4-1-2010

ABE FERMENTATION OF SUGAR IN BRAZIL

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Abstract

A fermentation plant was designed to ferment and process sugar cane juice into acetone, butanol, and ethanol (ABE) in Brazil. The plant was built to handle a feed of 40 tonnes of sugar per hour in 25% solution. The process runs continuously for 32 weeks out of the year, during the cane harvest, for 20 years.

The two main steps of the process are the fermentation and the separation of the ABE products into the desired 99.5% product purities. The fermentation section of the plant consists of nine 500,000 gallon fermenters that convert the bulk of the sugar cane into ABE products, as well as two 500,000 gallon fermenters that supply fresh cells to these fermenters and a series of smaller tanks that scale up cell concentrations from a test tube scale to the fermenter sizes used in this project. The separation section of the plant consists of a holding tank to store the ABE products, a gas stripper to remove most of the organics from water, a decanter to further separate the products into a butanol-rich phase and a water-rich phase, molecular sieves to remove the rest of the water from the butanol-rich phase, and two distillation columns to purify the products and prepare them for sale.

This design can be deemed a successful one with a 35.67% return on investment and a net present value $118,806,000. Also, the process as a whole was found to be significantly energy positive, with our combustible products having a fuel value of 3.36x108 BTU/hr and our utility inputs being only 2.14x106 BTU/hr. The main reason for our success on these two fronts was the use of a gas stripper and a molecular sieve, which allowed for most of the water in the separation step to be removed without needing to heat it.

Disciplines

Biochemical and Biomolecular Engineering

Senior Design Report (CBE)

University of Pennsylvania *Year 2010*

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ABE FERMENTATION OF SUGAR CANE IN BRAZIL

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Dr. Daniel Hammer

Professor Leonard Fabiano

Project Suggested by:

Mr. Bruce Vrana

Senior Design Report

April 13, 2010

University of Pennsylvania

Chemical and Biomolecular Engineering

University of Pennsylvania

School of Engineering and Applied Science

Department of Chemical & Biomolecular Engineering

220 South 33rd Street

Philadelphia, PA 19104

April 13th, 2010

Dear Professor Fabiano, Dr. Hammer, and Mr. Vrana,

The following is our proposed design process for the ABE fermentation of sugar cane in Brazil. Our process is split into two main sections: the fermentation of sugar cane and the separation of ABE into the final acetone, butanol, and ethanol products. The fermentation process handles the 160 tonnes of sugar cane juice being fed every hour, converting it to the desired products. The main challenge in the separation was to separate the organic products from the high amount of water and break the azeotropes of the mixtures. This process achieves the required purities of 99.5% for each of the organic products, as well as the goal of being completely self-sufficient in terms of water being used in the plant as specified in the problem statement.

The following report presents the design for the plant, the estimated equipment and utility costs, as well as a detailed economic analysis. The process is designed to operate for 20 years, for 32 weeks per year with the sugar cane harvest in Brazil.

Our process yields a Net Present Value of $118,806,000, an IRR of 37.03%, and a third-year ROI of 35.67%. A detailed economic analysis is included, as well as discussions for start-up, environmental and safety concerns, energy prices and the function of each piece of equipment being used.

Sincerely,

Marcelo C. Mansur	Maclyn K. O'Donnell
Matthew S. Rehmann	Mohammad Zohaib

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Abstract

Abstract

A fermentation plant was designed to ferment and process sugar cane juice into acetone, butanol, and ethanol (ABE) in Brazil. The plant was built to handle a feed of 40 tonnes of sugar per hour in 25% solution. The process runs continuously for 32 weeks out of the year, during the cane harvest, for 20 years.

The two main steps of the process are the fermentation and the separation of the ABE products into the desired 99.5% product purities. The fermentation section of the plant consists of nine 500,000 gallon fermenters that convert the bulk of the sugar cane into ABE products, as well as two 500,000 gallon fermenters that supply fresh cells to these fermenters and a series of smaller tanks that scale up cell concentrations from a test tube scale to the fermenter sizes used in this project. The separation section of the plant consists of a holding tank to store the ABE products, a gas stripper to remove most of the organics from water, a decanter to further separate the products into a butanol-rich phase and a water-rich phase, molecular sieves to remove the rest of the water from the butanol-rich phase, and two distillation columns to purify the products and prepare them for sale.

