July 2002 • NREL/SR-510-32381

Subcontractor Report

# Evaluation of the Potential for the Production of Lignocellulosic Based Ethanol at Existing Corn Ethanol Facilities

## Final Subcontract Report
## 2 March 2000—30 March 2002

Delta T Corporation
Williamsburg, VA

National Renewable Energy Laboratory
1617 Cole Boulevard
Golden, Colorado 80401-3393

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Contract No. DE-AC36-99-GO10337

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NREL Technical Monitor: A. Aden

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## GOALS

The goals of this study were:

- To provide the opportunity to explore the business potential provided by converting biomass to products such as ethanol.

- To take advantage of the grain-processing infrastructure by investigating the co-location of additional biomass conversion facilities at an existing plant site.

## SCOPE

Delta-T's task was to perform process assessments for three biomass conversion processes. The feasibility studies were used to evaluate the potential alternatives for plant integration of biomass feedstock with an existing corn to ethanol facility. Delta T has recommended a preferred technology, DDG Conversion using Dilute Acid hydrolysis, based upon its near-term commercial viability.

The three processes considered were:

- Gasification of Biomass to Synthesis Gas to Ethanol
- DDG Conversion
- Dilute Acid Hydrolysis

Gasification was to be reviewed and considered primarily as a stand-alone alternative approach to ethanol production and not necessarily for specific co-location and integration into an existing ethanol facility.

## OBJECTIVES

The technical objectives of the study were designed to evaluate the business opportunity for lignocellulosic biomass conversion for the specific processing site. The overall result of this study is as follows:

### GASIFICATION OF BIOMASS TO SYNTHESIS GAS TO ETHANOL

To date, production of fuel grade alcohol has been focused on converting feedstock from renewable sources such as corn and other grown starch containing materials or cellulose containing materials through biological processing. This approach, almost without exception, relies on “bugs” and materials that are susceptible to infection which result in either significantly reduced yields or greatly increased production costs for the alcohol product desired. An alternative approach to alcohol production is to consider a method of chemical processing to produce alcohol. Commercially in much of the world, carbonaceous materials are processed through partial oxidation (gasification) to generate a synthesis (“syngas”) gas consisting of carbon monoxide (CO) and hydrogen ($H_2$). This syngas is then processed through catalytic steps to produce a chemical basic material. Two bulk chemicals that come to mind that are produced in this fashion are ammonia and methanol. Sasol has, on a commercial basis for over 35 years now, used the approach of converting coal through gasification to a syngas from which gasoline and a wide variety of chemical basic and intermediate compounds are produced.

Since residual materials such as corn stover, rice straw and hull, etc. contain carbonaceous materials, they can, in principle, be processed in a complete (burning) or partial oxidation (gasification) system. Gasification offers the advantage over complete combustion of producing a syngas that has the potential of being converted to ethanol as well as heat that could be used for process steam. In addition, chemical synthesis of ethanol offers the following advantages over conventional biological approaches for ethanol production:

- High temperature processing eliminating conditions that support infections in typical biological processing.
- Elimination of the need for yeast and “bugs” for obtaining yield and conversion efficiencies.
- Elimination of large quantities and volume of tankage and piping & pumping required in a typical biomass unit, minimizing plant foot print requirements.

On the other hand, drawbacks of this approach are:

- Catalysis of syngas to chemical constituents is generally accomplished at high pressure, requiring special design considerations in plant & equipment design and significant compression requirements.
- Until recently, no catalyst had been identified that offers the potential for longevity and reasonable yield for alcohol production.

Accordingly, the effort associated with this evaluation was focused at identifying if any work exists that might lead to a near-term implementation of a gasification/syngas conversion approach for the production of alcohol. The following summarizes the results of this assessment.

A limited literature search was conducted to identify what work might have already been sponsored by DOE/NREL in the application of gasification to ethanol production. Under this category, some bench scale work had been performed in processing syngas in a liquid type fermentation approach. From Delta-T's review, the work performed resulted in an approach that rendered alcohol production susceptible to infection and large residence time (tankage) requirements similar to that in an acid hydrolysis process. In addition, work performed to-date had been at the bench scale level only. Significant research and development would be required to assess commercialization potential. As the criteria for assessing technologies is the potential for near term commercialization, this approach was deemed not consistent with this criteria for the purposes of this evaluation.

Additional investigation identified an ethanol production approach using catalytic conversion of syngas. The use of catalyst for chemical manufacturing represents an approach used commercially throughout the world today. Particularly, chemical synthesis of methanol using this approach is widely practiced. With respect to use of gasification for producing syngas from which chemicals are synthesized, Sasol represents a prime example where this has been practiced for over 35 years now. While an integrated gasification-to-ethanol system has not been demonstrated, individual steps of the process are commercially available. The following describes one such process.

The process entails using biomass feed stock such as corn stover, wood chips, or other carbonaceous material, prepared and introduced into a partial oxidation unit. High purity oxygen is fed with the feed material to generate both heat and the basic syngas. The syngas is cleaned; adjustment made to obtain the appropriate $H_2$/CO ratio, compressed, and fed through the catalytic conversion step. Dehydration of the Ecalene™ product is accomplished using molecular sieve technology and forwarded to storage. Heat recovery is obtained from the gasification process and downstream processing to provide process steam. A block diagram (attachment 5a) is attached summarizing this process approach.

Ecalene$^{TM}$ has the following advertised composition:

| Component | Weight % | Mole % |
|---|---|---|
| | | |
| Methanol | 0.3 | 0.4 |
| Ethanol | 75.0 | 81.9 |
| Propanol | 9.0 | 8.1 |
| Butanol | 7.0 | 4.8 |
| Pentanol | 5.0 | 2.8 |
| Hexanol & higher | 3.7 | 2.0 |
| Total | 100.0 | 100.0 |

Although the ASTM D 4806 requires a minimum of 92.1-volume% ethanol, Ecalene$^{TM}$ has been categorized by EPA as an acceptable gasoline additive. Bench scale testing has been performed to confirm effectiveness of the catalyst. A firm, Power Energy Technologies has obtained a patent for the Ecalene$^{TM}$ production catalyst. A 700-gallon per day demonstration skid unit has been constructed.

Based on published information, a conceptual capital and operating cost assessment was performed for a commercial size Ecalene$^{TM}$ production plant (attachments 5b, 5c, 5d). The general economic approach used was similar to that in NREL report "Corn Stover To Ethanol Process Design" dated 21May98. Cost estimates for sub-system within the Ecalene$^{TM}$ production facility were based on the referenced NREL report, information obtained from selected, available information on the web, and independent supplier contact. A 15 million gallon per year undenatured alcohol product capacity was selected. Based on this approach, the total annual cost using the Ecalene$^{TM}$ process approach with gasification is \$ 1.28/gal. Total installed equipment cost for this facility is approximately \$28.4 million and total capital investment, using the criteria from the referenced report above, of \$48.2 million.

The estimate prepared compares to the \$ 1.30/gal for the 58 million gallon per year capacity in the referenced NREL study. Assuming a capital scale factor of 0.6, the Ecalene$^{TM}$ process approach at the 58 million gallon per year capacity would result in a \$1.08/gal on an annual basis.

Based on the above, Delta-T believes this approach warrants further investigation. For instance, potential integration of this technology approach into an existing corn-to-ethanol plant could be considered to replace the steam generation equipment while producing ethanol. The unit would use corn stover as a feedstock and the product could be blended with corn plant alcohol directly. This effort would require a very specific detailed engineering study and was assessed to be beyond the scope of this effort.

In summary, the approach of gasification and catalytic conversion to an alcohol product for fuel grade material offers potential for renewable materials as a feedstock and should be investigated further.

## CONVERSION OF DDG TO SUGARS TO ETHANOL

Delta-T has looked at the inclusion of the Brelsford 2 stage dilute acid hydrolysis process at CVEC for the conversion of DDG to sugars to ethanol. This appeared to be a cost effective method of increasing the yield at an existing ethanol facility. Using the Brelsford process, a yield increase of up to 18.75% in ethanol production has been forecasted by Brelsford Engineering. The byproduct stream of Wet Distillers Grain would be reduced from 35,000 lbs/hr to 18,000 lbs/hr but the protein content has been assumed to remain the same. Therefore the revenue from this co-product is assumed to be unchanged and is not included in the analysis. The reduced cellulose content in the Wet Distillers Grain may in fact make it more marketable.

## DILUTE ACID HYDROLYSIS OF BIOMASS TO SUGARS TO ETHANOL

Both the NREL dilute acid hydrolysis process and an alternative dilute acid hydrolysis process had previously been considered earlier in the study. Further consideration of these approaches was not pursued because:

- Preliminary economic assessment did not indicate adequate commercial viability.
- Economic viability depended on conceptual, non-demonstrated processing steps for by-product recovery and basic process performance.

## NEAR TERM COMMERCIALIZATION STUDY

From a near term commercialization stand point, the conversion of DDG via dilute acid hydrolysis to sugars to ethanol appeared to be the most promising process and was therefore chosen by Delta-T as the preferred process for further study. This decision was made as technical and reaction model data was available from bench scale testing performed on DDGS for the Brelsford 2 stage process. Delta-T has investigated this process further and has proceeded with the study of the installation of a Brelsford two-stage dilute acid hydrolysis process at the Chippewa Valley conversion facility.

As part of this study the following information was produced:

### INCREMENTAL BLOCK BLOW DIAGRAM

See attachment 1a, b, and c.

### INCREMENTAL UTILITY REQUIREMENTS

| | |
|---|---|
| Steam to Brelsford Process | 8.5 MMBTU/Hr. |
| Electric Power | |
| New Chiller/Cooling Tower | 423KW |
| Brelsford Process | 420 KW |
| Lime Addition | 15KW |
| Seed fermentation | 31KW |
| Total Electric Power | 889 KW |

### CAPITAL EQUIPMENT MODIFICATIONS AND ADDITIONS TO THE EXISTING FACILITY

See attachment 2a, Based Case Equipment Estimate.

For the purposes of this study, it was assumed the existing dryer system could adequately handle the co product from the Brelsford process. The new co product of WDG is expected to have less fiber content and may be too dusty for processing in a direct fired dryer. This concern can only be addressed by testing and assessment of the existing dryer, which is beyond the scope of this study.

### ADDITIONAL INFRASTRUCTURE REQUIREMENTS

The addition of the Brelsford process would require additional steam at higher pressures than presently produced at the CVEC facility. An additional boiler and piping to the Brelsford process have been included in the capital estimate for the base case.

The new process will require additional chilled water capacity to cool the stream to fermentation from the Brelsford process. Costs have been included for an additional chiller, cooling tower cell, and pumps.

### PRODUCTION CAPACITY BEFORE AND AFTER THE INSTALLATION OF THE PROCESS

| | |
|---|---|
| Ethanol Production | |
| Before proposed modifications | 19.2 MMgal/yr |
| After proposed modifications | 22.8 MMgal/yr |
| | |
| DDGS Before proposed modifications | 114 Million #/yr. |
| DDGS After proposed modifications | 60 Million #/yr. |

(Note the total protein content of the DDGS produced before and after the modifications is assumed to be constant. Accordingly, the total value of the DDG was assumed to be unchanged from current plant revenues for this co product.)

## FEEDSTOCK DESCRIPTION

Wet Distillers Grain was the assumed feedstock for the addition of dilute acid hydrolysis process at this facility. This consists of the combination of the syrup and the wet distillers grain that normally is delivered to the dryer at this facility. The Brelsford process is expected to produce a Wet Distillers Grain co-product of lesser flow than the existing stream but it has been assumed for the purposes of this study that the total protein content will remain the same. A further assumption is that the revenue generated from the DDGS is based upon protein content. Therefore, the revenue generated from DDGS is considered to be unaffected and the cost of the feedstock is assumed to be zero.

*FEEDSTOCK COMPOSITION (WET CAKE AND SYRUP) LB/HR FLOW 35% DRY BASIS*

| Alpha Cellulose | Hemi Cellulose | Dextrin | Soluble Sugars | Protein | Fat | Ash | Lignin | Total Solids |
|---|---|---|---|---|---|---|---|---|
| 1187 | 2653 | 1130 | 1243 | 3562 | 1613 | 869 | 154 | 12411 |

## FACILITY DESCRIPTION

Chippewa Valley Ethanol Company, Benson, MN site specifications.

