Technical and Economic Evaluation Wood to Ethanol Process

A report prepared for Solar Energy Research Institute

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CHEM SYSTEMS

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I. SUMMARY

A. Introduction

The interest in using ethanol as an octane enhancer, fuel extender or neat liquid fuel has grown over the years. Recently there has been increased concern over the worsening situation of the environment, the national debt and the depletion of U.S. petroleum resources. The use of ethanol as a neat fuel can dramatically impact these problems and go a long way towards their resolution.

Today, 50 percent of U.S. petroleum consumption comes from oil imports. Not only do these oil imports make a serious contribution to the balance of payments deficit (between 30-40 percent) they also affect the strategic independence of U.S. government policy. The very fact that half the transportation fuels of the country are imported puts the country in very vulnerable position. In addition to the strategic aspect, the economic impact highlighted by the recent OPEC decision to raise oil prices by 15 percent will only cause the U.S. balance of payments deficit to increase.

Furthermore, there is increased concern over air quality standards in the nation stemming from the fact that half of all Americans live in areas that fail to meet Federal clean air standards. The combustion of conventional hydrocarbon transportation fuels is a major cause of the nation's ground level air pollution. It also contributes about 27 percent of the CO_2 released into the atmosphere in the United States each year. It is believed by many experts that the accumulation of CO_2 and other gases will lead to the warming of the earth thus having severe climatic, environmental and socio-economic consequences.

Using ethanol as a transportation fuel on a widespread basis will significantly improve the U.S. balance of payments by reducing oil imports and reducing the smog levels in cities because pure ethanol has a high combustion efficiency and together with ETBE, an oxygenated fuel additive, will reduce carbon monoxide emissions.

The Solar Energy Research Institute (SERI) has proposed the use of lignocellulosic materials (such as wood) to produce ethanol because of their low cost and their huge potential availability. Using a renewable feedstock source such as wood is a long term solution to the problem of dwindling petroleum reserves. It can also be argued that by using ethanol from lignocellulose there will be no net contribution of CO_2 to the atmosphere. This is because the CO_2 released during ethanol production and combustion will be absorbed during the growth of new biomass materials replacing those utilized during production.

Until now, the barrier to widespread ethanol use has been the lack of appropriate technology that would reduce the cost of ethanol to a reasonable level. Over the last five years, SERI has made significant improvements in the ethanol from lignocellulosic wood process and has developed a new process that incorporates these recent developments. The current process is a snapshot in time of the development work done to date.

As part of its ongoing program, SERI has developed a process design for large scale plant, based primarily on experimental data, to determine the economic feasibility of such a plant.

Accordingly, SERI has retained Chem Systems to provide an independent technical and economic evaluation of its plant design. Chem Systems' analysis is based on its extensive experience with commercial alcohol plants as well as familiarity with earlier work and ongoing developments in biomass to ethanol technology.

In addition, Chem Systems' awareness of the technology allowed for discussions with appropriate equipment manufacturers regarding equipment feasibility and current costs. A review of each process section of the SERI plant design is discussed below along with Chem Systems' detailed economic analysis.

B. Conclusions

Following the review of SERI's process design Chem Systems has concluded the following:

- The overall process concept appears to feasible and is generally supported by SERI and related laboratory data as well reasonable engineering judgement.

- The next step in the process development and scale-up program needs to be the construction of a pilot scale plant with all process steps integrated to verify data assumptions especially for a commercial scale plant.

- Vendor laboratory experiments are necessary to verify large scale equipment feasibility (e.g., disc refiner, impregnator, prehydrolysis reactor, etc.).

- Based on the current design, the economics for the production of the ethanol are much improved over previous (mid-1980s) designs. At the base case wood feed rate (1920 short tpd) and base case yield (68 percent), the price of ethanol is $1.27 per gallon including 20 percent capital charges.

- Initial laboratory results make it appear that improved overall yields are feasible. Many possibilities for yield improvements have been proposed. Assuming that the necessary research and development efforts will continue and that these yield improvements are proven, it could significantly reduce the cost of ethanol production from wood. For example, if carbohydrate yield can approach 90 percent, at this point the price of ethanol can be reduced to 96.5 cents per gallon at the 1920 tpd wood feed case. This assumes constant investment and wood cost, including 20 percent capital charges.

- If an analysis is made for a large plant (5 times the base case wood feed capacity) then the ethanol price is estimated to be 102.0 cents per gallon. If one assumes the same yield improvements as above, then the ethanol price is reduced to 78.1 cents per gallon. Both values include 20 percent capital charges.

- The effect of wood cost on ethanol price for various cases is illustrated in Figure I-B-1. Using the case of a large plant at 90 percent yield and a wood cost of $34/dry ton, which according to SERI is the production goal of Oak Ridge National Laboratory for energy crops, the ethanol price is reduced to 71.4 cents per gallon.

- Figure I-B-2 shows the effect of improved yield on ethanol price for the base case wood feed rate and the large plant including a capital investment sensitivity. With wood at $42/dry ton and yield at 90 percent, assuming a large plant with 15 percent investment reduction, the ethanol price would be 73.2 cents per gallon.

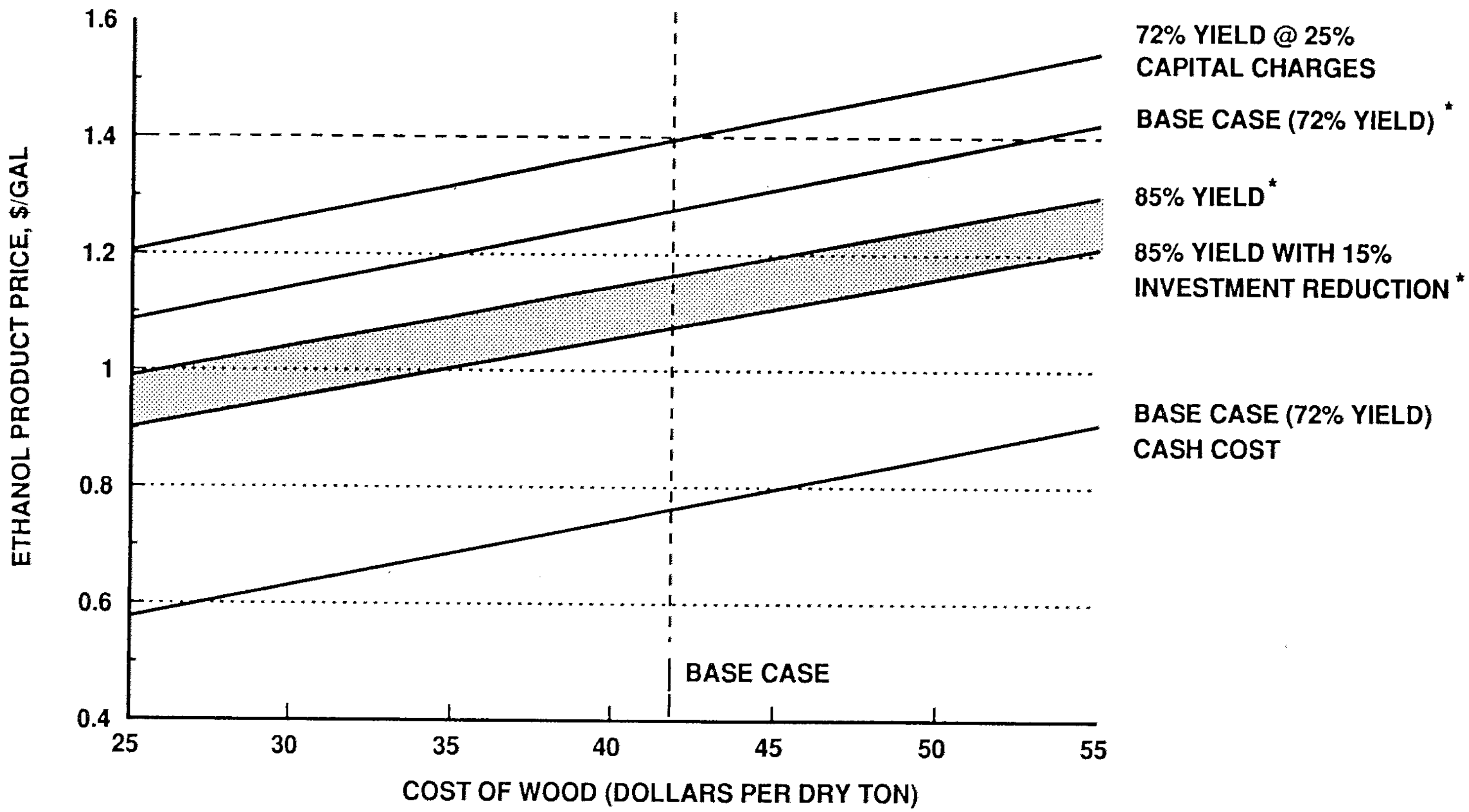
- If, in addition to the above improvements (yield, plant size, capital reduction and feedstock cost) efforts are made to reduce power consumption, optimize other aspects of the process, and increase the carbohydrate content of the feedstock, one could expect the ethanol price to be reduced even further.

? Doesn't say anything

They didn't determine the effect of yield on capital charges? This is crude analysis

FIGURE I-B-1

EFFECT OF WOOD COST ON ETHANOL PRODUCT PRICE AT VARYING CONDITIONS (USGC-4Q 1987) 160,000 LB/HR DRY WOOD FEED



* INCLUDES 20% CAPITAL CHARGES

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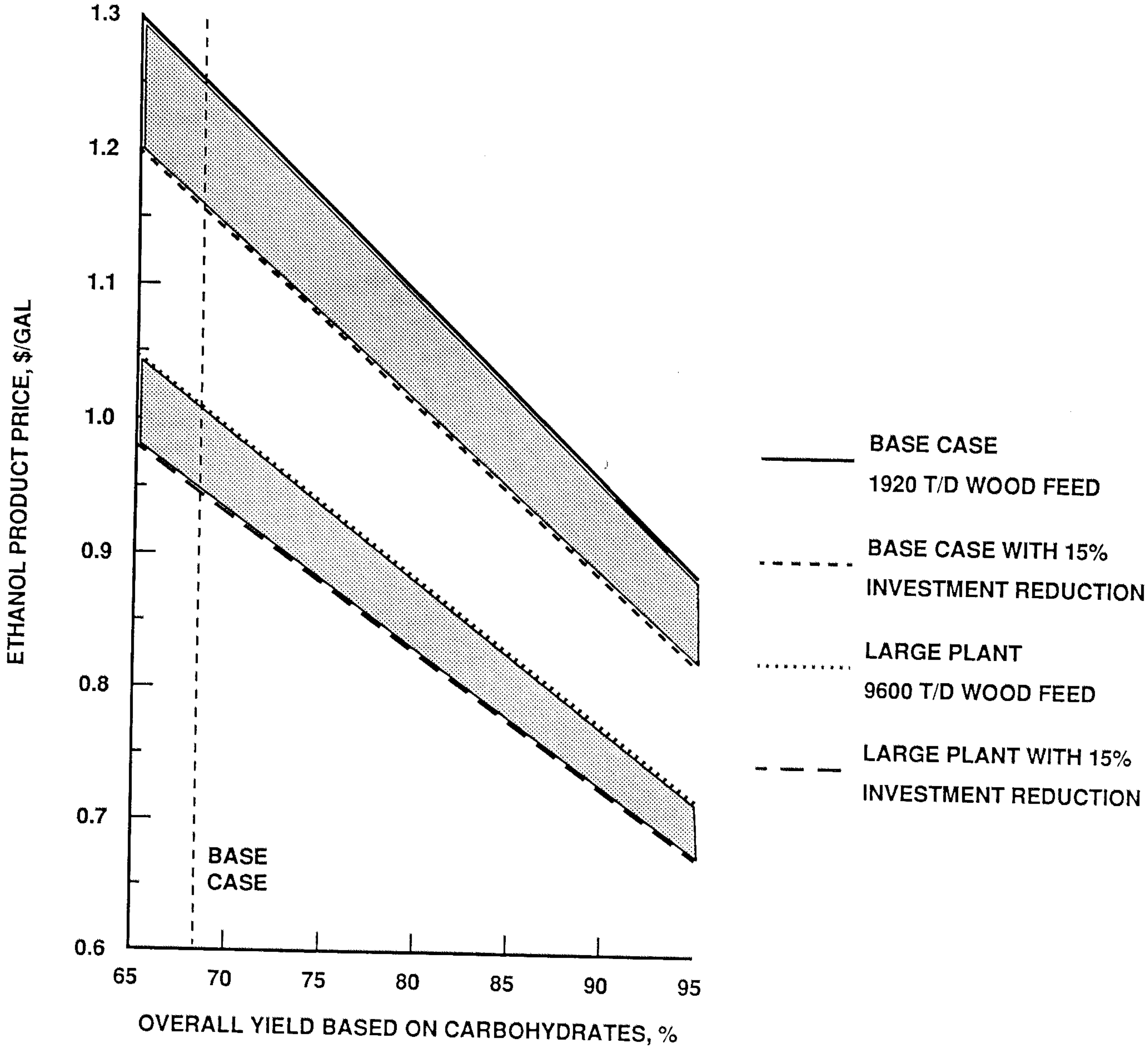
FIGURE I-B-2

EFFECT OF OVERALL YIELD ON ETHANOL PRODUCT PRICE
AT VARYING CONDITIONS
(USGC-4Q 1987, $42/DRY TON OF WOOD)
20% CAPITAL CHARGES



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II. PROCESS EVALUATION

A. Introduction

The current process of wood to ethanol proposed by SERI, is a simplified, straightforward process which contains significant improvements over processes developed in the early eighties. The major improvements lie in the following three areas:

- Xylose fermentation to ethanol
- Simultaneous Saccharification and Fermentation
- Elimination of numerous separation and concentration steps.

