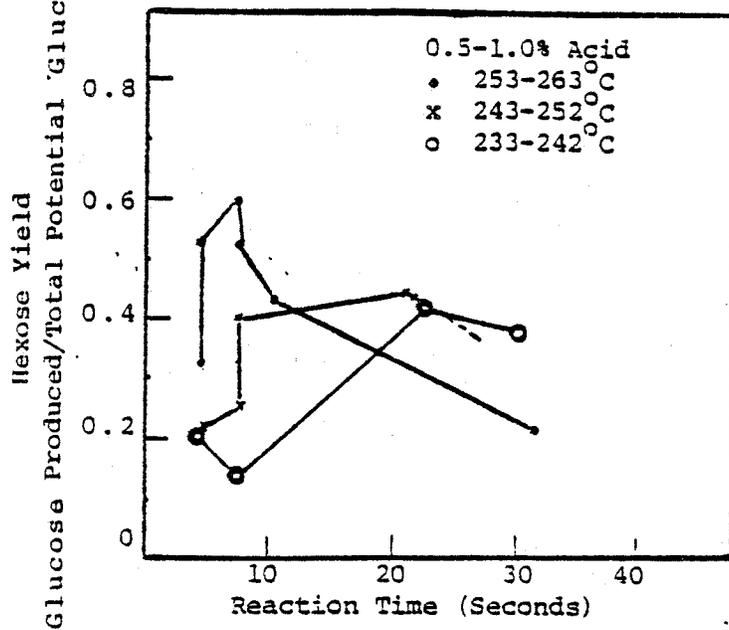
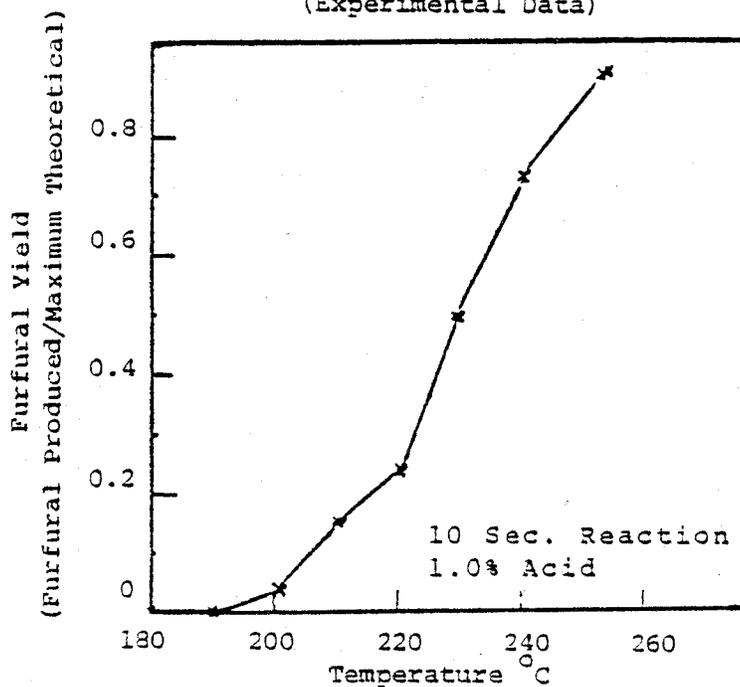


Figure 3.3  
Furfural Yield from Hardwood as a Function of Temperature (Experimental Data)



(5) Reactor Feed Solids Concentration

167°C  
In the commercial scale process design a wood slurry consisting of 20% solids by weight is made up in a slurry vessel. The temperature in this vessel is sustained at 332°F and the sulfuric acid content in the liquor is sustained at approximately 1%. The slurry is held for 5 minutes and, at these conditions, the wood softens and is partially hydrolyzed forming a "mud" which can be pumped using the twin screw pump specified. Work carried out during this study at Badger's Weymouth Laboratory with "through 20 mesh" aspen wood shows that 20% solids "slurries" can be readily softened at the above conditions to form a pumpable mud. Use of 20% solids in the reactor feed for the scaled-up process represents a major change from the small scale where the maximum solids concentration was 10%. However, the small scale operation did not use a slurry vessel for prehydrolysing the reactor feed.

The advantages of increased solids content in the reactor feed are a higher concentration of sugars in the hydrolyzate and reduced reactor energy requirements. (This is fully discussed in Reference 13).

(6) Reactor Steam Usage

The large scale hydrolysis reactor steam usage is based on a reactor temperature of 509°F. Reactor temperature is controlled by varying steam addition rate. Maximum use has been made of the available steam enthalpy by heat recovery at 150 psig from the hot reactor effluent. It is desirable for energy economy to attain as high a temperature as possible in the slurry vessel without causing significant hydrolysis of the cellulose or significant formation of furfural condensation products. A temperature of 332°F was chosen based on Dartmouth kinetics (Reference 1, 2, 3). A slurry vessel temperature of 332°F is also expected to give a sufficient degree of prehydrolysis as discussed in (5) above.

### The Alternative Hydrolysis Reaction Case

A different configuration for the hydrolysis system has been investigated in the alternative case design (Design Case II). In the alternative case, a two stage acid hydrolysis reaction is carried out instead of the single stage reaction previously described. The temperature in the first stage reactor is lowered resulting in less conversion of cellulose but greater yield of sugars from the cellulose converted. Unconverted cellulose is then fed to the second stage reactor where the hydrolysis is completed at a higher temperature. The sugar solution produced in the first stage is removed by centrifugation and washing. Sugars produced in the second stage are likewise removed in a separate centrifugation and washing step. Acid hydrolyzate streams from first and second stage are then combined in the neutralization step.

The two stage reactor design case has the potential advantages of lower overall capital costs, higher ethanol yield and less extraneous organic production and less tar production. but the disadvantage of producing less furfural byproduct, no excess electrical power and requiring supplementary fuel for the boiler in the form of wood chips partially offsetting the aforementioned yield advantage.

Although it was beyond the scope of this study it may well be possible to "optimize" the two stage design to increase the level of furfural production while maintaining the ethanol yield. Should this prove possible, the economics of Design Case II might be improved to the extent that the two stage system be preferred from an operational point of view.

### Wood Chip Grinding

A refiner grinder stage using 9000 HP total is used to produce "through 20 mesh" (840 micron) wood particles. This choice of unit operation is conservative for two reasons: firstly, the energy usage figure was obtained from production of "through 60 mesh" (250 microns) wood particles and should be lower for "through 20 mesh" (840 microns) particles; secondly, the use of the slurry vessel to precook the wood slurry may allow larger than "through 20 mesh" particles to be used. If the acceptable particle size were as large as say 2000 microns (2 mm) it may well be possible to use a hammer mill for chip grinding thus making significant savings in capital and energy costs.

### Lignin Solids Separation and Washing

Lignin solids separation and washing is carried out using three stages of nozzle disc centrifuges. These are special centrifuges which are operated so that wash water is introduced into the periphery of the bowl and flows countercurrently through the solids as they are forced outwards. This gives a high efficiency displacement wash system allowing 99% recovery of sugars in only three stages with a wash water to process flow ratio of 0.19 pounds wash water per pound of process flow. This system was designed and specified by the centrifuge manufacturers on the basis of laboratory trials carried out on Dartmouth hydrolyzate (see Appendix 4).

Other methods of lignin separation investigated during the study were dissolved air floatation followed by centrifugation, liquid hydroclones, and filtration. Although hydroclones at first appeared attractive, a laboratory test using hydrolyzate proved unsuccessful. These laboratory trials demonstrated that hydroclones do not give a good solid-liquid separation due to lignin floc disintegration in the high fluid shear regime within the hydroclone. This trial is the subject of an additional study appended to this report (see Appendix 4).

Hydrolyzate samples were sent to manufacturers of dissolved air flotation units. These manufacturers, specifically Linatex, Krofta Engineering, and Komline Sanderson, indicated that this was a suitable application. However, on the basis of capital cost figures obtained verbally a DAF/centrifugation system does not appear to be a cost effective method when compared with the disc nozzle washing centrifuge system chosen.

Filtration of acid hydrolyzates to remove lignin has been attempted by Dartmouth workers and at Badger Laboratories during this study. Experiments shows that the hydrolyzates are extremely slow filtering due to severe blinding of the filter cloths. Badger has measured the apparent specific resistance of residual lignin/tar cakes at  $2.1 \times 10^{-14} \text{ m}^{-1}$ .

### Hydrolysate Flashing

Hydrolysate Flashing is used to halt the hydrolysis reaction rapidly and to separate most of the furfural produced. The out-flow from the reactor at temperatures over 480°F is essentially completely liquid since all the cellulose and hemicellulose has been hydrolysed and dissolved and the temperature is above the glass point of lignin (Reference 25). Solidification of the lignin would be expected to take place on cooling, downstream of the flash valve. For this reason blocking of the reactor outlet valve should not be a problem. Blockage may be encountered however in the flash drum outlet (see Appendix III - Summary Report from Professor H.E. Grethlein).

Approximately 85% of the furfural produced in the hydrolysis reaction is separated in the flash vapors from the two stage flash system due to the very high activity coefficient of furfural in dilute aqueous solution. Flash vapors are condensed against incoming hydrolysis water for heat recovery. Total condensate from the two flashes contains furfural at a concentration of 6%.

### Neutralization

Neutralization to pH 10 of the acidic supernatant from the lignin centrifuges is carried out using calcium hydroxide. This over-neutralization with lime has been found by researchers at Dartmouth College to precipitate out soluble tars inhibitory to the fermentation.

### Gypsum Separation and Washing

Gypsum separation and washing is carried out using liquid cyclones. These allow many stages of washing and separation for relatively little capital cost. This reduces the need for excessive wash water for recovery of sugar from the cake liquors. In the present design eight stages of hydroclones and a final gypsum filtration are used resulting in 99.6% recovery of sugars using only 0.02 pounds wash water per pound of process flow. Hydroclones have been used on a commercial scale in the past for separation of gypsum from aqueous streams.

### Sugar Stripping

Sugar stripping using steam further reduces the levels of extraneous organics in the hydrolyzate. Most of the organics inhibitory to yeast fermentation are removed the previous steps of during hydrolyzate flashing and pH adjustment. The levels of these organics are further reduced by steam stripping. This produces a hydrolyzate stream with organic levels found by Dartmouth College researchers to be below that inhibitory to acclimatized yeast. The toxicity of extraneous organics to yeast in the JGC fermentation has been the subject of additional work carried out during the main study and is reported in Appendix 2. This work indicated that extraneous organics remaining in "treated" hydrolysates were not toxic to yeast used in the JGC process.

The Dartmouth work resulted in the following tentative specifications for acceptable levels of organics in the fermentation feed (Reference 25).

Furfural	0.1%
HMF	2.0%
Alpha-pinene	0.1%

### Unit 300 - Fermentation

In the fermentation section the sugar solution from the acid hydrolysis section is fed to the continuous fermenters where the fermentable sugars are converted to ethanol.

The fermentation system was designed by Japanese Gasoline Corporation (JGC) under subcontract. The JGC system is a continuous fixed bed fermentation with yeast immobilized within the fermenter. The yeast is entrapped in beads of photo-crosslinkable resin. The resin is a permeable polymer matrix in which the molecular structure is designed to allow free passage of substrate and product while preventing both escape of entrapped yeast or permanent establishment of any contaminants.

The JGC system was chosen for the following reasons:

- (1) Capital Cost - the high productivity obtained with the continuous system allows much smaller fermentation volume for a given ethanol production rate than required for batch fermentation.
- (2) Stability - the fixed bed nature of the system results in much greater stability to shock loads of inhibitory organics.
- (3) Durability - the JGC system uses a crosslinked polymer matrix of much higher compressive and tensile strength compared with natural polymers such as alginate or kappa-carageenan.

- (4) Hydrolyzate Fermentability - past work on hydrolyzate fermentability (Reference 5 and 6) has shown increased fermentation rate and yield when a very large inoculum has been used in batch fermentations or when continuous fermentation has been used (Reference 6).
  
- (5) Retention of Tolerance - work at Dartmouth College and elsewhere (Reference 6) shows that yeast strains need adaptation to hydrolyzates in order to give good fermentation rates and yields. The JGC system has the advantage of maximum retention of adapted, tolerant yeast within the fixed bed.

Samples of acid hydrolyzate produced at Dartmouth and further prepared by filtration, neutralization, concentration and steam stripping at Badger's Weymouth Laboratory have been sent to Japan for toxicity trials. In the trials yeast was exposed directly to hydrolysate of pH 5.2. Although no fermentation occurred yeast remained alive. These limited duration experiments suggest that the remaining extraneous organics in the hydrolyzate stream after the above operations were not toxic to the yeast nor corrosive to the entrapping polymer (see Appendix 2).

### Unit 400 - Ethanol Purification

Distillation calculations carried out during the present study show that an energy efficient method for ethanol separation from the dilute beer is to use two columns with the first column concentrating the ethanol to some intermediate point and the second column concentrating the ethanol to its azeotropic composition. In this way both columns have much lower reflux ratios (typically 0.4 and 1.6) resulting in much lower energy requirements than a single stage distillation column which requires reflux ratios of 15-20. This system has not been fully optimized and improved energy efficiency may be possible. For example, the distillation scheme based on innovative application of heat pumps operating between intermediate points on the column as described by Grethlein and Lynd could be applicable for very dilute ethanol beers (Reference 32). Further, there may well be alternative separation methods that give lower energy requirements. Low energy separation methods for dilute beers should relax water balance constraints on the hydrolysis section of the process.

### Multicomponent Vapor

Multicomponent vapor liquid equilibrium modelling done during the study show that the actual azeotropic composition is lowered and the equilibrium curve depressed from the simple binary model by the presence of extraneous organics such that the distillate from the second column is limited to 80 mole percent ethanol. This nearly doubles the amount of water that must be removed by the molecular sieve to obtain an anhydrous product.

### Molecular Sieves

Previous design studies by Badger have shown that molecular sieve dehydration offers several competitive advantages over the standard azeotropic entrainer method, including lower unit energy consumption and easier operation. For these reasons we have selected molecular sieves for the dehydration process. Although the comparison was made for a conventional azeotropic mixture (95% ethanol) it is believed that the conclusions still hold for azeotropic mixtures containing increased amounts of water.

### Fuel Grade Ethanol

The anhydrous ethanol product contains 0.5 percent by weight volatile components, most of which is acetaldehyde. No special provision has been made to concentrate and separate these components during the distillation process because of their insignificant amount. When the ethanol product is blended into engine fuels at ten percent by weight of total blend, acetaldehyde will be present at an unobjectionable 0.05 weight percent. This matter has been discussed with a commercial producer of fuel grade ethanol. The tentative conclusion was that the expected impurities were acceptable.

### Energy Integration

Concentrating ethanol is an energy intensive process and every opportunity for economically recovering heat should be utilized. Preheating feed with column overheads or bottoms and recompression of overhead vapor to supply reboiler heat are several standard methods. Conventional vapor recompression is found to be uneconomical in either of the two columns in the system because of the wide differences between overhead and bottoms temperatures and the minimum temperature differences required to transfer heat in the reboilers. What is needed is an overhead composition not so far removed from the bottoms composition. Such an opportunity presents itself by energy integration between units. The overhead vapor from the sugar stripper can be economically utilized to reboil the beer column using vapor recompression. This requires 2.2 mw of power but saves 17 pounds steam per gallon of ethanol product.

Unit 500 - Furfural Recovery

The design of the furfural recovery section is based on data developed by Smuk (Reference 8). Vapor-liquid equilibrium calculations have been based on a furfural-water binary mixture. For the purposes of the present study this approach is considered satisfactory. However, more rigorous calculations accounting for the effect of extraneous organics should be carried out as an additional study before a final design is developed.

Unit 600 - Tankage and Export

Denaturing Product Ethanol

Fuel alcohol, if transferred to another plant, need not be rendered unfit for beverage use, however, in the present study, it was assumed that not all customers would have their own denaturing facilities and therefore denaturing facilities have been included. Ethanol exported from the plant will have 5 gallons of gasoline added for every 100 gallons of pure ethanol product. This decision results in new capital equipment (for gasoline storage and blending) and adds 5% required capacity to the ethanol product tanks export pumps or around \$1 mm to the capital cost in the base case plant.

Fusel Oil

A fusel oil storage tank and export pump is provided to give flexibility in choosing the ultimate disposition of fusel oil byproduct. It may be burned onsite in the boiler, sold directly as is, or blended with ethanol product as fuel alcohol. In the present study provision has been made for its export. However, no credit has been taken for its sale.

Unit 700 - Environmental

Waste Water Treatment

Aqueous effluents produced from ethanol beer stripping contain approximately 40,000 pounds per hour of extraneous organics produced during the hydrolysis reaction and fermentation. These are treated by evaporation, burning of the concentrated organics as fuel and anaerobic digestion of part of the evaporate.

Anaerobic digestion alone was investigated as a possible alternative method of waste stillage treating. This would have the advantage of converting the organics to methane, a highly versatile fuel which could be burned, exported directly or even used as a feed for a methanol synthesis process. 339,840 cubic feet per hour (309 million BTU per hour) of methane from an anaerobic digestion system would be obtained; (resulting methanol production would be around 9000 tons per year). The anaerobic treatment system would, however, cost in the region of \$20 million (including aerobic polishing) and has the added disadvantage that it would be several times larger than any such system previously built.

The advantages of the evaporation option are that it produces a liquid solvent mixture which can be burned to produce steam in excess of evaporator requirements. Associated capital costs including anaerobic digestion of evaporate and extra boiler capacity are around one half of that for the large anaerobic digestion system and production of a concentrated mixed organics stream allows the future possibility of separation of the more valuable components such as hydroxymethyl furfural (HMF), levulinic acid, oleic acid or alpha pinene.

The condensate from the evaporators is of suitable quality to allow recycle of 50% of it to the hydrolysis section as process water. The remaining 50% is sent to anaerobic digestion and aerobic polishing to reduce the organic level to less than 5 ppm before being recycled to cooling tower make-up water.

The waste water treatment system allows total recycle of water and eliminates the need for any aqueous discharge to the environment.

Demineraliser regeneration and boiler blowdown streams are treated in the evaporation system.

#### Low Pressure Vents

Most of the low pressure process vents contain air or carbon dioxide saturated with water and small amounts of volatile organics. Air comes both from displacement during filling of vessels and from release of air trapped in pores in the raw wood feedstock. The organics are mostly ethanol and the extraneous organics such as terpenes and extractive resins. In process modeling terpenes and resins were characterized as alpha pinene and oleic acid respectively. These would present odor problem if vented directly to the atmosphere. Scrubbing with water would be inadequate since alpha pinene and oleic acid are insoluble. Recovering such small amounts of organics by chilling and condensing is uneconomical. Therefore, it was decided to incinerate the vents by mixing them with combustion air in the boilers.

Unit 800 - Utilities

Boilers

Two boilers are provided for the process. Each produces 50% of the total steam and 64% of the process needs. (Excess steam is produced in the base case design.) A further design criterion for the boilers is that either must be capable of burning all of the lignin/tar residue produced by the acid hydrolysis section. This capability avoids the necessity of storing or disposing of excess lignin/tar sludge produced during periods when one of the boilers is not in operation. (During these periods mixed solvents would be stored or exported for incineration.)

The fuel mix taken by the boiler under normal circumstances comprise:

	<u>% of Total Fuel Energy</u>
Lignin/Tar Solids	53%
Mixed Solvents	46%
Digester Gas	1%

The minimum boiler steam pressure was dictated by the hydrolysis reactor steam pressure requirement of 1100 psia.

### Flakt Drying System

Lignin sludge at 50% moisture from the hydrolysis process is dried using the Flakt drying system. This is a relatively new system but has several recent commercial applications in Europe and the U.S.