*FACILITY PRODUCTION CAPACITY*

| | |
|---|---|
| Anhydrous Ethanol-Production MMgal/yr. | 19.2 |
| Anhydrous Ethanol-gallons/Dry Bushel Corn | 2.7 |
| DDGS-lbs/Bushel corn | 16.3 |
| Design fermentable sugar to fermentation MM#/yr. | 213 |

*INFRASTRUCTURE DESCRIPTION (UTILITIES, WATER, WASTE DISPOSAL, ROADS, RAIL)*

**Power**

CVEC has 3 460V-1500KW transformers, 2 are fully loaded and 1 is currently loaded to 400KW. Substation capacity is 3700KW.

**Water**

CVEC is permitted for 75 million gallons per year. The year 2000 usage was 63.5 million gallons.

**Waste Disposal**

CVEC has 2 40000-gallon wastewater tanks that are permitted for only 12 days of use per year. Cooling tower and boiler blow down is permitted to go to an existing fire protection pond and overflow goes to a ditch leading to the Chippewa River. During overflow months, the flow is sampled and tested by the state. There are no other process sewer or wastewater connections. CVEC is a "zero effluent" plant therefore any process additions must minimize water emissions.

**Roads**

CVEC is located just to the West of Benson, MN. It is at the intersection of route. 9 and route 20.

**Rail**
CVEC has a rail spur for product load out to the north of the facility. This spur is off the Burlington Northern Railroad.

## PROCESS ADDITION DESCRIPTION

(See Brelsford Engineering, Inc. Distillers Grains Cellulose Hydrolysis to Fermentable Sugars for Production of Fuel Ethanol and Hi-Protein Feedstuff-Preliminary Engineering and Economics feasibility Study for a more detailed description)

Distillers Wet Grains and Syrup (DWGS) are extracted from the existing process by redirection of the flow from the existing centrifuges. An additional series of screw conveyors would direct this feedstock to a mix tank. In this mix tank the WDGS is mixed with back set from the existing process to yield an acceptable dilution ratio for pumping, estimated at 18% solids. This mixture is then pumped through a heat exchanger to a reaction temperature and pressure of 35 psia. and 275 deg. F via indirect steam heat. Sulfuric acid is added to the process stream to a concentration of 1.5%. The slurry is then pumped through a series of plug flow reactor tubes with a residence time of approximately 10 minutes. The hemi cellulose (HC) fraction and a part of the resistant cellulose fraction of the holocellulose are converted to their respective sugars.

At the end of first stage of the process, the cellulose hydrolyzate sugars in solution and unhydrolyzed cellulose-protein-fat residue are continuously flashed to lower pressures in two stages. The first slurry hydrolyzate flash tank is to stop degrading of the sugars in the hydrolyzate, and to recover process heat and remove water/furfural/methanol vapor. At the second flash, the vapor pressure is dropped to atmospheric pressure. In addition, the second flash tank provides partial clarification and thickening of the unhydrolyzed cellulose-protein-fat residue. The thickened slurry is separated into a liquid hydrolyzate stream and a dewatered residue wet cake by means of a rotary vacuum filter. (An alternate process improvement might be to utilize the existing or new centrifuges.) The hydrolyzate stream, consisting of the converted sugars, is sent on to the lime addition module for acid neutralization. The thickened wet cake from the first stage is sent to the second stage of the Brelsford process.

In the second stage of the Brelsford process the unhydrolyzed cellulose feedstock from the 1$^{st}$ stage and unhydrolyzed alpha-cellulose residue recycle from the 2$^{nd}$ stage are put through a reaction process similar to the first stage. In the 2$^{nd}$ stage of the process, the reaction temperature and pressure are raised to 356 deg. F and 160 psia. The reaction time is 13 minutes at an acid concentration of 1%. The resultant slurry from this second stage is separated into a high protein wet cake stream and a dilute acid and alpha cellulose hydrolyzate solution. The wet cake stream is sent back to the existing dryer system at the CVEC facility for processing and storage as high protein DDGS. The hydrolyzate solution from the second stage is recycled to the 1st stage of the Brelsford process for recovery of the hydrolyzate. Additional benefits of this recycle are reuse of acid, recovery of process heat, and the use of liquid for front end dilution of the Brelsford process feedstock (the existing WDG process stream).

The hydrolyzate from the 1st stage of the Brelsford process is sent on to the new lime addition module for acid neutralization. This process module is taken from the NREL study section A200 Lime addition.

After the acid neutralization and cooling down of the hydrolyzate sugar solution from the Brelsford process, the process stream is combined with the flow from the existing Saccharification Tank (now at a reduce liquid flow due to the reduction in backset sent to the slurry mix tank). The combined stream is sent to the existing Fermentation tanks for conversion of the sugars to ethanol.

The existing yeast fermentation at CVEC uses the Saccharamyces Cereviseae yeast. This yeast is unable to convert the Hemi cellulose hydrolyzate, which is the majority of the conversion from the Brelsford process, into ethanol. Delta- T has therefore included a module to convert the existing fermentation process to NREL's Zymomonas mobilis bacterium for ethanol production. An alternative for further investigation might be the use of the Purdue University's laboratory of Renewable Resources Engineering (LORRE) saccharomyces yeast strain 1400 (LNH-ST). This genetically engineered yeast is capable of fermenting the sugars produced from the Brelsford dilute acid process. Delta-T is not aware of any commercial use of either of these organisms. The conversion to untried fermentation at CVEC may be unacceptable. In addition, questions arise in the use of genetically altered organisms for the production of animal feedstock- the DDGS.

## REQUIRED MODIFICATIONS, PRODUCTION PARAMETERS, AND AVAILABILITY OF CAPITAL-EQUIPMENT AND INFRASTRUCTURE THAT WILL BE SHARED

The Proposed Brelsford process will be integrated into the existing CVEC production facility via a new building addition on the north side of the main process building. The two stage process would be located near the existing wet cake load out at the facility in order to minimize process runs for WDG to and from the process. Screw conveyors would tie into the existing routing of the WDG to the existing dryer and would also feed the new WDG co product from the Brelsford process. Process lines would be added from the existing pipe racks for backset, steam, condensate return, acid, and the new mash flow to the existing fermentation process. In order to support the Brelsford process steam and cool down requirements, additional boiler chiller, and cooling tower capacity has been included in the capital and operating cost estimate.

## ADDITIONAL EQUIPMENT

The addition of the Brelsford 2 stage dilute acid hydrolysis of WDG at CVEC would require the following equipment:

### *The Brelsford 2 stage process*

Each stage of the Brelsford process will consist of a mix tank with agitator, progressive cavity pump, heat exchanger, Double Tube Heat Exchanger reactor, flash vessel, slurry pump, flash cyclone, and an auto-vac filter.

### *Lime Addition to the resulting hydrolysate for acid neutralization*

The resulting Hydrolysate will be at 1½ % sulfuric acid concentration and therefore will need to be neutralized with lime addition. The equipment for this module was extracted from the NREL study ((NREL TP-580-26157) and adjusted for process flow. The equipment required for this module consists of Lime Unloading Blower, Lime Storage Bin, Lime solids feeder, Overliming Tank and agitator, pump, Reacidification tank and agitator, pump, and a Hydroclone and Rotary Drum filter. Gypsum produced by this process would be discharged into a dumpster to be sent to land fill.

### *Seed Fermentation Module*

The sugars produced by the Brelsford process result in approximately 10% hexoses and 90% pentoses sugars. The existing saccharomyces cereviseae yeast fermentation cannot convert the pentose sugars to ethanol. Additional equipment has been included to convert fermentation to zymomonas mobilis continuous fermentation. The process equipment design, sizes, and costs are scaled from the NREL study.

## MINIMUM FEEDSTOCK SUPPLY QUANTITIES AND EXPECTED QUALITY MIX

The overall feedstock to the Brelsford process will consist of the existing Wet Distillers Grain and syrup at a flow rate of 35,500 pounds per hour. Based upon information from CVEC, the composition of this stream is 35% solids and will, on a dry basis, consist of 50% carbohydrates, 29% protein, 13% fat, 7% ash, and 1% lignin.

## ETHANOL PRODUCTION RATE IN GAL/DAY AND SOLID BY-PRODUCT RATE

Based upon estimates from Brelsford Engineering, the increased ethanol production from this modification would be an additional 3.6 million gallons per year (9,912 gallons/day).

The solid co product, a high protein wet distillers grain, production rate would be 18,030 #/hr total on a wet basis. On a dry basis of 6310 #/hr., the DDGS has a protein content of 56.5%. This is a reduction in DDGS total dry mass flow but an unchanged production rate on a protein content basis.

## ENVIRONMENTAL EMISSION CHARACTERISTICS, IN TERMS OF QUANTITY EMITTED PER TON OF FEEDSTOCK PROCESSED.

The environmental emissions from this process would consist mainly of the gypsum produced in the overliming process, which is required to neutralize the hydrolysate to fermentation. The expected flow rate is 35 MM#/yr. of gypsum. A disposal cost from past NREL studies has been assumed.

In addition, a vapor stream of 15500 #/hr with concentrations of furfural would be generated. No additional equipment has been included for handling this process stream. Further study is required for the handling of this item.

## AREA REQUIREMENTS AND PREFERRED SHAPE

The area requirements for this process will consist of three main units.

*THE BRELSFORD 2 STAGE ACID HYDROLYSIS*

Given the size of the reactors from Brelsford Engineering, a 150'x 25' addition has been assumed. In order to facilitate access to the Wet Distillers Grain for infeed and discharge from this module, it has been located on the North side of the Main process building near the existing centrifuges and the dryer building.

*OVER LIME ADDITION MODULE*

The Lime Addition Module was sized from the NREL study dated July 1999 (NREL TP-580-26157). Using scaled down equipment from the NREL study, an area of 50' x 50' is shown on the plot plan to the south of the existing CVEC main process building.

*ZYMOMONAS MOBILIS SEED FERMENTATION MODULE*

The Seed fermentation Module for this study was sized from the referenced NREL study. Based upon the tank sizes for three stages of seed fermentation for 2 trains, a module area of 25' x 25' has been included in the CVEC layout. This area is shown as being located on the south side of the Main Process Building, next to the existing fermentation tanks.

## UTILITY AND CHEMICAL REQUIREMENTS (WATER, STEAM, FUEL, POWER, CHEMICALS)

### *Water Requirements*

Water for the Brelsford process is considered negligible. Make up to dilute the wet cake and syrup feedstock will come from the use of backset from the existing thin stillage surge tank. This liquid will be added back into the existing process in the form of the hydrolyzate from the Brelsford process. This hydrolyzate would be added back into the existing feed to the fermentation tanks. As a result of the use of the backset, there will be an increase in solid percentage content in the feed to cook tubes and saccharification tanks. This increase is assumed to be acceptable.

Water for Lime Addition Process is assumed to be negligible.

Water for the addition of the Seed Fermentation is based upon the information in the NREL study, is assumed to be negligible.

### *Steam Requirements*

Steam energy for the Brelsford process is required to achieve the high reaction temperatures in the two-stage process. A total input of 6762 #/hr of low-pressure steam (40 psia) and 7650 #/hr of high-pressure (150 psia) steam is required for the two stage process. The Brelsford process is based upon direct steam injection, but the addition of this water to the existing CVEC process would be unacceptable. Therefore additional work with Brelsford would be required to convert the process to indirect steam heating. Costs for indirect steam heat exchangers have been included in the estimate. In addition the reuse of flash steam heat from each stage will greatly reduce the steam needs for this process. The full load for steam and the high-pressure steam requirements could not be met with the existing CVEC boiler system. Therefore costs have been included for an additional boiler in capital cost estimate.

### *Fuel Requirements*

Fuel requirements for the process modifications have been included as the operating costs for steam.

### *Power Requirements*

Power for the Brelsford process has been scaled from the Brelsford Preliminary Engineering and Economics Feasibility Study based upon the total flow to the process and the number of stages. The Brelsford study quotes 138 KW for one stage at a flow of 7800 #/hr dry basis. The power requirement for these two modules was therefore estimated at 420 KW.

Power for the Lime Addition module was developed from the NREL study and adjusted for process flow. The resulting demand is 15 KW.