In previous designs, the xylan component of wood was hydrolyzed to xylose and then converted to furfural. Although a by-product credit was given to furfural, in the long term when ethanol production would be widespread, there would be a glut on the market and the value of this furfural by-product would be questionable. Today, SERI has a process that can ferment xylose to ethanol reducing the amount of xylose converted to furfural. Xylose fermentation by itself can increase the production of ethanol by 25 percent through the increased yield of ethanol.

The Simultaneous Saccharification and Fermentation (SSF) process has several advantages over the previous SHF (Separate Hydrolysis and Fermentation) process. The key advantage is in the reduction in end-product inhibition of the cellulase enzyme complex at high glucose concentrations. This no longer occurs because the glucose that is formed in a SSF reactor is converted very quickly to ethanol and therefore does not build-up in concentration.

This lack of inhibition allows for greatly reduced enzyme loading (from 33 to 7 IU/g cellulose) which dramatically cuts the cost of enzyme production. The overall cellulose to ethanol yield is also increased at the same enzyme loading.

The third improvement lies in the unit operations of the process. As an example, several costly separation steps have been completely eliminated. Gypsum is not separated after neutralization and is only separated after ethanol distillation (together with the lignin). Likewise, lignin flows from one process step to another and is only removed during the first stage of ethanol distillation. There is no multiple effect evaporation of sugar solutions prior to fermentation nor is there any furfural production step.

Wow! What a mess that must be

In general, the data on which the design conditions are based comes from SERI labs. The reported yields are not the best ever achieved but rather conservative and reproducible values that form a very reasonable basis for design. The yields are not optimum values but rather a snapshot in time reflecting the current state of process development. Improvements are expected as research and development work continues.

The major drawback in the design basis is the lack of actual experimental data from running the process on an integrated basis, i.e. running all the process steps in series using effluent from one step as the feed to the next step. The effect of leaving lignin in the feed stream to the xylose fermentation, or allowing gypsum to be present throughout the process or having an initial ethanol component in the feed to SSF has not been fully investigated. SERI plans to run an integrated process in the near future to demonstrate that the process will run as proposed.

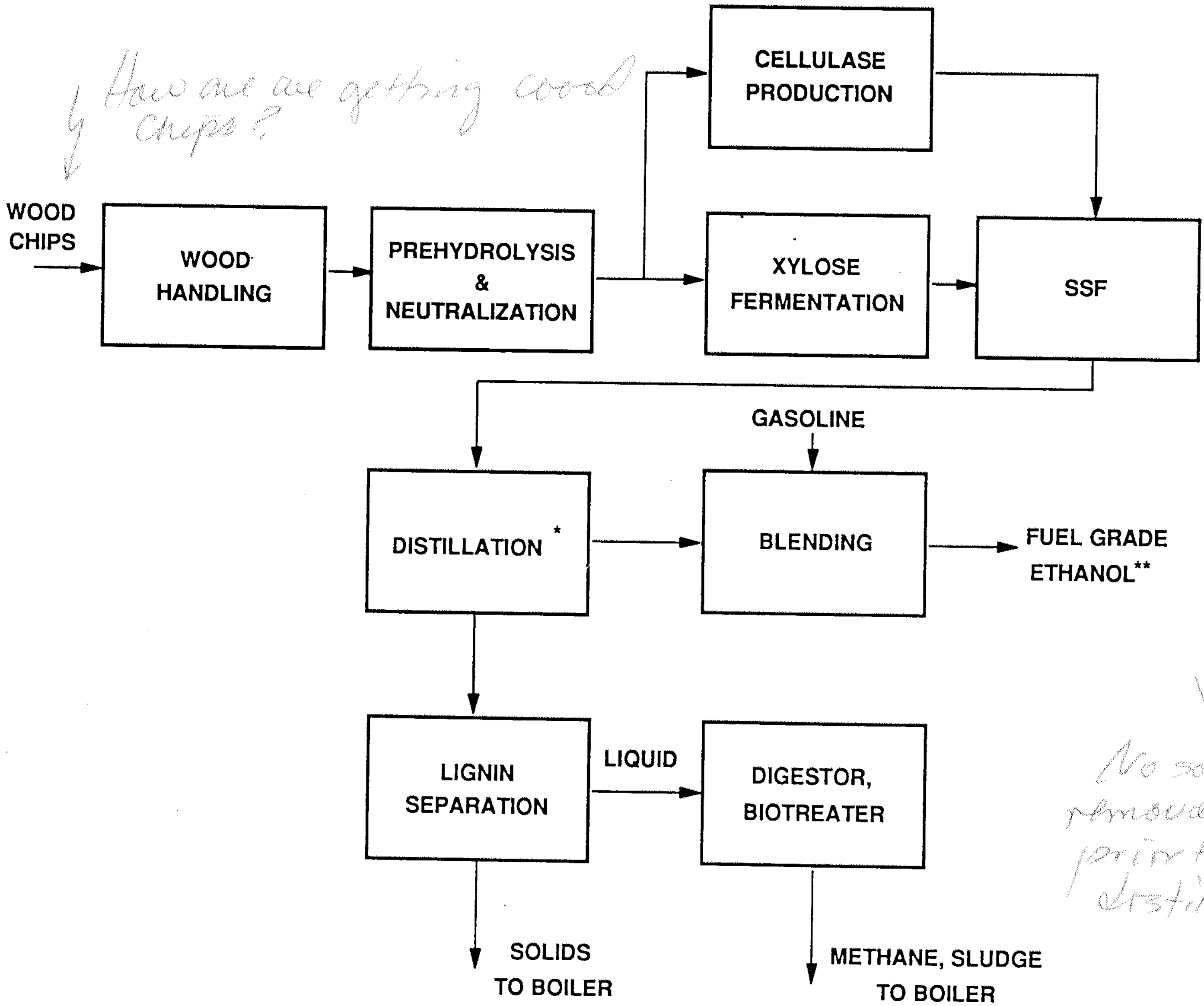
What follows is a description of the SERI wood to ethanol process. The attached block diagram (Figure II-A-1) outlines the main process units and flows. For each section there appears the design basis, a brief description, and Chem Systems' comments on the potential problems or possibilities that each process section contains. A separate block diagram appears for each section.

Is there any data on the effect of gypsum and lignin in each of these unit operations? Doesn't have to be "integrated" process data.

FIGURE II-A-1

BLOCK FLOW DIAGRAM
ETHANOL FROM WOOD PROCESS

How are we getting wood chips?



No solids removal prior to distill?

* SEPARATION TO AZEOTROPIC ETHANOL — what about breaking with membrane?

** 90.3 WT% ETHANOL
4.7 WT% WATER
5.0 WT% GASOLINE

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B. Wood Handling and Size Reduction (Section 100)

Design Basis: The composition of the wood used in the material balance is a typical hardwood with the following composition:

	Weight %
Cellulose:	46.2
Xylan:	24.0
Lignin:	24.0
Solubles:	5.6
Ash:	0.2
Total	**100**

The feed rate is 160,000 pounds of dry wood per hour. Wood chips are assumed to contain 50 percent water.

Wood chips of approximately one inch in size are delivered to the plant in large 23 ton trucks. An outside contractor will deliver the wood chips on a one to two shift a day, five day a week delivery schedule. There will be two weeks of storage on site.

Process Description: A block flow diagram is shown in Figure II-B-1. Wood chips are offloaded to a washing flume from three separate receiving stations with scales. The chips are transported from the water to a wood chip pile via a staking conveyor and are then fed to a disc refiner which reduces the size to 1mm (0.04 inches) - 3mm (0.12 inches).

Comments: Wood chip procurement is a complex operation and will require a dedicated system including reforestation, logging, debarking, chipping, handling, offsite storage and transportation to the ethanol plant on a large scale. Therefore, the smooth operation of the ethanol plant is dependent on achieving a reliable wood chip delivery system at a reasonable cost.

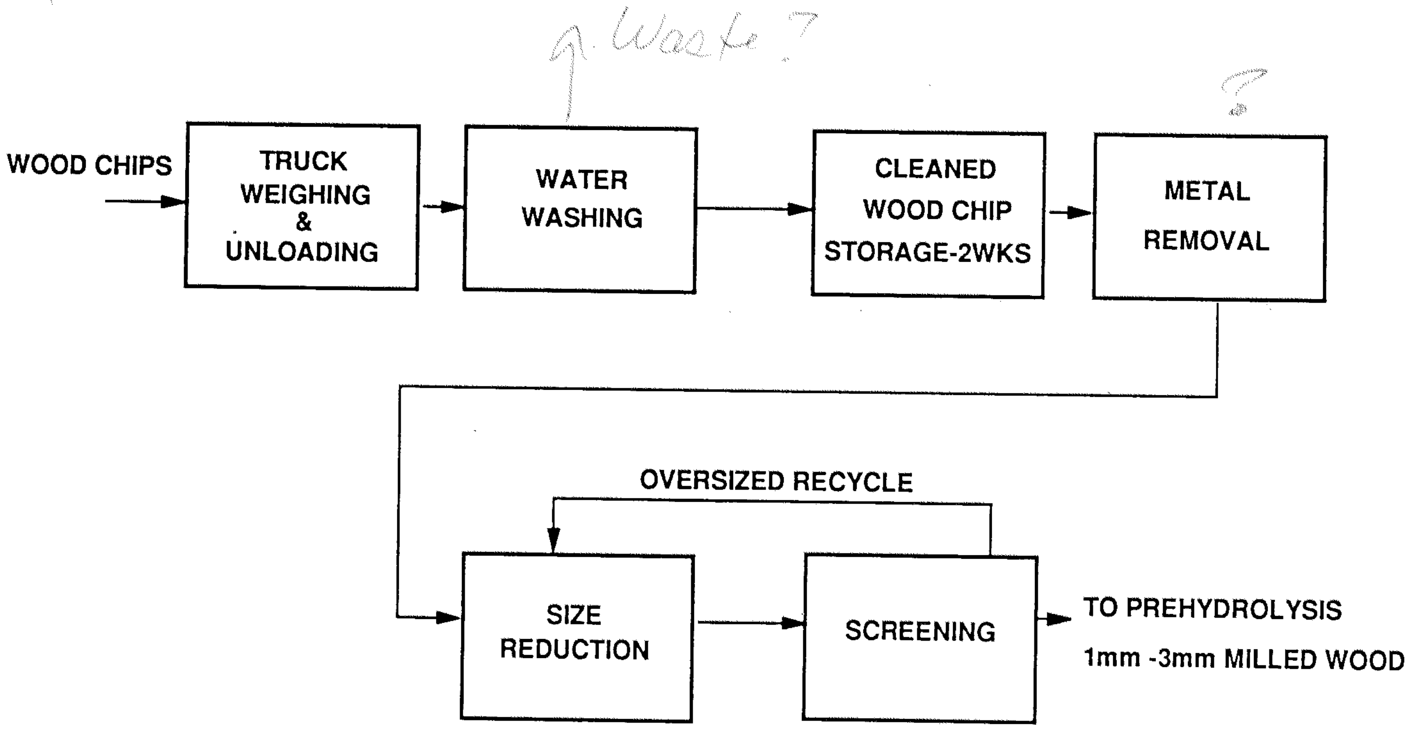
What they are really saying is that this has to be integral part of plant. Does the cost of raw matl's cover this?

FIGURE II-B-1

BLOCK FLOW DIAGRAM

SECTION 100

WOOD HANDLING



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Approximately one-third of the power requirement for the entire plant is used in the chip milling operation. The original SERI design incorporated three knife mills. Upon investigation it was found that these mills do not have the capacity required for a reasonable design. Instead of three knife mills, four disc refiners will be required (e.g. Sprout-Bauer Model 45-1B). The horsepower requirements have been adjusted accordingly. It is important to run trials using this system to verify process parameters for the desired product size including various methods of recycling oversized chips.

C. Prehydrolysis and Neutralization (Section 200)

Design Basis: The values for yields are based on SERI lab and pilot tests. Residence time and temperature have been adjusted to maximize xylan to xylose conversion. The system is not yet optimal. The design conditions for the prehydrolysis step are as follows:

Temperature:	160 °C
Residence time:	10 minutes
Xylan converted to xylose:	80 %
Xylan converted to furfural:	13 %
Xylan unconverted:	7 %
Cellulose to glucose:	3 %
Cellulose to HMF:	0.1 %
Cellulose unconverted:	96.9 %

Process Description: As shown in Figure II-C-1, milled chips from the disc refiner are fed into a screw feeder that feeds the wood into an impregnator where live steam and dilute sulfuric acid are injected. The residence time is 10 minutes at 100 °C. The impregnator discharges the wood through a rotary valve to the prehydrolysis reactor operating at the above conditions. Live steam under pressure is injected into this reactor to heat the material up to reaction temperature under 6 atm pressure. This step opens the wood to expose the cellulose for future hydrolysis and converts xylan to xylose.