A portion of the boiler flue gas is used to dry the sludge in a spray dryer system. Solids are dried to a powder form containing around 15% moisture. This results in much higher burner efficiency (around 89%) than firing wet solids. A conservative efficiency of 85% has been used in calculations for the present study.

The total flue gas stream from the boilers, including that used for solids drying, is processed for particulate emissions control.

### Turbogenerator

In order to determine the best steam pressure for economical design for the turbogenerator, a study was carried out using vendor-supplied data. The major trade-off was that of higher steam pressures giving increased power generation versus the possibility of higher capital cost generating equipment required for higher pressure steam.

For larger turbines of 8000 kw upwards steam inlet pressure has little effect on capital costs over the range 200-1600 psig. Therefore, at both the 25 and 5 million gallon scale the turbogenerators are specified to operate on the 1100 psia steam direct from the boilers.

PROCESS DESCRIPTION

This section contains a description of the process operation, process flow diagrams, plot plans, and equipment lists.

Unit 100	Feedstock Handling
Unit 200	Acid Hydrolysis
Unit 300	Fermentation
Unit 400	Ethanol Purification
Unit 500	Furfural Recovery
Unit 600	Offsite Tankage
Unit 700	Waste Treatment
Unit 800	Utilities

The process description as presented applies to both the 25 million gallon per year ethanol plant (base case) and the 5 million gallon per year plant with onsite power generation (Design Case III). A separate process description for the alternative case hydrolysis reaction (Design Case II) is given. No additional process description for the 5 million gallon plant with purchased electricity (Design Case IV) is given since there is no significant operational difference between this and Design Case III.

Process flow diagrams for Design Case I (base case) and for the hydrolysis reaction area of Design Case II (alternative case) are presented. The remaining units in Design Case II are identical to the base case.

Equipment lists for Design Case I, Design Case II, and Design Case III (5 million gallon with onsite power generation) are also provided. The equipment list for Design Case IV is not provided since it differs from Design Case III in only two items: firstly, elimination of the turbogenerator and, secondly, slight reduction in the size of the boilers.

Unit 100 - Feedstock Handling

Fresh wood chips (50% moisture) are delivered by rail (railcar type 60 short ton net capacity) or truck (truck trailer type 24 short ton net capacity). The receiving station consists of a hydraulic truck dumper, a truck weigh scale, a rail grating and a railcar weigh scale. Chips are offloaded into the washing flume which allows separation of heavy tramp material such as rocks and large metal items which sink to the bottom of the flume. Chips are swept by the flume pump (PP-101) and discharge to a bucket conveyor GS-101 which transfers the chips to the wood pile. Periodically, the flume is emptied and separated tramp material cleared. Water entrained with the chips as they are conveyed from the flume drains back from the conveyor.

The wood pile holds two weeks supply of wood chips and is managed on a first in first out basis by means of a front end loader truck GM-104A . Chips are fed from the pile through a vibratory chip pile screen (with 6" aperture) using a front end loader GM-104B to the belt conveyor GS-102 which starts below ground. Entrained small metallic items down to  $\frac{1}{2}$ " size are removed by a magnetic separator GS-105. The belt conveyor transfers the chips to the hydrolysis process, Section 200.

Unit 200 - Acid Hydrolysis

Chips from the belt conveyor are held in an activated bottom discharge chip feed bin MB-216 from where they are fed by means of chip screw feeders GS-201 A/B/C to the refiner grinders GG-202A/B/C. The grinders reduce the chip size to through 20 mesh (0.84 mm). The wood grinds are conveyed using conveyors GS-203A/B/C to a screen GY-229 which removes oversize wood of 1" or greater for regrinding. Wood grinds are fed from the ground wood bin MB-217 to weigh belts GS-231A/B. MB-217 is equipped with an activated bottom discharge to ensure continuous discharge to the weighfeeder. From the weighfeeder wood grinds fall into the rotary high pressure feeders GA-204A/B which feeds into the slurry vessels MS-201A/B.

Hydrolysis process water from storage tank MT-223 is pumped by PP-202 through exchanger TT-209 where it is heated by exchange with hot hydrolyzate to 141°F. It is further heated to 360°F by contact with low pressure hydrolyzate flash vapors (TT-211), and high pressure hydrolyzate flash vapors (TT-202 and TT-201). Sulphuric acid is metered using PD-219 and mixed, under flow ratio control, with the process water stream in static mixer GA-207 to give a dilute acid solution. The acidified process water is further heated to 413°F in TT-207 using high pressure steam (1100 psia) and is transferred under flow control to MS-201A/B where it is mixed with the ground wood to make a slurry of 20% solids concentration. Conical rotating screw mixers GA-205A/B ensure a uniform slurry in vessels MS-201A/B and prevent plugging of the bottom outlet.

Vessels MS-201A and B are maintained under a nitrogen pressure of 150 psig to prevent boiling and ensure a reasonable net positive suction head for pumps PD-201A/B. Vented vapors from MS-201A/B are condensed in TT-215 using cooling water. Condensate drains by gravity to second stage wash tank MS-225 and the remaining cooled vapor enters the low pressure vent header.

Wood slurry mud formed at the conditions in the slurry vessel flows out of MS-201A/B into the positive displacement screw pumps PD-201A/B and is forced under 1000 psia pressure into the hydrolysis reactors MR-202A/B. High pressure steam is injected into the hydrolysis reactors to raise the reactor temperature to 509°F. Pressure in the reactor is maintained by a back pressure control valve in the outlet line. Reactor temperature is controlled by varying the steam injection rate.

At the reactor conditions hexan polymer (cellulose) is hydrolyzed to soluble hexose sugars and other degradation compounds and pentosan polymer (hemicellulose) is converted to furfural and other degradation products.

Outflow from the reactors is flashed to 150 psig into the high pressure flash drum MS-203. Pressure is maintained in MS-203 by a pressure control valve on the vent outlet from TT-201. Liquid outflow from MS-203 on level control is further flashed to 10 psig in the low pressure flash drum MS-204. Pressure is maintained in MS-204 by a pressure control valve on the vent outlet from TT-211. Liquid from MS-204 consisting of a slurry of lignin and tar solids (6.9% by weight) in an aqueous solution of hexose sugars (6.5% by weight) and other miscellaneous organics is pumped on level control by lignin slurry pump PD-204 A/S and further cooled to 200°F by contact with incoming process water in slurry cooler TT-209.

Flash vapors from MS-203 containing mainly water and furfural are condensed against incoming process water in primary feedwater preheater TT-201 and high pressure flash vapor aftercooler TT-202 and further cooled in aqueous furfural cooler TT-205 by cooling water before being sent to the aqueous furfural drum MS-209. A portion of the high pressure flash vapors from MS-203 is also sent to the sugar stripper reboiler TT-203, condensate is returned from TT-203 to TT-202 using sugar stripper reboiler pump PP-226. Flash vapors from MS-204 are mainly condensed against incoming process water in secondary feedwater preheater TT-211. A small portion is used to reheat neutralized hydrolyzate slurry in hydrolyzate reheater TT-213. Condensate is returned to TT-212 by hydrolyzate reheater pump PP-227. Condensate from TT-211 and TT-213 is cooled in low pressure flash vapor aftercooler TT-212 and sent to MS-209.

Lignin residue is separated from the hydrolyzate in a three stage centrifuge system using lignin centrifuges GC-209A/B, GC-210A/B and GC-232A/B. Process water for counter current washing of the lignin solids to reduce losses of fermentable sugars is introduced into the third stage centrifuge GC-232 A/B. The washed lignin solids are dewatered to 50% moisture in solid bowl dewatering centrifuges GC-214A/B. The lignin sludge drops into lignin sludge tank MS-212 from where it is screw conveyed via GS-211A/B/C to drying and burning.

Overflow hydrolyzate from the first stage centrifuge flows to the neutralization reactor MR-206 where it is neutralized to pH 10 with lime slurry. Outflow from the neutralization reactor consisting of gypsum and precipitated tars (1.4% by weight total) in aqueous hydrolyzate is pumped using alkaline hydrolyzate pump PP-209 to gypsum hydroclone feed tank MS-207 where it is reacidified to pH 5 with sulfuric acid. Gypsum and precipitated tar solids are separated from the liquids and countercurrently washed using eight stages of hydroclones GC-215 and GC-216A-G.

Gypsum and tar solids are finally dewatered to 60% solids using gypsum rotary filter GF-220. To maximize precipitation of gypsum solids in order to prevent scaling in downstream process stages, the gypsum slurry is heated to 225°F with 10 psig flash vapors in TT-213 prior to separation and washing.

Overflow from the primary gypsum hydroclone GC-215 is cooled to 150°F in overflow cooler TT-214 and held in sugar stripper feed tank MT-222. MT-222 is sized for two hours hold time and is equipped with an internal baffle system and sump to allow separation of any entrained solids from upset operation of gypsum or lignin separations. Sump solids are returned periodically using sump pump PP-224 to the gypsum filter feed tank MS-214.

A small flow of hydrolyzate is taken from MT-222 via lime slurry liquor pump PP-216 and used for lime slurry make-up.

Residual extraneous organics inhibitory to the yeast fermentation are reduced to an acceptably low concentration by steam stripping in sugar stripper AS-201. AS-201 is fed on flow control by sugar stripper feed pump PP-215 and is reboiled by vapors from the high pressure flash drum MS-203. Condensate from TT-203 is returned under level control by PP-226. Overhead vapors from the sugar stripper are recompressed in PC-402 and used to reboil the beer column AS-401. Overhead vapors from the sugar stripper can also be directed to the furfural azeo column AS-501 in the event of a temporary shutdown of the beer column. Subcooled condensate from the beer column reboiler is returned to the aqueous furfural drum MS-209.

Sugar stripper bottoms are used to preheat the incoming feed in sugar stripper preheater TT-210, and further cooled to 122°F by exchanging against cold boiler feed water in hydrolyzate cooler TT-206 and are finally stored in fermentation feed tank MT-604 at a temperature of 122°F to inhibit contamination.

Aqueous furfural streams separated during hydrolyzate processing are collected in MS-209 and transferred to intermediate storage in the azeo column surge tank MT-606.

Lime slurry is made up in lime mix tank MS-218 using hydrolyzate liquors from MT-222. Lime powder is metered from bulk storage hopper MB-220 using star-valve GS-225 and weigh belt GS-226 to MS-218 where it is slurried with incoming hydrolyzate liquors. Slurry is circulated using lime mixing pump PP-220 and fed to lime slurry day tanks MT-219A/B. This operation is continued until the day tanks are full. Lime solids concentration in MT-219 is adjusted if necessary by recirculation to MS-218 using lime slurry pumps PP-221A/B.

Sulphuric acid is dosed directly to the process from day tank MT-221 using acid dosing pump PD-210 and sulfuric acid feed pump PD-219. MT-221 is filled from bulk storage tank MT-603.

Hydrolysis process water and fermenter off-gas scrubber water, is provided by MT-223. This tank is also used to collect recycled process water from stillage evaporation (MS-701) and furfural purification (MS-502) as well as make-up process water.

Unit 200 - Acid Hydrolysis (Alternative Case)

Chips from the belt conveyor are held in an activated bottom discharge chip feed bin MB-216 from where they are fed by means of chip screw feeders GS-201 A/B/C to the refiner grinders GG-202 A/B/C. The grinders reduce the chip size to through 20 mesh (0.85 mm). The wood grinds are conveyed using conveyors GS-203 A/B/C to a screen GY-229 which removes oversize wood of 1" or greater for regrinding. Wood grinds are fed from the ground wood bin MB-217 to weigh belts GS-231A/B. MB-217 is equipped with an activated bottom discharge to ensure continuous discharge to the weigh feeder.

From the weighfeeder wood grinds fall into the rotary high pressure feeders GA-204A/B which feed into the slurry vessels MS-201A/B.

Process water from storage tank MT-223 is pumped by PP-202 through exchanger TT-209 where it is heated by exchange with hot hydrolyzate to around 141°F. It is further heated to 360°F by contact with low pressure hydrolyzate flash vapors (TT-211), and high pressure hydrolyzate flash vapors (TT-202 and TT-201). Sulphuric acid is metered using PD-219 and mixed, under flow ratio control, with the process water stream in static mixer GA-207 to give a 1% acid solution. The acidified process water is further heated to 413°F in TT-207 using high pressure steam (900 psia). A portion of this stream is taken off to the second stage hydrolysis. The remainder is sent under flow control to MS-201A/B where it is mixed with the ground wood to make a slurry of 20% solids concentration. Conical rotating screw mixers GA-205A/B ensure a uniform slurry in vessels MS-201A/B and prevent plugging of the bottom outlet.

Vessels 210 A and B are maintained under a nitrogen pressure of 150 psig to prevent boiling and ensure a reasonable net positive suction head for pumps PD-201A/B. Vented vapors from MS-201A/B are condensed in TT-215. Wood slurry mud formed at high temperature flows out of MS-201A/B into the positive displacement screw pumps PD-201A/B and is forced under 800 psia into the hydrolysis reactors MR-201A/B. High pressure steam (900 psia) is injected into the hydrolysis reactors to raise the reactor temperature to 482°F. Pressure in the reactor is controlled by varying the steam injection rate. At the reactor conditions a portion of the hexan polymer (cellulose) is hydrolyzed to soluble hexose sugars and other degradation compounds and the pentosan polymer is hydrolyzed to pentose sugars, furfural and other degradation compounds.

Outflow from the reactors is flashed to 150 psig into the high pressure flash drum MS-203. Liquid outflow from MS-203 is further flashed to 10 psig in the low pressure flash drum MS-204. Liquid from MS-204 consisting of a slurry of lignin, unconverted cellulose, and tar solids (9.1% by weight) is further cooled to 200°F by contact with incoming process water in TT-209. Flash vapors from MS-203 containing mainly water and furfural are condensed against incoming process water in TT-201 and TT-202 and further cooled in TT-205 before being sent to the aqueous furfural drum MS-209. A portion of the high pressure flash vapors from MS-203 is also sent to the sugar stripper reboiler TT-203. Condensate is returned from TT-203 to TT-202 using PP-226. Flash vapors from MS-204 are mainly condensed against incoming process water in TT-211. A smaller portion is used to reheat neutralized hydrolyzate slurry in TT-213.

Lignin residue and unconverted cellulose is separated from the hydrolyzate in a three stage centrifuge system using centrifuges GC-209A/B, GC-210A/B and GC-232A/B. Process water for countercurrent washing of the lignin and cellulose solids to reduce losses of fermentable sugars is introduced into the third stage centrifuge GC-232. The washed solids are dewatered to 50% moisture in solid bowl centrifuges GC-214A/B. Overflow hydrolyzate from the first stage centrifuge GC-209A/B flows into the neutralization reactor MR-206. The sludge from GC-214 containing lignin residue precipitated tars and unconverted cellulose is discharged into the sludge dilution vessel MS-259 where it is diluted with process water to 35% solids before being pumped to the second stage slurry vessel MS-251 where it is mixed with hot acidified process water from TT-207. The temperature in the slurry vessel is 263°F and it is maintained at a pressure of 150 psig with a nitrogen blanket to prevent boiling and to ensure a reasonable net positive suction head for second stage hydrolysis feed pump PD-250. Lignin/cellulose slurry is pumped under 800 psia pressure into the second stage reactor which operates at 482°F. Here, remaining cellulose is converted to hexose sugars and other degradation products.

Outflow from the reactor is flashed to 150 psig in second stage high pressure flash drum MS-252 and is further flashed to 10 psig in low pressure flash drum MS-253. Slurry from MS-253 is cooled to 200°F and residual lignin and tar solids are separated and washed in a two stage centrifuge system GC-251A/B and GC-252. Process water for washing is fed under flow control to GC-252. Discharged solids from GC-252 are finally dewatered in GC-253 which produces a sludge of 50% moisture content. The sludge is discharged to lignin sludge tank MS-257 and is transferred to drying and burning via sludge screws GS-254 A/B/C.

Flash vapors from MS-252 flow to the vapor outlet of MS-203 to be condensed in TT-201 and flash vapors from MS-253 flow to MS-204.

Hydrolyzate overflow from the second stage primary lignin centrifuge GC-251 is sent to the neutralization reactor MR-206 for neutralization along with hydrolyzate liquors from the first stage.

From the neutralization reactor onwards the process is essentially the same as for the base case (one stage reactor) process.

Unit 300 - Fermentation

The fermentation section is designed to continuously produce alcohol from raw material consisting of saccharified wood, by means of a bioreactor in which immobilized yeast is used. The following covers the startup, normal operation, sludge removal and decontamination, regeneration of immobilized yeast, and emergency shutdown.

(1) Startup

41,000 gallons (total) of immobilized yeast beads are introduced from an immobilized yeast production unit into fermenters MR-301A and MR-301B. Subsequently, 61,500 gallons of medium is introduced there into. Preculture is started by injecting aseptic air into these reactors using an air compressor PC-308. Usually, the initial temperature of fermentation in these reactors is constant, but it rises due to heat of fermentation.

When the temperature reaches 89.6°F, it is maintained by cooling the broth in the broth coolers (TT-302 and TT-303), using the broth cycle pumps (PP-301 and P-302). After the fermentation has been completed, the broth is poured into a pit and the medium is again introduced into reactors A and B. The same operation as above-mentioned is repeated two or three times.

Also, the broth from the second and subsequent operation can be fed directly to the distillation section, because its fermentation has almost been completed.

Since the yeast concentration of the immobilized yeast will attain normal levels as a result of the above operation, the broth need not be taken out of the system after the final batch operation and normal operation can proceed.

(2) Normal Operation

Medium from the preparation section is supplied at the rate of 1344 gpm to two trains of fermentation tanks, cooled at 89.6°F by medium cooler, and introduced into reactor A. Carbon dioxide gas formed by fermentation is passed through a scrubber AS-301, and a part of the carbon dioxide gas is introduced into reactor A using the recycle blower and the immobilized yeast is fluidized. When the amount of carbon dioxide gas formed is small at the early stage of fermentation, it is fluidized with aseptic air. Also, it is cooled and kept at 89.6°F by the broth cooler using the broth recycle pump to prevent the temperature in the reactors from rising due to the heat of fermentation. A part of the recycle broth is introduced into reactor B, fermented in the same manner, and fed into the distillation section as fermented broth.

(3) Sludge Removal and Decontamination

This fermentation system is controlled at pH = 4.0, and contamination due to various bacteria, etc., which hinder alcohol fermentation, is slight. Also, this fermentation system is designed so that it can sufficiently cope with such contamination by means of the following operations. First, 10 wt.% sodium polyphosphate is prepared in the pH (potassium bisulfite-metal) tanks MS-305A/B in advance. After the broth in the reactors A and B has been fermented, the fermented broth is poured into the pit and fed to the distillation section.

Subsequently, sodium polyphosphate of about 1 wt.% is obtained by mixing the above-mentioned 10 wt.% sodium polyphosphate and process water in the piping, it is introduced into reactors A and B, and sludge adhering to the surface of the beads is removed, while fluidizing the immobilized yeast for about one hour using air. After putting the washing water into the pit, the sludge is almost completely removed by carrying out the same operation once more. Therefore, the fermentation is restarted by introducing the medium from reactor A to reactor B in accordance with the normal procedure. On the other hand, PBM of about 10 wt.% is prepared in the PBM tanks, the flow rate is controlled with the PBM pump (PD-307) so that the concentration of the above-mentioned 10 wt.% PBM becomes about 1,000 ppm in the medium, and this PBM is introduced into the reactors. Various bacteria in the fermentation system can be almost completely removed and normal fermentation speed can be maintained by continuously carrying out this procedure for about ten days.