Power for the Seed Fermentation module is estimated at 31 KW based upon estimates from the NREL study.

Power requirements for the additional chiller, cooling tower, and pumps are based upon Richardson and are estimated at 423 KW.

*CHEMICAL REQUIREMENTS*

Chemical requirements for the Brelsford process will consist of the addition of Sulfuric Acid. Based upon information from Brelsford Engineering, the two stage process will required the addition of 1037 #/ hour of acid.

Chemical requirements for the Lime addition module were estimated from the NREL study. The lime addition requirements were based upon the acid flow to the lime addition module from the Brelsford process. Lime requirements were estimated at 1155 #/hr. Pricing for the lime was based upon a cost from the Brelsford study at $60.00 per ton.

Incremental Chemical Requirements for the Seed Fermentation module were assumed to be negligible.

*SPECIAL TRANSPORTATION REQUIREMENTS (TRUCK, WATER, RAIL LINE)*

Other than as mentioned above, it is assumed there will be no additional transportation requirements. The existing truck receiving and rail service is assumed to be adequate for the modifications proposed to the CEVC facility.

*SPECIAL STORAGE REQUIREMENTS FOR FEEDSTOCK, BY-PRODUCTS, AND CHEMICALS.*

**Storage for Feedstock**

There will be no additional storage requirements for the feedstock.

**Storage for Co Product WDG and DDGS**

There will be no additional storage requirements for the feedstock for the Brelsford process. The existing wet cake pad, dryers and DDGS storage facility are assumed to be adequate for the co product. One concern is the fact that the new co product will have less fiber content and therefore will be a more powdery product. This may cause problems with the existing dryer system and material handling procedures. For the purposes of this study, these concerns have not been addressed and are held for further study. A proforma case has been run with the cost of a replacement dryer included.

**Storage for Chemicals**

Chemical Storage for the Brelsford process will consist of an additional Acid storage tank and delivery system. CVEC has an existing acid tank, but from a logistics standpoint, an additional tank has been included in the capital equipment estimate.

Chemical Storage for the Lime addition module will consist of additional receiving and storage for lime. No specific site in the layout has been shown for this receiving equipment or storage. The sizing for the lime storage and the costs are based upon the NREL study.

Chemical storage for the Seed fermentation module is assumed to be adequate.

## CAPITAL AND OPERATING COSTS

Delta T has developed capital and operating costs for the recommended process based on process considerations. These are included in the proforma economics.

For the purposes of the study, the cost of the incremental feedstock, the Wet Distillers Grain and syrup from the existing facility was assumed to be zero.

## INCREMENTAL FINANCIAL PRO FORMA

Delta T has prepared three incremental financial evaluations for the installation of the recommended process at the CVEC facility. The assumptions and the rationale used in the Pro Forma are as listed below:

### *BASE CASE (ATTACHMENT 2A, 2B)*

Assumptions

The Brelsford process capital equipment cost is as supplied by Brelsford Engineering. An adjustment was made to the reactor cost to adjust from 304-carbon steel to Hastelloy C-2000 by increasing cost by 50%. This is the same adjustment factor as used in the referenced NREL cost estimate.

### *FEEDSTOCK COSTS:*

This process is expected to increase the yield of ethanol per bushel of corn. The base case assumes the facility could increase throughput accordingly.

Case 2 looks at decreasing the corn feedstock to maintain the hydraulic flow to fermentation. It was assumed that the typical facility would be running at capacity and the net result of this process improvement would be a decrease in corn feedstock.

### *ETHANOL PRODUCTION VALUE*

Ethanol production was estimated to increase by 3.6 million gallons per year for an increase in revenue of $3,960,000.

*Byproducts Production Value:*
The Brelsford process will result in a net decrease in the production of WDG but the protein content is assumed to remain constant. For the purposes of this study, the value of the WDG is based upon the protein and is therefore considered to be unchanged.

*Payback period*
A division of the capital cost for this case divided by the expected net increased revenue yields a payback period of 3.42 years.

## SENSITIVITY ANALYSIS

*Case 2- Reduced Infeed to the existing Process (attachment 3a, 3b)*
Delta T has identified that the existing process could not handle the new flow to the distillation and evaporation modules. Therefore the corn feedstock was adjusted in this case by 15%. The feedstock costs and variable costs of the existing CVEC operation were adjusted by 15%. In addition the operating costs and production increases for the Brelsford process addition were adjusted by 15%. The capital costs for this case were considered identical to the base case above.

Payback period
Unfortunately, the decrease in DDGS production resulted in a net decrease in the revenue generated, resulting in a negative payback.

*Case 3-Equipment Cost Estimate with New Dryer and Centrifuges (attachment 4a, 4b)*
The Brelsford process uses a low cost vacuum filter process for solid liquid separation. In this case costs were included for the addition of new centrifuges for the process. In addition, costs for a new dryer have been included in this case.

Payback Period
The addition of capital equipment in this case resulted in a further reduction in revenue generated, resulting in a negative payback.

## RECOMMENDATIONS FOR FURTHER WORK

The Brelsford Process
In summarization of the Brelsford process, the increased hydraulic load of the returned liquid to fermentation causes a bottleneck in the distillation process of an existing facility. The facility would have to be derated, resulting in a negative affect on the revenue generation. In addition the following items would require additional investigation:

The Brelsford process design is based upon 304 ss 1' diameter reactors. The material may have to be adjusted to Hastelloy C-2000 as was used in the NREL study for the reactor M-202. A 50% increase in the cost of the Brelsford reactor could be expected with this adjustment. Further study of the materials of composition for this process is recommended.

The steam usage in the Brelsford process is based upon unproven assumptions for the reuse of high-pressure steam between the stages. The best-case cost would be the reuse of the flash steam in the Brelsford process. The worst-case cost would be the cost of no flash recovery of steam. The handling of furfural and other gases generated in the process requires further study. Delta-T looked at condensing the stream and sending it back to the process or disposing of it. CVEC is a zero effluent facility and no method of disposing of the stream is available. In addition, Delta-T looked at venting the stream with the dryer stack gas. This also appeared unacceptable from an emissions standpoint. The resolution of the disposal of this stream is outside the scope of this investigation.

The assumption has been made that the existing fermentation can be switched over to Z-mobilis with the addition of seed fermentation. This conversion may be unacceptable to the existing operation. Further study is required to identify the best method to convert the sugars produced from the Brelsford process.

Gasification of Biomass

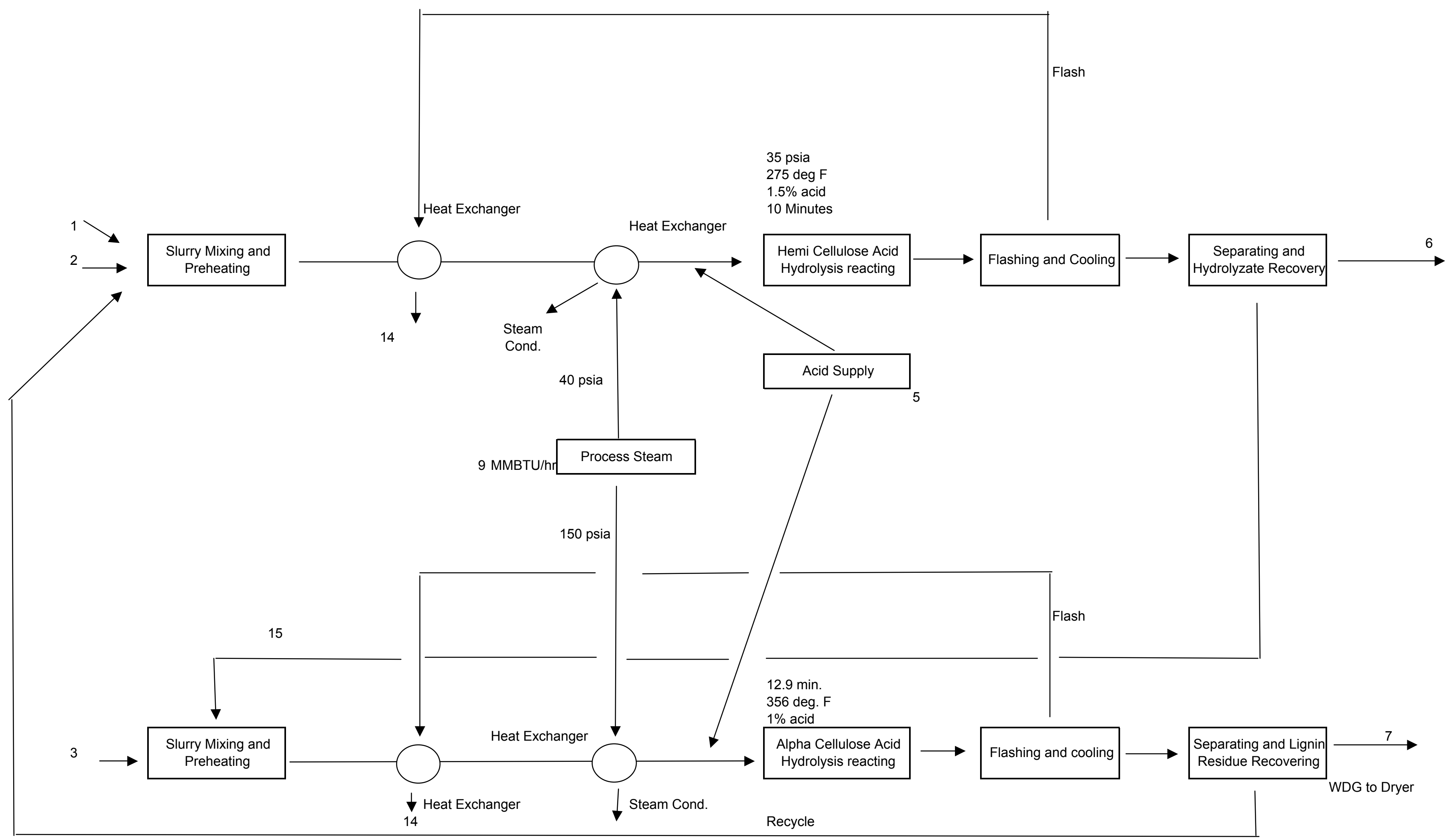
The Web based search of gasification and catalytic conversion to an alcohol product for fuel grade material as a standalone or add on to a facility offers potential for renewable materials as a feedstock and should be investigated further. The preliminary economics of this study are favorable and it is recommended that a more detail study now be considered for this approach.