This design can be deemed a successful one with a 35.67% return on investment and a net present value $118,806,000. Also, the process as a whole was found to be significantly energy positive, with our combustible products having a fuel value of $3.36x10^8$ BTU/hr and our utility inputs being only $2.14x10^6$ BTU/hr. The main reason for our success on these two fronts was the use of a gas stripper and a molecular sieve, which allowed for most of the water in the separation step to be removed without needing to heat it.

Introduction

Introduction

Because of increasing worldwide energy consumption, diminishing fuel reserves, and increasing concerns about the environment, the need to find alternative energy sources is vital. Biofuels present a particularly effective way to fix these problems, since they can be used with the existing infrastructure. The problem with most biofuel processes is that they only produce ethanol, which is already fairly oxidized and thus has a relatively low heat value (19.6 MJ/L). There is an alternative process that uses a different cell line (typically a species of *Clostridia*) that produces acetone, butanol, and ethanol, or ABE, instead of just ethanol. This process makes a fuel mixture that has more industrial value than a process producing just ethanol, since butanol has a higher energy value than ethanol (29.2 MJ/L), and acetone is an important solvent and base material for making polymers. However, it produces a dilute fuel mixture that is difficult and expensive to separate. For this reason, the process has been mostly avoided by industry since the 1950s. We are reexamining the ABE process now since there are new technologies, like molecular sieves and gas strippers, that can make this process profitable. This process will be compared to a similar process which makes only ethanol and is widely used in Brazil (where our plant will be located as well).

The plant will operate for 224 days a year while the cane juice is being produced. The process begins with a fermentation step where we mix the cane juice with the cells and allow the cells to produce our product. This process uses seven fermenters each day, with two extra to account for cleaning time. All the fermenters are run in 24 hour batches, the maximum time before the cane juice goes bad, and then the liquids are removed using a bank of centrifuges. Since the production phase does not generate significant amounts of new cells, the cells from the centrifuges are then reused to seed the next fermenter. This is done five times before the cells are exhausted and are removed and dried out to be sold as fertilizer. New cells are generated in two growth fermenters, which keep conditions in the growth phase for the cells and are able to generate enough cells to seed a new fermenter every 24 hours.

To obtain acetone, butanol, and ethanol at the necessary purities, a separation train is needed. The first step of this process is where we have made a major change from how this process was done in the past. Since the stream coming out the fermenter is over 97% water, separating the water from the ABE by distillation was extremely expensive, due to the energy required to heat the water. Instead, we remove the ABE from the water using a gas stripper, which enabled us to get the water content down to 21%, with only a small loss of the ABE product and at a significantly lower cost than a column. This product stream is then sent to a decanter and to molecular sieves which get the water content down to under 300ppm. With all the water removed the rest of the products can be easily separated with two distillation columns, since the ABE mixture does not form any azeotropes. Due to our changes in the separation train we were able to take the ABE fermentation process from something generally avoided by industry to a potentially very profitable investment.

Due to the nature of this project particular attention was paid to the environmental impact of the factory. Our process recycles all the water it needs and sends a significant portion of the water back to the cane mill. In addition, this process needs to be as energy efficient as possible, and energy used should be no more than 35,000 BTU per gallon of total product, which this plant successfully accomplished. This is based on the assumed energy need to produce a gallon of ethanol.

Project Charter

Project Name	Fermentation of Sugar Cane to Acetone, Butanol, and Ethanol
Project Champions	Mr. Bruce Vrana, Professor Leonard Fabiano, Professor Daniel Hammer
Project Leaders	Marcelo Mansur, Maclyn O'Donnell, Matthew Rehmann, Mohammad Zohaib
Specific Goals	Development of a profitable ABE fermentation process that has a net positive energy balance with a novel *Clostridium* strain
Project Scope	In-scope: • Process design that is able to handle 40 tonnes of sugar cane per hour and use the sugar cane within 24 hours of its arrival at the plant • Profitable fermentation and separation to 99.5% acetone, 99.5% ethanol, and 99.5% butanol products • Net positive energy balance so that butanol and ethanol can be used as alternative fuels • Zero-discharge process design, so that all water is purified and recycled • Environmentally acceptable and economical disposal of large amounts of H_2 and CO_2 product Out-of-scope: • Experimental and laboratory process optimization work • Upstream milling and processing to generate sugar cane juice solution • Detailed consideration of *Clostridium* nutritional requirements • Batch scheduling of fermentation section
Time Line	Completed design in four months