After pressure letdown in the flash tank, the hydrolyzate is neutralized with lime. Prior to neutralization, recycled water is added to dilute the stream down to a slurry of 12 percent solids. After neutralization, the stream flows straight to xylose fermentation without any gypsum removal. A 2 percent slipstream is taken off at this point and sent to cellulase production. 12 % Solids

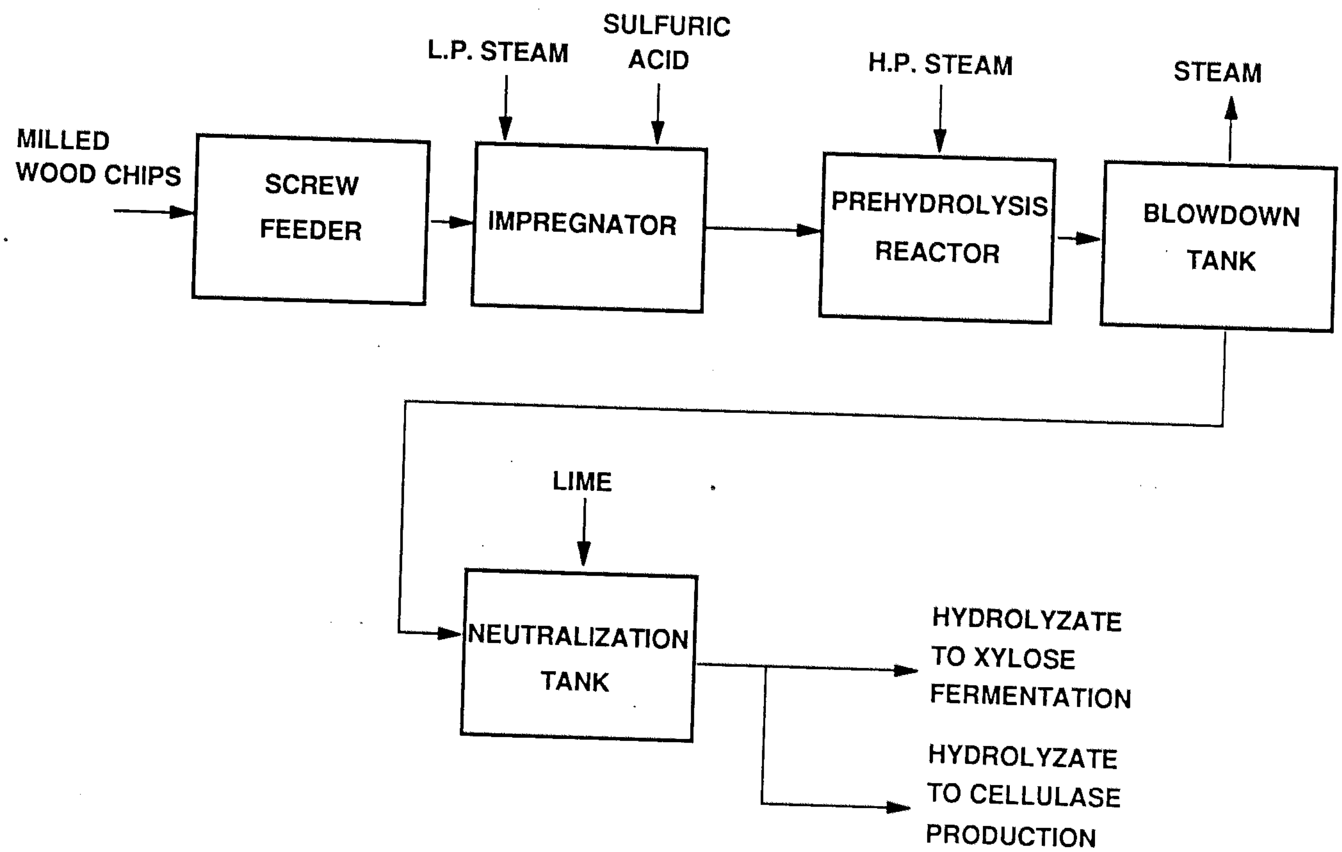
Comments: Most of the work at SERI on the impregnation and prehydrolysis steps was done on a batch reactor with a standard agitator. Although some work has been done on continuous prehydrolysis, both units should be run continuously at the process conditions to confirm the batch results. Nevertheless, xylose yields as high as 90 percent

FIGURE II-C-1

BLOCK FLOW DIAGRAM

SECTION 200

PREHYDROLYSIS AND NEUTRALIZATION



have been achieved on the lab scale. While this higher value is currently not constantly reproducable, achieving this value on a regular basis is not unreasonable.

While gypsum was removed in previous designs, it is not removed in this section. SERI contends that gypsum would be inert to the "bugs" and not affect the biological processes. Since there is no multiple- effect evaporator running at elevated temperatures, the risk of plating out gypsum at higher temperatures is small. The only place where the temperature is higher than 100 °C, the temperature at which neutralization takes place, is in the bottoms of the beer column (the first of two columns in the ethanol distillation system) and there the temperature is only 107 °C.

Chem Systems has confirmed that the type of prehydrolysis equipment used is based on existing equipment used in the pulp and paper industry and manufactured by companies such as Black Clawson. However, for a plant of this capacity, the vendor recommends two separate lines. This would be feasible although each proposed reactor would be larger than current operating equipment. Because of the economic consideration and vendor recommendation, Chem Systems' cost estimate is based on two prehydrolyis lines.

The material in the neutralization tank following prehydrolysis has a solids content of 12 percent. SERI has mixed and pumped 10 percent material and believe the 12 percent solids level should not present any difficulties. This needs to be confirmed.

Wow!

D. Xylose Fermentation (Section 300)

Design Basis: The data for this step is based on SERI lab runs (5 liter reactors) using purchased xylose. The system used is based on work done at the University of Florida using E.coli.. Also, T.V.A. labs ran the xylose fermentation on actual prehydrolysis effluent with the cellulose and lignin present and found no problems operating with all those solids present. The design basis is as follows:

Method of operation:	Continuous cascade ?
Xylose available:	95 %
Xylose converted:	90 %
Fermentation time:	2 days
pH:	7.0
Temperature:	37 °C

Continuous? Needs to be demonstrated. Contamination?

Process Description: The feed from neutralization enters the xylose fermenters directly, as shown in Figure II-D-1. Beforehand, a small (3 percent) slipstream is fed to the xylose seed fermenters as a nutrient for growing the E. coli.. The seed fermenters continuously feed cell mass into the xylose fermenters. These are operated as a series of CSTR's. The fermenters (eight total) are large (750,000 gallons) agitated vessels. Flow from tank to tank is by gravity. Temperature is maintained using cooling coils in the tank.

Wow!

Comments: The inclusion of this step is a major advance in the ethanol from wood process, increasing the ethanol production by 25 percent over earlier cases. The assumption that 95 percent of the xylose is available is reasonable and yields are high. There isn't a great deal of room for improvement on the yields since selectivity is almost 100 percent. An optimum yield may be around 90 percent.

SERI has to run this fermentation on actual effluent from its prehydrolysis/neutralization steps. This is needed in order to confirm that the high yields can be maintained in a 12

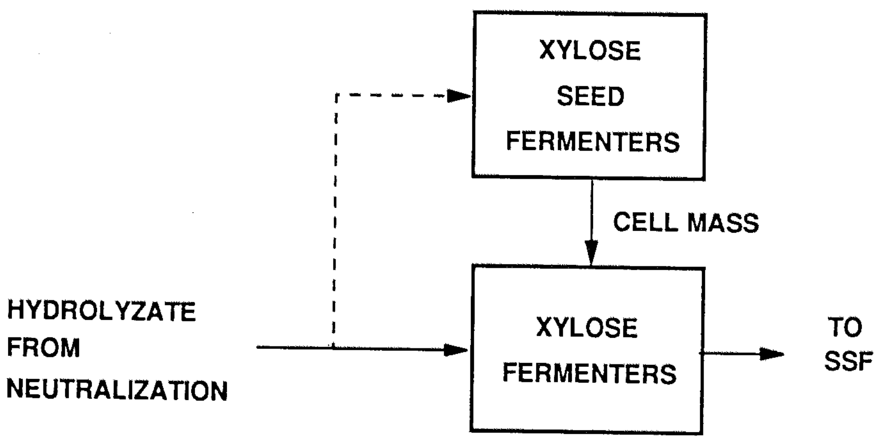
I agree.

FIGURE II-D-1

BLOCK FLOW DIAGRAM

SECTION 300

XYLOSE FERMENTATION



percent solids slurry feedstream, containing unconverted xylan, gypsum, furfural, hydroxymethyl furfural and other components.

An area of optimization would be to provide a system to recover ethanol from the CO_2/ethanol offgas stream. The current loss is nearly 1 percent of the total ethanol production.

Could be significant

E. Cellulase Production (Section 400)

Design Basis: The data are based on SERI lab experiments run on a batch basis. The parameters reflect the average performance and are considered by SERI to be conservative. The design basis is as follows:

Method of operation:	Batch
Temperature:	28 °C
Pressure:	10 psig
Fermentation time:	5.5 days
Cellulase yield:	202 IU/g cellulose

Process Description: As shown in Figure II-E-1, the slipstream is fed to a batch fermenter for a six day total batch cycle. Seed fermenters feed the main fermenter with cell mass. Nutrients and fermentation air enter separately. Chilled water is used to cool the fermenters which are agitated 250,000 gallon vessels. The cellulase fluid is held in a hold tank and is fed continuously to the next step which is SSF. Big

Comments: Although the parameters are an average literature value, the lab experiments run by SERI were on ideal substrates. These experiments must be run on actual substrates containing lignin and other soluble solids in order to confirm the design basis. SERI will attempt to make improvements in cellulase yield and growth rate. However, improvements in cellulase production though, will not impact seriously on the cost of production.

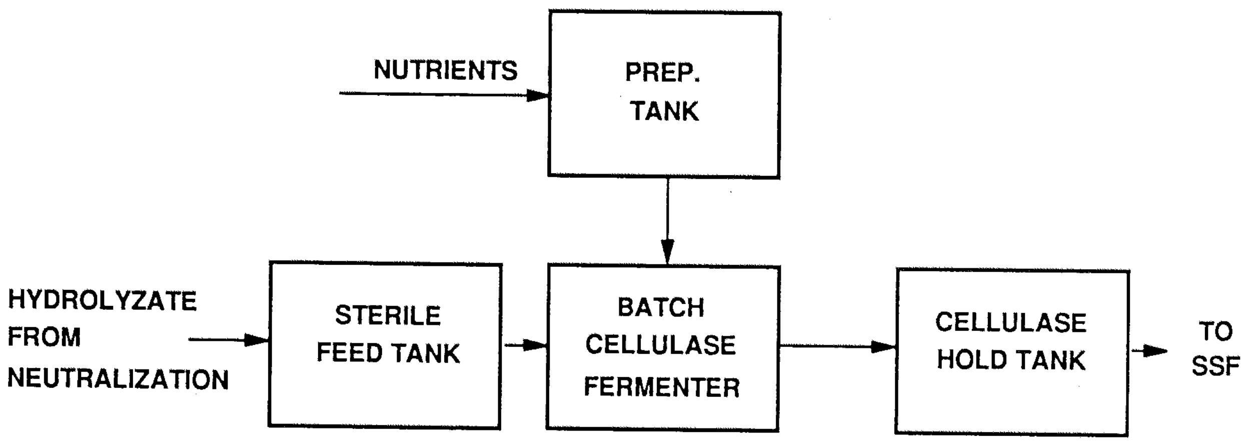
This design seems premature.

FIGURE II-E-1

BLOCK FLOW DIAGRAM

SECTION 400

CELLULASE PRODUCTION



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F. SSF- Simultaneous Saccharification and Fermentation (Section 500)

Design Basis: The SSF step is the key step in SERI's process. The design parameters are based on SERI batch lab experiments. The feed used in the lab was lignin and cellulose separated from the liquid after the prehydrolysis step and run on the 50 gallon scale. The solids were reconstituted to the appropriate concentration before being used in the experiment. The design basis is as follows:

Mode of operation:	Continuous cascade
Temperature:	37 °C
Residence time:	7 days
Cellulose converted	
to ethanol:	72.0 %
to fusel oils:	0.1 %
to glycerol and acetaldehyde:	4.9 %
to the cells:	10.0 %
Cellulose unconverted:	13.0 %

How does gravity flow work?

incredible!