(4) Regeneration of Immobilized Yeast

Though normal operation leads to practically no deactivation of the immobilized yeast in reactor A, it gradually deactivates that in reactor B. Therefore, in order to maintain stable fermentation activity, it can be said to be effective to occasionally regenerate the immobilized yeast in reactor B. Though the frequency of regeneration depends on the state of deactivation, one cycle per month is considered to be desirable. One method of regeneration is to transfer about one-fourth of the immobilized yeast in reactor B into the catalyzer regeneration tank MS-303A/B using the IMY transfer pump PP-304A/B, and after putting the broth into the broth storage tank, fresh medium is added thereto, and aeration and agitation is conducted.

After repeating this procedure twice a day, this broth is put into reactor B using the broth transfer pump PP-303A/B. It is possible to maintain stable fermentation for long periods of time by repeating this procedure once a week.

(5) Emergency Shutdown

Though an emergency situation is not foreseen at the fermentation section, it is necessary to shutdown the operation in the event of an emergency at the downstream sections such as distillation section and waste treatment section. The method of coping with an emergency in the fermentation system is to take measures to preserve the immobilized yeast in a live condition for a long period of time.

In this process, it is considered best to cope with an emergency by stopping the supply of medium, switching the cooling water of broth cooler A (TY-302A/B) and broth cooler B (TY-303A/B) to that of about 5°C if possible, and lowering the temperature of reactors A and B to 10-5°C. Therefore, it is necessary to separately provide a facility which can supply cooling water of 5°C for emergency use.

### Immobilized Yeast Section

This section is designed to comprehensively immobilize yeast having alcoholic fermentability, using photo-crosslinkable resin, and to continuously produce immobilized yeast in bead form. The following covers the basic operation of this section.

#### (1) Feed Preparation

Usually, the photo-crosslinkable resin (ENTG-3800) prepolymer is packed in drums and should preferably be kept at 20°C or less. When producing immobilized yeast, the photo-crosslinkable resin (ENTG-3800) prepolymer is fed into the resin tank MS-311 using the ENTG-pump PP-311 and the specified amount of polymerization initiator (KW-30) is added. These are then thoroughly mixed, using a mixer. As this resin solution remains stable over a relatively long period of time, no special precautions are required. Na-alginate, which is an immobilizing aid, is dissolved with aseptic water in the Na-alginate tank MS-313, to the specified concentration. The temperature is raised with steam, as necessary, when dissolving the Na-alginate. On the other hand, it is necessary to keep the yeast at 5°C or less, as an aseptic water suspension. The yeast liquid is fed into the yeast liquid tank MS-312 using the yeast pump PP-312 and, when necessary, its concentration is adjusted using aseptic water.

The resin solution, Na-alginate solution, and yeast suspension prepared in the above-mentioned manner are fed at their respective flow rates to the line mixer GA-311 using the feed pump PD-313, and after being uniformly mixed, they are introduced into the next process, namely, the immobilization unit.

(2) Immobilization

The liquid mixture fed from the line mixer is supplied to the atomizer of the continuous immobilizing unit. It is dripped from the nozzle of the atomizer to the header of the  $\text{CaCl}_2$  solution as particles of 2-2.5 mm size. The  $\text{CaCl}_2$  solution is prepared beforehand in the Ca-tank (MS-314) to the specified concentration, and is introduced into the header using the Ca-pump (PP-315). In addition, the solution is recycled, using the Ca feed pump to return it through four trains of lamp units and separators, to the header. The flow rate of each train at this recycling must be set to its optimum value necessary for immobilization. The mixed solution of resin and yeast dripped from the atomizer is then solidified in the form of beads by reaction with  $\text{CaCl}_2$ . In addition, it is fed together with the fluidized  $\text{CaCl}_2$  solution into the lamp units, and dropped into the separators as completely immobilized yeast obtained by photo-crosslink reaction.

The immobilized yeast accumulates in the lower part of the separators, and is controlled using a sequencer so that it automatically flows into an intermediate storage tank at a constant rate. As  $\text{CaCl}_2$  is recycled from the upper part of the separators through one of two strainers, the strainer must be switched to the other one to permit cleaning, whenever necessary. Also, a part of the  $\text{CaCl}_2$  solution is drained from the recycle line to the drain tank.  $\text{CaCl}_2$  solution equivalent to the drained  $\text{CaCl}_2$  solution is automatically supplied from the Ca tank. Immobilized yeast accumulated in the intermediate storage tank is cleaned with process water to wash away  $\text{CaCl}_2$ . The sequencer is controlled so that the cleaned immobilized yeast flows into an immobilized yeast storage pit at a constant rate. The immobilized yeast storage pit is kept at  $5^\circ\text{C}$  using a chiller GY-313, and the product immobilized yeast is stored in this pit until it is to be supplied to an alcohol production plant.

(3) Cleaning

Usually, the aforesaid immobilization operation is carried out continuously about 16 hours a day. After the continuous operation is completed, it is necessary to clean the inside of the spiral tube of the lamp unit. For this purpose, the  $\text{CaCl}_2$  solution is drained out of the immobilization system and cleaning liquid, prepared in advance to the specified concentration in the cleaning liquid tank, is introduced into the spiral tube of the lamp unit, and is held there for several hours. Subsequently, the immobilization operation can be resumed by returning the cleaning liquid to the cleaning tank and flushing the inside of the spiral tube with process water.

### Unit 400 - Ethanol Purification

Dilute beer containing 3.3 weight percent ethanol from fermentation (Unit 300) is stored in dilute beer feed tank MT-406 at 90°F to prevent contamination. From there it is pumped by dilute beer feed pump PP-409A/S on flow control to feed preheater TT-413 to exchange heat with hot aqueous furfural previously condensed in the shell side of the beer column reboilers. The fresh feed along with recycled bottoms from the rectification column, then enters degasser drum MS-401 where some of the dissolved CO<sub>2</sub> leaves as flash vapor. Degasser drum condenser TT-402 uses cooling water to condense and recover the light organic compounds and some water from the flash vapor and these liquids return to MS-401 through a loop seal. The remaining cooled vent vapor passes into the low pressure vent header. The degassed feed then passes through a loop seal and enters beer column AS-401 on tray twelve.

AS-401 is an atmospheric column with 16 sieve trays and is designed to concentrate the ethanol to 40 weight percent in the overhead while losing negligible amounts in the stillage. Heat is supplied to the column by beer column reboilers TT-403A/B. These are thermosiphon reboilers and the heating medium is overhead vapor from sugar stripper AS-201 compressed by sugar stripper vapor compressor PC-402 to 24 psig. The flow of compressed vapor is sent to the shell side of TT-403A/B and controlled by the column bottoms pressure. Excess vapor flow is sent directly to Azeo column AS-501. The process condensate from TT-403A/B is further cooled in aqueous furfural cooler TT-412 and feed preheater TT-413. The flow of hot aqueous furfural to TT-413 is controlled by the temperature of the dilute beer feed to MS-401.

The remainder of the hot aqueous furfural is cooled in TT-412 using cooling water. The combined cooled aqueous furfural streams are sent to aqueous furfural drum MS-209.

The flow of stillage from the bottom of the beer column, is controlled by the column liquid level and is pumped by beer column bottoms pump PP-401A/S to first effect evaporator GE-701.

Overhead vapor from AS-401 is condensed in beer column condenser TT-405 using cooling water and collected in beer column reflux drum MS-402. Vapor from MS-402 is further condensed in beer column vent condenser TT-406 using cooling water and the liquid is returned to the drum. The remaining cooled vent vapor passes into the low pressure vent header.

Reflux to AS-401 is pumped by beer column reflux pump PP-403A/S on flow control reset by reflux drum level control. Distillate product is fed to tray 14 of rectification column AS-402 by PP-403A/S on flow control reset by temperature control from a tray near the top of AS-401.

AS-402 is an atmospheric column with 24 sieve trays and is designed to concentrate the ethanol to its azeotropic composition (91.2 weight percent) in the overhead. The bottoms stream, containing 3.5 weight percent ethanol in water, is pumped by rectification column bottoms pump PP-407A/S on level control and recycled as feed to the beer column through MS-401. Heat is supplied to this column by rectification column reboiler TT-409. This is a thermosiphon reboiler and the heating medium is 50 psig steam controlled by the column bottoms pressure.

Overhead vapor from AS-402 is condensed in rectification column condenser TT-410 using cooling water and collected in rectification column reflux drum MS-405. Vapor from MS-405 is further condensed in rectification column vent condenser TT-411 using cooling water and the liquid is returned to the drum. The remaining cooled vent vapor passes into the low pressure vent header.

Reflux to AS-402 is pumped by rectification column reflux pump PP-408A/S on flow control reset by reflux drum level control. Distillate product is pumped to molecular sieve feed tank MT-404 by PP-408A/S on flow control reset by temperature control from a tray near the top of AS-402.

Several components, chiefly amyl, isoamyl, and propyl alcohols, termed fusel oils, and furfural tend to accumulate within the rectification column and would eventually affect the column's performance if not removed. A small continuous liquid sidedraw purge removes these components by gravity flow. The sidedraw stream is mixed with process water, cooled in fusel oil cooler TT-407 by cooling water, and enters fusel oil decanter MS-403. There the oily phase is decanted and pumped by fusel oil pump PP-405A/S on level control to fusel oil storage tank MT-605. Water and dissolved alcohols are returned to AS-402 on flow control by wash return pump PP-404A/S.

Azeotropic ethanol is pumped from MT-404 by molecular sieve feed pump PP-406A/S on flow control to molecular sieve GZ-401. This package unit, consisting of three towers and a regeneration skid, will dehydrate the ethanol to 199<sup>+</sup> proof via adsorption into the dessicant. The dessicant (molecular sieves) preferentially absorbs water over ethanol as the ethanol passes through the dessicant bed. The spent (saturated) dessicant is regenerated by simultaneously applying heat via embedded steam tubes and pulling a vacuum on the bed using a condenser and a vacuum pump. Dehydrated ethanol product is denatured and sent to ethanol product tanks MT-601A/B. Waste water is sent to equalization tank MT-703.

Unit 500 - Furfural Recovery

Aqueous furfural from intermediate storage in MT-606 is fed to the azeo column AS-501. This produces an overhead vapor stream of composition close to the azeotropic furfural water mixture (35 percent by weight furfural). This is condensed in TT-503 and subcooled in TY-502. The cooled stream splits into two immiscible phases, a furfural - rich phase and a watery phase. These are separated in decanter MS-501. The furfural rich phase from the decanter containing 84% furfural is sent to the dehydration column AS-502 from where a 99.5% furfural stream is taken as the bottoms product. This is cooled by heat exchange with incoming feed from the decanter and further cooled to 100°F in TY-507 and sent to product storage in MT-602A/B. The watery phase from the decanter is recycled as reflux to the azeotrope column. Azeotrope column bottoms consisting of water and extraneous organics are cooled to 100°F against incoming azeo column feed and sent to decanter MS-502 where alpha-pinene which separates out of solution at this temperature, is removed as a separate layer and sent to mixed solvents storage tank MT-702.

Aqueous furfural column bottoms from the decanter are recycled to the hydrolysis process water tank MT-223.

Unit 600 - Offsite Tankage

Offsite tankage is provided for bulk storage of intermediate and final liquid products, sulfuric acid, gasoline and diesel fuel, and fire water.

Ethanol product from molecular sieve package unit GZ-401 is pumped to ethanol product tanks MT-601A/B. Prior to entering the tank, the ethanol is denatured by blending with gasoline in the proportion five gallons of gasoline per one hundred gallons of ethanol product. Gasoline delivered by truck or railcar is stored in gasoline storage tank MT-610 which has two days capacity. The gasoline is blended into the ethanol by gasoline blending pump PP-610A/S on flow control. Denatured ethanol is transported to the shipping point by ethanol export pumps PP-601 A/B/S.

Furfural product from furfural product cooler TY-507 is pumped to furfural product tanks MT-602A/B. Furfural is transported to the shipping point by furfural product pumps PP-602 A/B/S.

Fusel oil product is pumped by fusel oil pump PP-405A/S to fusel oil storage tank MT-605. Fusel oil is transported to the shipping point by fusel oil export pump PP-605A/S.

Slops from the collection header in the acid hydrolysis unit are stored in slops tank MT-608 and transferred on flow control by slops transfer pump PP-608 back to sugar stripper feed tank MT-222. Slops are off-spec material collected after shutdown or upset conditions and are to be gradually reworked through the system.

Refined hydrolyzate from hydrolyzate cooler TT-206 is pumped to fermentation feed tank MT-604 at 122°C, a temperature selected to inhibit contamination by unwanted microbial growth. Air vent filter GF-604 mounted on the vent of MT-604 also prevents contamination from the atmosphere. Hydrolyzate is transported to Number 1 Fermenters MR-301A/B by fermentation feed pump PP-604A/S via fermentation feed cooler TT-604 which cools the hydrolyzate to 84°F using cooling water.

Aqueous furfural from MS-209 is transported by aqueous furfural pump PP-218A/S to azeo column surge tank MT-606. Aqueous furfural is then fed to azeo column preheater TY-501 by azeo column feed pump PP-606A/S.

Concentrated sulfuric acid delivered by truck or railcar is stored in sulfuric acid storage tank MT-603. Moisture from the atmosphere is kept from the acid by dessicant air filter GF-603 mounted on the vent of MT-603. Sulfuric acid transfer pump PP-603A/S transports acid to sulfuric acid day tank MT-221.

Diesel fuel delivered by truck is stored in diesel fuel tank MT-609 which has a one month capacity. Fuel is delivered to mobile equipment by diesel fuel pump PP-609A/S.

Fire water tank MT-607 and fire water pump PP-607A/S are remotely located from the process units and offsite tankage.

Unit 700 - Waste Treatment

The environmental unit consists of three parts; an evaporation section, an anaerobic digestion and aerobic treatment section, and a low pressure vent system.

In the evaporation section, a three-stage evaporator is used to produce mixed solvents at 60% concentration in water to be burned as liquid fuels to produce steam. About half of the recovered condensate is recycled to the acid hydrolysis unit and about half is sent to anaerobic digestion.

Beer column bottoms pump PP-401A/S transports stillage to first effect evaporator GE-701 running at 14 psig. Heat for the first effect is provided by first effect heater TT-701 using 50 psig steam. Vapors from the first effect are condensed in second effect heater TT-702 which provides heat for second effect evaporator GE-702 running at 11 psia. Liquid is pumped from the first effect to the second effect on level control by first effect pump PP-701A/S. Similarly, second effect pump PP-702A/S feeds third effect evaporator GE-703 on level control and vapors from the second effect condense and provide heat for the third effect in third effect heater TT-703. Vapors from the third effect are condensed in evaporator condenser TT-704 using cooling water.

The pressure in the third effect is controlled at 3 psia by evaporator vacuum system PE-704 using 50 psig steam as the motive source and a pressure control bypass which allows steam back to the inlet to TT-704.

Liquid from the third effect, containing a mixture of about sixty percent organics and forty percent water, is pumped on level control to rotary filter GF-704 where precipitated salts are separated and sent to disposal. Solid-free liquid is then sent to mixed solvent tank MT-702 for storage.

Boiler feed water demineralizer regeneration effluent from GU-803A/B and boiler water blowdown from MS-802 are also treated by evaporation.

Condensate from the evaporators is collected in Hotwell Decanter MS-701 where liquid phase separation can occur. The organic layer overflows a baffle and is pumped on level control by solvent pump PP-706 A/S to solvent cooler TP-706, cooled by cooling water, and sent to MT-702. Waste water is pumped on interface level control by evaporator waste water pump PP-705 A/S to stillage cooler TT-705, cooled by cooling water. The temperature is controlled using a bypass line around TT-705. About half of the waste water is recycled to the acid hydrolysis unit as process water and the remaining waste water is sent to anaerobic treatment.

MT-702 contains mixed solvents collected from MS-502, MS-701, and GF-704. Solvent is transported by mixed solvent pump PP-707A/S to boilers HB-801A/B where the solvent is burned to produce high pressure steam.

In the anaerobic treatment section, the Celrobic process is used. Waste water (evaporator condensate from TT-705, flume washings from PP-101, and molecular sieve runoff from GZ-401) is collected in equalization tank MT-703 and kept blended by equalization tank eductors PE-711. Reactor feed pumps PP-708A/S pump waste water on flow control to reactor feed heater GU-705 which directly injects low pressure steam on temperature control. After mixing with recycled reactor effluent, the waste water is fed to anaerobic reactor MT-704. Ninety percent of the process effluent COD is removed in this high rate anaerobic digester. The digester produces methane gas which passes through off-gas suction knock-out pot MS-706 to off-gas blower PB-710. Off-gas is compressed and sent to off-gas knock-out drum MS-709. A split range pressure controller maintains the reactor outlet pressure by recycling compressed off-gas via off-gas cooler TT-701, cooled by cooling water, or making up with nitrogen. Excess system pressure will open a back pressure control valve allowing off-gas to pass into off-gas burner GQ-706 where it is flared.

Liquid effluent from MT-704 overflows into reactor surge drum MS-705 and is pumped on level control by reactor recycle pump PP-709 to biotreater MT-707. A portion is recycled to the reactor inlet.

The remaining COD is removed in the aerobic treatment section. Effluent from anaerobic treatment is combined with recycled underflow from the secondary clarifier and filtrate from the filtrate sump and fed to biotreater MT-707. The biotreater is aerated by aeration blowers PB-712A/S. Effluent from the biotreater is combined with sand filter backwash and enters secondary clarifier GV-708. Overflow from the clarifier is used as cooling tower makeup and is pumped on level control by final effluent pumps PP-717A/S.

A portion of the clarifier underflow is recycled on flow control as feed to the biotreater and the remainder is pumped on flow control by sludge pumps PD-713A/S to filter feed tank MS-708. Filter feed pump PD-714A/S mixes sludge with polymer mixer GA-711 and belt filter press separates solids for disposal. Filtrate is collected in filtrate sump MF-709 and recycled to the biotreater by filtrate sump pump PP-715.

Low pressure vents from all units are drawn into low pressure vent knock-out drum MS-710 to remove entrained liquids and transported by low pressure vent blower PB-717A/S to the firebox of boilers HB-801A/B where it is mixed with combustion air. The suction pressure of PB-717A/S is controlled using a pressure control bypass line.

#### Low Pressure Vents

Most of the low pressure process vents contain air or carbon dioxide saturated with water and small amounts of volatile organics. The air comes both from displacement during filling of vessels and from release of air trapped in pores in the wood. The organics are mostly ethanol and the extraneous material from wood such as terpenes and extractive resins. In our process modeling we characterized terpenes as alpha pinene and resins as fatty acids, specifically oleic acid. These would present odor problems if vented directly to the atmosphere. Scrubbing with water would be inadequate since alpha pinene and oleic acid are insoluble. Recovering such small amounts of organics by chilling and condensing is uneconomical. Therefore, it was decided to incinerate the vents by mixing them with combustion air in the boilers.

### Unit 800 - Utilities

The utilities unit consists of the boilers and steam distribution system, the boiler feedwater system, the turbogenerator system, the cooling water system, the process water system, and the plant and instrument air system.

#### Boilers and Steam Distribution

The boilers HB-801A/B are designed to burn solid, liquid, and gaseous fuel byproducts from the process units in order to generate 1100 psia steam (900 psia, Alternative Case) for use in the process and for electrical power generation. Process requirements amount to 78% of the total available steam. Solid fuels comprise about 53% of total byproduct fuel heating value. Liquid fuel, comprising about 46% of total fuel heating value, is a mixture of solvents, sugar, acids, dissolved organic salts, and soluble tars in a 36% water mixture. Fuel gas comes from anaerobic digestion and comprises less than 1% of total heating value.

Gas and liquid fuels are burned directly but lignin and tar sludge from the acid hydrolysis unit is fed to a Flakt type drying system which dries and fluidizes the solids into the burners using boiler flue gas. No other solid fuel (such as coal) will be utilized as there is always an excess of byproduct fuel available over normal process unit energy requirements. Diesel fuel is used for startup.