# APPENDIX

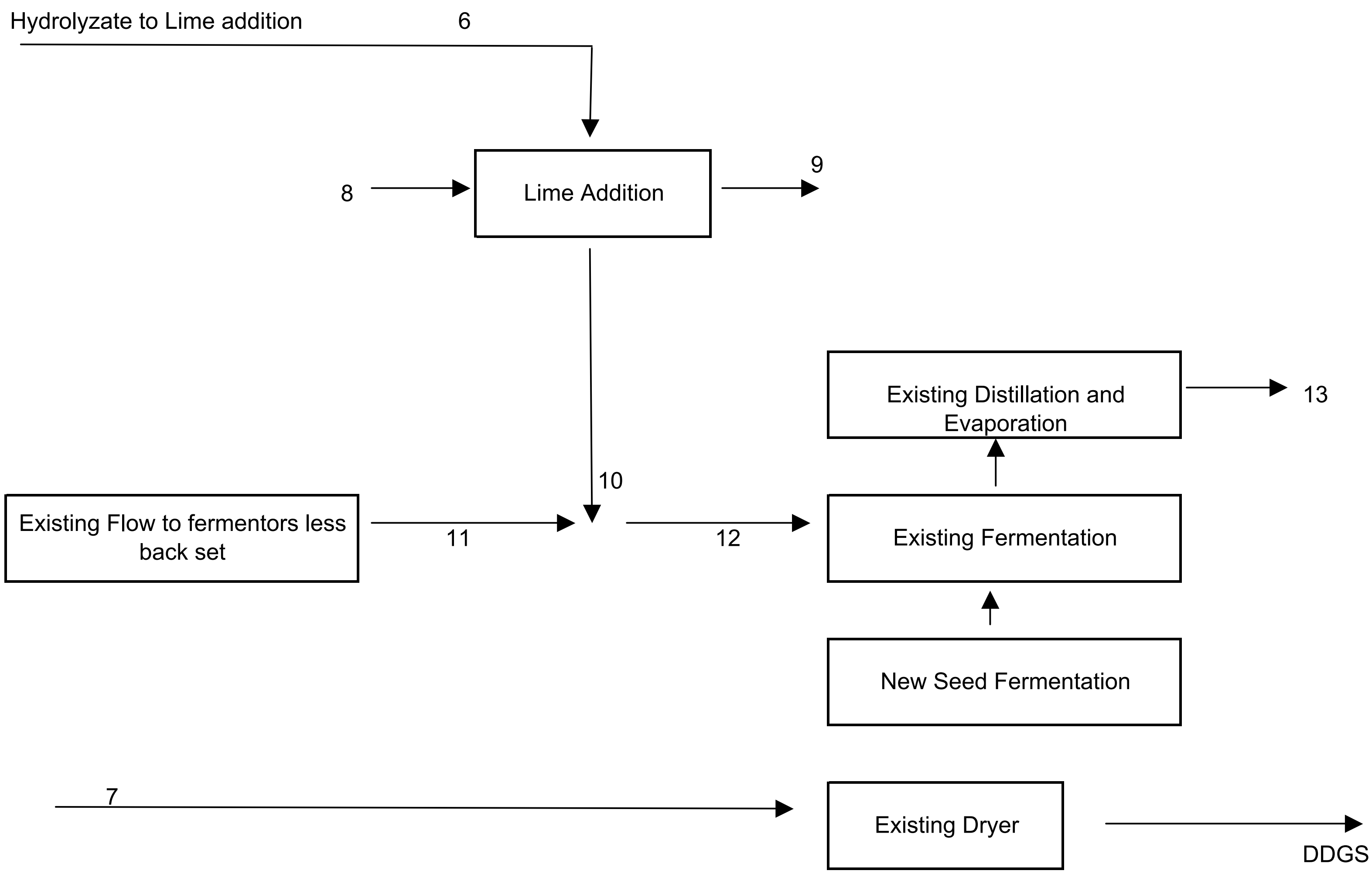
**Attachment 1a**

**Incremental Process Block Flow Diagram**

**Brelsford Process Addition, CVEC, Benson, MN.**



## Attachment 1b
## Incremental Process Block Flow Diagram
## Brelsford Process Addition, CVEC, Benson, MN.



# Attachment 1c
# Block Flow Diagram Stream Legend
# Incremental Stream Flows

| | | | | | Flow #/hr. | | | | | | |
|---|---|---|---|---|---|---|---|---|---|---|---|
| Stream | Description | Total Flow | Liquid | Dry matter/ Vapors | HCH | ACH | Carbo. | Protein | Fat | Ash | Lignin |
| 1 | Feedstock-WDG and Syrup | 35500 | 23075 | 12425 | | | 6213 | 3566 | 1615 | 870 | 162 |
| 2 | Backset Make Up to 1st Stage | 7574 | 7574 | 682 | | | | | | | |
| 3 | Backset Make Up to 2nd Stage | 20700 | 20700 | 1863 | | | | | | | |
| 5 | Sulfuric Acid | 1037 | 1037 | | | | | | | | |
| | **Total input** | 64811 | 52386 | 14970 | | | | | | | |
| | | | | | | | | | | | |
| 14 | Flash Cond to WDG (see comment box) | 15482 | 12767 | 5260 | | | | | | | |
| 6 | Hydrolyzate From Brelsford Process | 31300 | 27900 | 3400 | 2500 | 900 | | | | | |
| 7 | WDG from Brelsford Process | 18029 | 11719 | 6310 | | | | 3562 | | | |
| | **Outputs from Brelsford Process** | 49329 | 39619 | 9710 | | | | | | | |
| | | | | | | | | | | | |
| 8 | Lime Addition | 1155 | 0 | 1155 | | | | | | | |
| 9 | Gypsum to Landfill | 3936 | 787 | 3149 | | | | | | | |
| 10 | **Net Flow to fermentation** | 46548 | 38831 | 7716 | | | | | | | |
| | | | | | | | | | | | |
| 11 | Existing Flow to Fermen. Less Back Set | 122644 | | | | | | | | | |
| 12 | New Flow to fermentation with Brelsford Process | 169192 | | | | | | | | | |
| | Derating of Plant for Hydraulic Flow Limit | 85% | | | | | | | | | |
| | | | | | | | | | | | |
| 13 | Additional Ethanol Production | 2682 | 18.00% | | | | | | | | |

**Attachment 2a**
**Base Case Equipment Estimate**
**Capital Equipment Additions for Brelsford Process**

| Quantity | Equip. Number | Equipment Description | Source | Size Ratio Current/ Base | Original Cost | Base Year | Scaling Exponent | Scaled Cost in Base Year | Install Factor | Installed Cost In Base Year | Installed Cost in 1999 |
|---|---|---|---|---|---|---|---|---|---|---|---|
| **Infrastructure Requirements** | | | | | | | | | | | |
| 1 | | Boiler | Richardso | 0.40 | $ 195,000 | 1994 | 0.40 | $ 135,163 | 1.5 | $ 202,745 | $ 235,037 |
| 1 | | Chiller | Richardso | 1.00 | $ 92,649 | 1994 | 1.00 | $ 92,649 | 1.50 | $ 138,974 | $ 161,108 |
| 1 | | Cooling Tower | Richardso | 1.00 | $ 18,084 | 1994 | 1.00 | $ 18,084 | 2.80 | $ 50,635 | $ 58,700 |
| 1 | | Chiller Pump | Richardso | 1.00 | $ 4,800 | 1994 | 1.00 | $ 4,800 | 2.80 | $ 13,440 | $ 15,581 |
| 1 | | Condens. Pump | Richardso | 1.00 | $ 4,800 | 1994 | 1.00 | $ 4,800 | 2.80 | $ 13,440 | $ 15,581 |
| | | | | | | | | | | **Total Plant Modifications** | **$ 486,007** |
| **Brelsford Stage 1** | | | | | | | | | | | |
| 1 | CF-1 | Wet Grains Feed conveyor | Brelsford | 1.38 | $ 4,080 | 1999 | 0.51 | $ 4,804 | 1.2 | $ 5,765 | $ 5,765 |
| | | Slurry Mix Tank | Delta-T | 1.00 | $ 21,136 | 1995 | 1.00 | $ 21,136 | 1.2 | $ 25,363 | $ 29,168 |
| 1 | M-1 | Conveyor | Brelsford | 1.38 | $ 8,300 | 1999 | 0.51 | $ 9,773 | 1.2 | $ 11,728 | $ 11,728 |
| 1 | D-1 | Disintegrator | Brelsford | 1.38 | $ 18,900 | 1999 | 0.79 | $ 24,343 | 1.3 | $ 31,647 | $ 31,647 |
| 1 | P-1 | Progressive cavity Pump | Brelsford | 1.38 | $ 8,800 | 1999 | 0.79 | $ 11,335 | 2.8 | $ 31,737 | $ 31,737 |
| 1 | HH-1 | Hydro Heater | Brelsford | 1.38 | $ 4,850 | 1999 | 0.79 | $ 6,247 | 2.8 | $ 17,491 | $ 17,491 |
| 1 | HX-1 | Heat Exchanger | Atlas | 1.00 | | | | | | | $ 18,100 |
| 1 | HX-2 | Heat Exchanger | Atlas | 1.00 | | | | | | | $ 17,050 |
| 2 | PFR-1 | Double Tube reactor | Brelsford | 1.38 | $ 331,000 | 1999 | 0.79 | $ 426,333 | 1.5 | $ 639,500 | $ 639,500 |
| 1 | T-1 | Flash vessel | Brelsford | 1.38 | $ 11,000 | 1999 | 0.71 | $ 13,810 | 1.4 | $ 19,333 | $ 19,333 |
| 1 | P-1 | Slurry Pump | Brelsford | 1.38 | $ 2,500 | 1999 | 0.79 | $ 3,220 | 2.8 | $ 9,016 | $ 9,016 |
| 1 | Cy-1 | Flash Cyclone | Brelsford | 1.38 | $ 11,000 | 1999 | 0.79 | $ 14,168 | 1.4 | $ 19,835 | $ 19,835 |
| 1 | S-1 | Screen | Brelsford | 1.38 | $ 5,500 | 1999 | 0.79 | $ 7,084 | 2.8 | $ 19,835 | $ 19,835 |
| 1 | RF-1 | Rotary Filter | Brelsford | 1.38 | $ 54,800 | 1999 | 0.79 | $ 70,583 | 2.8 | $ 197,633 | $ 197,633 |
| 1 | T-2 | Surge Tank | Brelsford | 1.38 | $ 13,600 | 1999 | 0.79 | $ 17,517 | 1.4 | $ 24,524 | $ 24,524 |
| 1 | | Conveyors to second stage | Brelsford | 1.38 | $ 5,500 | 1999 | 0.79 | $ 7,084 | 1.2 | $ 8,501 | $ 8,501 |
| 1 | | Pump to Lime Addition | Brelsford | 1.38 | $ 2,500 | 1999 | 0.79 | $ 3,220 | 2.8 | $ 9,016 | $ 9,016 |
| | | | | | | | | | | **Total Brelsford Stage 1** | **$ 1,109,879** |
| **Brelsford Stage 2** | | | | | | | | | | | |
| 1 | T-201 | Acid Storage Tank | NREL | 0.24 | $ 42,500 | 1997 | 0.51 | $ 20,389 | 1.2 | $ 24,467 | $ 25,957 |
| 1 | P-201 | Acid Storage Tank Pump | NREL | 0.24 | $ 8,000 | 1997 | 0.79 | $ 2,564 | 2.8 | $ 7,180 | $ 7,617 |
| 1 | A-201 | Acid Mixer | NREL | 0.23 | $ 1,900 | 1997 | 0.48 | $ 935 | 1.0 | $ 935 | $ 992 |
| 1 | P-1 | Progressive cavity Pump | Brelsford | 1.38 | $ 8,800 | 1999 | 0.79 | $ 11,335 | 2.8 | $ 31,737 | $ 31,737 |
| 1 | HH-1 | Hydro Heater | Brelsford | 1.38 | $ 4,850 | 1999 | 0.79 | $ 6,247 | 2.8 | $ 17,491 | $ 17,491 |
| 1 | HX-1 | Heat Exchanger | Atlas | 1.00 | | | | | | | $ 18,100 |
| 1 | HX-2 | Heat Exchanger | Atlas | 1.00 | | | | | | | $ 17,050 |
| 1 | PFR-1 | Double Tube reactor | Brelsford | 1.38 | $ 165,500 | 1999 | 0.79 | $ 213,167 | 1.5 | $ 319,750 | $ 319,750 |
| 1 | T-1 | Flash vessel | Brelsford | 1.38 | $ 11,000 | 1999 | 0.71 | $ 13,810 | 1.4 | $ 19,333 | $ 19,333 |
| 1 | P-1 | Slurry Pump | Brelsford | 1.38 | $ 2,500 | 1999 | 0.79 | $ 3,220 | 2.8 | $ 9,016 | $ 9,016 |
| 1 | Cy-1 | Flash Cyclone | Brelsford | 1.38 | $ 11,000 | 1999 | 0.79 | $ 14,168 | 1.4 | $ 19,835 | $ 19,835 |
| 1 | S-1 | Screen | Brelsford | 1.38 | $ 5,500 | 1999 | 0.79 | $ 7,084 | 2.8 | $ 19,835 | $ 19,835 |
| 1 | RF-1 | Rotary Filter | Brelsford | 1.38 | $ 54,800 | 1999 | 0.79 | $ 70,583 | 2.8 | $ 197,633 | $ 197,633 |
| 1 | T-2 | Surge Tank | Brelsford | 1.38 | $ 13,600 | 1999 | 0.79 | $ 17,517 | 1.4 | $ 24,524 | $ 24,524 |
| 1 | | Conveyors to dryer | Brelsford | 1.38 | $ 5,500 | 1999 | 0.79 | $ 7,084 | 1.2 | $ 8,501 | $ 8,501 |
| 1 | | Slurry Pump to 1st Stage | Brelsford | 1.38 | $ 2,500 | 1999 | 0.79 | $ 3,220 | 2.8 | $ 9,016 | $ 9,016 |
| | | | | | | | | | | **Total Brelsford Stage 2** | **$ 746,388** |
| **Lime Addition Module** | | | | | | | | | | | |
| 1 | T-301 | Lime Storage Bin | NREL | 1.00 | $ 69,200 | 1997 | 0.46 | $ 69,200 | 1.3 | $ 89,960 | $ 95,439 |
| 1 | S-301 | Lime Dust Baghouse | NREL | 1.00 | $ 32,200 | 1997 | 1.00 | $ 32,200 | 1.5 | $ 48,300 | $ 51,198 |
| 1 | C-301 | Lime Solids Feeder | NREL | 0.15 | $ 3,900 | 1997 | 1.00 | $ 3,900 | 1.3 | $ 5,070 | $ 5,374 |
| 1 | T-300 | Overliming Tank | NREL | 0.15 | $ 71,000 | 1997 | 0.71 | $ 18,188 | 1.4 | $ 25,463 | $ 26,991 |
| 1 | A-300 | Overliming Tank Agitator | NREL | 0.15 | $ 19,800 | 1997 | 0.51 | $ 7,444 | 1.3 | $ 9,677 | $ 10,258 |
| 1 | P-300 | Overlimed Hydrolyzate Pump | NREL | 0.15 | $ 10,700 | 1997 | 0.79 | $ 2,351 | 2.8 | $ 6,583 | $ 6,978 |
| 1 | S-302 | Rotary Drum Filter and Hydroclone | NREL | 0.15 | $ 165,000 | 1998 | 0.39 | $ 78,089 | 1.4 | $ 109,324 | $ 112,604 |
| 1 | P-302 | Overlimed Hydrolyzate Liquor Pump | NREL | 0.15 | $ 10,800 | 1997 | 0.79 | $ 2,373 | 2.8 | $ 6,644 | $ 6,844 |
| 1 | T-303 | Reacidification Tank | NREL | 0.15 | $ 147,800 | 1997 | 0.51 | $ 55,566 | 1.2 | $ 66,679 | $ 68,680 |
| 1 | A-303 | Reacidificaiton Tank Agitator | NREL | 0.15 | $ 65,200 | 1997 | 0.51 | $ 24,512 | 1.2 | $ 29,415 | $ 30,297 |
| 1 | P-302 | Fermentation Feed Pump | NREL | 0.19 | $ 61,368 | 1998 | 0.70 | $ 19,122 | 2.8 | $ 53,541 | $ 55,147 |
| 1 | A-304 | In-line Acid Mixer | NREL | 0.15 | $ 2,600 | 1997 | 0.48 | $ 1,035 | 1.0 | $ 1,035 | $ 1,098 |
| 1 | B-300 | Lime unloading Blower | NREL | 0.15 | $ 47,600 | 1998 | 0.50 | $ 18,242 | 1.4 | $ 25,539 | $ 26,305 |
| 1 | P-303 | Reacidification Liquor Pump | NREL | 0.15 | $ 10,800 | 1997 | 0.79 | $ 2,373 | 2.8 | $ 6,644 | $ 6,844 |
| 1 | | Heat Exchanger | Atlas | 1.00 | | | | | | | $ 18,100 |
| | | | | | | | | | | **Total Lime Addition Module** | **$ 522,155** |
| **Seed Fermentation** | | | | | | | | | | | |
| 2 | A-304 | Seed Vessel Agitator | NREL | 0.91 | $ 11,700 | 1997 | 0.51 | $ 22,295 | 1.2 | $ 26,754 | $ 28,360 |
| 2 | A-305 | Seed Vessel Agitator | NREL | 0.91 | $ 10,340 | 1996 | 0.51 | $ 19,704 | 1.2 | $ 23,644 | $ 25,772 |
| 2 | F-301 | 1st SSCF Seed Fermentor | NREL | 1.00 | $ 14,700 | 1997 | 0.91 | $ 29,400 | 2.8 | $ 82,320 | $ 87,259 |
| 2 | F-302 | 2nd SSCF Seed Fermentor | NREL | 1.00 | $ 32,600 | 1997 | 0.93 | $ 65,200 | 2.8 | $ 182,560 | $ 193,514 |
| 2 | F-303 | 3rd SSCF Seed Fermentor | NREL | 1.00 | $ 81,100 | 1997 | 0.93 | $ 162,200 | 2.8 | $ 454,160 | $ 481,410 |
| 2 | F-304 | 4th SSCF Seed Fermentor | NREL | 0.91 | $ 39,500 | 1997 | 0.93 | $ 72,331 | 1.2 | $ 86,798 | $ 92,005 |
| 1 | H-301 | SSCF Seed Hydrolyzate Cooler | NREL | 0.91 | $ 15,539 | 1998 | 0.78 | $ 14,414 | 2.1 | $ 30,270 | $ 31,178 |
| 3 | H-302 | SSCF Hydrolyzate Cooler | NREL | 0.98 | $ 25,409 | 1998 | 0.78 | $ 75,201 | 2.1 | $ 157,922 | $ 162,660 |
| 1 | H-304 | Seed Fermentor Coils | NREL | 0.92 | $ 3,300 | 1997 | 0.83 | $ 3,066 | 1.2 | $ 3,679 | $ 3,899 |
| 2 | P-302 | Seed Transfer Pump | NREL | 0.91 | $ 54,088 | 1998 | 0.70 | $ 101,228 | 1.4 | $ 141,720 | $ 145,971 |
| | | | | | | | | | | **Total Seed Fermentation Module** | **$ 1,252,028** |
| | | | | | | | | | | Eng./Const.Mgt. | **$ 1,975,899** |
| | | | | | | | | | | **Total Incremental Capital Equipment Cost** | **$ 6,092,356** |