Process Description: Figure II-F-1 presents a block flow diagram for this section. The effluent from the last xylose fermenter flows directly to the first of twenty-seven (27) SSF fermenters (each with 750,000 gallon capacity). Each tank is agitated using very low power requirements (0.1 HP per 1000 gals.) and the residence time per tank is approximately over 6 hours. Cell mass is continuously fed into the first fermenter from the SSF seed fermenters. There is no recycle of cell mass and flow from tank to tank is by gravity. The effluent from the last SSF fermenter enters the ethanol distillation section.

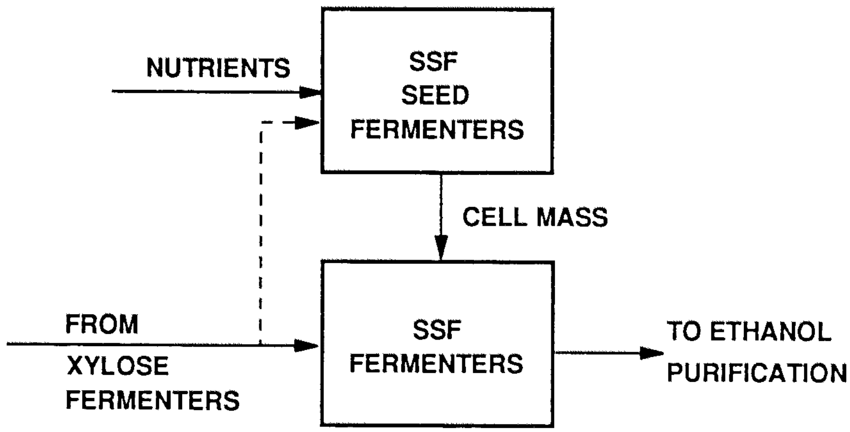
Comments: As in the previous section, the major issue with the SERI lab data is that actual material that passed through all the process stages is not used. The SSF experiments done by SERI did not contain xylan, soluble solids, xylose, gypsum (soluble and insoluble), xylose fermentation cell mass and ethanol (1.5 percent) produced in the xylose fermentation step.

FIGURE II-F-1

BLOCK FLOW DIAGRAM

SECTION 500

SIMULTANEOUS SACCHARIFICATION AND FERMENTATION (SSF)



What do they mean by that? Either it was or it wasn't there. Pretty obnoxious.

SERI claims that the key material that could interfere with the fermentation, lignin, was present in the lab runs. They believe that gypsum should be inert to the process and have no effect. The initial ethanol concentration, according to SERI, should not affect the process since ethanol inhibition occurs at higher concentrations. In addition, the SERI experiment ranged from 0 percent ethanol to 4.5 percent ethanol and therefore covers the actual operation conditions of 1.5- 4.5 percent. The other materials mentioned above are believed by SERI to not affect the results.

I agree

Running an integrated system in a continuous mode, though, should be a key priority as SERI continues its research.

The yields used are reasonable and reproducible. The major area for process improvement lies in increasing the SSF yield. This is because the yield (72 percent) is relatively low compared to the yields for other process steps and the impact on the cost of production is significant for every percentage point increase in SSF yield.

The enzyme loading (7 IU) is relatively low so that a cost analysis should be made to evaluate the relative benefit to cost ratio of increasing enzyme loadings (increasing cost) to achieve increase cellulose yields (cost savings). SERI feels confident that improvements in yield will be achieved in the future as they have done so albeit not on a reproducible basis. There are lots of comments about data which are not shown.

?

The hydraulic data used in the design comes from information received by SERI researchers during several visits to commercial corn to ethanol producers. These producers today operate large CSTR tanks in series, using gravity feed only, and using very low HP ratings for the agitators. These facts serve as a sound basis for design. However, the ability to operate this particular system via gravity flow should be verified by engineering design. Yes. This system is probably physically different.

This makes me uncomfortable. I would like to have seen a document which says our data fed into the design to arrive @ basis

As in the xylose fermentation, an area for improvement to be investigated is recovery of ethanol from the CO_2 vent stream. The total ethanol loss is about 1.5 percent of total ethanol production.

This is also an environmental issue.

G. Ethanol Purification and Solids Separation (Section 600)

Design Basis: SERI has not performed any lab experiments on this portion of the process since the unit operations are straightforward. They are relying on previously engineered systems. Yes but this design needs to accomodate high level of solids.

The ethanol distillation process scheme is well defined in the industry. SERI uses a two column design: the first column (called the beer column) produces a distillate with a composition of 40 wt percent ethanol and 60 percent water and the second column (rectification column) distills overhead the ethanol-water azeotrope composition.

SERI is assuming that the 4 percent lignin feed to the beer column will not affect the column's operation. SERI is using a 95 percent solids recovery from the solids separation step. They are also assuming a 50 percent solids content in the centrifuge cake.

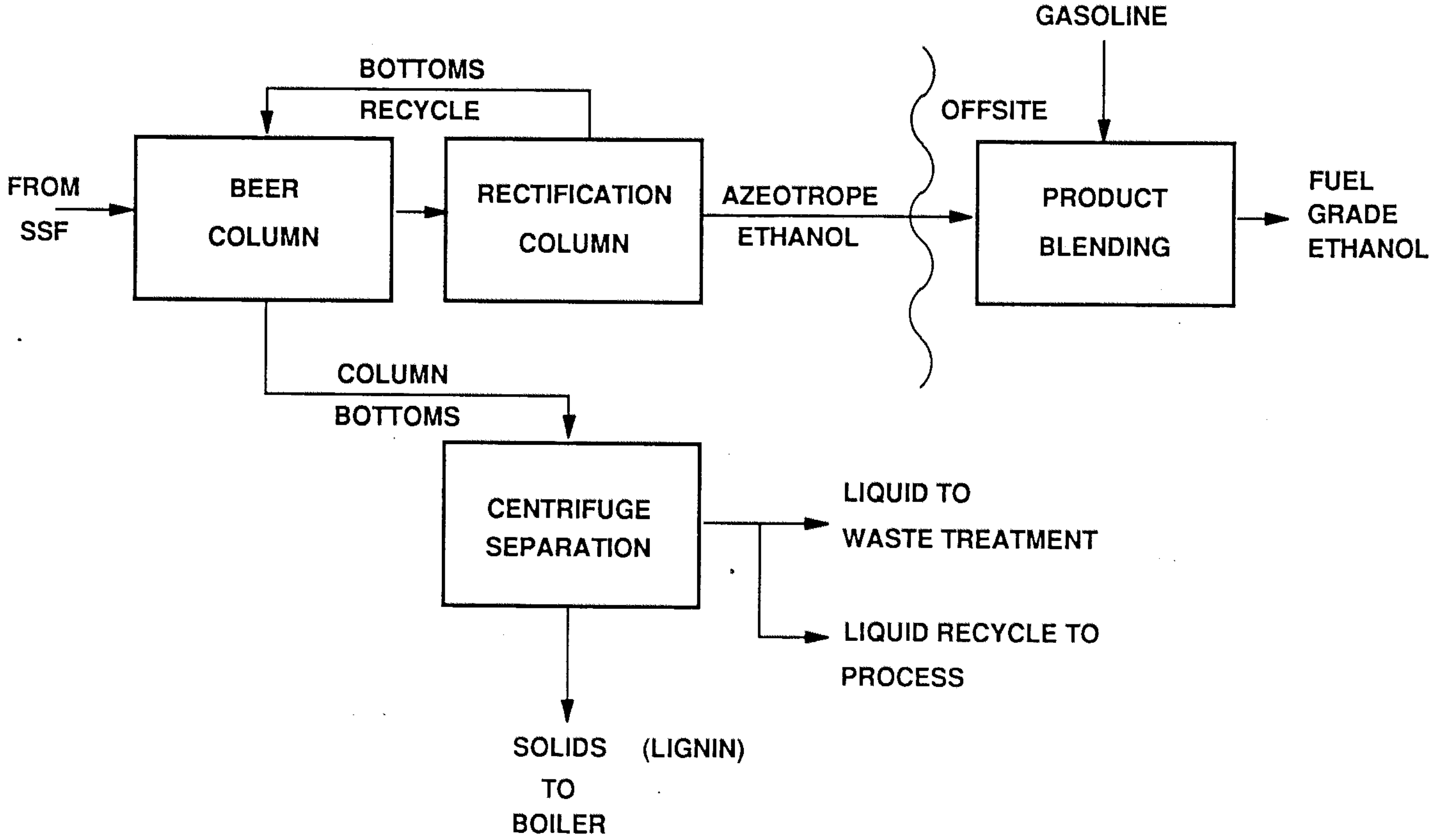
Process Description: The SSF effluent stream is heated to 100°C and fed to a degasser drum as shown in Figure II-G-1. The CO_2 is vented and the liquid is fed directly into the beer column. This stream contains 1 percent cellulose and 4 percent lignin. The column is operated under a slight pressure and the distillate is 40 percent ethanol, 60 percent water. The overheads are fed to the rectification column where azeotrope ethanol is removed overhead and the bottoms, which contain water and 4 percent ethanol, are recycled to the beer column via the degasser drum. The ethanol which is recovered from the knockback condensers on the xylose and SSF fermenters, is also added to the top of the rectification column. The overhead ethanol-water mixture is mixed in the offsite tank area with gasoline to make the final fuel grade product.

The bottoms, containing all the suspended and dissolved solids, are fed to three centrifuges in parallel, which through the use of a supernatant recycle scheme recover 95 percent of the solids. The solids leaving the centrifuge have a water content of 50 percent. Two screws feed a sludge to a special boiler where the solids are burned as fuel. The liquid from the centrifuge is divided into two approximately equal streams. The

Even after these condensers, still have a 1.5% loss?

FIGURE II-G-1

BLOCK FLOW DIAGRAM
SECTION 600
ETHANOL PURIFICATION AND SOLIDS SEPARATION



first stream is sent to waste water treatment and the second is recycled to the process.

Comments: The main issue in this process section is the nature of the distillation feed, i.e. the 5 percent solids content of the feed. The second issue is the low ethanol concentration (4 percent). SERI is basing its design on current practice in corn-to-ethanol plants. According to SERI, the percentage of solids that are fed to distillation towers operating today is considerably higher and have worse characteristics (sticky and more apt to clog the piping) than the finely divided lignin particles. Because there is no commercial operation of a distillation tower with 4 percent lignin in the feed, the viability of this step has to be proven on the pilot scale. Although there is an alternate scheme if the above does not work; separating the lignin before distillation; this might require a more complex washing cycle to reduce the amount of ethanol adhering to the solid particles which otherwise would be lost to the product.

What about solid substrate fermentation ?

The lower ethanol concentration in the feed is a result of the amount of solids in the stream leaving the neutralization tank and the yield in the SSF process step. Increasing the solids concentration entering the xylose fermentation will increase the final ethanol concentration. Here, limitations due to fluid flow at higher solids content and the impact on the rate and yield of xylose fermentation have to be explored. Increasing the yield of ethanol during SSF is more realistic in the short term and can increase the ethanol concentration slightly to the 4.5 percent range. However, this increase in yield will positively impact the production economics far more in areas of raw material cost and capital than it will in steam savings.

The current design basis assumes that the capital cost savings more than offsets higher steam cost incurred by not having process steps to increase ethanol feed concentration. This appears reasonable.

In addition to the comments made above concerning the lack of running this process on an integrated basis, it should be pointed out that this section requires the use of a recycle stream, specifically the stream that comes off the lignin separation. Although this stream does not contain solids, it does contain soluble solids, xylose, soluble gypsum, glycerol, and other materials. It is imperative that when this process is operated on an integrated basis that this recycle stream be also included to ensure that there is no build-up of materials in the equipment that can adversely affect the process.

H. Waste Treatment (Section 800)

Design Basis: The design of this section is based on commercially available technology. It contains three process systems: anaerobic digestion, aerobic digestion, and a low pressure vent system. The liquid from the lignin separation first flows to the anaerobic digester for conversion to methane. The design basis for this process is as follows:

Conversion of soluble solids:	90%
Conversion of xylose:	90%
Conversion of furfural:	90%
Conversion of glycerol:	90%

The methane is sent directly to the boiler as fuel. The liquid from the digester is sent to an aerobic digester. Here, all the remaining dissolved solids are assumed to be digested.