Low pressure vents are collected, compressed, and injected into the boilers to mix with combustion air prior to burning. None of these vents contain an explosive mixture either singly or in combinations. They contain mostly air or carbon dioxide saturated with water and with very small amounts of organics.

Process users of 1100 psia steam are fed directly from the high pressure steam header. Users of 150 and 50 psig steam are supplied with steam letdown from 1100 psia through a turbogenerator. All recoverable condensate is returned to the boiler feedwater system.

#### Boiler Feedwater System

Boiler feedwater preparation consists of makeup process water demineralization, condensate polishing, deaeration, and chemical dosing.

Recoverable condensate is collected in condensate collection system GU-805 (package unit). Flashed low pressure steam is used in the deaerator. Condensate is then pumped through condensate polisher GU-804A/S to condensate surge drum MS-804. Makeup process water is introduced on level control via demineralizers GU-803A/B, a package unit consisting of anion and cation resin beds with intermediate degasifier and regeneration equipment. Deaerator GV-806 is fed by deaerator feed pump PP-809A/S on flow control reset by level control. The deaerator runs at 10 psig and expels air and steam to the atmosphere. Steam to the deaerator is fed on flow control and is supplied by flashes from recovered condensate and from boiler blowdown flash drum MS-802 and made up from the 50 psig steam header. Boiler blowdown is sent to evaporator GE-701 by blowdown pump PP-806A/S.

Demineralizer regenerant effluent is sent to evaporator GE-701.

Deaerated boiler feedwater is dosed with hydrazine and ammonia made up in hydrazine addition unit GU-807 and ammonia unit GU-808. BFW pump PP-808A/S feeds boilers HB-801A/B. The boiler steam drums are dosed with phosphate made up in phosphate addition unit GU-809. A shed encloses the three chemical additions units. Hydrazine is transferred from hydrazine drum MS-803 by hydrazine transfer pump PP-807 to be mixed with condensate in GU-807. Liquid ammonia from rented tanks is mixed with condensate in GU-808. Phosphate dumped from bags is mixed with condensate in GU-809. BFW chemicals are introduced into the boiler system by variable flow metering pumps.

#### Turbogenerator

Turbogenerator package unit GZ-811 is rated to take 1100 psia, 825°F steam and generate 45 mw of electricity. The unit includes a condenser, vacuum ejector set, and controls. 4% of the steam is let down in a topping cycle to 150 psig, 66% is let down to 50 psig, and the remainder is condensed at 89 mm Hg(A). Turbine condensate pump PP-810A/S returns condensate to the boiler feedwater system.

NOTE: The Alternative Case uses 900 psia steam.

### Cooling Water System

Cooling tower GT-812 provides cooling water at 88°F with an allowable temperature rise of 30°F. Cooling water distribution is maintained at a pressure of 60 psig by cooling water pumps PP-812A-H. Cooling tower blowdown containing dissolved salts at 2000 ppm is discharged directly to the environment. Cooling tower blowdown contains zero COD.

### Process Water System

Process water preparation consists of filtering well water pumped by well water pumps PP-813 A/B/S through a sand filter and anthracite bed GF-801. Process water transfer pump PP-802A/S feeds process water tank MT-801. Backwash feed pumps PP-804A/B provides a daily backwash flow to GF-801. Backwash overflow is collected in backwash transfer tank MF-805 and is transported by backwash transfer pumps PP-805A/S to clarifier GV-708. Process water is distributed via a process water ring main pressurized to 60 psig by process water circulating pump PP-803A/S.

### Plant and Instrument Air Systems

Plant and instrument air are provided by air compressor PC-811 operating on pressure control from plant air receiver MS-806. Instrument air receiver MS-807 is fed via instrument air dryer GY-810.

## ECONOMIC ASSESSMENT

### Introduction

This section presents the economic analysis of the process for fuel ethanol manufacture from wood. The four design cases investigated are summarized in Table 2.1. Two ethanol selling prices are calculated for each design case using two different sets of economic assessment parameters: One set requested by SERI and one set normally used by Badger.

The section covers the following:

- o Raw material, and byproduct price analysis.
- o Discussion of the economic assessment parameters used in the calculations.
- o Presentation of the estimated capital costs for the four design cases.
- o Presentation of the operating costs for each of the four design cases.
- o Presentation of the ethanol selling prices for each design case calculated using the two different sets of parameters.

- o Summary tables for revenues and costs of sales for each design case using both sets of parameters.
- o Sensitivity analyses showing ethanol price variation with feedstock cost, capital cost and furfural byproduct price for design cases I and II for both sets of parameters.
- o Sensitivity analysis showing the effect of the major economic assessment parameters on the internal rate of return (IRR) for design case I.

Feedstock, Byproduct and Labor Costs

Wood Chip Price

A woodchip price of \$42 per dry metric ton has been taken as the basis for economic calculations. This figure is based on a study commissioned by Burlington Electric Company, from discussions with two chip supply companies, and from published reports (References 14 to 29). The prices quoted have been converted into equivalent 1984 prices, assuming an average inflation rate of 7% per annum, and are listed in Table 5.1.

Table 5.1 Wood Feedstock Price Data

<u>Source</u>	<u>Price Quoted \$/Dry Metric Ton</u>	<u>Date of Quotation</u>	<u>Equivalent Mid-1984 Price</u>
Burlington Electric	32	1980	42
Chip Broker No. 1	40*	Nov 1983	41
Chip Broker No. 2	37.5*	Nov 1983	39
Reference 13	44**	Apr 1984	44
Reference 16	40	May 1984	40
Reference 17	37.5	Nov 1982	41
Reference 28	30.5	Mar 1981	37.5

\*Debarked

\*\*Aspen Wood

The major components of the wood chip price are given in Table 5.2 (Reference 14):

Table 5.2 Components of wood Feedstock Price

<u>Component</u>	<u>Approximate %</u>
Harvest Equipment Fixed Costs	40
Harvest Labor	25
Stumpage	7
Transportation (Up to 50 Miles)	25
Miscellaneous Costs	3
	<u>100</u>

The experience of the Burlington Electric Company in gathering wood chips for their McNeil Generating Station suggests that in order to maintain a stable group of wood supply subcontractors, it is necessary to pay a price for wood chips FOB the plant gate such that the subcontractor not only recovers his cost but also makes a reasonable profit. A price of \$42 per ton is believed to be one that, for 1984, will meet these criteria and ensure the stable supply of wood chips required to feed a 25 million gallon per year ethanol from wood facility (Design Case I).

Although the \$42 per ton wood chip price has also been used for the 5 million gallon per year case in the economic analysis which follows, the cost of wood for such a facility could be lower since it is possible that the smaller wood supply requirements of the 5 million gallon per year case could be supplied from mill waste or other scrap wood source. Finding such a source of waste wood would improve the economics since the economics are sensitive to feedstock cost as shown in Table 5.10.

Alternatively, a reduced capacity plant could, with minor modifications, be fed from the residue of a local agricultural operation. These residues include sugar cane bagasse, corn stover, and corn cobs. A further source of feedstock which could be utilized might be cellulosic domestic refuse. Feedstock obtained under such local site-specific conditions would be expected to be available for \$30 to \$35 per dry metric ton on average.

The sensitivity of ethanol selling price to feedstock cost has been investigated for the 25 million and 5 million gallon plants. These results are reported later in this section.

#### Furfural Price

Byproduct furfural has been priced at 55¢ per pound for the short range market (0-5 years) and 15¢ per pound for the long range market (10-20 years). It is anticipated that in the early stages of introduction of this technology, i.e. demonstration and first generation commercial operation, that the quantities of furfural produced will not greatly affect the market price. For example, a first commercial size plant production 5 million gallons ethanol per year would produce 26 million pounds per year of furfural or around 13% of the existing U.S. market.

The present published market price of furfural is 66¢ per pound (Reference 20) and two or three such plants producing furfural at a selling price of 55¢ per pound would be expected to displace a portion of the current U.S. production (Reference 10). Domestic furfural production has been on the decline in recent years as a result of uneconomic production methods.

In the longer term (10 to 20 years), since furfural is an extremely versatile intermediate chemical, an increase in its use would be expected in response to lowering of its price. Existing commodity chemicals adipic acid, maleic anhydride, tetrahydrofuran, and butanol currently manufactured using petrochemical routes can be produced from furfural using known technology (Reference 9, 10, and 11). A quantitative analysis of the potential furfural market is given in Section 6.0.

Based on this analysis a furfural price of 15¢ per pound has been taken as the long term market price. At this price, the analysis predicts a market demand of 2.2 billion pounds per year (References 9 and 10). The sensitivity of the ethanol selling price to furfural price is calculated for the 25 million gallon plant.

#### Electricity Selling Price

The co-generated export electricity selling price has been taken as 5¢ per kilowatt hour for the purposes of the study. Electric power has a guaranteed market under the provisions of the PURPA Act (Reference 21) provided generating capacity does not exceed 80 MW. The price is negotiable and varies considerably from state to state. It is based on the "avoided cost" of electricity production by the local utility. Prices quoted range from 5¢ to 10¢ per kilowatt hour depending on local utility costs as well as credits available in some states for biomass derived energy (Reference 22 and 23). Generating capacity for the 25 million gallon scale plant is 44.5 MW.

### Ethanol Denaturant

Gasoline used as ethanol denaturant is sold along with the ethanol as "denatured ethanol" (a blend of 5% by volume gasoline and 95% ethanol). A marginal benefit to the economics is gained by purchasing the gasoline (87 octane) at the wholesale prices of 85¢ per gallon and selling it at the calculated ethanol price. (This presumes, of course, that the calculated denatured ethanol price is above 85¢ per gallon.)

### Fusel Oils

No revenue has been assumed for fusel oils from either direct sales or burning for power generation. In the process, fusel oil is exported.

### Landfill

The cost of disposal of gypsum, ash and other separated salts is taken at a typical value of \$5.5 per metric ton.

### Other Raw Material Prices

The following sources have been used for other raw material prices:

- Chemical Marketing Reporter
- Oil & Gas Journal (Gasoline)
- Badger In-house Prices (Celrobic)
- Vendor Information (Molecular Sieve Packing)

### Labor Rates

Operating labor rates are based on those published by the U.S. Department of Commerce, Bureau of Economic Analysis.

Assumptions Underlying the Economic Analysis

Badger has conducted an economic analysis of a conceptual acid hydrolysis based ethanol plant using two different sets of basic assumptions or parameters. One set is provided by SERI to give a common basis for comparing this analysis with others. The second set Badger normally uses in this type of study to avoid influencing the outcome with difficult to predict factors. The major differences between the two sets of parameters are discussed in the following pages and summarized in Table 5.3.

Table 5.3 Parameters for Parallel Economic Analysis

	<u>SERI</u>	<u>BADGER</u>
Capacity Factor (Hours/Year)	8000	8000
Inflation Rate (%)	7	Zero
Base Year for Dollars	1984	1984
Year of Cost Information	1984	1984
Construction Start Date	F/Q 1985	F/Q 1985
Investment Drawdown	Monthly (Straight Line)	Monthly (Straight Line)
First Commercial Operation	Mid-1987	Mid-1987
Operating Lifetime (Years)	15	15
Tax Depreciation Schedule	ACRS*	ACRS*
Taxes and Insurance (% of Investment)	2 Total	2 Total
Income Tax Rate (%)	45	45
Investment Tax Credit (%)	10	10
Energy Tax Credit (%)	Zero	10
Multiple or Single Venture Company (%)	Single	Multiple
Ratio of Debt to Capitalization (%)	30	Zero
Annual Rate of Return on Debt (%)	13	--
Annual Rate of Return on Equity (%)	15 (25)** (After Tax)	15 (After Tax)
Escalation Rate on Capital (%)	1	Zero
Escalation Rate on Operations (%)	1.5	Zero
Operating Capacity Schedule	100% After Startup	50% First 12 Months 80% 12-24 Months 100% Thereafter
Salvage Value	Zero	10% (Of Investment)

\*ACRS - Accelerated Cost Recovery System allowed by law over 5 years.  
Yearly percent write-offs are: 15, 22, 21, 21, 21 (Reference 24).

\*\*Ethanol selling price at 25% IRR as well as at 15% on Equity, is calculated using SERI parameters.

### Discounted Cash Flow Internal Rate of Return (IRR)

A IRR of 15% on equity has been requested by SERI. This figure may well be acceptable to companies in the long term once the technology is proven and established. It has therefore been used in calculating an ethanol selling price for both the 25 and 5 million gallon plants. However, in the nearer term it is believed that this IRR may not be sufficient to encourage companies to utilize the fuel ethanol from wood technology. Consequently, the 5 million gallon plant economics calculated using SERI parameters have also been presented using a 25% IRR.

The base year for discounting cash flows in the IRR calculation is the start of construction year (1985).

### Debt to Equity Ratio

The SERI based economic calculations are based on 30% leveraged debt, 70% equity participation. This represents a "special case" since normally economic analysis at this stage would be carried out using 100% equity financing. The parallel Badger analysis uses 100% equity.

In economic analyses involving leveraged debt capital, interest payable during the construction phase has been capitalized.

### Investment Tax Credit

An investment tax credit of 10% of total investment has been used in both the SERI and Badger based analysis. Energy tax credits (10% of total investment) normally available to this type of project have not been included in the SERI based analysis, but have been assumed for the Badger analysis (Reference 29).

Removing 10% Energy Tax Credit has the effect of increasing the ethanol selling price by ¢ 15 per gallon to \$1.95 for the base case plant.

#### Single Versus Multiple Venture Company

Whether or not the company operating the plant is a single or multiple venture company affects the economic calculations. For multiple venture companies, tax credits and operating loss tax write-offs can be taken as soon as they are available, but with single venture companies such write-offs must be deferred until they can be written-off against an operating profit in later years. The assumption of a single venture company is, of course, the more conservative since there is an economic benefit to being able to take early advantage of available tax allowances. In the SERI based analysis a single venture company has been assumed, and in the Badger analysis a multiple venture company has been assumed.

#### Operating Capacity Schedule

An operating capacity schedule of 100% of design immediately after startup has been specified by SERI. This seldom occurs in actual practice and, consequently, Badger normally uses a more conservative basis that assumes the plant will operate at 50% of design for the first 12 months, 80% of design capacity for the next year (months 12-24), and at 100% of design capacity thereafter.

### Inflation and Escalation

SERI has requested that the economic analysis be based upon an overall inflation rate of 7% per year. In addition, an escalation rate of 1% on capital costs and 1.5% on operating costs has also been specified. This in effect gives an inflation rate on capital costs of 8% and operating costs of 8.5%. The effect of inflation is to raise the calculated IRR roughly in proportion to the yearly inflation rate. The effect of inflation rate has been calculated for the base case economics and is shown in Figure 5.10.

### Working Capital

Working capital required has been estimated by the following method:

- 30 days accounts receivable
- +Cash (10% accounts payable)
- +15 days inventories
- +30 days maintenance and spare parts
- 30 days wages
- 30 days raw materials
- 30 days maintenance and labor
- 90 days taxes

The effect of working capital on the IRR is shown in Figure 5.10.

Distribution Expenses

Distribution expenses are a site-specific item and depends on specific customer bases not included in the present calculation. Production costs have therefore been calculated FOB plant.

A full description of the complete Badger method of economic assessment is given in Appendix VI.

Capital Costs

Estimated capital costs for the four design cases have been prepared and are presented in Table 5.4.

Table 5.4 Capital Costs of the Four Design Cases

	<u>Design Case</u>	<u>Capital Cost \$MM</u>
I	25 Million Gallons (Base)	139.78
II	25 Million Gallons (Alternative)	134.88
III	5 Million Gallons	69.76
IV	5 Million Gallons (Purchased Power)	58.13

The estimates of capital cost in Table 5.4 have been developed as is described in Appendix VII.

The components of the capital cost for each design case is given below in Table 5.5.

Table 5.5 Components of Capital Cost for Each Design Case

Design Case	I (Base)	II (Alternative)	III (5 MM Gal)	IV (5 MM Gal Purchased Power)
Unit 100 (Feed Handling)	0.97	0.98	0.52	0.53
Unit 200 (Hydrolysis)	21.74	23.35	9.36	9.66
Unit 300 (Fermentation)	5.38	5.44	2.91	3.00
Unit 400 (Ethanol Purification)	9.22	9.31	4.09	4.22
Unit 500 (Furfural Recovery)	2.58	2.61	1.28	1.32
Unit 600 (Tankage)	6.36	6.43	3.83	3.95
Unit 700 (Environmental)	8.39	8.48	4.14	4.27
Unit 800 (Utilities)	48.76	42.70	22.87	13.25
<b>Total (USGC)*</b>	<b>103.4</b>	<b>99.30</b>	<b>49.00</b>	<b>40.20</b>
Overhead at 50% Plus G/A at 15% of Engineering Costs	5.82	5.94	4.06	4.02
Construction Taxes	1.77	1.70	0.80	0.66
Spare Parts	2.10	1.94	0.80	0.65
Startup Fees	1.00	1.00	0.70	0.70
Transfer Plant to Typical Mid-west Location	17.69	17.00	8.40	6.60
Contingency**	8.00 (6%)	8.00	6.00 (10%)	5.30 (10%)
<b>Total</b>	<b>139.80</b>	<b>134.90</b>	<b>69.30</b>	<b>58.10</b>

\*U.S. Gulf Coast estimate.

\*\*See discussion on contingency in Section 6.0

The cost of land is included separately in the economic analysis and does not appear in the preceding figures.

The base case capital cost is \$139.78 million. This is based on a U.S. Gulf Coast cost estimate of \$103.4 million. The final capital cost of \$139.8 million accounts for engineering overhead plus general and administrative costs, construction taxes, spares, startup fees, extra costs associated with a mid-west location and a contingency based on the estimating method which corrects the base estimate to the "most likely" figure.

The capital cost of the alternative design case for the large scale plant (two stage hydrolysis) is \$134.88 million. The difference between this and the base case is due mainly to reduction in the boiler and turbogenerator capacity. As can be seen from the capital cost component analysis the two stage hydrolysis section, as expected, is slightly more expensive than in the base case.

The reduced capacity plant for 5 million gallon per year ethanol costs \$69.76 million. This plant is designed on the same basis as for the base case 25 million gallon plant except that the hydrolysis unit in the smaller plant is single train.

The 5 million gallon scale plant with purchased electrical power costs \$58.13 million. The reduction in cost of \$8.8 million being due to elimination of the turbogenerator and slight reduction in the size of the boilers.

The above capital costs have been generated for a typical mid-west location. For further discussion of the geographical location see Section 2.0.

The two major components of capital cost are the hydrolysis unit (21%) and the utilities section (47%). The high capital cost of the hydrolysis section reflects what is believed to be a conservative design considering the present state of development of the process. (Further consideration of this aspect of technical risk is given in Section 6.0.)

By far the greatest component of capital cost is that of the plant utilities section. This is dominated by the cost of the boiler and turbogenerator. For this reason, quotations from a vendor for both these items were obtained and considerable effort was expended on defining these costs. However, it is believed that there may be significant cost savings to be made in this area.

The high capital cost of the boiler system is primarily due to the need to burn lignin, waste solvents and digester off-gas in the same unit. There is also the additional cost of the Flakt lignin drier in the boiler system.

The boiler cost is further increased by the specification of two parallel boilers each producing 60% of the total steam (70% of process needs) and each sized to burn all of the lignin/tar residue.

Economic analysis discussed later in this section show that it is cost effective to include a turbogenerator to generate power onsite, even for the reduced capacity plant. For the 25 million gallon plant onsite power generation has an even stronger justification.

Operating Costs

The following pages give the operating requirements and production figures for the four design cases. Costs are for base year 1984. The production rate of denatured ethanol (27.5 million gallons per year) is composed of the basic design production rate of fuel grade ethanol 26.2 million gallons per year in addition to 87 octane gasoline 1.3 million gallons per year (blended at 5% by volume) as a denaturant. Yearly revenues for a 15% IRR are given for both the SERI and the Badger calculations.