Notes

1. Brelsford Module Costs based upon estimates from Breslford Engineering "Distiller grains Cellulose Hydrolysis to Fermentatable sugars for Production of Fuel Ethanol and Protein Feedstuff" Preliminary Engineering and Economics Feasibility Study.
2. Lime Addition Module Costs based upon July 1999 NREL/TP-580-26157 Study Lignocellulosic Biomass to Ethanol Process Design and Economics Utilizing Co-Current Dilute Acid Prehydrolysis and Enzymatic Hydrolysis current and Futuristic Scenarios. Costs were adjusted for sulfuric acid flow rates.

3. Seed fermentation Module Costs based upon July 1999 NREL/TP-580-26157.

# Attachment 2b
# Base Case Proforma Economics

| | Units | Unit Price $ | Quantity/yr | $/Yr. |
|---|---|---|---|---|
| **Revenue** | | | | |
| Increased Ethanol | gallons | 1.10 | $ 3,600,000 | $ 3,960,000 |
| | | | | |
| **Operating Cost** | | | | |
| Feedstock | | | | na |
| Sulfuric Acid | Tons | 96.88 | 4542 | $ 440,035 |
| Lime | Tons | 60.00 | 5059.33 | $ 303,560 |
| Gypsum Removal | $/# | 0.01 | 34479975 | $ 344,800 |
| Steam | MM BTU | 5.31 | 74460 | $ 395,383 |
| Electric Power | $/KWh | 0.04 | 7787640 | $ 280,355 |
| Shift operators | 1/shift | 28000.00 | 4 | $ 112,000 |
| Maint.Cost-5% Capital | | | | $ 304,618 |
| | | **Total Incremental Operating Cost** | | $ 2,180,750 |
| | | | **Net Increased Revenue** | $ 1,779,250 |
| | | | **Capital Cost** | $ 6,092,356 |
| | | | **Capital payback period** | 3.42 |

Notes
1. Increased ethanol based upon estimate from Brelsford Engineering of and additional 3.6 million gallons/yr of increased production. Unit price based upon Fiscal year 2000
2. Feedstock cost changes assumed to be zero.
3. Sulfuric Acid cost based upon CVEC unit cost for FY 2000. Sulfuric Acid use based upon estimates from Brelsford Engineering.
4. Lime addition costs and requirements based upon NREL TP-580-26157 study adjusted for Brelsford study flows.
5. Gypsum removal costs and requirements based upon NREL TP-580-26157 study adjusted for Brelsford study flows.
6. Steam costs based upon CVEC unit costs for FY 2000. Steam use based upon estimates from Brelsford Engineering.
7 Electric Power costs based upon CVEC unit costs for FY 2000