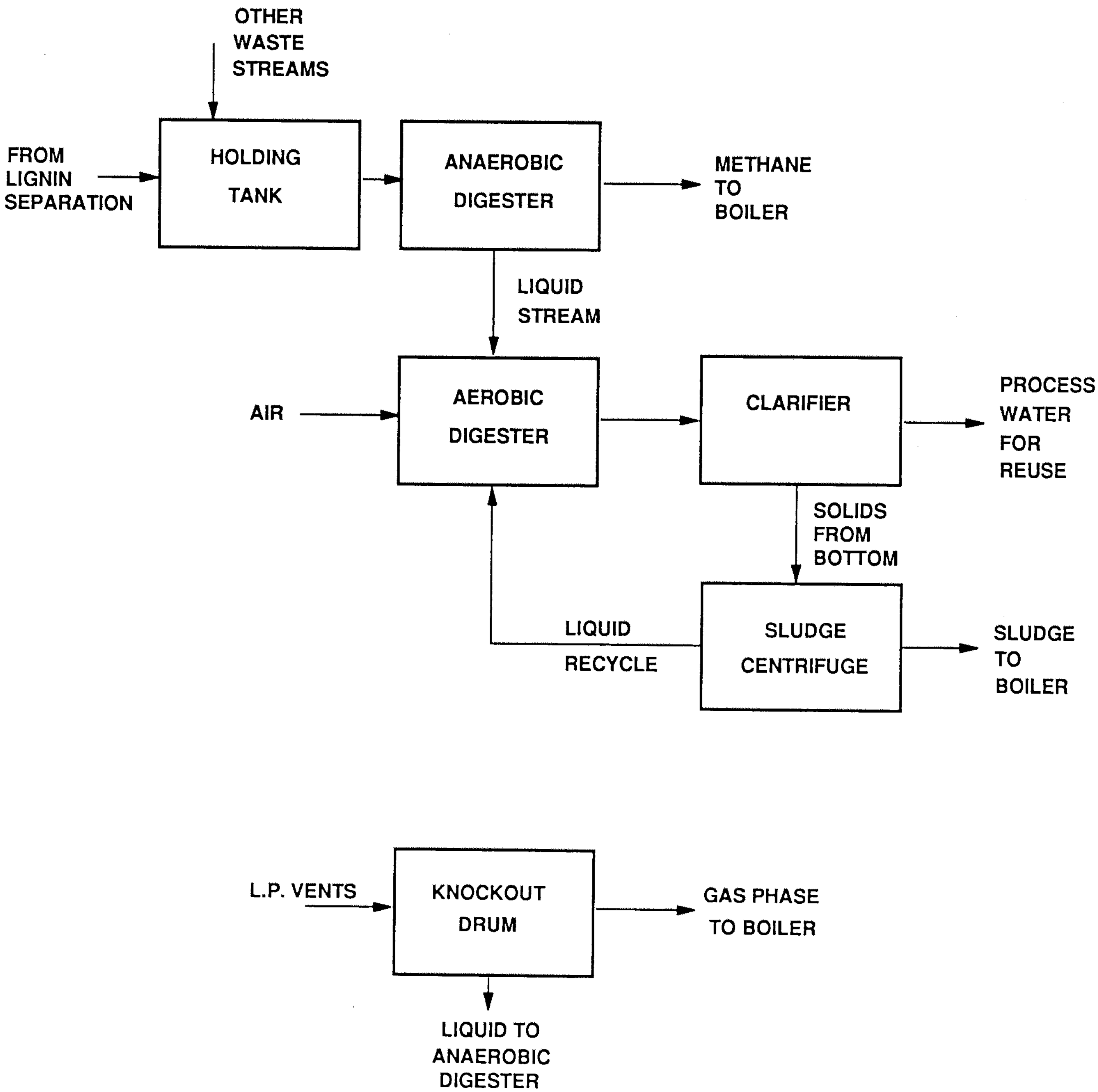
Process Description: As shown in Figure II-H-1, about half of the supernatant liquid from the centrifuge in the lignin/solids separation step after the beer column is sent to waste treatment. Here it enters a holding tank where other streams are mixed before flowing to the anaerobic digester. Ninety percent of the soluble solids, xylose, furfural and glycerol are converted to methane in this digester. The methane produced supplies a substantial amount of the heat produced in the boiler. The remaining liquid is sent to an aerobic digester. Lignin, insoluble gypsum and cell mass are not converted here. After the digester the liquid is sent to a clarifier where clear water is separated and sent to the process water tank for reuse. The bottoms of the clarifier are sent to a sludge centrifuge where they are concentrated to 15 percent solids. This sludge is sent to the boiler for burning.

All the vents from the plant are fed into a knockout drum with demister before allowing the vapors to be sent to the boiler. The entrained liquid is sent to the anaerobic digester.

Comments: The design is based on commercially available technology. However, tests should be made of actual material to confirm yields, throughputs, etc.

FIGURE II-H-1

BLOCK FLOW DIAGRAM
SECTION 800
WASTE TREATMENT



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I. Utilities (Section 900)

The equipment and production rates are discussed below in the utility summary.

III. ECONOMICS

A. Basis

Chem Systems has developed investment and cost of production estimates for a plant producing 57.9 million gallons per year of an ethanol fuel mixture based on a wood feedstock. The fuel mixture is composed of 90.3 weight percent ethanol, 4.7 weight percent water and 5.0 weight percent gasoline. The plant is based on SERI's process design as described above and the estimate is for a U.S. first quarter 1987 time frame (to be consistent with SERI previous studies).

The investment cost has been developed by determining base equipment cost for each piece of equipment. For the major equipment items, which constitute about 80 percent of total equipment costs, prices are based on current vendor estimates. The other items are based on Chem Systems' and SERI's internal data bases, primary derived from the ICARUS cost estimating computer program as well as vendor data.

From the bare equipment cost, the total investment estimate has been determined using installation factors. These factors are based on vendor information, specific plant data for fermentation type plants (as provided to SERI) and Chem Systems experience.

To be consistent with SERI's format a total fixed investment for the total complex has been estimated without distinguishing between processing units typically considered inside battery limits (ISBL) and auxiliary and supporting facilities typically designated as offsites.

Detailed estimates of utilities, capital investment and production costs are described below.

B. Utilities

1. Cogeneration System

The plant is designed with a boiler/power cogeneration system which allows for the coproduction of steam and electricity in a high pressure steam turbine. The 1,100 psia steam boiler is designed to burn, gaseous, and solid fuels derived from the various organic waste streams in the process. Methane and lignin account for the bulk of the energy value in the fuel stream fed to the boiler.

Gaseous fuels are burned directly but wet solids are first sent to a drying system that dries and fluidizes the solids into the boiler using boiler flue gas.

The steam and power generation capacities are sized in accordance with the wood feed rate. The steam turbine is an extracting type which allows for extraction of 50 psig and 150 psig steam to meet internal process requirements with the balance condensed to maximize turbine output.

Based on a steam turbine feed rate of 433.9 thousand pounds per hour of 300°F superheated 1,100 psia steam, and extraction of 41.4 thousand pounds per hour of 150 psig and 222.9 thousand pounds per hour of 50 psig steam, 36.1 megawatts of power are generated (A.B.B. vendor calculation). With total plant demand of 22.7 megawatts, this results in a 13.3 megawatt power surplus.

2. Utility Requirements

Chem Systems has performed a heat balance based on SERI's process design and material balance to determine the utility requirements which are discussed below.

a. Electricity

All of the power requirements in the plant are provided by cogenerated power. A summary of electricity requirements by plant section is shown in Table III-B-1.

TABLE III-B-1

PLANT ELECTRICITY REQUIREMENTS

Area No.	Section	Electricity Consumed (KW)
100	Wood handling	7,690
200	Pretreatment	410
300	Xylose fermentation	597
400	Cellulase production	746
500	SSF	2,237
600	Ethanol purification	485
700	Offsite tanks	89
800	Waste treatment	373
900	Utilities	10,126
	Total electricity consumed	**22,753**
	Electricity produced, kw	**36,100**
	Surplus power produced	**13,347**

The bulk of the electricity requirements are for the disc refiner (Section 100) and the air compressors and refrigeration system (Section 900). The disc refiner requirement is based on 5 Hp per dry ton per day of wood feed. This is a vendor ((Sprout-Bauer) estimate for a similar wood product. However, laboratory testing is necessary to confirm the proper morphology of the wood as an feed to prehydrolysis. Although undefined, it is conceivable that the actual power requirement of the refiner could increase substantially.

b. Steam

The proposed process requires approximately 222.9 thousand pounds per hour of 50 psig steam and 41.4 thousand pounds per hour of 150 psig steam. The steam requirements by section are shown in Table III-B-2.

TABLE III-B-2

PLANT STEAM REQUIREMENTS

		Thousand Lbs/Hr.	
		50 psig	150 psig
100	Wood handling	-	-
200	Pretreatment	30.6	41.4
300	Xylose fermentation	-	-
400	Cellulase production	0.12	-
500	SSF	-	-
600	Ethanol purification	171.1	-
700	Offsite tanks	-	-
800	Waste treatment	-	-
900	Utilities	1.1	-
	Miscellaneous	20.0	-
Total		**222.92**	**41.4**

Steam is used primarily in the impregnator and prehydrolysis reactor in the treatment (Section 200) and in the reboiler in the beer column and rectification column in ethanol purification (Section 600).

c. Cooling/Process Water

Cooling water is available from the cooling tower at 90°F. A temperature rise of 27°F has been assumed for the process users. Cooling water requirements are estimated at 19.55

million pounds per hour (39.1 thousand gpm). Table III-B-3 summarizes the cooling water requirements by sections.

TABLE III-B-3

PLANT COOLING WATER REQUIREMENTS

Area No.	Section	Cooling Water Consumed (GPM)
100	Wood handling	400
200	Pretreatment	7,935
300	Xylose fermentation	-
400	Cellulase production	-
500	SSF	-
600	Ethanol purification	10,400
700	Offsite tanks	-
800	Waste treatment	3,865
900	Utilities	16,500
	Total cooling water consumed	
	GPM	**39,100**
	Thousand lbs/hr.	**19,550**

Primarily requirements for cooling water are for the steam turbine condenser and the condensers on the ethanol purification columns. Although other alternatives exists for condensing the steam exiting the steam turbine (i.e., forced air), Chem Systems believes that a water condenser is the most economical alternative.

In a departure from the original SERI proposed design, cooling tower water for xylose fermentation and simultaneous saccharification and fermentation (SSF), has been replaced by well water. Based on Chem Systems' experience, a 3°F temperature rise for cooling water (which is necessary in the summer months) for the xylose fermentation and SSF units has proven unfeasible in some similar commercial operations.

Utilizing well water (55°F) for xylose fermentation and SSF cooling, allows for a 7.5 percent reduction of cooling water flow. The heated (assuming a 35°F rise) well water can then be utilized for process water requirements. This scheme allows for better heat integration and results in a slight reduction in capital investment for the cooling tower system.

Overall, well water requirements are calculated at 2,400 gpm (1.2 million pounds per hour). If well water were not available at a particular site, this requirement would need to be included in the chilled water system.

d. Chilled Water

Chilled water at 50°F is provided by a chilled water package with a 4,000 ton per hour refrigeration capacity. The chilled water requirements for each section are shown in Table III-B-4.

For chilled water, two different temperature rises are assumed. The lower temperature rise (3.6°F) is used to maximize the condensation of streams leaving the knock out drums in Sections 300-500. In the cellulase production (Section 400), chilled water with a 27 degree temperature rise is used to cool the fermenters and in the utilities (Section 900) chilled water is required for the interstage cooling of the air compressors.

e. Fermentation Air

Process air (45 psig) is required for the seed fermenters and the fermentation processes in xylose fermentation., cellulase production and SSF. The air requirements for these sections are shown in Table III-B-5.

In the SERI design, a pressure of 45 psig is stated. Based on vendor information (Ingersoll Rand), three compressors would be required to delivered the desired air flow. However, based on information supplied by the vendor, if it is feasible to design for a 35 psig pressure requirement, this would allow for a different machine design and the use of a single compressor. Reducing the number of compressors from three to one (excluding spares), would result in a substantial cost savings. Additionally, it is the vendor's belief that some fermentation processes utilize lower pressure air (i.e. 35 psig).

TABLE III-B-4

CHILLED WATER REQUIREMENTS

		Chilled Water @ 50°F Consumed (GPM)	
Area No.	Section	3.6°F delta, T	27°F delta, T
100	Woodhandling	-	-
200	Pretreatment	-	-
300	Xylose fermentation	405	-
400	Cellulase production	1,840	1,435
500	SSF	1,490	-
600	Ethanol purification	-	-
700	Offsite tanks	-	-
800	Waste treatment	-	-
900	Utilities	-	1,385
Total cooling water consumed			
GPM		**3,735**	**2,820**
Thousand lbs/hr.		**1,868**	**1,410**

TABLE III-B-5

PLANT PROCESS AIR REQUIREMENTS

Area No.	Section	Process Air Lb/Hr.
100	Wood handling	-
200	Pretreatment	-
300	Xylose fermentation	39,800
400	Cellulase production	48,400
500	SSF	158,300
600	Ethanol purification	-
700	Offsite tanks	-
800	Waste treatment	-
900	Utilities	-
	Total plant air	
	Pounds per hour	**246,500**
	SCFM	**56,000**

C. Capital Summary

The breakdown of the total investment cost for the ethanol plant based on the design described above is summarized in Table III-C-1.

The overall installation factor includes bulk installations, construction labor, site development, buildings, roads control room, laboratory as well as indirect costs, home office costs, engineering and design, etc. Note that the cost of the steam boiler package, with was obtained from a vendor estimate, is reported on a separate line item.

The total cost is estimated at $138 million. To be consistent with previous studies, this cost includes items such as owners costs, land, start-up and commissioning/expenses, but does not include financing and any licensing fees.

For this analysis the capital cost is put on a 1987 basis to be consistent with other SERI studies.