Design Case I  
(Base Case)

Costs

Raw Materials

<u>Item</u>	<u>Units</u>	<u>Usage Units/Yr</u>	<u>Cost \$/Unit</u>	<u>Cost \$/Yr</u>
Wood Chips	Metric Tons	590,400	42	24,797,000 ✓
Sulfuric Acid	Metric Tons	18,155	65	1,180,000 ✓
Calcium Hydroxide	Metric Tons	18,383	35	643,000 ✓
Yeast and IMY Chemicals				136,000
Diesel	U.S. Gallons	10,000	1	10,000
Nitrogen				12,000
Boiler Water Chemicals				17,000
Molecular Sieve Packing				85,000
Cooling Water Chemicals				5,000
Anaerobic Digester Chemicals				355,000
Gasoline (Denaturant)	U.S. Gallons	1.31x10 <sup>6</sup>	0.85	1,114,000 ✓
<b>TOTAL</b>				<b>28,354,000</b>

<u>Labor</u>	<u>Number</u>	<u>Cost \$/Yr</u>
Operators	60	1,280,000
Foreman	8	200,000
Supervisors	2	60,000
<b>TOTAL</b>		<b>1,540,000</b>
Direct Overhead (40% of Labor)		616,000

**TOTAL LABOR** 2,156,000

<u>Maintenance</u>	(4% of Total Investment)	5,592,000
<u>General Overhead</u>	(60% of Labor and Maintenance)	4,648,000
<u>Royalties</u>		250,000
<u>Taxes and Insurance</u>	(1% of Investment each)	2,796,000
<u>Landfill Disposal</u>	(\$5.5 per Tonne)	348,000

**TOTAL OPERATING COST** 44,144,000

<u>Byproduct Credits</u>	<u>Units</u>	<u>Units/Year</u>	<u>\$/Unit</u>	
Furfural	Pounds	130.2 MM	0.15	19,530,000
Electricity	Megawatts	176,000	50	8,800,000
<b>TOTAL</b>				<b>28,330,000</b>

Design Case II  
(Alternate Case)

Costs

Raw Materials

<u>Item</u>	<u>Units</u>	<u>Usage Units/Yr</u>	<u>Cost \$/Unit</u>	<u>Cost \$/Yr</u>
Wood Chips	Metric Tons	528,000	42	22,176,000
Sulfuric Acid	Metric Tons	25,919	65	1,685,000
Calcium Hydroxide	Metric Tons	22,509	35	788,000
Yeast and IMY Chemicals				136,000
Diesel	U.S. Gallons	10,000	1	10,000
Nitrogen				14,000
Boiler Water Chemicals				11,000
Molecular Sieve Packing				85,000
Cooling Water Chemicals				5,000
Anaerobic Digester Chemicals				300,000
Gasoline (Denaturant)	U.S. Gallons	1.31x10 <sup>6</sup>	0.85	1,114,000
<b>TOTAL</b>				<b>26,324,000</b>

<u>Labor</u>	<u>Number</u>	<u>Cost \$/Yr</u>
Operators	60	1,280,000
Foreman	8	200,000
Supervisors	2	60,000
<b>TOTAL</b>		<b>1,540,000</b>

Direct Overhead (40% of Labor) 616,000

**TOTAL LABOR** 2,156,000

<u>Maintenance</u>	(4% of Total Investment)	5,395,000
<u>General Overhead</u>	(60% of Labor and Maintenance)	4,531,000
<u>Royalties</u>		250,000
<u>Taxes and Insurance</u>	(1% of Investment each)	2,698,000
<u>Landfill Disposal</u>	(\$5.5 per Tonne)	350,000

**TOTAL OPERATING COST** 41,704,000

<u>Byproduct Credits</u>	<u>Units</u>	<u>Units/Year</u>	<u>\$/Unit</u>	
Furfural	Pounds	93.16 MM	0.15	13,974,000
Electricity	Megawatts	-	50.	-
			<b>TOTAL</b>	<b>13,974,000</b>

Design Case III  
(5 Million Gallon - Onsite Electricity Generation)

Costs

Raw Materials

<u>Item</u>	<u>Units</u>	<u>Usage Units/Yr</u>	<u>Cost \$/Unit</u>	<u>Cost \$/Yr</u>
Wood Chips	Metric Tons	118,080	42	4,959,000
Sulfuric Acid	Metric Tons	3,631	65	236,000
Calcium Hydroxide	Metric Tons	3,677	35	129,000
Yeast and IMY Chemicals				272,000
Diesel	U.S. Gallons	2,000	1	2,000
Nitrogen				2,000
Boiler Water Chemicals				3,000
Molecular Sieve Packing				17,000
Cooling Water Chemicals				1,000
Anaerobic Digester Chemicals				71,000
Gasoline (Denaturant)	U.S. Gallons	262,000	0.85	223,000
<b>TOTAL</b>				<b>5,915,000</b>

Labor

	<u>Number</u>	<u>Cost \$/Yr</u>
Operators	40	853,000
Foreman	8	200,000
Supervisors	2	60,000
<b>TOTAL</b>		<b>1,113,000</b>
Direct Overhead (40% of Labor)		445,000

**TOTAL LABOR**

1,558,000

Maintenance

(4% of Total Investment)

2,790,000

General Overhead

(60% of Labor and Maintenance)

2,609,000

Royalties

250,000

Taxes and Insurance

(1% of Investment each)

1,395,000

Landfill Disposal

(\$5.5 per Tonne)

70,000

**TOTAL OPERATING COST**

**14,587,000**

Byproduct Credits

	<u>Units</u>	<u>Units/Year</u>	<u>\$/Unit</u>	<u>Credit</u>
Furfural	Pounds	26.04 MM	0.55	14,322,000
Electricity	Megawatts	35,200	50.0	1,760,000
<b>TOTAL</b>				<b>16,082,000</b>

Design Case IV  
(Purchased Electricity)

Costs

Raw Materials

<u>Item</u>	<u>Units</u>	<u>Usage Units/Yr</u>	<u>Cost \$/Unit</u>	<u>Cost \$/Yr</u>
Wood Chips	Metric Tons	118,080	42	4,959,000
Sulfuric Acid	Metric Tons	3,631	65	236,000
Calcium Hydroxide	Metric Tons	3,677	35	129,000
Yeast and IMY Chemicals				272,000
Diesel	U.S. Gallons	2,000	1	2,000
Nitrogen				2,000
Boiler Water Chemicals				3,000
Molecular Sieve Packing				17,000
Cooling Water Chemicals				1,000
Anaerobic Digester Chemicals				71,000
Gasoline (Denaturant)	U.S. Gallons	262,000	0.85	223,000
<b>TOTAL</b>				<b>5,915,000</b>
<u>Purchased Electricity</u>	Megawatts	32,800		1,968,000

Labor

	<u>Number</u>	<u>Cost \$/Yr</u>
Operators	40	853,000
Foreman	8	200,000
Supervisors	2	60,000
<b>TOTAL</b>		<b>1,113,000</b>
Direct Overhead (40% of Labor)		445,000

**TOTAL LABOR**

1,558,000

Maintenance

(4% of Total Investment)

2,325,000

General Overhead

(60% of Labor and Maintenance)

2,330,000

Royalties

250,000

Taxes and Insurance

(1% of Investment each)

1,163,000

Landfill Disposal

(\$5.5 per Tonne)

70,000

**TOTAL OPERATING COST**

**15,580,000**

Byproduct Credits

	<u>Units</u>	<u>Units/Year</u>	<u>\$/Unit</u>	
Furfural	Pounds	26.04 MM	0.55	14,322,000
Electricity	Megawatts	-	50	-
			<b>TOTAL</b>	<b>14,322,000</b>

Calculated Ethanol Selling Price

The calculated ethanol selling price for each design case using the SERI and the Badger economic parameters is shown in Table 5.6 below. (Details of the calculations are given in Appendix VI.)

Table 5.6 Selling Price of Denatured Ethanol for the Four Design Cases

Design Case	Ethanol Selling Price \$ Per Gallon		
	S.E.R.I Parameters		Badger Parameters
	15% IRR	25% IRR*	15% IRR
I 25 Million Gallon (Base)**	1.23		1.80
II 25 Million Gallon (Alternative)	1.65		2.23
III 5 Million Gallon***	1.42	2.15	2.83
IV 5 Million Gallon (Purchased Power)	1.63		2.85

\*For rationale behind using 25% DCFIROR for 5 million gallon per year cases, see page 9, Section 5.

\*\*Furfural priced at ¢ 15/lb.

\*\*\*Furfural priced at ¢ 55/lb.

The calculated ethanol selling price using SERI supplied economic assessment parameters is \$1.23 per gallon for the base case plant. This is the price required to give a 15% IRR (after tax). The plant capital cost is \$139.78 million, furfural price is 15¢ per pound, electricity 5¢ per kwh and feedstock price is \$42 per dry metric ton. The parallel calculation using Badger economic assessment parameters gives a selling price of \$1.80 per gallon required to give a 15% IRR after tax under the same conditions of capital and feedstock cost and byproduct revenues.

Alternative case (Design Case II) at a capital cost of \$134.88 million gives an ethanol selling price of \$1.65 (SERI) and \$2.23 (Badger).

The reduced capacity 5 million gallon plant (Design Case III) at a capital cost of \$69.78 million requires a selling price of \$1.42 per gallon of ethanol to give a 15% IRR and a selling price of \$2.15 per gallon to give a 25% IRR (both figures based on SERI calculation). The parallel Badger calculation shows a required selling price of \$2.83 per gallon to give a 15% IRR.

For Design Case IV (5 million gallon plant with purchased electric power) at a capital cost of \$58.13 million the required ethanol selling price to give a 15% IRR is \$1.63 using the SERI parameters and \$2.85 with Badger parameters.

Inspection of Table 5.6 shows that Design Case I (single stage hydrolysis) is more attractive than Design Case II (two stage hydrolysis reaction) despite the fact that Design Case II has a lower capital cost and higher ethanol yield on wood. This is due to increased revenues, in Design Case I, from furfural and exported electricity.

For the 5 million gallon scale plant using SERI parameters it is clearly cost effective to generate electric power onsite rather than to purchase it. Onsite generation gives a price advantage of 22¢ per gallon for the SERI calculation. For the Badger calculation, the decision is not so clear, a marginal advantage 2¢ being gained by onsite generation.

There is a considerable difference between the selling price based on the SERI economic parameters and that based on the Badger parameters. This is primarily due to removing the effects of inflation, and leveraged debt capital and lowering the operating capacity schedule. This indicates that at the present state of the technology a "specially" leveraged financing arrangement such as that specified in the SERI parameters would almost certainly be necessary for the first commercial plants.

There is also a considerable difference between the assumption of 15% IRR and 25% IRR. This suggests that to encourage initial introduction of the technology on the commercial-scale a subsidy of some type will be required. On the longer term, once the technology is established, it should be possible to reduce or eliminate these subsidies.

Summary of Annual Revenues and Costs

Based on the foregoing operating requirements and production figures, yearly costs and revenues have been calculated for the plant operation for each design case.

Two sets of figures are given: one based on 100% capacity factor (SERI), and the other based on the capacity factor schedule assumed in the Badger calculation.

The price shown for ethanol in each case is that calculated to give a 15% IRR after tax. A summary of the calculated prices is given on page 22 of this section.

### Ethanol Selling Price Sensitivity Analysis

The ethanol selling price sensitivity to wood feedstock price, plant capital costs was calculated for both the base case plant (Design Case I), and the 5 million gallon plant (Design Case III) and sensitivity to furfural price for Design Case I, using both sets of economic parameters. The results are shown Tables 5.7 to 5.10 and in graphical form in Figures 5.1 to 5.9.

The purpose of the sensitivity analysis was threefold: firstly, to investigate the effects of factors which could change in response to location, i.e., capital cost and feedstock cost; secondly, in the case of capital cost, to show the effect of cost growth due to process modifications unforeseen at this level of development (or cost reduction due to improvements in the technology); and thirdly, to show how the market price of furfural affects the required ethanol selling price.

Ethanol Price Sensitivity Analysis Using SERI Parameters

25 Million Gallon Plant

The economic analysis for Design Case I using SERI parameters shows that with wood chip feedstock at an average price of \$42 per dry metric ton, an estimated plant capital cost of \$139.79 million and a byproduct furfural price of 15¢ per pound, an ethanol selling price of \$1.23 per gallon is required to give a 15% IRR.

Feedstock price shows a strong influence on ethanol selling price, for example, a reduction in average price to \$35 per ton results in a 12¢ per gallon reduction in ethanol selling price to \$1.10. Increasing the feedstock price to \$52 per ton (perhaps due to a longer transportation distance) would necessitate raising the ethanol price to \$1.45 per gallon. The average sensitivity of ethanol price to feedstock price is approximately 1¢ per gallon per percent change in feedstock price.

The sensitivity to capital costs for Design Case I using SERI parameters shows an ethanol selling price of 93¢ per gallon for a \$100 million capital cost rising to \$1.68 per gallon for a \$200 million capital cost. Considerable variation in capital cost occurs with location. For example, the difference in capital cost between a U.S. Gulf Coast location and a typical mid-west location is \$17.69 million. This difference in capital cost translates into a selling price decrease of around 13¢ per gallon for the Gulf Coast location.

The relative effects of feedstock price and capital cost on selling price suggests that there may well be a "trade off" between location and transportation distance. Feedstock at \$50 per ton due to extra transportation distance may be acceptable if a capital cost reduction is possible at a more attractive location.

The average sensitivity of selling price to capital cost is approximately 1¢ per gallon per percent change.

#### 5 Million Gallon Plant

Ethanol selling price for the 5 million gallon design case using SERI parameters shows almost the same relative sensitivity to feedstock cost as for the 25 million gallon case. A reduction in the average feedstock cost to \$35 per ton results in a selling price of \$1.21 per gallon for the ethanol. It may well be possible to obtain feedstock at this price in some locales. An increase in feedstock price to \$52 per ton results in an ethanol selling price of \$1.62 per gallon. The average sensitivity of ethanol selling price to feedstock cost for the small scale commercial plant is 1¢ per gallon per percent change.

The ethanol selling price shows high sensitivity to the plant capital cost varying from 62¢ per gallon at a capital cost of \$50 million to \$2.20 at a capital cost of \$90 million. As found the 25 million gallon plant capital cost varies considerably with location.

The difference between a U.S. Gulf Coast and a typical mid-west location being \$8.4 million translating into an ethanol price difference of around 33¢ per gallon. The average sensitivity is nearly 3¢ per gallon per percent change in capital cost.

Here again, these may well be a beneficial trade off between location and feedstock cost. Expensive feedstock may be acceptable if the location results in capital cost savings.

Table 5.7 Sensitivity Analysis for 25 Million Gallon Plant  
(Design Case I)

(a) Ethanol Selling Price Versus Wood Chip Cost  
(SERI Parameters)

<u>Wood Chip Cost</u> <u>\$/Dry Metric Ton</u>	<u>Ethanol Selling Price \$/Gallon</u> <u>at 15% DCFIROR on Equity,</u> <u>Capital Cost - \$139.78 Million</u>
35	1.10
40	1.18
42	1.23
44	1.28
47	1.35
52	1.45

Average Sensitivity \$0.86 Per Gallon Per Percent Change

(b) Ethanol Selling Price Versus Plant Capital Cost  
(SERI Parameters)

<u>Plant Capital Cost</u> <u>\$MM</u>	<u>Ethanol Selling Price \$/Gallon</u> <u>at 15% DCFIROR on Equity,</u> <u>Feedstock \$42/Tonne</u>
100	0.93
120	1.10
139.78	1.23
150	1.33
175	1.53
200	1.68

Average Sensitivity \$1.05 Per Gallon Per Percent Change

Table 5.8 Sensitivity Analysis for 5 Million Gallon Plant  
(Design Case III)

(a) Ethanol Selling Price Versus Wood Chip Cost  
(SERI Parameters)

<u>Wood Chip Cost \$/Dry Metric Ton</u>	<u>Ethanol Selling Price \$/Gallon at 15% DCFIROR on Equity, (Capital Cost - \$69.76 Million)</u>
35	1.21
40	1.37
42	1.42
44	1.43
47	1.47
52	1.62

Average Sensitivity \$1/01 Per Gallon Per Percent Change

(b) Ethanol Selling Price Versus Plant Capital Cost  
(SERI Parameters)

<u>Plant Capital Cost \$MM</u>	<u>Ethanol Selling Price at 15% DCFIROR on Equity (Feedstock - \$42/Tonne)</u>
50	0.62
60	1.00
70	1.42
80	1.81
90	2.20

Average Sensitivity \$2.76 Per Gallon Per Percent Change

Figure 5.1 Ethanol Selling Price Versus Feedstock Cost at 15%  
DCFIROR on Equity 25 Million Gallon Base Case Plant  
(SERI Parameters)

FIGURE 5.1  
ETHANOL SELLING PRICE  
VS.  
FEEDSTOCK COST  
@  
15% DCFIROR ON EQUITY  
FOR  
25 MILLION GALLON  
BASE CASE PLANT  
(SERI PARAMETERS)

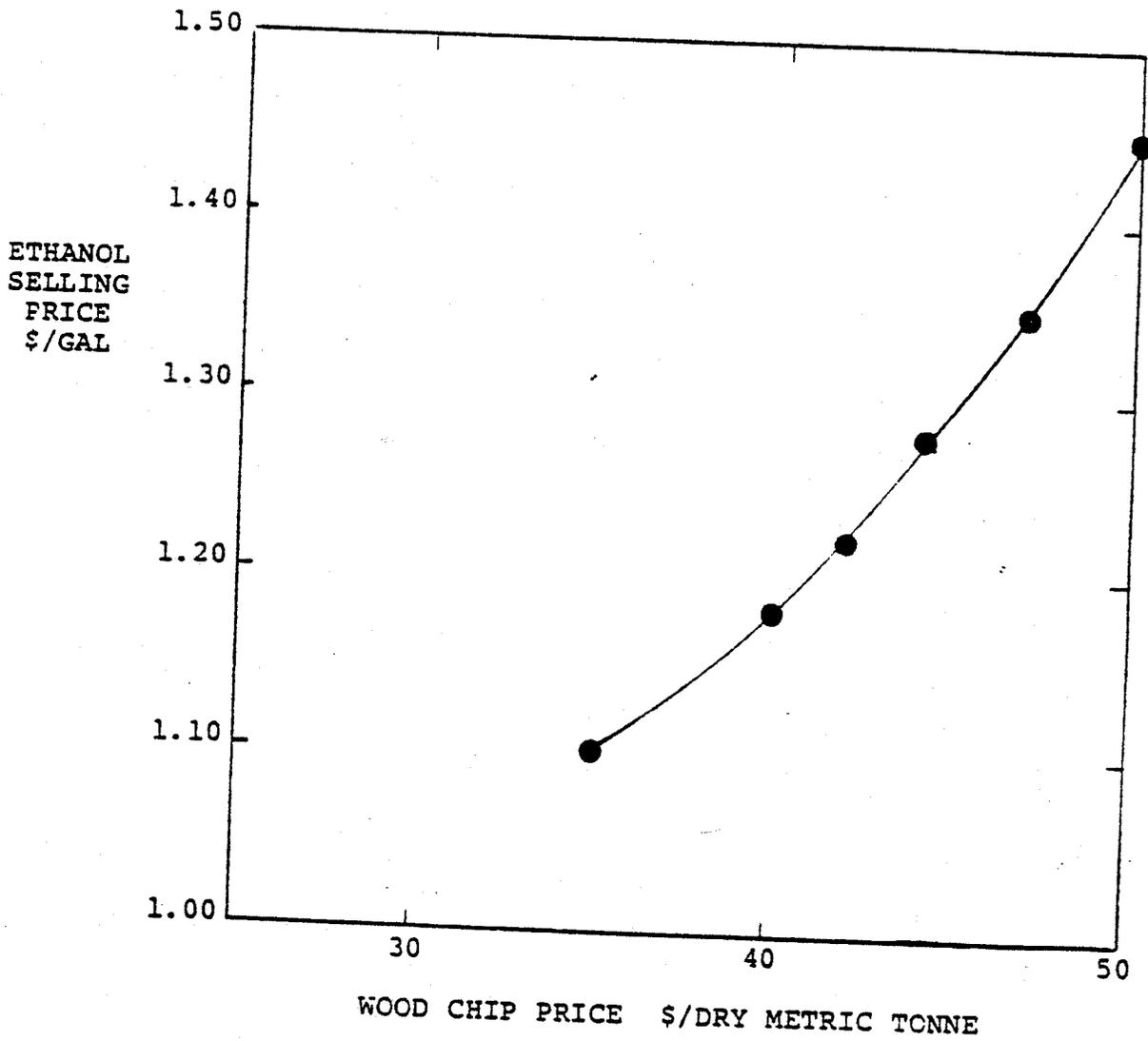


Figure 5.2 Ethanol Selling Price Versus Capital Cost at  
15% DCFIROR on Equity for 25 Million Gallon  
Base Case Plant (SERI Parameters)