Technology-Readiness Assessment

Fermentative production of ABE was discovered over 100 years ago: In 1861, Pasteur discovered that anaerobic bacteria could produce butanol, and, in 1905, Schardinger discovered that these bacteria could also produce acetone (Jones and Woods, 1986). In 1912, Weizmann isolated the bacterium *Clostridium acetobutylicum*, which was particularly effective at producing acetone and butanol from starchy substances, resulting in much greater solvent yields than originally discovered (Jones and Woods, 1986). This discovery led to the first successful industrial ABE fermentation process in 1916, when Great Britain used the technology to produce acetone and methyl ethyl ketone (from butanol) for use in World War I (Jones and Woods, 1986).

ABE fermentation technology was used extensively from 1916 to 1945 to produce acetone and butanol, but the rapid growth of the petrochemical industry in the 1950s and 1960s caused industrial fermentative butanol production to completely cease in North America and Europe (Chiao and Sun, 2007). Because the batch fermentation processes generally produced solutions containing 98% water or more (by weight), product recovery by distillation made the fermentation process cost-prohibitive. In addition, these early plants suffered from lack of sterility, and contamination issues often caused the loss of products. The preference for petrochemical solvent production continued until the mid 1970s, when the oil crisis generated a renewed interest in ABE fermentation technology (Jones and Woods, 1986). Much of this renewed interest has centered around making the fermentation process more energy-efficient and cost-efficient.

The main idea behind these new approaches to ABE product recovery is to run the fermentation in a semi-continuous manner, feeding in concentrated sugar solution and constantly removing products as the fermentation progresses. This is because the *Clostridia* used in the fermentation exhibit both substrate inhibition and product inhibition. Because of substrate inhibition, a typical batch fermentation process uses about 60 g/L of sugar; continuously feeding sugar into the reactor allows the bacteria to ferment up to 500 g/L of sugar, at least in a laboratory-scale reactor (Ezeji, Qureshi, and Blaschek, 2004). Because the bacteria are also inhibited by the products of the fermentation, using a sugar solution of this concentration also necessitates continuous product removal. On a laboratory scale, this has been successfully accomplished by gas stripping, liquid-liquid extraction, and pervaporation (Qureshi, Maddox, and Friedl, 1992). Indeed, using a continuously fed concentrated sugar solution and *in situ* product recovery techniques can, in theory, make the ABE fermentation process economically competitive with the petrochemical synthesis of these solvents (Qureshi and Blaschek, 2001).