**Attachment 3a**
**Reduced Infeed Equipment Estimate**
**Capital Equipment Additions for Brelsford Process**

| | Equip. Number | Equipment Description | Source | Size Ratio Current/ Base | Original Cost | Base Year | Scaling Exponent | Scaled Cost in Base Year | Install Factor | Installed Cost In Base Year | Installed Cost in 1999 |
|---|---|---|---|---|---|---|---|---|---|---|---|
| **Infrastructure Requirements** | | | | | | | | | | | |
| 1 | | Boiler | Richardson | 0.40 | $ 195,000 | 1994 | 0.40 | $ 135,163 | 1.5 | $ 202,745 | $ 235,037 |
| 1 | | Chiller | Richardson | 1.00 | $ 92,649 | 1994 | 1.00 | $ 92,649 | 1.50 | $ 138,974 | $ 161,108 |
| 1 | | Cooling Tower | Richardson | 1.00 | $ 18,084 | 1994 | 1.00 | $ 18,084 | 2.80 | $ 50,635 | $ 58,700 |
| 1 | | Chiller Pump | Richardson | 1.00 | $ 4,800 | 1994 | 1.00 | $ 4,800 | 2.80 | $ 13,440 | $ 15,581 |
| 1 | | Condens. Pump | Richardson | 1.00 | $ 4,800 | 1994 | 1.00 | $ 4,800 | 2.80 | $ 13,440 | $ 15,581 |
| | | | | | | | | | | **Total Plant Modifications** | **$ 486,007** |
| **Brelsford Stage 1** | | | | | | | | | | | |
| 1 | CF-1 | Wet Grains Feed conveyor | Brelsford | 1.38 | $ 4,080 | 1999 | 0.51 | $ 4,804 | 1.2 | $ 5,765 | $ 5,765 |
| | | Slurry Mix Tank | Delta-T | 1.00 | $ 21,136 | 1995 | 1.00 | $ 21,136 | 1.2 | $ 25,363 | $ 29,168 |
| 1 | M-1 | Conveyor | Brelsford | 1.38 | $ 8,300 | 1999 | 0.51 | $ 9,773 | 1.2 | $ 11,728 | $ 11,728 |
| 1 | D-1 | Disintegrator | Brelsford | 1.38 | $ 18,900 | 1999 | 0.79 | $ 24,343 | 1.3 | $ 31,647 | $ 31,647 |
| 1 | P-1 | Progressive cavity Pump | Brelsford | 1.38 | $ 8,800 | 1999 | 0.79 | $ 11,335 | 2.8 | $ 31,737 | $ 31,737 |
| 1 | HH-1 | Hydro Heater | Brelsford | 1.38 | $ 4,850 | 1999 | 0.79 | $ 6,247 | 2.8 | $ 17,491 | $ 17,491 |
| 1 | HX-1 | Heat Exchanger | Atlas | 1.00 | | | | | | | $ 18,100 |
| 1 | HX-2 | Heat Exchanger | Atlas | 1.00 | | | | | | | $ 17,050 |
| 2 | PFR-1 | Double Tube reactor | Brelsford | 1.38 | $ 331,000 | 1999 | 0.79 | $ 426,333 | 1.5 | $ 639,500 | $ 639,500 |
| 1 | T-1 | Flash vessel | Brelsford | 1.38 | $ 11,000 | 1999 | 0.71 | $ 13,810 | 1.4 | $ 19,333 | $ 19,333 |
| 1 | P-1 | Slurry Pump | Brelsford | 1.38 | $ 2,500 | 1999 | 0.79 | $ 3,220 | 2.8 | $ 9,016 | $ 9,016 |
| 1 | Cy-1 | Flash Cyclone | Brelsford | 1.38 | $ 11,000 | 1999 | 0.79 | $ 14,168 | 1.4 | $ 19,835 | $ 19,835 |
| 1 | S-1 | Screen | Brelsford | 1.38 | $ 5,500 | 1999 | 0.79 | $ 7,084 | 2.8 | $ 19,835 | $ 19,835 |
| 1 | RF-1 | Rotary Filter | Brelsford | 1.38 | $ 54,800 | 1999 | 0.79 | $ 70,583 | 2.8 | $ 197,633 | $ 197,633 |
| 1 | T-2 | Surge Tank | Brelsford | 1.38 | $ 13,600 | 1999 | 0.79 | $ 17,517 | 1.4 | $ 24,524 | $ 24,524 |
| 1 | | Conveyors to second stage | Brelsford | 1.38 | $ 5,500 | 1999 | 0.79 | $ 7,084 | 1.2 | $ 8,501 | $ 8,501 |
| 1 | | Pump to Lime Addition | Brelsford | 1.38 | $ 2,500 | 1999 | 0.79 | $ 3,220 | 2.8 | $ 9,016 | $ 9,016 |
| | | | | | | | | | | **Total Brelsford Stage 1** | **$ 1,109,879** |
| **Brelsford Stage 2** | | | | | | | | | | | |
| 1 | T-201 | Acid Storage Tank | NREL | 0.24 | $ 42,500 | 1997 | 0.51 | $ 20,389 | 1.2 | $ 24,467 | $ 25,957 |
| 1 | P-201 | Acid Storage Tank Pump | NREL | 0.24 | $ 8,000 | 1997 | 0.79 | $ 2,564 | 2.8 | $ 7,180 | $ 7,617 |
| 1 | A-201 | Acid Mixer | NREL | 0.23 | $ 1,900 | 1997 | 0.48 | $ 935 | 1.0 | $ 935 | $ 992 |
| 1 | P-1 | Progressive cavity Pump | Brelsford | 1.38 | $ 8,800 | 1999 | 0.79 | $ 11,335 | 2.8 | $ 31,737 | $ 31,737 |
| 1 | HH-1 | Hydro Heater | Brelsford | 1.38 | $ 4,850 | 1999 | 0.79 | $ 6,247 | 2.8 | $ 17,491 | $ 17,491 |
| 1 | HX-1 | Heat Exchanger | Atlas | 1.00 | | | | | | | $ 18,100 |
| 1 | HX-2 | Heat Exchanger | Atlas | 1.00 | | | | | | | $ 17,050 |
| 1 | PFR-1 | Double Tube reactor | Brelsford | 1.38 | $ 165,500 | 1999 | 0.79 | $ 213,167 | 1.5 | $ 319,750 | $ 319,750 |
| 1 | T-1 | Flash vessel | Brelsford | 1.38 | $ 11,000 | 1999 | 0.71 | $ 13,810 | 1.4 | $ 19,333 | $ 19,333 |
| 1 | P-1 | Slurry Pump | Brelsford | 1.38 | $ 2,500 | 1999 | 0.79 | $ 3,220 | 2.8 | $ 9,016 | $ 9,016 |
| 1 | Cy-1 | Flash Cyclone | Brelsford | 1.38 | $ 11,000 | 1999 | 0.79 | $ 14,168 | 1.4 | $ 19,835 | $ 19,835 |
| 1 | S-1 | Screen | Brelsford | 1.38 | $ 5,500 | 1999 | 0.79 | $ 7,084 | 2.8 | $ 19,835 | $ 19,835 |
| 1 | RF-1 | Rotary Filter | Brelsford | 1.38 | $ 54,800 | 1999 | 0.79 | $ 70,583 | 2.8 | $ 197,633 | $ 197,633 |
| 1 | T-2 | Surge Tank | Brelsford | 1.38 | $ 13,600 | 1999 | 0.79 | $ 17,517 | 1.4 | $ 24,524 | $ 24,524 |
| 1 | | Conveyors to dryer | Brelsford | 1.38 | $ 5,500 | 1999 | 0.79 | $ 7,084 | 1.2 | $ 8,501 | $ 8,501 |
| 1 | | Slurry Pump to 1st Stage | Brelsford | 1.38 | $ 2,500 | 1999 | 0.79 | $ 3,220 | 2.8 | $ 9,016 | $ 9,016 |
| | | | | | | | | | | **Total Brelsford Stage 2** | **$ 746,388** |
| **Lime Addition Module** | | | | | | | | | | | |
| 1 | T-301 | Lime Storage Bin | NREL | 1.00 | $ 69,200 | 1997 | 0.46 | $ 69,200 | 1.3 | $ 89,960 | $ 95,439 |
| 1 | S-301 | Lime Dust Baghouse | NREL | 1.00 | $ 32,200 | 1997 | 1.00 | $ 32,200 | 1.5 | $ 48,300 | $ 51,198 |
| 1 | C-301 | Lime Solids Feeder | NREL | 0.15 | $ 3,900 | 1997 | 1.00 | $ 3,900 | 1.3 | $ 5,070 | $ 5,374 |
| 1 | T-300 | Overliming Tank | NREL | 0.15 | $ 71,000 | 1997 | 0.71 | $ 18,188 | 1.4 | $ 25,463 | $ 26,991 |
| 1 | A-300 | Overliming Tank Agitator | NREL | 0.15 | $ 19,800 | 1997 | 0.51 | $ 7,444 | 1.3 | $ 9,677 | $ 10,258 |
| 1 | P-300 | Overlimed Hydrolyzate Pump | NREL | 0.15 | $ 10,700 | 1997 | 0.79 | $ 2,351 | 2.8 | $ 6,583 | $ 6,978 |
| 1 | S-302 | Rotary Drum Filter and Hydroclone | NREL | 0.15 | $ 165,000 | 1998 | 0.39 | $ 78,089 | 1.4 | $ 109,324 | $ 112,604 |
| 1 | P-302 | Overlimed Hydrolyzate Liquor Pump | NREL | 0.15 | $ 10,800 | 1997 | 0.79 | $ 2,373 | 2.8 | $ 6,644 | $ 6,844 |
| 1 | T-303 | Reacidification Tank | NREL | 0.15 | $ 147,800 | 1997 | 0.51 | $ 55,566 | 1.2 | $ 66,679 | $ 68,680 |
| 1 | A-303 | Reacidificaiton Tank Agitator | NREL | 0.15 | $ 65,200 | 1997 | 0.51 | $ 24,512 | 1.2 | $ 29,415 | $ 30,297 |
| 1 | P-302 | Fermentation Feed Pump | NREL | 0.19 | $ 61,368 | 1998 | 0.70 | $ 19,122 | 2.8 | $ 53,541 | $ 55,147 |
| 1 | A-304 | In-line Acid Mixer | NREL | 0.15 | $ 2,600 | 1997 | 0.48 | $ 1,035 | 1.0 | $ 1,035 | $ 1,098 |
| 1 | B-300 | Lime unloading Blower | NREL | 0.15 | $ 47,600 | 1998 | 0.50 | $ 18,242 | 1.4 | $ 25,539 | $ 26,305 |
| 1 | P-303 | Reacidification Liquor Pump | NREL | 0.15 | $ 10,800 | 1997 | 0.79 | $ 2,373 | 2.8 | $ 6,644 | $ 6,844 |
| 1 | | Heat Exchanger | Atlas | 1.00 | | | | | | | $ 18,100 |
| | | | | | | | | | | **Total Lime Addition Module** | **$ 522,155** |
| **Seed Fermentation** | | | | | | | | | | | |
| 2 | A-304 | Seed Vessel Agitator | NREL | 0.91 | $ 11,700 | 1997 | 0.51 | $ 22,295 | 1.2 | $ 26,754 | $ 28,360 |
| 2 | A-305 | Seed Vessel Agitator | NREL | 0.91 | $ 10,340 | 1996 | 0.51 | $ 19,704 | 1.2 | $ 23,644 | $ 25,772 |
| 2 | F-301 | 1st SSCF Seed Fermentor | NREL | 1.00 | $ 14,700 | 1997 | 0.91 | $ 29,400 | 2.8 | $ 82,320 | $ 87,259 |
| 2 | F-302 | 2nd SSCF Seed Fermentor | NREL | 1.00 | $ 32,600 | 1997 | 0.93 | $ 65,200 | 2.8 | $ 182,560 | $ 193,514 |
| 2 | F-303 | 3rd SSCF Seed Fermentor | NREL | 1.00 | $ 81,100 | 1997 | 0.93 | $ 162,200 | 2.8 | $ 454,160 | $ 481,410 |
| 2 | F-304 | 4th SSCF Seed Fermentor | NREL | 0.91 | $ 39,500 | 1997 | 0.93 | $ 72,331 | 1.2 | $ 86,798 | $ 92,005 |
| 1 | H-301 | SSCF Seed Hydrolyzate Cooler | NREL | 0.91 | $ 15,539 | 1998 | 0.78 | $ 14,414 | 2.1 | $ 30,270 | $ 31,178 |
| 3 | H-302 | SSCF Hydrolyzate Cooler | NREL | 0.98 | $ 25,409 | 1998 | 0.78 | $ 75,201 | 2.1 | $ 157,922 | $ 162,660 |
| 1 | H-304 | Seed Fermentor Coils | NREL | 0.92 | $ 3,300 | 1997 | 0.83 | $ 3,066 | 1.2 | $ 3,679 | $ 3,899 |
| 2 | P-302 | Seed Transfer Pump | NREL | 0.91 | $ 54,088 | 1998 | 0.70 | $ 101,228 | 1.4 | $ 141,720 | $ 145,971 |
| | | | | | | | | | | **Total Seed Fermentation Module** | **$1,252,028** |
| | | | | | | | | | | Eng./Const.Mgt. | 1975899 |
| | | | | | | | | | | **Total Incremental Capital Equipment** | **$ 6,092,356** |

Notes

1. Brelsford Module Costs based upon estimates from Breslford Engineering "Distiller grains Cellulose Hydrolysis to Fermentatable sugars for Production of Fuel Ethanol and Hi-Protein Feedstuff" Preliminary Engineering and Economics Feasibility Study.

2. Lime Addition Module Costs based upon July 1999 NREL/TP-580-26157 Study Lignocellulosic Biomass to Ethanol Process Design and Economics Utilizing Co-Current Dilute Acid Prehydrolysis and Enzymatic Hydrolysis current and Futuristic Scenarios. Costs were adjusted for sulfuric acid flowrates.

3. Seed fermentation Module Costs based upon July 1999 NREL/TP-580-26157.

# Attachment 3b
## Reduced Infeed Proforma Economics

| | Units | Unit Price $ | Quantity/yr | Infeed Reduction | | $/Yr. |
|---|---|---|---|---|---|---|
| **Revenue** | | | | | | |
| Increased Ethanol | gallons | 1.10 | 180,000 | 85% | $ | 198,000 |
| Reduction in DDGS Value | Tons | 67.24 | -8559.75 | 85% | $ | (575,558) |
| | | | Revenue Reduction | | | (377558) |
| **Operating Cost** | | | | | | |
| Existing Operation Reduction | | | | | | |
| Corn Feedstock reduction | bushels | 1.67 | -1051273.05 | | $ | (1,755,626) |
| Reduction in Variable Cost | $/bushel | 0.27 | -1051273.05 | | $ | (283,591) |
| | | | | | | |
| Brelsford Process Addition | | | | | | |
| Sulfuric Acid | Tons | 96.88 | 4542 | 85% | $ | 374,030 |
| Lime | Tons | 60.00 | 5059.33 | 85% | $ | 258,026 |
| Gypsum Removal | $/# | 0.01 | 34479975 | 85% | $ | 293,080 |
| Steam | MM BTU | 5.31 | 74460 | 85% | $ | 336,075 |
| Electric Power | $/KWh | 0.04 | 7787640 | 85% | $ | 238,302 |
| Shift operators | 1/shift | 28000.00 | 4 | | $ | 112,000 |
| Maint.Cost-5% Capital | | | | | $ | 304,618 |
| | | **Total Incremental Operating Cost** | | | $ | (123,087) |
| | | | | **Net Increased Revenue** | $ | (254,470) |
| | | | | **Capital Cost** | $ | 6,092,356 |
| | | | **Capital payback period yrs.** | | | negative payback |

Notes

1. Reduced Infeed Rate is Based upon holding total flow to existing fermentation at 144000 #/hr. This results in a reduction in feedstock and chemical costs of 15%. Power and Steam requirements are held constant for the existing facility.

2. Brelsford process, lime addition, and seed fermentation operating costs have been adjusted to 85% of base case to adjust for reduction in infeed of facility.