FIGURE 5.2  
ETHANOL SELLING PRICE  
VS.  
PLANT CAPITAL COST  
@  
15% DCFROR ON EQUITY  
FOR  
25 MILLION GALLON  
BASE CASE PLANT  
(SERI PARAMETERS)

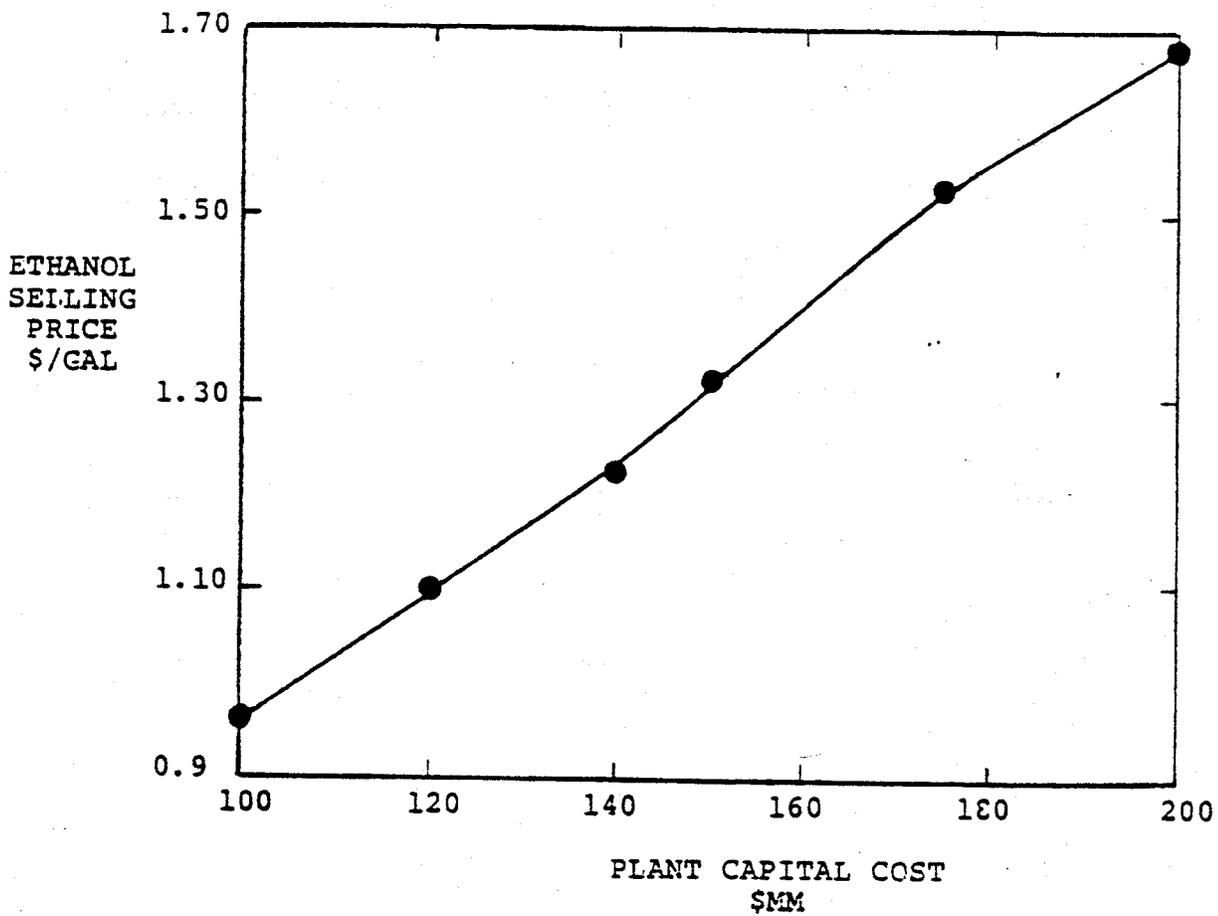


Figure 5.3 Ethanol Selling Price Versus Feedstock Cost at  
15% DCFIROR on Equity for 5 Million Gallon Plant  
Design Case III (SERI Parameters)

FIGURE 5.3  
ETHANOL SELLING PRICE  
VS.  
FEEDSTOCK COST  
@  
15% DCFROR ON EQUITY  
FOR  
5 MILLION GALLON PLANT  
DESIGN CASE III  
(SERI PARAMETERS)

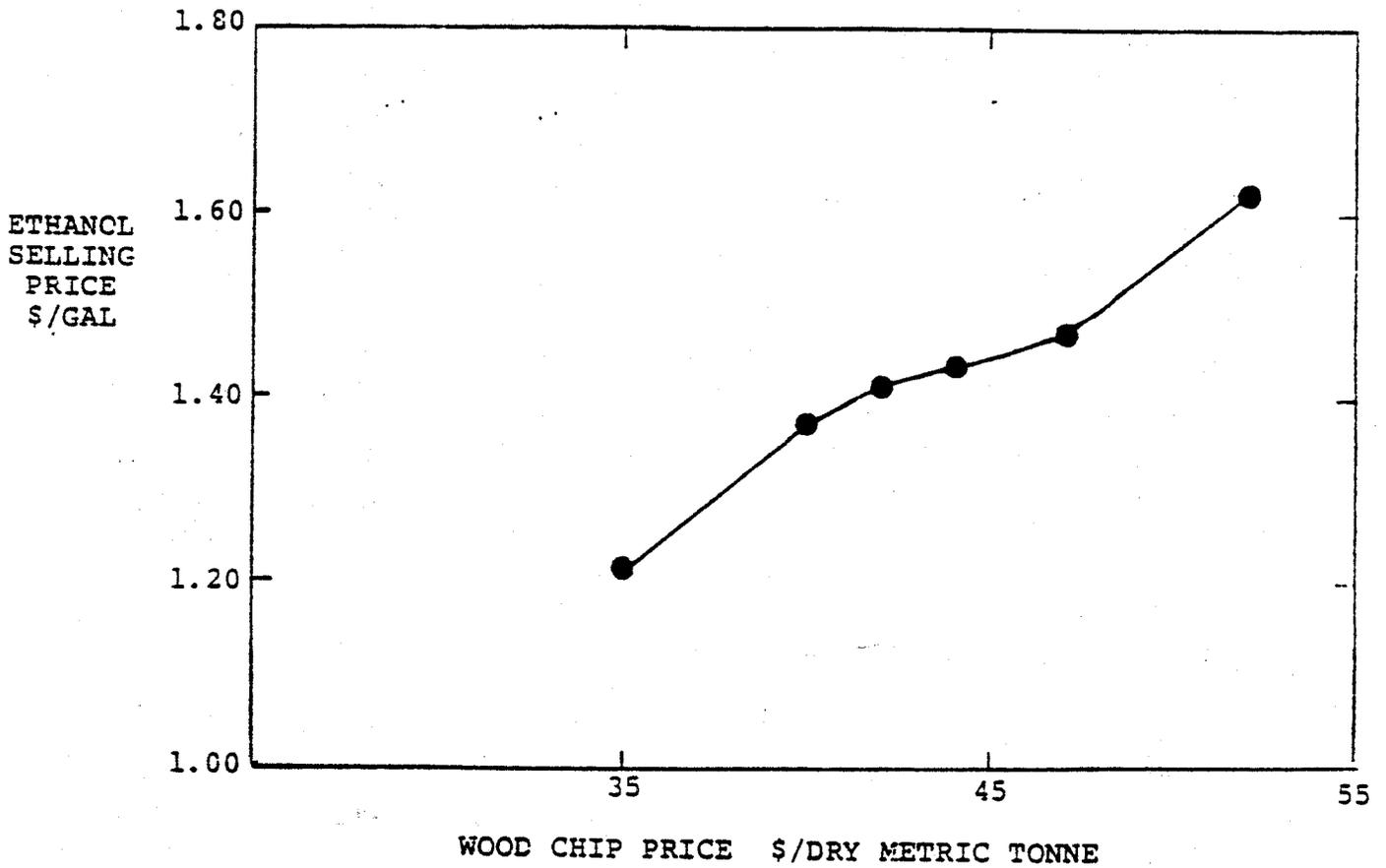
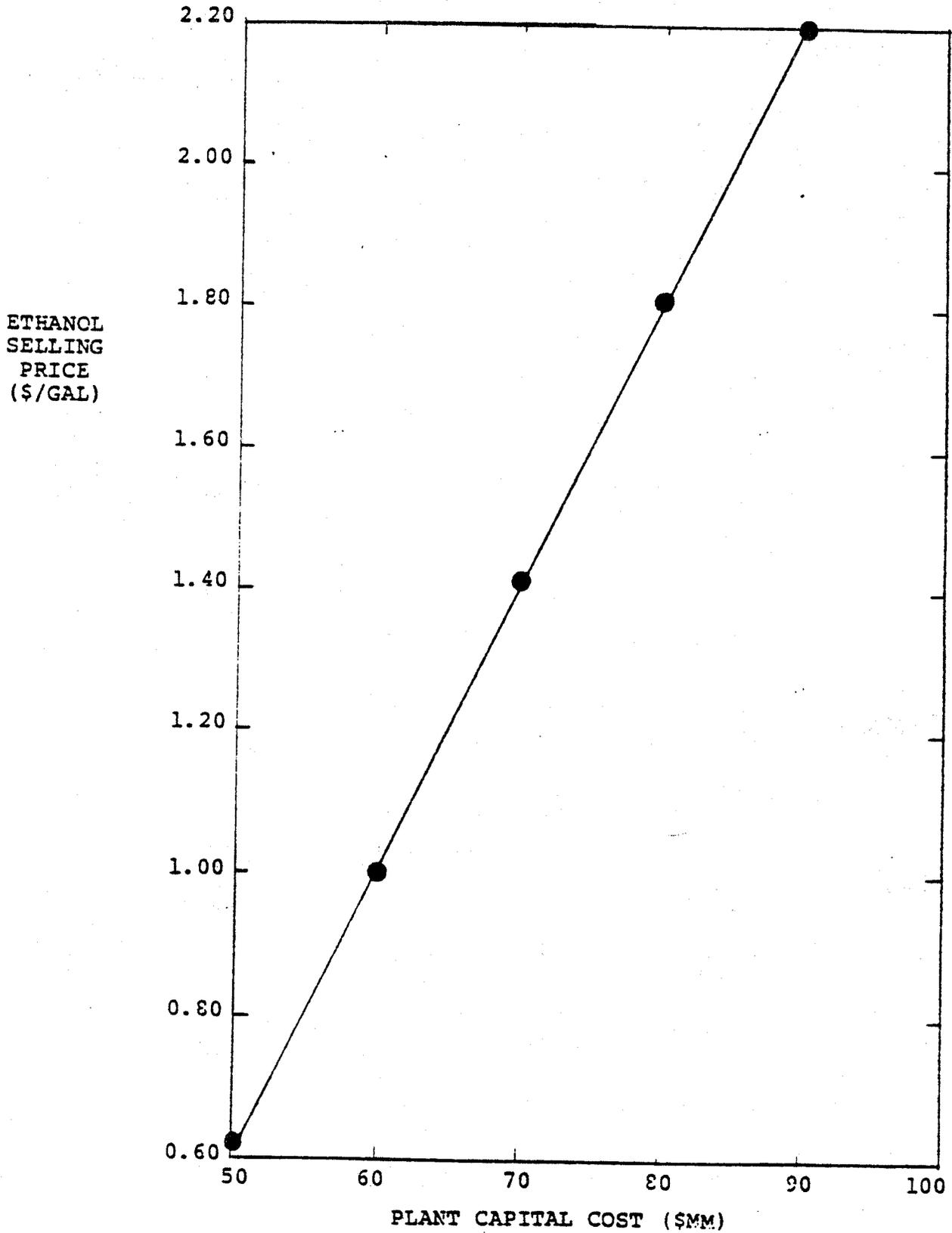


Figure 5.4 Ethanol Selling Price Versus Plant Capital Cost at  
15% DCFIROR on Equity for 5 Million Gallon Plant  
Design Case III (SERI Parameters)

FIGURE 5.4

ETHANOL SELLING PRICE VS. PLANT CAPITAL COST  
@ 15% DCFIROR ON EQUITY FOR 5 MILLION GALLON PLANT  
DESIGN CASE III (SERI PARAMETERS)



## Ethanol Price Sensitivity Analysis Using Badger Parameters

### 25 Million Gallon Plant

The economic analysis for Design Case I using Badger parameters shows that with wood chip feedstock priced at \$42 per dry metric ton, an estimated plant capital cost of \$139.78 million and a byproduct furfural price of 15¢ per pound the required ethanol selling price to achieve a 15% IRR is \$1.80.

Feedstock price shows a marked influence on ethanol price, for example, if feedstock were available for \$35 per dry metric ton then the selling price of ethanol would fall to \$1.63 per gallon. If feedstock were priced at \$52 per dry metric ton the ethanol price required would be \$2 per gallon. The average sensitivity to feedstock cost is 1¢ per gallon per percent change.

The sensitivity of ethanol price to plant capital cost shows that if the plant capital cost were reduced to \$100 million then the required ethanol selling price would be \$1.35 per gallon. At a capital cost of \$200 million the selling price would be \$2.47 per gallon. The average sensitivity over the range investigated is \$1.5 per gallon per percent change in capital cost.

### 5 Million Gallon Plant

Ethanol selling price for the 5 million gallon design case using Badger parameters shows the same relative sensitivity to feedstock price (1¢ per gallon per percent change) as for the 25 million gallon case. Feedstock available for \$35 per dry metric ton would allow a selling price of the ethanol of \$2.66 per gallon. Feedstock at \$52 per dry metric ton would lead to a selling price of \$2.92 per metric ton.

The ethanol selling price for this scale of operation is extremely sensitive to capital cost showing an average sensitivity of 4.5¢ per gallon per percent change. At a capital cost of \$50 million the selling price is \$1.67 and at \$80 million the selling price is \$3.60. This indicates that a 5 million gallon plant cost overrun can have serious consequences on the plant profitability. Conversely, at this scale cost savings should give a major benefit. Such cost saving could come from a favorable location or elimination of turbogenerator capacity in an area if cheap electric power or from improvements in the process during development.

These figures suggest that there may be an optimum location with regard to feedstock transportation distance and capital cost.

Table 5.9 Sensitivity Analysis for 25 Million Gallon Plant  
(Design Case I)

(a) Ethanol Selling Price Versus Wood Chip Cost  
(Badger Parameters)

<u>Wood Chip Cost \$/Dry Metric Ton</u>	<u>Ethanol Selling Price \$/Gallon at 15% DCFIROR on Equity, Capital Cost - \$139.78 Million</u>
35	1.63
40	1.75
42	1.80
44	1.84
47	1.90
52	2.00

Average Sensitivity \$0.91 Per Gallon Per Percent Change

(b) Ethanol Selling Price Versus Plant Capital Cost  
(Badger Parameters)

<u>Plant Capital Cost SMM</u>	<u>Ethanol Selling Price \$/Gallon at 15% DCFIROR on Equity, (Feedstock \$42/Tonne)</u>
100	1.35
120	1.57
139.78	1.80
160	2.01
180	2.23
200	2.47

Average Sensitivity \$1.56 Per Gallon Per Percent Change

Table 5.10 Sensitivity Analysis for 5 Million Gallon Plant  
(Design Case III)

(a) Ethanol Selling Price Versus Wood Chip Cost  
(Badger Parameters)

<u>Wood Chip Cost \$/Dry Metric Ton</u>	<u>Ethanol Selling Price \$/Gallon at 15% DCFIROR on Equity, (Capital Cost - \$69.76 Million)</u>
30	2.60
35	2.66
40	2.82
42	2.83
44	2.88
47	2.92

Average Sensitivity \$0.91 Per Gallon Per Percent Change

(b) Ethanol Selling Price Versus Plant Capital Cost  
(Badger Parameters)

<u>Plant Capital Cost \$MM</u>	<u>Ethanol Selling Price \$/Gallon at 15% DCFIROR on Equity, (Feedstock \$42/Tonne)</u>
50	1.75
60	2.33
70	2.83
80	3.50

Average Sensitivity \$4.5 Per Gallon Per Percent Change

Figure 5.5 Ethanol Price Versus Feedstock Cost Design Case I  
(Badger Parameters)

FIGURE 5.5

ETHANOL PRICE VS. FEEDSTOCK COST  
@ 15% DCFIROR ON EQUITY  
FOR 25MM GALLON  
BASE CASE PLANT (DESIGN CASE I)  
(BADGER PARAMETERS)

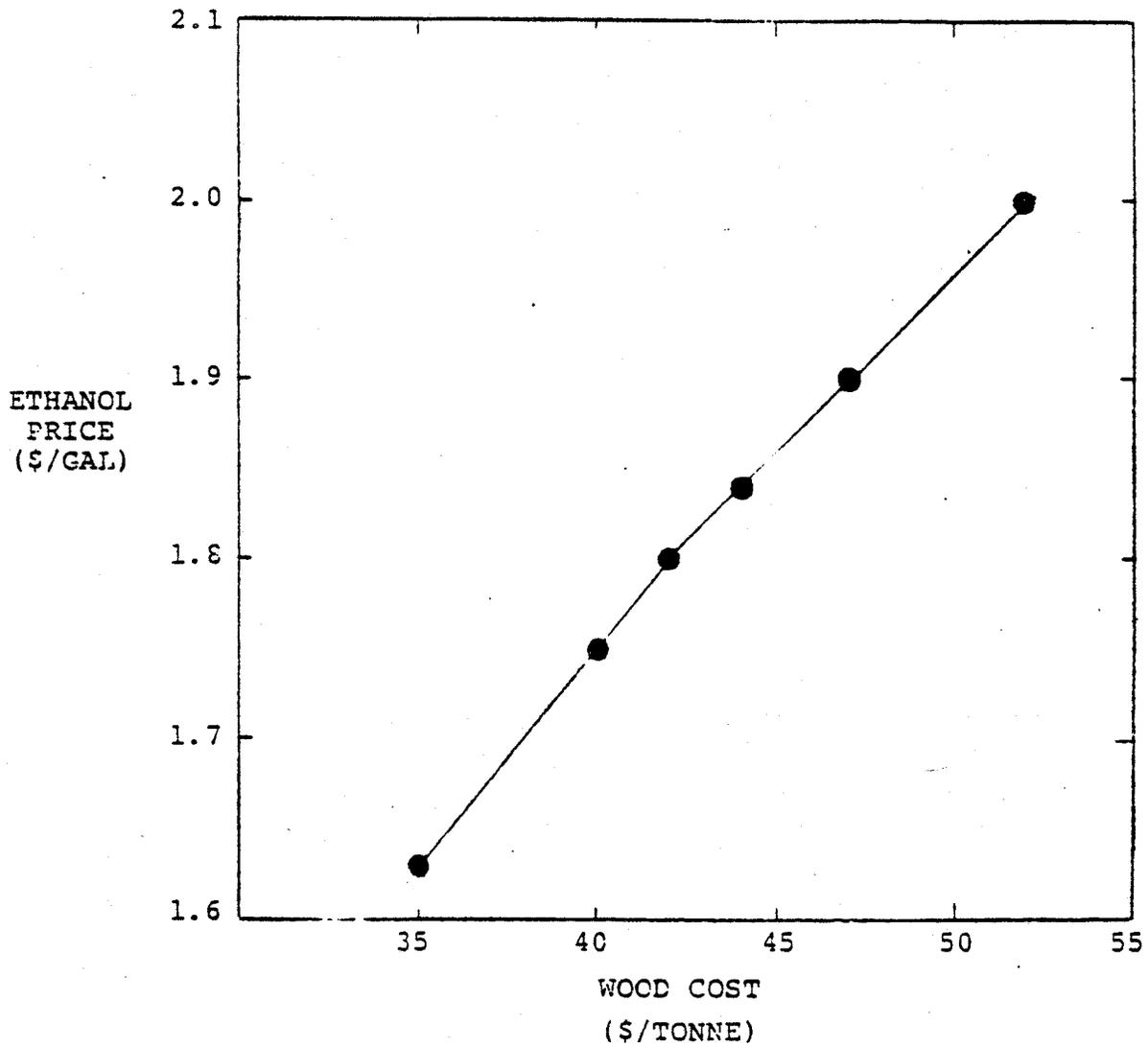


Figure 5.6 Ethanol Price Versus Capital Cost Design Case I  
(Badger Parameters)