TABLE III–C–1

CAPITAL COST SUMMARY

	Purchased Cost $MM/ Unit	No. Req'd	Total Purchased, $MM	Source *
			$ 1990	
Wood Handling (Section 100)				
Major Equipment				
Disc Refiners	0.37	4	1.48	Sprout Bauer
Front End Loaders	0.16	3	0.47	
Belt Conveyor	0.19	1	0.19	
Other			0.36	
Section Total			2.50	
Prehydrolysis (Section 200)				
Major Equipment				
Impregnator/Prehydrolysis system	3.66	2	7.32	Black Clawson
Screw feeder	0.28	2	0.55	Black Clawson
Other			0.28	
Section Total			8.15	
Xylose Fermentation (Section 300)				
Major Equipment				
Xylose Fermenters	0.20	8	1.60	CBI
Other			0.58	
Section Total			2.18	
Cellulase Production (Section 400)				
Major Equipment				
Cellulase Fermenter	0.07	3	0.20	
Fermenter Agitator	0.08	3	0.23	
Feed Tank Agitator	0.13	1	0.13	
Other			0.39	
Section Total			0.94	
SSF Fermentation (Section 500)				
Major Equipment				
SSF Fermenter (501A)	0.20	27	5.40	CBI
SSF Fermenter (501B)	0.31	1	0.31	CBI
Seed Hold Tank (501A)	0.12	1	0.32	CBI
Seed Hold Tank (501B)	0.19	1	0.41	CBI
SSF Fermenter Agitators	0.03	27	0.81	
Other			0.56	
Section Total			7.81	

TABLE III-C-1 (Cont'd)

CAPITAL COST SUMMARY

	Purchased Cost $MM/ Unit	No. Req'd	Total Purchased, $MM	Source *
			$ 1990	
Distillation (Section 600)				
Major Equipment				
Centrifuge	0.23	3	0.68	
Beer Column	0.17	1	0.17	
Rectification Column	0.16	1	0.16	
Other			0.36	
Section Total			1.36	
Offsite Tankage (Section 700)				
Major Equipment				
Ethanol Product Tank	0.25	2	0.49	CBI
NH3 Storage Tank	0.09	2	0.17	
Fire Water Tank	0.14	1	0.14	
Other			0.30	
Section Total			1.10	
Environmental and Wastewater (Section 800)				
Major Equipment				
Secondary Clarifier	0.26	1	0.26	
LP Vent Blower	0.07	2	0.15	
Equalization Tank	0.24	1	0.24	
Other			0.80	
Section Total			1.44	
Boiler and Steam Distribution (Section 900)				
Major Equipment				
Cooling Tower System	0.73	1	0.73	
Demineralizers	0.31	2	0.62	
Condensate Polisher	0.10	2	0.20	
Turbo Generator	6.50	1	6.50	A.B.B.
Air Compressor Package	0.45	4	1.80	Ingersoll-Rand
Chilled Water Package	0.60	1	0.60	
Other			1.20	Trane
Section Total			11.65	

TABLE III–C–1 (Cont'd)

CAPITAL COST SUMMARY

		Total Purchased, $MM		Source *
		$ 1990	$ 1987	
Plant Sub–Total		37.12		
Miscellaneous Equipment		2.23		
Total – Bare Equipment		39.35		
Total Installed Cost – Factor:=	2.85	112.14		
Steam Boiler Package (w/ predryer)	installed	19.80		A.B.B.
Total Plant Investment		131.94	120.07	
Owner's Cost, Fees and Profit		13.19	12.01	
Startup Cost			6.00	
Grand Total Plant Investment			138.07	

* On a purchased equipment basis, about 80 percent of the equipment is based on current vendor budgetary estimates.

D. Production Costs

1. Base Case

A summary of the cost of production estimate for a plant producing about 58 million gallons per year of ethanol-based fuel is presented in Table III-D-1. This is based on about 2,000 STPD wood feed. The economics are on a 1987 U.S.G.C. basis. The total capital, as discussed above is estimated at $138 million.

TABLE III-D-1

ETHANOL BASED FUEL
ECONOMIC SUMMARY
(1987 - U.S. Gulf Coast)

Investment	$138 MM
Working capital	9 MM
Production cost, $/gal	
Net raw materials	0.60
Utilities	(0.05)
Direct cash cost	0.11
Allocated cash cost	0.10
Full cash cost	**0.76**
Cost plus 20% capital charges	$1.27

Cost of production can be divided into several categories:

- Raw materials
  - Primarily wood
- Utilities
  - Electricity and well water
- Operating costs
  - Includes labor for operating the plant as well as materials and labor for annual maintenance costs

- Overhead expenses
  - Includes plant overheads, taxes, and insurance

Raw materials and utilities are considered variable costs since they are function of the plant operating capacity. Operating costs and overhead expenses are fixed costs since they are independent of operating rate. The sum of variable and fixed costs is usually termed the cash cost of production. This is the actual out-of-pocket cost an owner incurs before considering depreciation of the capital investment and profits.

A summary of the cost basis is shown in Table III-D-2 and the detailed cost estimate is presented in Table III-D-3.

As in earlier reports evaluating natural gas and coal to methanol, and a biomass gasification to methanol a capital charge of 20 percent of total fixed investment plus working capital is taken as an overall capital recovery factor. This is equivalent to approximately a 10 percent discounted cash flow after tax rate of return.

In the ethanol cost estimates, raw material cost which is the largest component of the production cost is estimated at a net cost of $0.60 per gallon of fuel. The major component is wood, taken at $42 per short ton on a dry basis.

Because the plant has a cogeneration system, which uses waste materials as fuel to the boiler, the plant is a net producer of power and the only external utility is well water.

Based on a net utility credit of $0.054 per gallon of fuel product, the total variable cost of the material is estimated at $0.55 per gallon. The actual power consumption in the wood mill section is subject to actual vendor testing verification.

TABLE III-D-2

BASES FOR ETHANOL PRODUCTION COSTS

- 4th quarter, 1987
- Operating factor: 91 percent, 8,000 hours per year
- Direct overhead at 45 percent of labor and supervision
- General plant overhead at 65 percent of operating costs
- Maintenance at 3 percent of total fixed investment
- Insurance and property taxes at 1.5 percent of total fixed investment
- Working capital is recovered at the end of the life of the project and is calculated as the sum of the following three items:

  1. Accounts receivable - one month's gross cost of production (COP).*
  2. Cash - one week gross cost of production less depreciation.*
  3. Warehouse/spares - three percent replacement inside battery limits (ISBL).

  Less a fourth item:

  4. Accounts payable - one month's supply of raw materials at delivered prices.

- Capital charges at 20 percent of total capital requirements (fixed plus working capital). This charge covers depreciation recovery and a return on capital.

*Gross COP = net COP (ex. IWC) less by-product credit.

The unit cost of exported electricity is taken at 3 cents per kilowatt hour which is one cent less than purchased power. This discount is very dependent on the site (utility company) but believed to be typical for 1987 transactions.

Direct cash cost, including labor, maintenance and direct plant overheads, totals about $0.11 per gallon of product. The plant operating labor is based on 9 men per shift. This

TABLE III-D-3

COST OF PRODUCTION ESTIMATE FOR : Denatured Fuel (90.25% Ethanol)

PROCESS : SERI - Wood to Ethanol (68% Yield)

		Capital Cost	MILLION U.S. $ ORIG	BOOK	REPL
Plant Startup	1987				
Analysis Date	4Q-1987				
Location:	U.S.				
Capacity:	57.91 million gallons/yr				
	173,378 metric tons/yr	Total Fixed Inv.	138.1	138.1	138.1
On Stream Time:	8,000 hours per year	Working Capital			9.1
Throughput:	57.91 million gallons/yr				

	PRODUCTION COST SUMMARY	UNITS PER GAL	PRICE, U.S. $ /UNIT	U.S. $ PER GAL	ANNUAL U.S. $ COST MM U.S. $	PER MET TON
RAW MATERIALS	Wood (dry), ST	0.0111	42.000	0.464	26.88	
	Sulfuric Acid, lb	0.3976	0.032	0.013	0.74	
	Lime, lb	0.2940	0.023	0.007	0.38	
	Ammonia, lb	0.6296	0.041	0.026	1.50	
	Nutrients, lb	0.0181	0.115	0.002	0.12	
	Corn Steep Liquor, lb	0.0633	0.100	0.006	0.37	
	Corn Oil (Antifoam), lb	0.0039	0.240	0.001	0.05	
	Glucose, lb	0.0496	0.510	0.025	1.46	
	Gasoline/Diesel, gal	0.0570	0.770	0.044	2.54	
	Catalyst & Chemicals		0.010	0.010	0.58	
	TOTAL RAW MATERIALS			0.598	34.62	200
BY-PRODUCT CREDITS	Solids Disposal, ton	(0.00034)	20.000	0.007	0.40	
	TOTAL BY-PRODUCT CREDITS			0.007	0.40	2
NET RAW MATERIALS				0.605	35.02	202
UTILITIES	Power, KWH	(1.85300)	0.030	(0.056)	(3.22)	
	Well Water, M Gal	0.01987	0.100	0.002	0.12	
	TOTAL UTILITIES			(0.054)	(3.10)	(18)
VARIABLE COST OF PRODUCTION				0.551	31.91	184
DIRECT CASH COSTS	Labor, 41 Men	29.80 Thousand U.S. $		0.021	1.22	
	Foremen, 9 Men	34.00 Thousand U.S. $		0.005	0.31	
	Super., 1 Men	40.00 Thousand U.S. $		0.001	0.04	
	Maint., Material & Labor	3.00 % of ISBL		0.072	4.14	
	Direct Overhead	45 % Labor & Supervision		0.012	0.71	
	TOTAL DIRECT CASH COSTS			0.111	6.42	37
ALLOCATED CASH COSTS	General Plant Overhead	65 % Labor & Maintenance		0.064	3.71	
	Insurance, Property Tax	1.5 % Total Fixed Investment		0.036	2.07	
	TOTAL ALLOCATED CASH COSTS			0.100	5.78	33
FULL CASH COST OF PRODUCTION				0.762	44.11	254
NET COST OF PRODUCTION				0.762	44.11	254
	COST PLUS 20 % CAPITAL CHARGES			1.270	73.55	424

Direct cash cost, including labor, maintenance and direct plant overheads, totals about \$0.11 per gallon of product. The plant operating labor is based on 9 men per shift. This consists of 1 man in the control room, 2 in wood handling, 2 in the process, 1 in the tank farm and blending, 1 for waste handling and 2 for utilities.

Allocated cash cost which includes general plant overhead and insurance and local property taxes, contribute a total of \$0.10 to the cost of production. This results in a total cash cost of \$0.76 per gallon of fuel ethanol.

Adding a 20 percent capital recovery charge, reflecting both depreciation and return on investment, the required ethanol fuel price would be \$1.27 per gallon of fuel.

This production cost estimate does <u>not</u> include items such as shipping, packaging, research and development expenses, general, sales and administrative costs, royalties, etc.

2. <u>Sensitivities</u>

The base case economics have been based as evaluation of the SERI experimental data. Accordingly, a number of cost sensitivities have been carried out to illustrate the effect of various parameters and are discussed below.

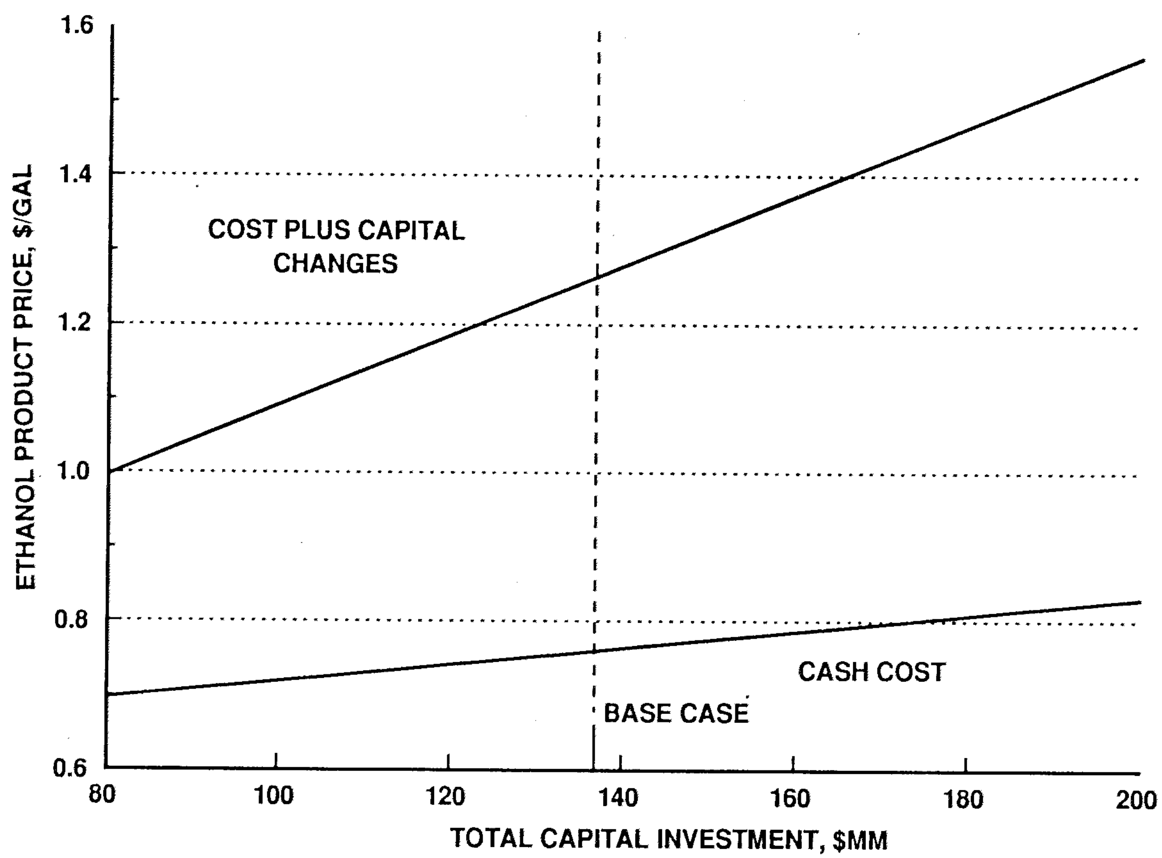
a. <u>Capital Sensitivity</u>

The investment cost has been based on a factored estimate and may vary depending on the design philosophy used in the plant. This type of budgetary estimate is considered plus or minus 30 percent accurate. Figure III-D-1 illustrates the effect on investment cost of ethanol production economics.