**Attachment 4a**
**Equipment Cost Estimate with New Dryer and Centrifuges**
**Capital Equipment Additions for Brelsford Process**

| Quantity | Equip. Number | Equipment Description | Source | Size Ratio Current/Base | Original Cost | Base Year | Scaling Exponent | Scaled Cost in Base Year | Install Factor | Installed Cost In Base Year | Installed Cost in 1999 |
|---|---|---|---|---|---|---|---|---|---|---|---|
| **Infrastructure Requirements** | | | | | | | | | | | |
| 1 | | Boiler | Richardson | 0.40 | $ 195,000 | 1994 | 0.40 | $ 135,163 | 1.5 | $ 202,745 | $ 235,037 |
| 1 | | Chiller | Richardson | 1.00 | $ 92,649 | 1994 | 1.00 | $ 92,649 | 1.50 | $ 138,974 | $ 161,108 |
| 1 | | Cooling Tower | Richardson | 1.00 | $ 18,084 | 1994 | 1.00 | $ 18,084 | 2.80 | $ 50,635 | $ 58,700 |
| 1 | | Chiller Pump | Richardson | 1.00 | $ 4,800 | 1994 | 1.00 | $ 4,800 | 2.80 | $ 13,440 | $ 15,581 |
| 1 | | Condens. Pump | Richardson | 1.00 | $ 4,800 | 1994 | 1.00 | $ 4,800 | 2.80 | $ 13,440 | $ 15,581 |
| | | | | | | | | | | **Total Plant Modifications** | **$ 486,007** |
| **Brelsford Stage 1** | | | | | | | | | | | |
| 1 | CF-1 | Wet Grains Feed conveyor | Brelsford | 1.38 | $ 4,080 | 1999 | 0.51 | $ 4,804 | 1.2 | $ 5,765 | $ 5,765 |
| | | Slurry Mix Tank | Delta-T | 1.00 | $ 21,136 | 1995 | 1.00 | $ 21,136 | 1.2 | $ 25,363 | $ 29,168 |
| 1 | M-1 | Conveyor | Brelsford | 1.38 | $ 8,300 | 1999 | 0.51 | $ 9,773 | 1.2 | $ 11,728 | $ 11,728 |
| 1 | D-1 | Disintegrator | Brelsford | 1.38 | $ 18,900 | 1999 | 0.79 | $ 24,343 | 1.3 | $ 31,647 | $ 31,647 |
| 1 | P-1 | Progressive cavity Pump | Brelsford | 1.38 | $ 8,800 | 1999 | 0.79 | $ 11,335 | 2.8 | $ 31,737 | $ 31,737 |
| 1 | HH-1 | Hydro Heater | Brelsford | 1.38 | $ 4,850 | 1999 | 0.79 | $ 6,247 | 2.8 | $ 17,491 | $ 17,491 |
| 1 | HX-1 | Heat Exchanger | Atlas | 1.00 | | | | | | | $ 18,100 |
| 1 | HX-2 | Heat Exchanger | Atlas | 1.00 | | | | | | | $ 17,050 |
| 2 | PFR-1 | Double Tube reactor | Brelsford | 1.38 | $ 496,500 | 1999 | 0.79 | $ 639,500 | 1.5 | $ 959,249 | $ 959,249 |
| 1 | T-1 | Flash vessel | Brelsford | 1.38 | $ 11,000 | 1999 | 0.71 | $ 13,810 | 1.4 | $ 19,333 | $ 19,333 |
| 1 | P-1 | Slurry Pump | Brelsford | 1.38 | $ 2,500 | 1999 | 0.79 | $ 3,220 | 2.8 | $ 9,016 | $ 9,016 |
| 1 | Cy-1 | Flash Cyclone | Brelsford | 1.38 | $ 11,000 | 1999 | 0.79 | $ 14,168 | 1.4 | $ 19,835 | $ 19,835 |
| | | Centrifuge | Delta-T | 1.00 | $ 530,000 | 1995 | 1.00 | $ 530,000 | 1.2 | $ 636,000 | $ 715,824 |
| 1 | S-1 | Screen | Brelsford | 1.38 | | 1999 | 0.79 | $ - | 2.8 | $ - | $ - |
| 1 | RF-1 | Rotary Filter | Brelsford | 1.38 | | 1999 | 0.79 | $ - | 2.8 | $ - | $ - |
| 1 | T-2 | Surge Tank | Brelsford | 1.38 | $ 13,600 | 1999 | 0.79 | $ 17,517 | 1.4 | $ 24,524 | $ 24,524 |
| 1 | | Conveyors to second stage | Brelsford | 1.38 | $ 5,500 | 1999 | 0.79 | $ 7,084 | 1.2 | $ 8,501 | $ 8,501 |
| 1 | | Pump to Lime Addition | Brelsford | 1.38 | $ 2,500 | 1999 | 0.79 | $ 3,220 | 2.8 | $ 9,016 | $ 9,016 |
| | | | | | | | | | | **Total Brelsford Stage 1** | **$ 1,927,984** |
| **Brelsford Stage 2** | | | | | | | | | | | |
| 1 | T-201 | Acid Storage Tank | NREL | 0.24 | $ 42,500 | 1997 | 0.51 | $ 20,389 | 1.2 | $ 24,467 | $ 25,957 |
| 1 | P-201 | Acid Storage Tank Pump | NREL | 0.24 | $ 8,000 | 1997 | 0.79 | $ 2,564 | 2.8 | $ 7,180 | $ 7,617 |
| 1 | A-201 | Acid Mixer | NREL | 0.23 | $ 1,900 | 1997 | 0.48 | $ 935 | 1.0 | $ 935 | $ 992 |
| 1 | P-1 | Progressive cavity Pump | Brelsford | 1.38 | $ 8,800 | 1999 | 0.79 | $ 11,335 | 2.8 | $ 31,737 | $ 31,737 |
| 1 | HH-1 | Hydro Heater | Brelsford | 1.38 | $ 4,850 | 1999 | 0.79 | $ 6,247 | 2.8 | $ 17,491 | $ 17,491 |
| 1 | HX-1 | Heat Exchanger | Atlas | 1.00 | | | | | | | $ 18,100 |
| 1 | HX-2 | Heat Exchanger | Atlas | 1.00 | | | | | | | $ 17,050 |
| 1 | PFR-1 | Double Tube reactor | Brelsford | 1.38 | $ 165,500 | 1999 | 0.79 | $ 213,167 | 1.5 | $ 319,750 | $ 319,750 |
| 1 | T-1 | Flash vessel | Brelsford | 1.38 | $ 11,000 | 1999 | 0.71 | $ 13,810 | 1.4 | $ 19,333 | $ 19,333 |
| 1 | P-1 | Slurry Pump | Brelsford | 1.38 | $ 2,500 | 1999 | 0.79 | $ 3,220 | 2.8 | $ 9,016 | $ 9,016 |
| 1 | Cy-1 | Flash Cyclone | Brelsford | 1.38 | $ 11,000 | 1999 | 0.79 | $ 14,168 | 1.4 | $ 19,835 | $ 19,835 |
| | | Centrifuge | Delta-T | 1.00 | $ 530,000 | 1995 | 1.00 | $ 530,000 | 1.2 | $ 636,000 | $ 715,824 |
| 1 | S-1 | Screen | Brelsford | 1.38 | | 1999 | 0.79 | $ - | 2.8 | $ - | $ - |
| 1 | RF-1 | Rotary Filter | Brelsford | 1.38 | | 1999 | 0.79 | $ - | 2.8 | $ - | $ - |
| 1 | T-2 | Surge Tank | Brelsford | 1.38 | $ 13,600 | 1999 | 0.79 | $ 17,517 | 1.4 | $ 24,524 | $ 24,524 |
| 1 | | Conveyors to dryer | Brelsford | 1.38 | $ 5,500 | 1999 | 0.79 | $ 7,084 | 1.2 | $ 8,501 | $ 8,501 |
| 1 | | Slurry Pump to 1st Stage | Brelsford | 1.38 | $ 2,500 | 1999 | 0.79 | $ 3,220 | 2.8 | $ 9,016 | $ 9,016 |
| | | | | | | | | | | **Total Brelsford Stage 2** | **$ 1,244,743** |
| **Lime Addition Module** | | | | | | | | | | | |
| 1 | T-301 | Lime Storage Bin | NREL | 1.00 | $ 69,200 | 1997 | 0.46 | $ 69,200 | 1.3 | $ 89,960 | $ 95,439 |
| 1 | S-301 | Lime Dust Baghouse | NREL | 1.00 | $ 32,200 | 1997 | 1.00 | $ 32,200 | 1.5 | $ 48,300 | $ 51,198 |
| 1 | C-301 | Lime Solids Feeder | NREL | 0.15 | $ 3,900 | 1997 | 1.00 | $ 3,900 | 1.3 | $ 5,070 | $ 5,374 |
| 1 | T-300 | Overliming Tank | NREL | 0.15 | $ 71,000 | 1997 | 0.71 | $ 18,188 | 1.4 | $ 25,463 | $ 26,991 |
| 1 | A-300 | Overliming Tank Agitator | NREL | 0.15 | $ 19,800 | 1997 | 0.51 | $ 7,444 | 1.3 | $ 9,677 | $ 10,258 |
| 1 | P-300 | Overlimed Hydrolyzate Pump | NREL | 0.15 | $ 10,700 | 1997 | 0.79 | $ 2,351 | 2.8 | $ 6,583 | $ 6,978 |
| 1 | S-302 | Rotary Drum Filter and Hydroclone | NREL | 0.15 | $ 165,000 | 1998 | 0.39 | $ 78,089 | 1.4 | $ 109,324 | $ 112,604 |
| 1 | P-302 | Overlimed Hydrolyzate Liquor Pump | NREL | 0.15 | $ 10,800 | 1997 | 0.79 | $ 2,373 | 2.8 | $ 6,644 | $ 6,844 |
| 1 | T-303 | Reacidification Tank | NREL | 0.15 | $ 147,800 | 1997 | 0.51 | $ 55,566 | 1.2 | $ 66,679 | $ 68,680 |
| 1 | A-303 | Reacidificaiton Tank Agitator | NREL | 0.15 | $ 65,200 | 1997 | 0.51 | $ 24,512 | 1.2 | $ 29,415 | $ 30,297 |
| 1 | P-302 | Fermentation Feed Pump | NREL | 0.19 | $ 61,368 | 1998 | 0.70 | $ 19,122 | 2.8 | $ 53,541 | $ 55,147 |
| 1 | A-304 | In-line Acid Mixer | NREL | 0.15 | $ 2,600 | 1997 | 0.48 | $ 1,035 | 1.0 | $ 1,035 | $ 1,098 |
| 1 | B-300 | Lime unloading Blower | NREL | 0.15 | $ 47,600 | 1998 | 0.50 | $ 18,242 | 1.4 | $ 25,539 | $ 26,305 |
| 1 | P-303 | Reacidification Liquor Pump | NREL | 0.15 | $ 10,800 | 1997 | 0.79 | $ 2,373 | 2.8 | $ 6,644 | $ 6,844 |
| 1 | | Heat Exchanger | Atlas | 1.00 | | | | | | | $ 18,100 |
| | | | | | | | | | | **Total Lime Addition Module** | **$ 522,155** |
| **Seed Fermentation** | | | | | | | | | | | |
| 2 | A-304 | Seed Vessel Agitator | NREL | 0.91 | $ 11,700 | 1997 | 0.51 | $ 22,295 | 1.2 | $ 26,754 | $ 28,360 |
| 2 | A-305 | Seed Vessel Agitator | NREL | 0.91 | $ 10,340 | 1996 | 0.51 | $ 19,704 | 1.2 | $ 23,644 | $ 25,772 |
| 2 | F-301 | 1st SSCF Seed Fermentor | NREL | 1.00 | $ 14,700 | 1997 | 0.91 | $ 29,400 | 2.8 | $ 82,320 | $ 87,259 |
| 2 | F-302 | 2nd SSCF Seed Fermentor | NREL | 1.00 | $ 32,600 | 1997 | 0.93 | $ 65,200 | 2.8 | $ 182,560 | $ 193,514 |
| 2 | F-303 | 3rd SSCF Seed Fermentor | NREL | 1.00 | $ 81,100 | 1997 | 0.93 | $ 162,200 | 2.8 | $ 454,160 | $ 481,410 |
| 2 | F-304 | 4th SSCF Seed Fermentor | NREL | 0.91 | $ 39,500 | 1997 | 0.93 | $ 72,331 | 1.2 | $ 86,798 | $ 92,005 |
| 1 | H-301 | SSCF Seed Hydrolyzate Cooler | NREL | 0.91 | $ 15,539 | 1998 | 0.78 | $ 14,414 | 2.1 | $ 30,270 | $ 31,178 |
| 3 | H-302 | SSCF Hydrolyzate Cooler | NREL | 0.98 | $ 25,409 | 1998 | 0.78 | $ 75,201 | 2.1 | $ 157,922 | $ 162,660 |
| 1 | H-304 | Seed Fermentor Coils | NREL | 0.92 | $ 3,300 | 1997 | 0.83 | $ 3,066 | 1.2 | $ 3,679 | $ 3,899 |
| 2 | P-302 | Seed Transfer Pump | NREL | 0.91 | $ 54,088 | 1998 | 0.70 | $ 101,228 | 1.4 | $ 141,720 | $ 145,971 |
| | | | | | | | | | | **Total Seed Fermentation Module** | **$ 1,252,028** |
| **Replacement Dryer** | | | | | | | | | | | |
| | | Dryer | Delta T | 1.00 | $ 1,015,529 | 1,995 | 0.80 | $ 1,015,529 | 1.20 | $1,218,635 | **$ 1,371,584** |
| | | | | | | | | | | Eng./Const.Mgt. | $ 3,266,161 |
| | | | | | | | | | | **Total Incremental Capital Equipment** | **$ 10,070,662** |

Notes

1. Additional costs to base case equipment estimate have been added for inclusion of new centrifuges on each Brelsford module. In addition costs have been included for replacement of the dryer for a unit that might better handle the Brelsford process low fiber DDGS.