FIGURE 5.6

ETHANOL PRICE VS. CAPITAL COST  
@ 15% DCFI ROR ON EQUITY  
FOR 25MM GALLON  
BASE CASE PLANT (DESIGN CASE I)  
(BADGER PARAMETERS)

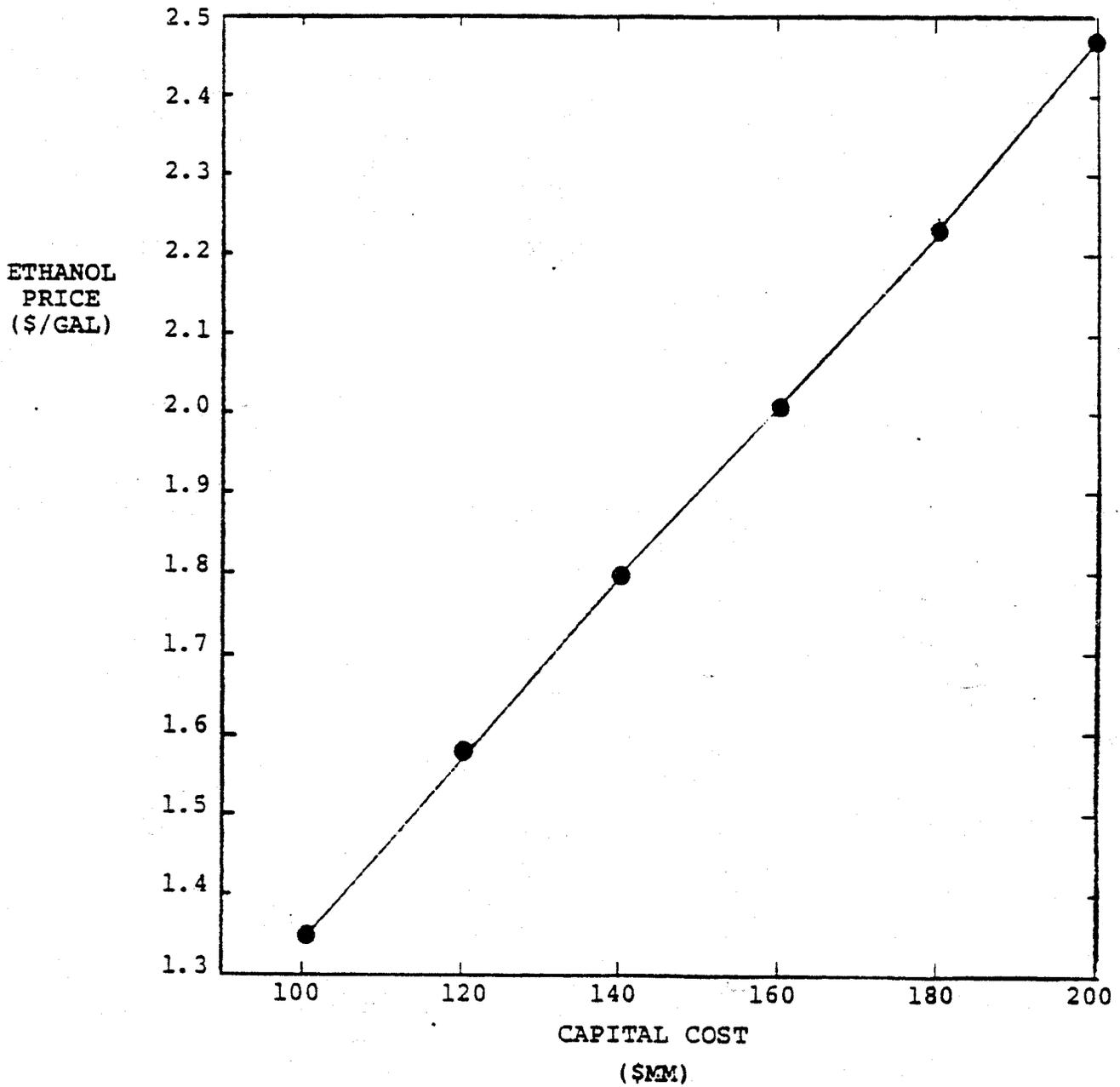


Figure 5.7 Ethanol Price Versus Wood Cost Design Case III  
(Badger Parameters)

FIGURE 5.8

ETHANOL PRICE VS. CAPITAL COST  
@ 15% DCFI ROR ON EQUITY  
FOR 5 MM GALLON PLANT  
(DESIGN CASE III)  
(BADGER PARAMETERS)

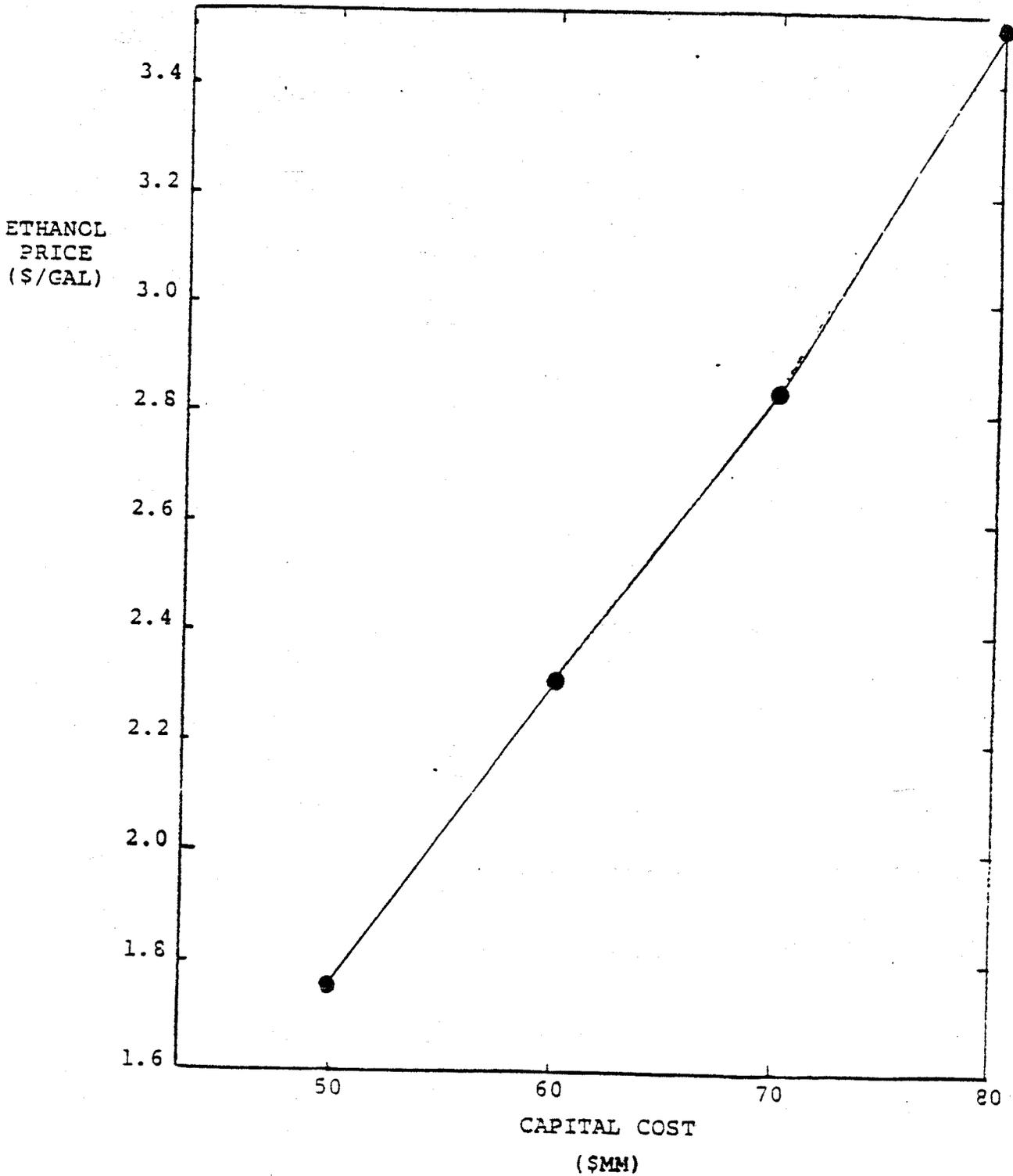


Figure 5.8 Ethanol Price Versus Capital Cost Design Case III  
(Badger Parameters)

Sensitivity of Ethanol Selling Price to Furfural Selling Price  
(Badger and SERI Parameters)

The sensitivity of ethanol selling price to furfural price is given in Table 5.11 and shown in Figure 5.9 for the 25 million gallon plant. The calculation was carried out using both SERI and Badger parameters.

This shows that the selling price of ethanol is very sensitive to furfural market price. The average sensitivity over the range investigated is 0.8¢ per gallon per percent change in furfural price (SERI) and 0.71¢ per gallon per percent change (Badger). If furfural sells at 33¢ per pound the threshold for adipic acid production (see Section 6.0) ethanol can be sold for 0.90¢ per gallon (Badger) and with furfural at 9¢ per pound the threshold for butanol production (see Section 6.0) ethanol must sell for \$1.50 per gallon (SERI basis); \$2.05 per gallon (Badger basis) to achieve a 15% IRR.

These figures suggest that early large scale plants will be very attractive ventures supplying a medium term furfural market at a price of around 25¢-35¢ per pound.

Table 5.11 Ethanol Selling Price Versus Furfural Price

<u>Furfural Price ¢ Per Pound</u>	<u>Ethanol Price \$ Per Gallon</u>	
	<u>SERI</u>	<u>Badger</u>
0.10	1.48	2.05
0.15	1.22	1.80
0.20	1.00	1.55
0.25	0.70	1.32
0.35	--	0.85
Average Sensitivity ¢ Per Gallon per Percent Change	0.80	0.72

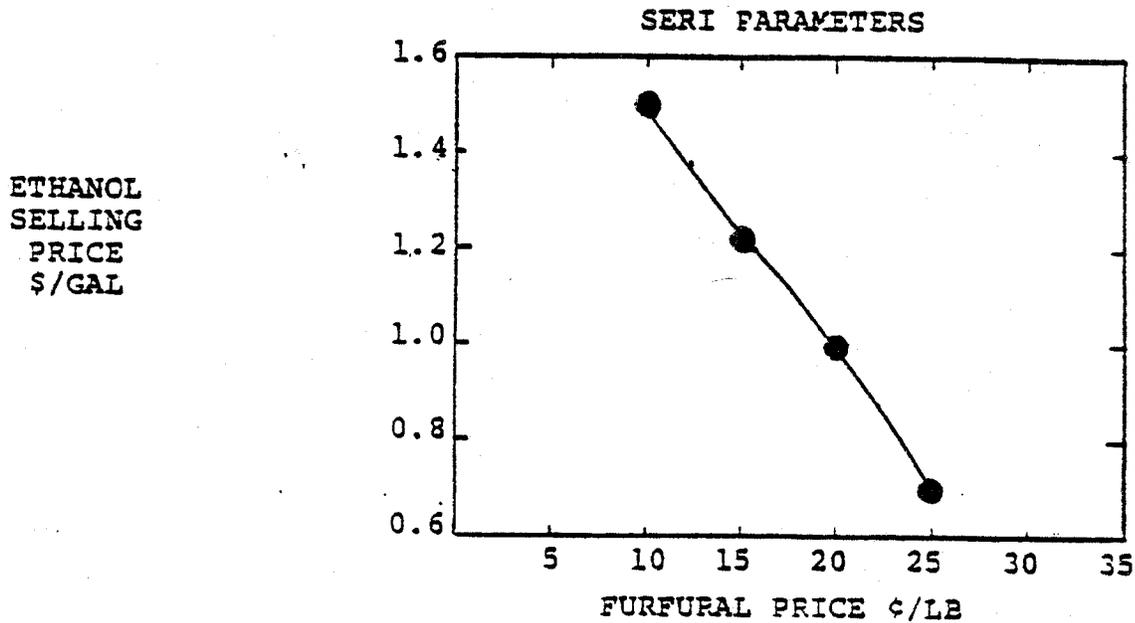
Figure 5.9 Ethanol Selling Price Versus Furfural Price for  
25 Million Gallon Plant

FIGURE 5.9

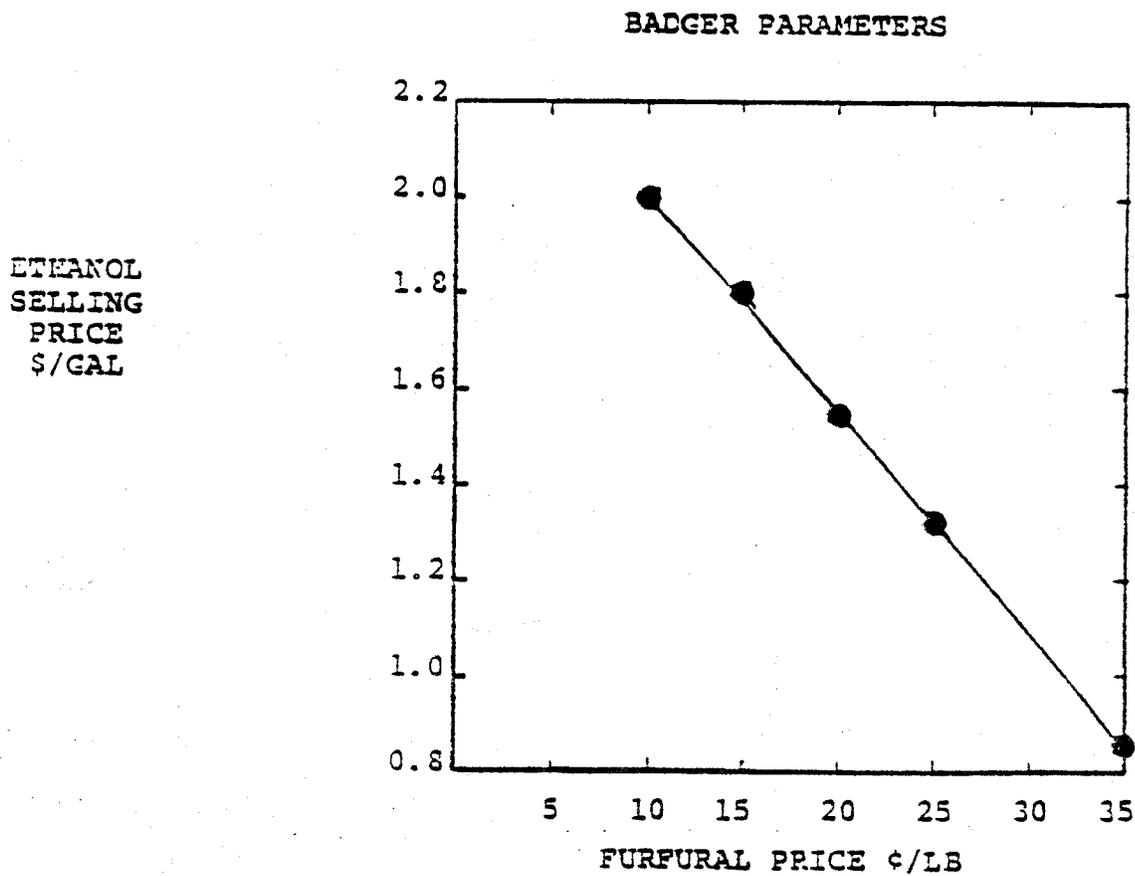
ETHANOL SELLING PRICE VERSUS FURFURAL PRICE FOR  
25 MILLION GALLON PLANT  
(BASE CASE)

(A)

@ 15% DCFIROP ON EQUITY



(B)



Effect of DCFIROR of Economic Parameters

The relative effect on the IRR of various economic parameters for the base case plant is given for both SERI and Badger parameters on Figure 5.10. The index shown is a relative number and can be used to compare the individual effects. The index is the % change in IRR caused by a % change in parameter.

The calculation is performed by holding all the other variables fixed at their normal value and varying the particular parameter over a reasonable range.

Figure 5.10 for Design Case I (Base Case) shows that the IRR is generally more sensitive in the SERI based calculation than in the Badger calculation. For example, a 1% change in ethanol price gives a 0.25% change in IRR for the SERI case and 0.15% change for the Badger calculation. The major influences on the IRR in both calculations are ethanol price, capital cost, feedstock cost and furfural price. The effects of electricity selling price and inflation are also significant. Percentage equity also has a significant negative effect in the SERI calculation. This is because the IRR calculated using the SERI parameters is higher than the interest rate on debt, giving an advantage to leveraged capital. In the Badger calculation, however, the effect of equity is positive since the calculated IRR is below the interest rate on debt.

The effect of initial working capital in both calculations is not significant over the range investigated. The data used in the sensitivity analyses for both calculations is given in Appendix VI.

The implications of the effects of feedstock cost, capital cost and furfural price on the process economics are discussed earlier in this section.

Figure 5.10    Sensitivity of DCFIROR to SERI and Badger Economic  
Parameters for 25 Million Gallon Plant  
(Design Case I)

The assessment of the potential furfural market is based on published information by Chem Systems Inc., SERI, and Mitre Corporation (References 9, 10, and 11). These studies discuss the technology available for conversion of furfural into commodity chemicals such as adipic acid, maleic anhydride, butanol, tetrahydrofuran, and ethylbenzene. The main conclusion is that large markets could exist for furfural should it be available at a price of 35¢ per pound or less. The elasticity of the furfural market, based on the SERI report, is calculated allowing an optimum price/sales volume to be derived. This price is used in the economic assessment.

#### Technical Risk Assessment

Use of the Dartmouth plug flow reactor to produce fermentable sugars by acid hydrolysis of lignocellulose represents a considerable deviation from past commercial practice where the hydrolysis reaction was carried out in batch (Reference 6) or semicontinuous (Reference 7) mode.

The present process based on the Dartmouth reactor has several areas of major technical uncertainty which may lead to greater than estimated capital cost or lower than expected operating performance in the first generation of commercial-scale plants. The major technical risks identified in the present design study are associated with the following:

- o Continuous stable operation of the large-scale 10" diameter plug flow reactor over periods of 200 to 300 hours without plugging and with adequate heat transfer capacity.
- o Mixing and pumping an acidified wood slurry of 20% solids consistency.