FIGURE III-D-1

EFFECT OF CAPITAL INVESTMENT COST ON ETHANOL PRODUCT PRICE

(US-4Q 1987) 160,000 LB/HR DRY WOOD FEED



b. Wood Cost

The wood feed system is a complex system and a major component of the plant operational feasibility. Accordingly, the actual cost of wood for an operation of this size could vary from the base estimate. Figure III-D-2 shows ethanol fuel production cost as a function of wood (dry) cost.

c. Yield

The overall yield for the base case is 68 percent based on carbohydrates (hemicellulose and cellulose). If improvements can be made in the various processing steps, the overall yield can increase substantially, SERI believes that there is a strong basis for improved yields and cites the following points:

- Xylan to xylose yields in the lab have been as high as 90 percent which proves that they can be achieved. These base case uses 80.5 percent.

- Xylose used for cellulase production and for seed fermenters can be reduced.

- Xylose to ethanol conversion using existing better organisms which have been tested in other labs can boost yield to between 90-95 percent instead of the 85.5 percent assumed in the base case.

- Cellulose for cellulase production and seed fermenters can be reduced.

- SERI claims experimental evidence that indicates that ethanol yield from cellulose during SSF can reach as high as 95 percent instead of a yield of 72 percent assumed in the base case.

Estimates have been made of the effect of increased yield on ethanol cost at a constant wood feed rate resulting in increased capacity at the same capital cost.

FIGURE III-D-2

EFFECT OF WOOD COST ON ETHANOL PRODUCT PRICE (US-4Q 1987) 1920 SHORT TONS/DAY DRY WOOD FEED

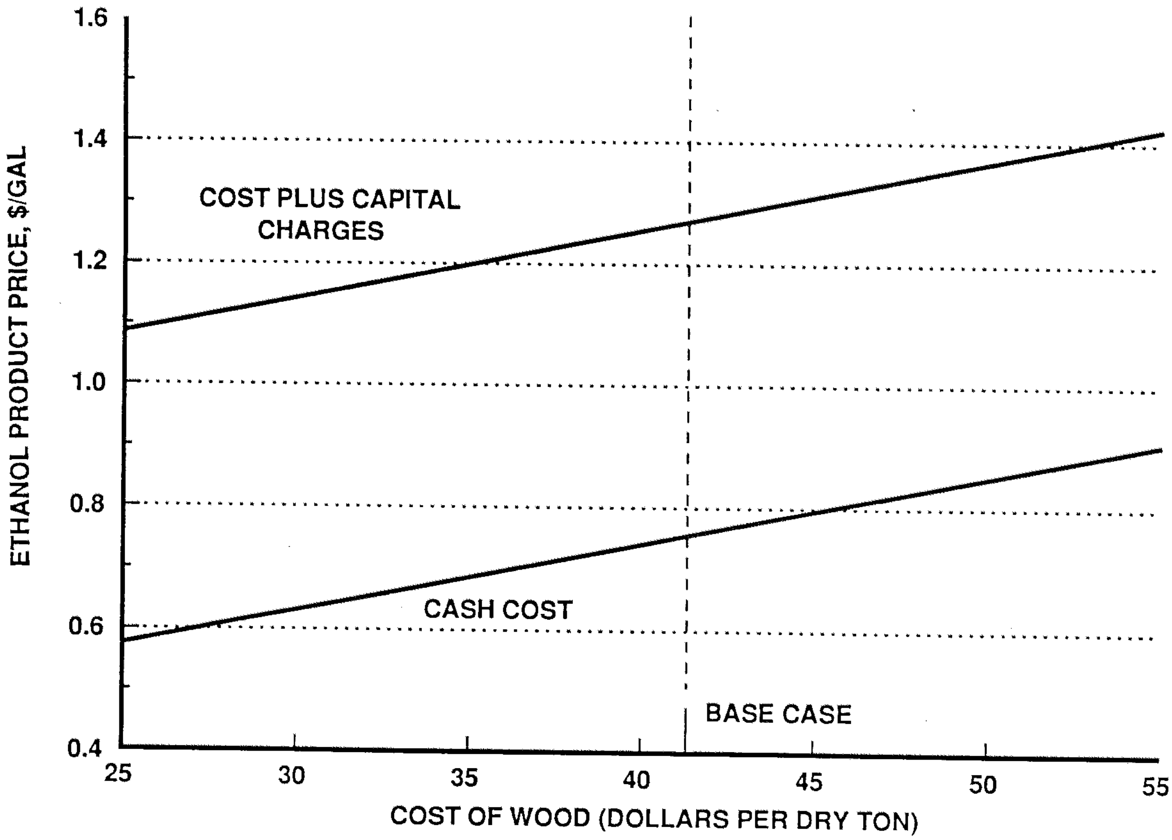


Table III-D-4 presents a cost of production estimate for a case at 90 percent overall carbohydrate yield. It should be pointed out that as the yield increases the amount of carbohydrate not converted to ethanol decreases so that the amount available for use as a boiler fuel is also reduced. Thus the improved yield is accompanied with a reduction, albeit small, of the amount of exported electricity. The captive energy requirements, both steam and electricity, can still be satisfied by the lignin component of the wood feed stock.

The effect of overall yield on ethanol prices is shown in Figure III-D-3. Capacity for the base case at various yields increases and is listed below.

Yield (%)	Capacity (millions of gal/yr)
68	57.9
75	64.6
85	73.2
90	77.5

As mentioned before, all yields and potential improvements need to be verified on actual runs made on material that flows from one process section to another. This is essential in order to take into account the effect of by-products formed during upstream steps and carried forward to subsequent steps throughout the complete process.

d. Plant Size

In order to evaluate the effects of increased capacity on ethanol price, an analysis was made for a plant having a capacity of five times the base case, or 9,600 short tons per day of dry wood feed. This case was chosen to match previous evaluations on production from methanol natural gas, coal and biomass. There is a need, of course, to verify the costs and logistics of wood collection, delivery, and renewal for such a large plant. This is beyond the scope of this study.

TABLE III-D-4

COST OF PRODUCTION ESTIMATE FOR : Denatured Fuel (90.25% Ethanol)

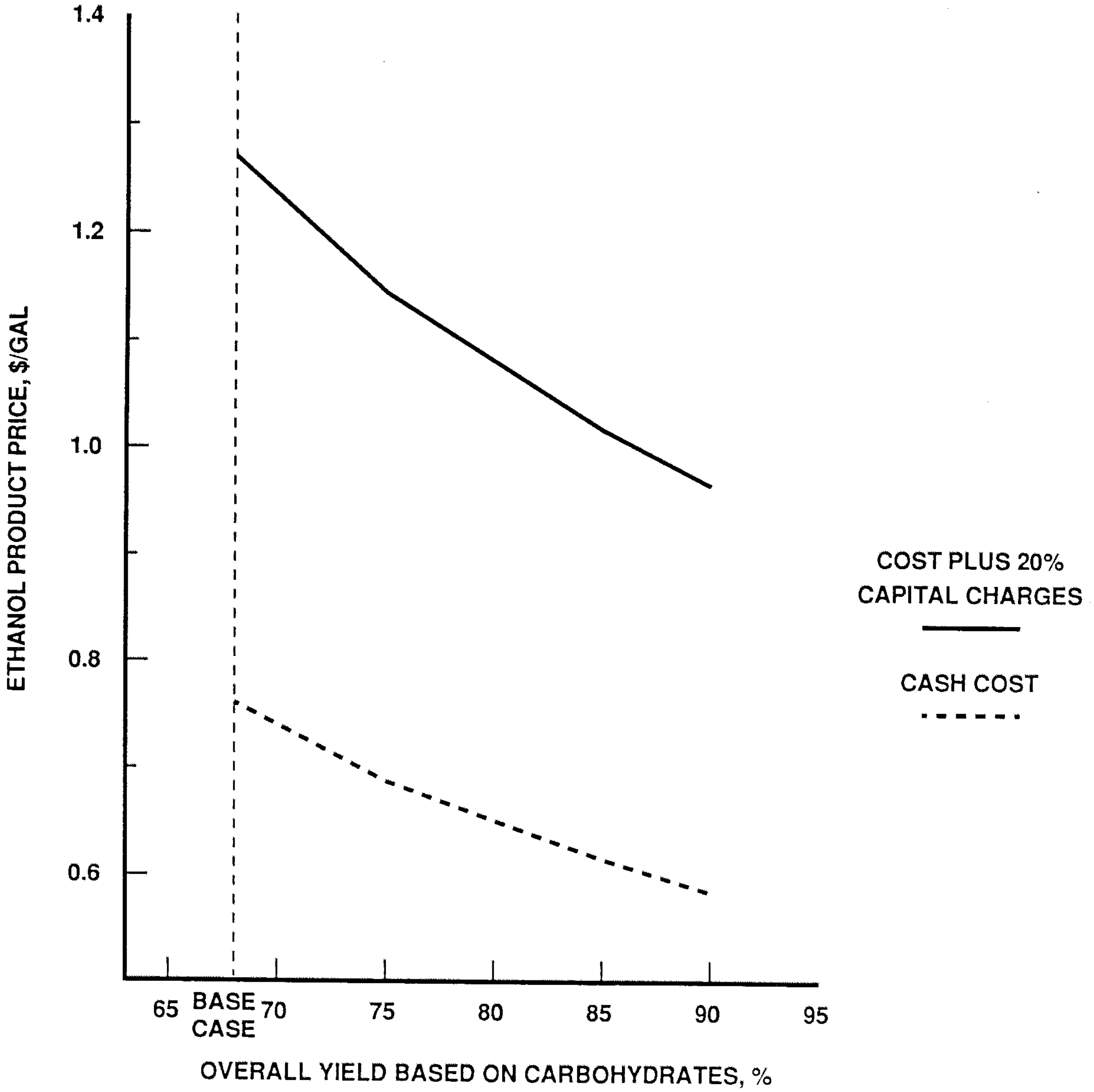
PROCESS : SERI - Wood to Ethanol - 90 % yield

			MILLION U.S. $		
		Capital Cost	ORIG	BOOK	REPL
Plant Startup	1987				
Analysis Date	4Q-1987	Battery Limits	138.1	138.1	138.1
Location:	U.S.	Offsites	0.0	0.0	0.0
Capacity:	77.50 million gallons/yr				
	232,015 metric tons/yr	Total Fixed Inv.	138.1	138.1	138.1
On Stream Time:	8,000 hours per year	Working Capital			9.2
Throughput:	77.50 million gallons/yr				