# Attachment 4b

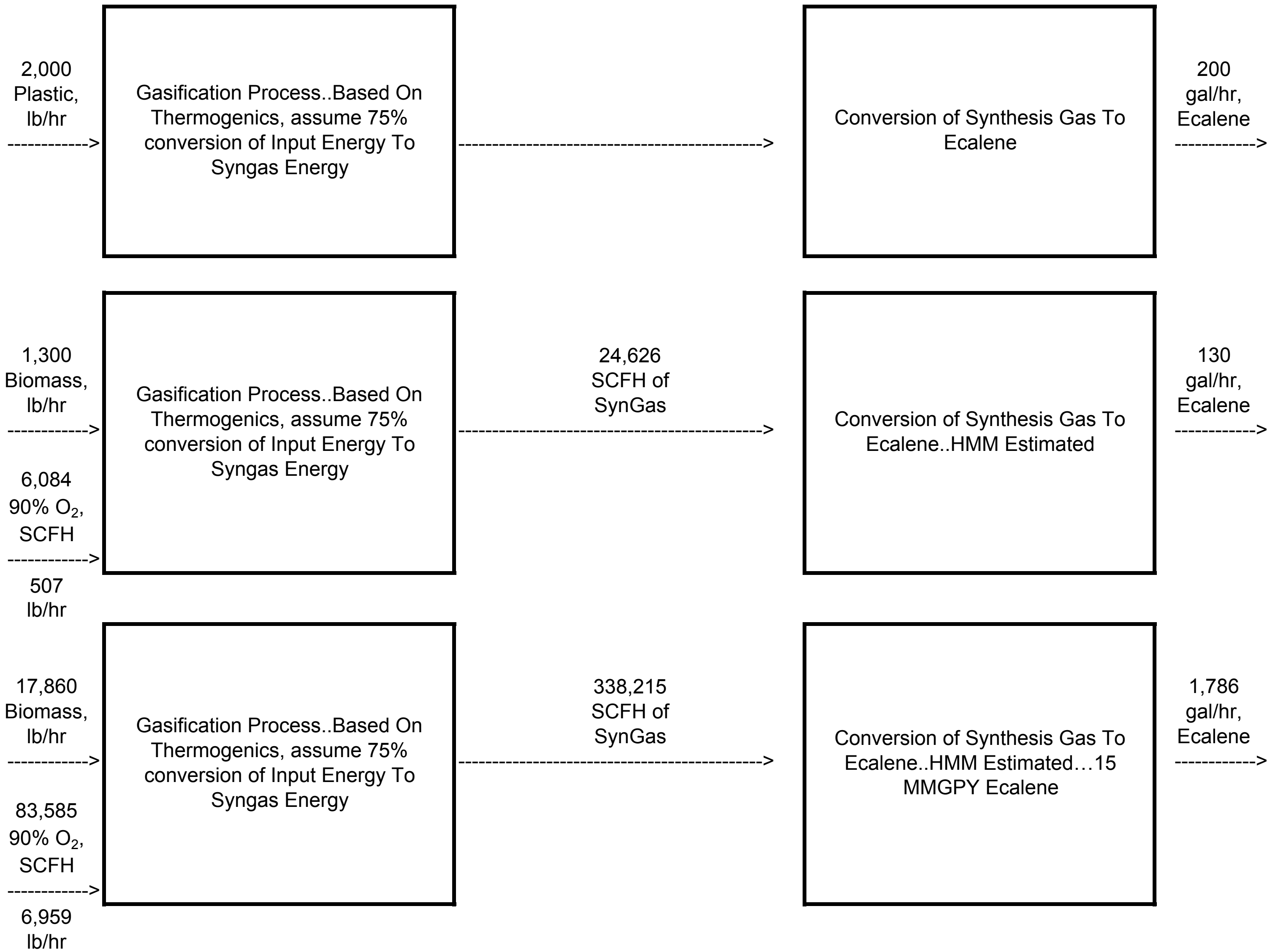
## Equipment Cost with New Dryer and Centrifuges Proforma Economics

| | Units | Unit Price $ | Quantity/yr | Infeed Reduction | $/Yr. |
|---|---|---|---|---|---|
| **Revenue** | | | | | |
| Increased Ethanol | gallons | 1.10 | 180,000 | 85% | $ 198,000 |
| Reduction in DDGS Value | Tons | 67.24 | -8559.75 | 85% | $ (575,558) |
| | | | Revenue Reduction | | (377558) |
| **Operating Cost** | | | | | |
| Existing Operation Derating | | | | | |
| Corn Feedstock reduction | bushels | 1.67 | -1051273.05 | | $ (1,755,626) |
| Reduction in Variable Cost | $/bushel | 0.27 | -1051273.05 | | $ (283,591) |
| | | | | | |
| Brelsford Process Addition | | | | | |
| Sulfuric Acid | Tons | 96.88 | 4542 | 85% | $ 374,030 |
| Lime | Tons | 60.00 | 5059.33 | 85% | $ 258,026 |
| Gypsum Removal | $/# | 0.01 | 34479975 | 85% | $ 293,080 |
| Steam | MM BTU | 5.31 | 74460 | 85% | $ 336,075 |
| Electric Power | $/KWh | 0.04 | 7787640 | 85% | $ 238,302 |
| Shift operators | 1/shift | 28000.00 | 4 | | $ 112,000 |
| Maint.Cost-5% Capital | | | | | $ 503,533 |
| | | | **Total Incremental Operating Cost** | | $ 75,828 |
| | | | | **Net Increased Revenue** | $ (453,385) |
| | | | | **Capital Cost** | $ 10,070,662 |
| | | | | **Capital payback period yrs.** | negative payback |

Notes

1. Additional costs to base case equipment estimate have been added for inclusion of new centrifuges on each Brelsford module. In addition costs have been included for replacement of the dryer for a unit that might better handle the Brelsford process low fiber DDGS.

**Attachment 5a**
**Conceptual Feed Alternatives..Ecalene Production**



## Attachment 5b

## Conceptual Capital Cost Estimate..15 MMGPY Ecalene Production

| **Capital Cost** | Base Capacity | Capacity Units | Base Cost, Current USD | Base Cost Source | Current Capacity | Capacity Units | Cost, Current USD | Source |
|---|---|---|---|---|---|---|---|---|
| Feed Handling & Processing | 70 | DSTPH | 5,200,000 | NREL 21May98 Report | 9 | DSTPH | 1,511,670 | Delta-T Conceptual Estimate |
| Oxygen Plant | | | | Vendor Budget Price | 83,585 | SCFH | 3,000,000 | Vendor Budget Price |
| SynGas Generation & Clean-up | 300 | DSTPD | 5,000,000 | Delta-T | 182 | DSTPD | 3,706,690 | Delta-T Conceptual Estimate |
| Ecalene Production | 1 | USGAL/YR | 1.00 | Vendor Budget Price | 15,002,400 | 1.00 | 15,002,400 | Delta-T Conceptual Estimate |
| Ecalene Dehydration & Processing | 3,630 | USGPH | 2,800,000 | Delta-T | 1,786 | USGPH | 1,829,550 | Delta-T Conceptual Estimate |
| Ecalene Storage & Loading | 7,250 | USGPH | 2,300,000 | NREL 21May98 Report | 1,786 | USGPH | 992,330 | Delta-T Conceptual Estimate |
| Utilities | 7,250 | USGPH | 5,000,000 | NREL 21May98 Report | 1,786 | USGPH | 1,078,620 | Delta-T Conceptual Estimate |
| Waste Water Treatment | 1,225 | USGPM | 9,000,000 | NREL 21May98 Report | 48 | USGPM | 1,280,850 | Delta-T Conceptual Estimate |
| Total Installed Equipment Cost | | | | | | | 28,402,110 | |
| Added Cost To Get To Total Project Investment | | | | 73% of sum of Area Capital | | | 19,798,520 | NREL 21May98 Report |
| **Total Capital Cost** | | | | | | | 48,200,630 | |

## Attachment 5c

## Conceptual Operating Cost Estimate..15 MMGPY Ecalene Production

**Operating hours/yr:** **8400**
**Interest Rate:** **10.0%**
**Time Period, Yrs:** **8**
**Capital Factor:** **0.1874**
**Annual UDA ETOH:** **15,000,000**

| **Variable Annual Operating Cost** | Annual Quantity | Capacity | Unit Cost | Cost, Current | Source |
|---|---|---|---|---|---|
| | | | | | |
| Feedstock | 75,012 | DST | 15.00 | 1,125,180 | Unit Cost NREL 21May98 Report |
| Solids Disposal | 3,380 | DST | 30.00 | 101,400 | Unit Cost NREL 21May98 Report |
| Electrical Power Consumption | 58 | $10^6$ KWH | 45,000.00 | 2,608,200 | Unit Cost NREL 21May98 Report |
| Denaturant | - | USGAL | 1.00 | - | Not required for product |
| Chemicals & Catalyst | | | | 2,606,483 | Delta-T Estimate |
| Make-up Water | 94 | $10^6$ USGAL | 1,065.00 | 100,110 | Unit Cost NREL 21May98 Report |
| **Total Annual Variable Operating Cost** | | | | 6,541,373 | |
| | | | | | |
| **Fixed Annual Operating Cost** | | | | | |
| Operating Labor | 10 | EA | 37,120.00 | 371,200 | Unit Cost NREL 21May98 Report |
| Operating Foreman | 2 | EA | 42,532.56 | 85,065 | Unit Cost NREL 21May98 Report |
| Supervision | 1 | EA | 52,200.00 | 52,200 | Unit Cost NREL 21May98 Report |
| Direct Overhead | 1 | EA | 45% of L&M | 228,809.30 | Unit Cost NREL 21May98 Report |
| Maintenance Material & Labor | 1 | EA | 4% of Fixed | 1,928,025 | Unit Cost NREL 21May98 Report |
| Plant Overhead | 1 | EA | 65% of L&M | 125,322 | Unit Cost NREL 21May98 Report |
| Insurance & Taxes | 1 | EA | 1.5% of Fixed | 723,009 | Unit Cost NREL 21May98 Report |
| **Total AnnualFixed Operating Cost** | | | | 3,513,631 | |
| **Total Annual Operating Cost** | | | | 10,055,004 | |
| **Total Annualized Capital Cost** | | | | 9,034,920 | |
| **Total Annual Cost** | | | | 19,089,923 | |
| **Annual Cost/Undenatured Gal ETOH** | | | | 1.273 | |

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Delta T Corporation
Williamsburg, Virginia

8. PERFORMING ORGANIZATION REPORT NUMBER
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