- o The adequacy of grinding the reactor feed solids down to a particle size of "through 20 mesh."
- o Maintenance of the fermentable sugar and furfural yield over extended operating periods.
- o The fermentability of the hydrolyzate by yeast to give ethanol in acceptable yield.
- o Presence in the ethanol of extraneous organics which may prove unacceptable in motor fuel.
- o The separation and working of the major solid residue, lignin, from the hydrolyzate.
- o The stability of the hydrolyzate sugars produced to process temperatures in the presence of extraneous organics in the hydrolyzate.

These areas of major technical risk are all confined to the hydrolysis unit. The remaining processing units: wood chip handling, fermentation, ethanol and furfural recovery as included in the design, do not represent significant deviations from existing commercial technology and therefore would be expected to work well providing that the hydrolysis unit performs as planned.

The above areas of concern have been addressed in two ways. Firstly, considerable additional development work has been performed at Dartmouth College, Badger's Weymouth Laboratory, equipment vendors' laboratories, and at Japanese Gasoline Corporation's laboratories in Japan. Further experimental development is planned on a scaled-up reactor at the SERI laboratories in the near future. At the time of writing, no details of the SERI system are available. Secondly, extra processing steps have been included where it was felt that these would significantly reduce the degree of uncertainty in the design. The above measures are discussed in the following pages (see also Section 3.0).

(1) Reactor Stability

The operational stability of the reactor was investigated by attempting to operate the Dartmouth reactor stably over an extended period. Operation for 4 to 6 hours without blockage was targeted as a reasonable demonstration, considering the operating constraints of the Dartmouth laboratory such as limited manpower, feedstock storage, effluent disposal, limited reactor velocity, small scale of equipment ( $\frac{1}{2}$ " diameter reactor), rudimentary process controls, and limited steam pressure.

In the test, reasonably stable operation was obtained over a period of approximately two hours before blockage occurred. Although the results of these stability trials reduce to some extent the degree of uncertainty over this issue, reactor stability has not yet been adequately demonstrated and must remain an area of serious technological risk. (A full report of this experimental work is included in Appendix III).

In the present process design an installed spare reactor and feed pump has been included to allow isolation and cleaning of one reactor without affecting the operation of the rest of the plant. Several installed spare reactors and pumps could be included without a major effect on overall capital cost should blockage prove to be a frequent occurrence. However, this practice is obviously limited from an operational point of view (if not from a capital cost point of view). Operation of the new scaled-up version of the Dartmouth reactor at the SERI laboratories should increase the degree of confidence in this aspect of the process.

Improved operational stability may be achieved with a two-stage reaction or by prehydrolysis and removal of pentose sugars prior to the main hydrolysis. This second option may reduce tar formation by exclusion of furfural from the main hydrolysis.

(2) High Consistency Slurry Handling

A wood slurry of 20% solids consistency exhibits very little free water and demonstrates poor flow properties. A prehydrolysis vessel has been included in the design to improve the pumpability of the wood slurry. At the conditions specified for the prehydrolysis (332°F, 1% acidity, 5 minutes residence time) considerable hemicellulose hydrolysis and defibration of the wood has been observed. The resultant slurry has the flow characteristics of a mud. The effect of prehydrolysis is supported with experimental work at Badger's Weymouth Laboratory where small batches of through-20-mesh acidified wood slurry were heated to 332°F for 5 minutes, resulting in a viscous mud. Inclusion of a prehydrolysis step in the process should reduce the concern about slurry handling.

(3) Reactor Feed Particle Size

Although calculations show that particles with a smallest dimension of up to 1000 microns are suitable for use in the hydrolysis reactor, this has not been demonstrated experimentally. Work to date at Dartmouth has used wood flour of 250 micron average diameter.

Reactor feed particle size has a significant bearing on the process economics because of the energy required for grinding and the capital cost of the refiner grinding equipment. Prehydrolysis may relieve the constraints on particle size since it results in considerable breakdown of the wood fibers before they enter the main reactor. Operation of the new scaled-up SERI reactor should establish an upper limit of feed particle size if sufficient prehydrolysis is included in the SERI feed preparation.

(4) Fermentable Sugar and Furfural Yields

Fermentable sugar and furfural yields greater than those assumed in the design basis were obtained in the 2-hour extended reactor operation discussed above. This suggests that the process design basis is conservative and that this issue is of less concern than the other areas. However, operation of the new SERI reactor using larger particle size and more concentrated feeds should establish the expected sustainable yield with greater confidence.

(5) Hydrolyzate Fermentability

The fermentability of hydrolyzates from the Dartmouth reactor has been demonstrated on a small scale by researchers there. This work showed that with a specially adapted common yeast strain, good fermentation yields of over 95% could be expected. The hydrolyzate was prepared by neutralization to pH 10 and reacidification to pH 5 whilst boiling.

In the present process design, hydrolyzate is flashed to remove furfural, the pH is adjusted as above, and, as an additional final step, steam-stripped to remove volatile organics that may be inhibitory. Inclusion of steam stripping as an extra step should increase the degree of confidence in the fermentability over that established by the Dartmouth work. Degradation of fermentable sugars during the stripping still remains uncertain.

The toxicity to the yeast of the residual extraneous organics in the hydrolyzate was tested at Japanese Gasoline Corporation's laboratories in Japan. The tests showed that hydrolyzate prepared using the Dartmouth methods did not contain any compounds acutely toxic to yeast. (A full report of this toxicity trial is given in Appendix II.) They should provide a basis for further work on yeast acclimatization to extraneous hydrolyzate organic components.

(5) Ethanol Product Contaminants

The presence of undesirable contaminants in the fuel-grade ethanol has been investigated by attempting to characterize all the extraneous organic substances produced during the hydrolysis reaction and the fermentation. The ethanol purification system design was based on a feed containing identified extraneous organics. Multicomponent distillation calculations identified acetaldehyde as the only significant contaminant remaining in the product. This conclusion, however, remains to be demonstrated in practice, and the calculations depend on the accuracy of the characterization of extraneous organics produced in the hydrolysis.

Although there are no full published specifications for fuel-grade ethanol, this matter has been discussed with a commercial producer in an attempt to formulate a target specification. Although the tentative conclusion was that the expected impurities were acceptable, significant technical uncertainty remains.

(6) Lignin Separability and Washability

The separability of the lignin/tar solid residue from the hydrolyzate was tested at a vendor's laboratory. These tests showed that the residue could be separated easily using centrifuges. Based on these tests a special type of washing centrifuge has been specified. A high degree of confidence is associated with the design of this step, but the separability and sugar recovery by washing for a continuous large scale system remains to be demonstrated .

(7) Stability of Hydrolyzate Sugars

Work on Dartmouth hydrolyzates at Badger's Weymouth Laboratory showed that the fermentable sugars degraded on prolonged exposure to boiling temperatures of 215°F. On boiling the hydrolyzate for a period of three hours, 100% loss was observed. In the process design, the hydrolyzate is maintained at high temperature for minimal time. This is an area of uncertainty since yield losses of sugars at shorter exposure times has still to be quantified. A further source of yield loss may be repolymerisation of sugars under acidic conditions and at higher concentrations. This loss also still remains to be quantified.

### Quantitative Analysis of Technical Risk

Although measures have been taken to address perceived areas of uncertainty, it is obvious from the above discussion that considerable technical risk is still attached to this new technology. It is desirable therefore to quantify this risk in terms of potential capital cost growth (or extra capital cost required to correct for errors in the design basis assumptions) and performance shortfalls (or the degree to which the new process does not reach expected operational performance).

This risk assessment would give guidance in planning future expenditures and controlling capital cost growth and provide an assessment of operating capacity in the early life of the new process. It would also provide a basis for judging the relative importance of future development work by reducing the level of uncertainty associated with the technology. An estimating contingency has already been applied to the capital estimates, and in the Badger economic analysis a conservative operating capacity schedule has been included; however, contingency is based on the estimating method itself, and the operating schedule is based on general experience with new technology rather than on the nature of the technology itself.

The technical risk assessment is based on a study by Rand Corporation entitled "Understanding Cost Growth and Performance Shortfalls in Pioneer Process Plants," September 1981, R-2569-DOE. This study by Rand was done for the U.S. Department of Energy to identify the factors responsible for cost growth and for failure to meet design performance rates in pioneer process plants. Thirty-four firms in the process industries provided data to support a statistical analysis of the cost growth and performance shortfalls in pioneer plants.

The data base contained information on forty-four commercial-scale process plants build over the past fifteen years in five generic categories (olefins and derivatives, aromatics and derivatives, refinery products, chlorine chemicals, minerals processing).

The Rand Report developed a cost growth model where the relative influence of a set of independent variables was statistically estimated by regression analysis. The result is a linear equation relating cost growth to the effects of process uncertainty, plant complexity, estimate inclusiveness, project definition, and development stage. Cost growth is defined as the ratio of the estimated cost at any particular stage in the project development to the final completed commercial plant cost. External factors such as inflation, scope changes, bad weather, and strikes were removed from the analysis.

Badger does not consider that the Rand methodology should be used to adjust estimates quantitatively in the early stages of development, nor do we consider as inevitable the cost growth predicted by the methodology. We do view the methodology as a powerful tool to measure cost growth risk and, more importantly, as a tool to assist research and development engineers and designers in controlling cost growth and ensuring that plants meet their design production rates.

Badger has applied the Rand Report methodology to the wood acid hydrolysis development project. The first objective was to assess the present situation - the present level of process development and the cost estimate and definition of the commercial plant in the study. The second objective was to predict the reductions in cost growth risk of several courses of action: (1) building and operating a demonstration unit, (2) completing a site-specific study, and (3) doing both. The detailed calculations including the assessment of the independent variables are given in Appendix V.

The elements of a "site-specific" study are outlined in Appendix V.

The results of Badger's calculations are shown below in Table 6.1.

Table 6.1 Likely Capital Cost Growth Risk Associated with  
Various Development Options

<u>Case</u>	<u>Calculated Cost Growth Ratio</u>	<u>Reduction in Cost Growth Risk</u>
Present Estimate	0.62	--
Building and Operating Demonstration Unit	0.85	0.23
Completing Site-Specific Study	0.81	0.19
Both Demonstration Unit and Site-Specific Study	0.90	0.28

The issue of cost should be addressed from two standpoints. Firstly, before proceeding into a commercial plant design, the cost for the plant should be established as accurately as practical. Secondly, every practical effort should be made to minimize cost growth.

The Badger analysis using the Rand methodology indicates that a cost estimate made subsequent to both the demonstration plant operation and a site-specific study would have much less risk of cost growth associated with it than would be the case without a study. In fact, although the cost growth risk is very sensitive to whether a site-specific study has been made, the results suggest that both the demonstration unit and the site-specific study are necessary. Neither is sufficient alone to reduce the risk of cost growth to a level where a major financial commitment for the first commercial plant can be made with confidence.

A further conclusion of the analysis is that a considerable reduction in cost growth uncertainty (19%) can be achieved by carrying out a site-specific study alone.

#### Operating Shortfall

Similar Rand methodology was applied to prediction of the risk of plant performance shortfall. Plant performance is also a critical parameter when firms do market planning and when the government hopes that synthetic fuels will relieve the nation's difficulties. For example, in the present process an increase in capacity from 50% to 70% in months 7 to 12 after startup results in a decrease in the required selling price of ethanol from \$1.80 to \$1.74 per gallon, and an increase in the same operating period to 100% of design capacity results in a selling price reduction to \$1.60 (Badger parameters.)

Process development factors influencing performance of the new plant were the number of process steps new in commercial use, the confidence in the heat and material balance figures, difficulty with waste handling and whether or not the plant processes solids.

Badger also applied the Rand methodology to the wood-acid hydrolysis project for performance shortfall. As with the cost growth analysis, the first objective was to assess the present situation and the risk of poor performance associated with the present state of process development. Similarly, the second objective was to investigate reductions in the risk due to several possible courses of action: (1) building and operating the demonstration unit, (2) completing a site specific study, (3) doing both. The detailed calculations are given in Appendix V.

The calculation gives the most likely average production rate as a percentage of plant design capacity in months 7 to 12 after startup.

The results of the calculation are shown in the following Table 6.2.

Table 6.2 Performance Shortfall Risk Associated with Various  
Development Options

<u>Case</u>	<u>Calculated Expected Performance % of Design Months 7 to 12</u>	<u>Increase In Likely Performance % of Design</u>
Present Estimate	51.4	--
Building and Operating a Demonstration Unit	93.1	41.7
Completing Site-Specific Study	53.4	2
Both Demonstration Unit and Site-Specific Study	97.6	46.2

The above analysis shows that at the present stage of development a commercial plant could reasonably be expected to operate at no more than one-half of design in the 7 to 12 month period after startup.

A design based on demonstration plant data could be expected to result in operation at greater than 90% of design in the same period, and a design based on site-specific information would be expected to permit operation at just over one-half of design for the same period. Finally, a design based on both demonstration unit data and site-specific conditions could reasonably be expected to result in operation at nearly 100% of design. This analysis clearly shows the benefit of building and operating a demonstration unit because it results in a very high degree of confidence in the predicted operating performance. Performing a site-specific study has very little effect on the ability to predict performance.

The Badger analysis using the Rand methodology indicates that an economic analysis made subsequent to both the demonstration plant operation and a site-specific study would have much less risk of cost growth and performance shortfall. Significant reduction in cost growth risk can be attained by performing a site-specific process design study alone; however, this study would not significantly lend confidence to the predicted performance.

Construction and operation of a integrated process demonstration unit containing all the major process sections greatly reduces both cost growth and performance shortfall risks, and it appears to be an indispensable step in the approach to commercialization of this particular technology.

#### Assessment of the Potential Furfural Market

Data on the potential future market for furfural for various derivatives which can be competitively produced at various furfural prices have been published by SERI (Reference 10). The data are given in the table below. Prices in the table have been calculated for 1984 assuming an average escalation rate of 7% per year.

<u>Furfural Derivative</u>	<u>Furfural Price \$/Pound</u>	<u>Total Furfural Demand Metric Ton/Year</u>	<u>Total Revenue \$MM</u>
Furfural	0.57	63,492	79.8
Adipic Acid	0.33	517,913	377
Maleic Anhydride	0.12	723,810	192
Butanol	0.09	1,521,088	302

This information is plotted in Figure 6.1.

The price given above is the threshold price of furfural at which production of the derivative using furfural becomes economic. Based on the above figures an optimum furfural selling price of 15¢ per pound can be calculated. At the above price, Figure 6.1 shows that the predicted long term furfural market is around 2 billion pounds per year. This market demand would support the output from fifteen or sixteen large-scale plants each with a capacity of 25 million gallons per year ethanol. Although this represents a limited quantity of ethanol in terms of the gasoline market, it is expected that later generations of the process will have lower capital costs, higher ethanol yields, and better energy usage efficiency as a result of technology advances and operating experience. Furfural could then be sold at a lower price that would lead to increased market volume.

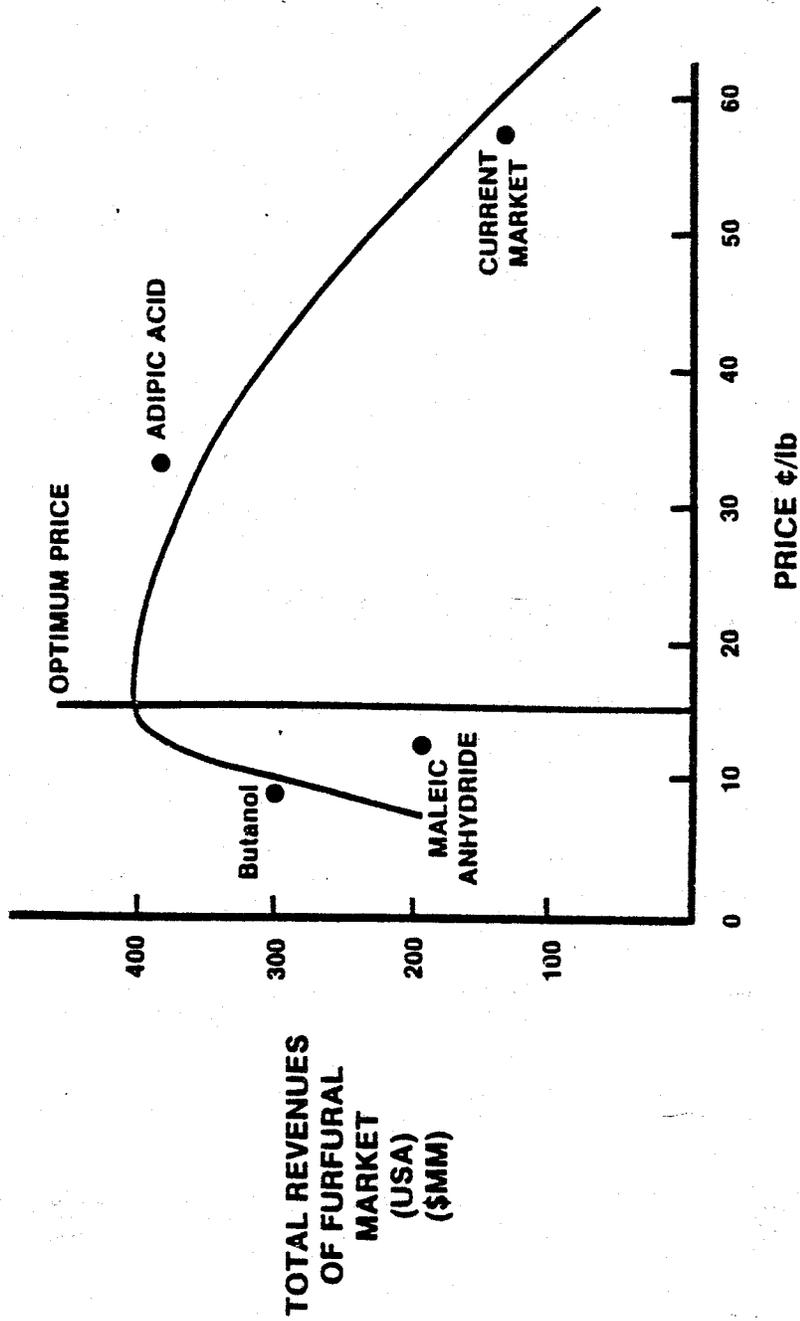
Ultimately furfural could be priced at its fuel value of around 5¢ per pound and burned to produce electric power. In this case, the electric power generated would increase to around 55 MW which is still below the maximum size of 80 MW covered by the PURPA Act. It must be emphasized here, with respect to the question of the furfural market development, that it would be unrealistic to expect an overnight change in the technology base of a number of major chemical industries. The establishment of a major furfural market under normal circumstances would happen gradually over a period of say 10 to 15 years after initial commercialization of the process. The price of furfural would gradually fall as more is produced. This implies that early plants will benefit from a higher furfural price than indicated in the base case economics. The sensitivity analysis of ethanol price to furfural price shows that early large scale processes selling furfural for say 33¢ per pound (the threshold price for adipic acid production) be able to sell ethanol for \$0.90 per gallon at 15% DCFIROR.

An unforeseen increase in the price of crude oil with resultant increases in the price of petrochemical feedstocks could alter this scenario and significantly accelerate the growth of the furfural market.

A second long-term possibility is that the fermentability of the pentose sugars to ethanol could be improved. This would, of course, remove the dependency of the process economics on limited byproduct markets. In this case a two-stage hydrolysis system would be required to remove pentose sugars prior to the main hydrolysis.

Figure 6.1 Projected Furfural Market Revenues Versus Selling  
Price (Reference 10)

**FIG. 6.1 - PROJECTED FURFURAL REVENUES VS. SELLING PRICE**



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