	PRODUCTION COST SUMMARY	UNITS PER GAL	PRICE, U.S. $ /UNIT	U.S. $ PER GAL	ANNUAL U.S. $ COST MM U.S. $	PER MET TON
RAW MATERIALS	Wood (dry), ST	0.0083	42.000	0.347	26.88	
	Sulfuric Acid, lb	0.2970	0.032	0.010	0.74	
	Lime, lb	0.2190	0.023	0.005	0.38	
	Ammonia, lb	0.4704	0.041	0.019	1.49	
	Nutrients, lb	0.0181	0.115	0.002	0.16	
	Corn Steep Liquor, lb	0.0633	0.100	0.006	0.49	
	Corn Oil (Antifoam), lb	0.0039	0.240	0.001	0.07	
	Glucose, lb	0.0370	0.510	0.019	1.46	
	Gasoline/Diesel, gal	0.0570	0.770	0.044	3.40	
	Catalyst & Chemicals		0.010	0.010	0.78	
	TOTAL RAW MATERIALS			0.463	35.85	155
BY-PRODUCT CREDITS	Solids Disposal, ton	(0.00034)	20.000	0.007	0.53	
	TOTAL BY-PRODUCT CREDITS			0.007	0.53	2
NET RAW MATERIALS				0.470	36.39	157
UTILITIES	Power, KWH	(1.48400)	0.030	(0.045)	(3.45)	
	Well Water, M Gal	0.01987	0.100	0.002	0.15	
	TOTAL UTILITIES			(0.043)	(3.30)	(14)
VARIABLE COST OF PRODUCTION				0.427	33.09	143
DIRECT CASH COSTS	Labor, 41 Men	29.80 Thousand U.S. $		0.016	1.22	
	Foremen, 9 Men	34.00 Thousand U.S. $		0.004	0.31	
	Super., 1 Men	40.00 Thousand U.S. $		0.001	0.04	
	Maint., Material & Labor	3.00 % of ISBL		0.053	4.14	
	Direct Overhead	45 % Labor & Supervision		0.009	0.71	
	TOTAL DIRECT CASH COSTS			0.083	6.42	28
ALLOCATED CASH COSTS	General Plant Overhead	65 % Labor & Maintenance		0.048	3.71	
	Insurance, Property Tax	1.5 % Total Fixed Investment		0.027	2.07	
	TOTAL ALLOCATED CASH COSTS			0.075	5.78	25
FULL CASH COST OF PRODUCTION				0.584	45.29	195
NET COST OF PRODUCTION				0.584	45.29	195
	COST PLUS 0 % RETURN ON TOTAL BOOK INV. PLUS WC			0.584	45.29	195
	COST PLUS 20 % RETURN ON TOTAL BOOK INV. PLUS WC			0.965	74.75	322
	COST PLUS 30 % RETURN ON TOTAL BOOK INV. PLUS WC			1.155	89.48	386

FIGURE III-D-3

EFFECT ON IMPROVED OVERALL YIELD ON ETHANOL PRODUCT PRICE (US-4Q 1987) 1920 SHORT TONS/DAY DRY WOOD FEED



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Assuming the viability of a 10,000 tons per day plant, a cost of production estimate was prepared for this plant at various overall yields. Tables III-D-5 and III-D-6 present the economics for such a plant at base case yield and 90 percent overall yield. Figure III-D-4 illustrates the effect of overall yield on ethanol product price for this large capacity case.

How did they scale up.

TABLE III-D-5

COST OF PRODUCTION ESTIMATE FOR : Denatured Fuel (90.25% Ethanol)

PROCESS : SERI - Wood to Ethanol, Large Plant (68% Yield)

Plant Startup	1987
Analysis Date	4Q-1987
Location:	U.S.
Capacity:	289.55 million gallons/yr
	866,838 metric tons/yr
On Stream Time:	8,000 hours per year
Throughput:	289.55 million gallons/yr

Capital Cost (MILLION U.S. $)	ORIG	BOOK	REPL
Battery Limits	466.7	466.7	466.7
Offsites	0.0	0.0	0.0
Total Fixed Inv.	466.7	466.7	466.7
Working Capital			35.2

	PRODUCTION COST SUMMARY	UNITS PER GAL	PRICE, U.S. $ /UNIT	U.S. $ PER GAL	ANNUAL COST MM U.S. $	U.S. $ PER MET TON
RAW MATERIALS	Wood (dry), ST	0.0111	42.000	0.464	134.39	
	Sulfuric Acid, lb	0.3976	0.032	0.013	3.68	
	Lime, lb	0.2940	0.023	0.007	1.92	
	Ammonia, lb	0.6296	0.041	0.026	7.47	
	Nutrients, lb	0.0181	0.115	0.002	0.60	
	Corn Steep Liquor, lb	0.0633	0.100	0.006	1.83	
	Corn Oil (Antifoam), lb	0.0039	0.240	0.001	0.27	
	Glucose, lb	0.0496	0.510	0.025	7.32	
	Gasoline/Diesel, gal	0.0570	0.770	0.044	12.70	
	Catalyst & Chemicals		0.010	0.010	2.90	
	TOTAL RAW MATERIALS			0.598	173.09	200
BY-PRODUCT CREDITS	Solids Disposal, ton	(0.00034)	20.000	0.007	2.00	
	TOTAL BY-PRODUCT CREDITS			0.007	2.00	2
NET RAW MATERIALS				0.605	175.09	202
UTILITIES	Power, KWH	(1.85300)	0.030	(0.056)	(16.10)	
	Well Water, M Gal	0.01987	0.100	0.002	0.58	
	TOTAL UTILITIES			(0.054)	(15.52)	(18)
VARIABLE COST OF PRODUCTION				0.551	159.57	184
DIRECT CASH COSTS	Labor, 82 Men	29.80 Thousand U.S. $		0.008	2.44	
	Foremen, 18 Men	34.00 Thousand U.S. $		0.002	0.61	
	Super., 2 Men	40.00 Thousand U.S. $		0.000	0.08	
	Maint., Material & Labor	3.00 % of ISBL		0.048	14.00	
	Direct Overhead	45 % Labor & Supervision		0.005	1.41	
	TOTAL DIRECT CASH COSTS			0.064	18.55	21
ALLOCATED CASH COSTS	General Plant Overhead	65 % Labor & Maintenance		0.038	11.14	
	Insurance, Property Tax	1.5 % Total Fixed Investment		0.024	7.00	
	TOTAL ALLOCATED CASH COSTS			0.063	18.14	21
FULL CASH COST OF PRODUCTION				0.678	196.25	226
NET COST OF PRODUCTION				0.678	196.25	226
	COST PLUS 0 % RETURN ON TOTAL BOOK INV. PLUS WC			0.678	196.25	226
	COST PLUS 20 % RETURN ON TOTAL BOOK INV. PLUS WC			1.024	296.64	342
	COST PLUS 30 % RETURN ON TOTAL BOOK INV. PLUS WC			1.198	346.83	400

TABLE III-D-6

COST OF PRODUCTION ESTIMATE FOR : Denatured Fuel (90.25% Ethanol)

PROCESS : SERI-Wood to Ethanol-Large Plt.-90 % yield MILLION U.S. $

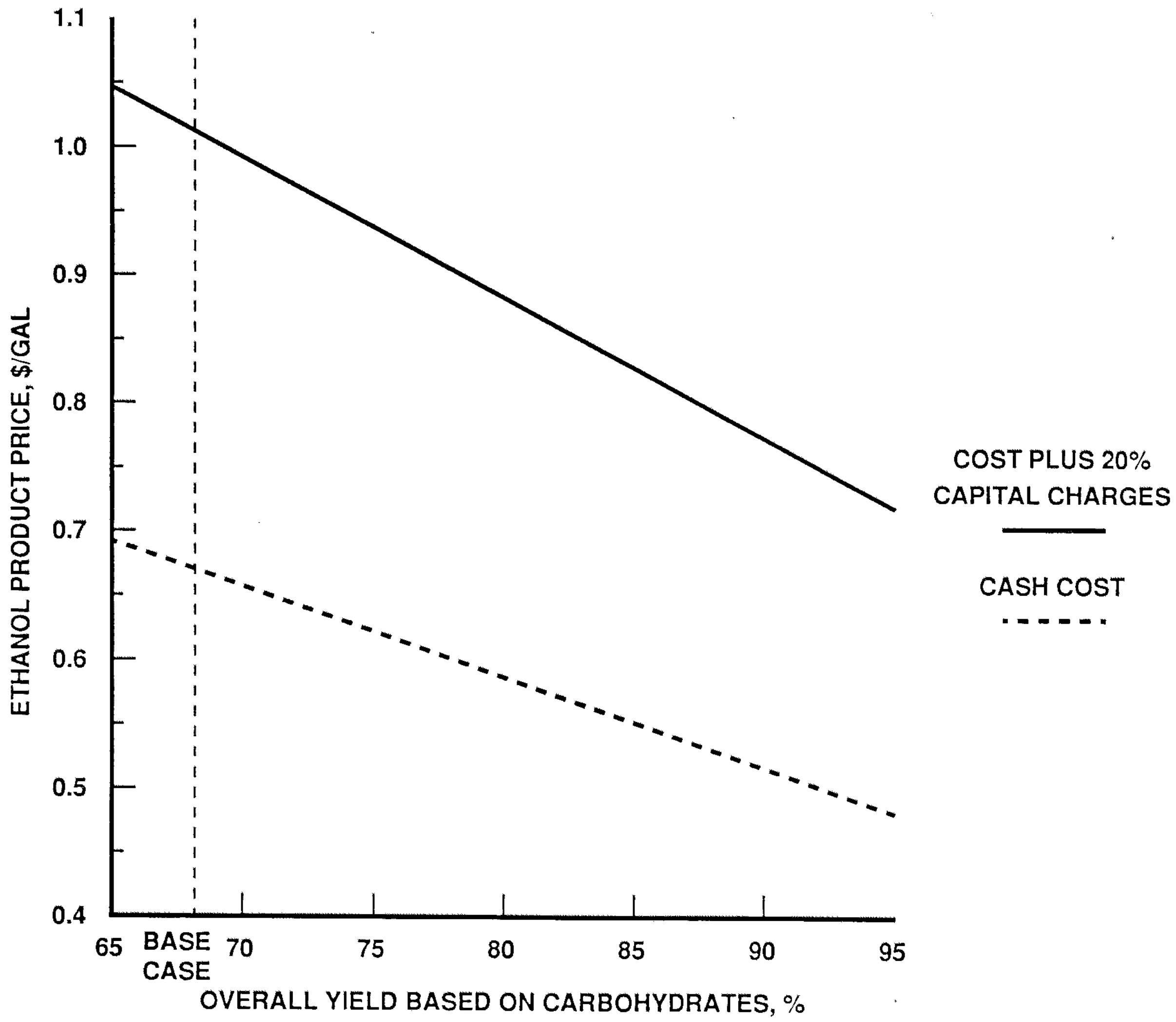
		Capital Cost	ORIG	BOOK	REPL
Plant Startup	1987				
Analysis Date	4Q-1987	Battery Limits	466.7	466.7	466.7
Location:	U.S.	Offsites	0.0	0.0	0.0
Capacity:	387.50 million gallons/yr				
	1,160 thousand metric tons/yr	Total Fixed Inv.	466.7	466.7	466.7
On Stream Time:	8,000 hours per year	Working Capital			35.8
Throughput:	387.50 million gallons/yr				

	PRODUCTION COST SUMMARY	UNITS PER GAL	PRICE, U.S. $ /UNIT	U.S. $ PER GAL	ANNUAL U.S. $ COST MM U.S. $	PER MET TON
RAW MATERIALS	Wood (dry), ST	0.0083	42.000	0.347	134.40	
	Sulfuric Acid, lb	0.2970	0.032	0.010	3.68	
	Lime, lb	0.2190	0.023	0.005	1.91	
	Ammonia, lb	0.4704	0.041	0.019	7.47	
	Nutrients, lb	0.0181	0.115	0.002	0.81	
	Corn Steep Liquor, lb	0.0633	0.100	0.006	2.45	
	Corn Oil (Antifoam), lb	0.0039	0.240	0.001	0.36	
	Glucose, lb	0.0370	0.510	0.019	7.31	
	Gasoline/Diesel, gal	0.0570	0.770	0.044	17.00	
	Catalyst & Chemicals		0.010	0.010	3.88	
	TOTAL RAW MATERIALS			0.463	179.27	155
BY-PRODUCT CREDITS	Solids Disposal, ton	(0.00034)	20.000	0.007	2.67	
	TOTAL BY-PRODUCT CREDITS			0.007	2.67	2
NET RAW MATERIALS				0.470	181.94	157
UTILITIES	Power, KWH	(1.48400)	0.030	(0.045)	(17.25)	
	Well Water, M Gal	0.01987	0.100	0.002	0.77	
	TOTAL UTILITIES			(0.043)	(16.48)	(14)
VARIABLE COST OF PRODUCTION				0.427	165.46	143
DIRECT CASH COSTS	Labor, 82 Men	29.80 Thousand U.S. $		0.006	2.44	
	Foremen, 18 Men	34.00 Thousand U.S. $		0.002	0.61	
	Super., 2 Men	40.00 Thousand U.S. $		0.000	0.08	
	Maint., Material & Labor	3.00 % of ISBL		0.036	14.00	
	Direct Overhead	45 % Labor & Supervision		0.004	1.41	
	TOTAL DIRECT CASH COSTS			0.048	18.55	16
ALLOCATED CASH COSTS	General Plant Overhead	65 % Labor & Maintenance		0.029	11.14	
	Insurance, Property Tax	1.5 % Total Fixed Investment		0.018	7.00	
	TOTAL ALLOCATED CASH COSTS			0.047	18.14	16
FULL CASH COST OF PRODUCTION				0.522	202.15	174
NET COST OF PRODUCTION				0.522	202.15	174
	COST PLUS 0 % RETURN ON TOTAL BOOK INV. PLUS WC			0.522	202.15	174
	COST PLUS 20 % RETURN ON TOTAL BOOK INV. PLUS WC			0.781	302.65	261
	COST PLUS 30 % RETURN ON TOTAL BOOK INV. PLUS WC			0.911	352.90	304

FIGURE III-D-4

EFFECT OF OVERALL YIELD ON ETHANOL PRODUCT PRICE FOR A LARGE CAPACITY PLANT

(US-4Q 1987) 9600 SHORT TONS/DAY DRY WOOD FEED



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