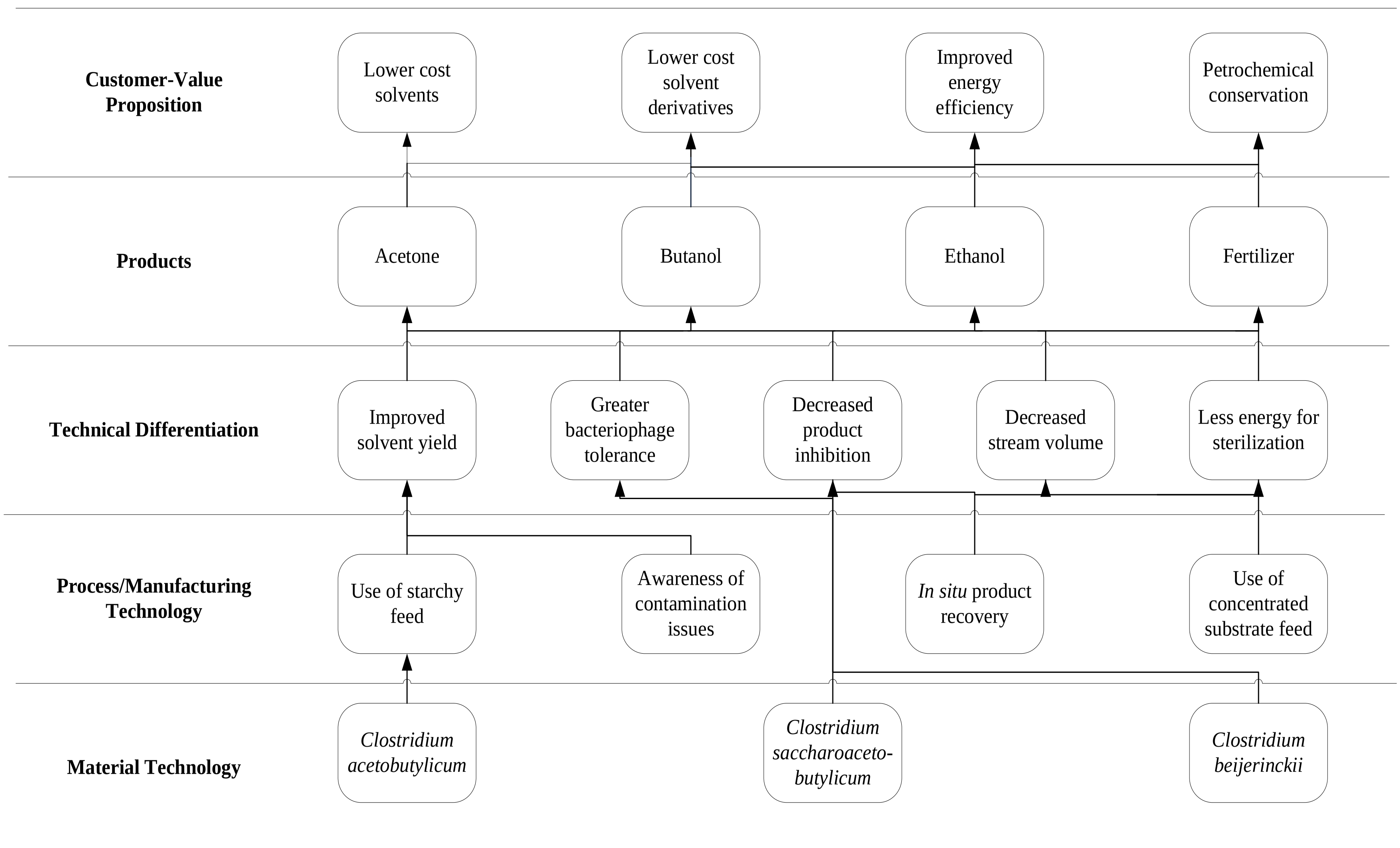
Unfortunately, the continuous product removal techniques are not quite well-enough understood to be employed on an industrial scale without an experimental component. Ezeji, Qureshi, and Blaschek, who have had a tremendous part in development of these product recovery technologies (see References), admit that "several recent advances have been made including the development of microbial cultures, process technologies, and use of waste substrates; however, these advances will need to be further developed to run a fermentation-based biobutanol industry" (2007).

When incorporated into a batch ABE fermentation process, however, both gas stripping and liquid-liquid extraction are well-characterized unit operations that can be modeled by a process simulator, such as ASPEN PLUS. Using either as the first operation in an ABE separation train can greatly decrease the concentration of water in the product solution. For example, gas stripping can selectively remove the solvents from the water solution, increasing the ABE concentration in the product solution from about 2% to about 75% (Wu et al., 2007). This results in a tremendous decrease in the energy requirement for distillation (which causes a decrease in cost); for comparison, Ezeji et al. (2004) cite a study by Phillips and Humphrey, who claim increasing the butanol concentration from 1% to just 4% decreases the energy requirement of the distillation train by a factor of six.

A few other innovations also promise to make ABE fermentation processes more economical. For instance, many laboratory studies have replaced *C. acetobutylicum*, the bacterial species used in traditional ABE fermentation, with *C. beijerinckii*, which has less stringent nutritional requirements and a greater tolerance to sugars, butanol, and variations in pH (Ezeji, Qureshi, and Blaschek, 2004). Many think that *C. beijerinckii* has greater potential for economical industrial production, and our problem statement specifies that the company has developed a novel strain of *Clostridium* that has operating parameters similar to *C. beijerinckii*. Other work has focused on improving the productivity of immobilized cell reactors, which would eliminate the need for mixing and agitation in the fermenters (Ezeji, Qureshi, and Blaschek, 2004).

These technological developments promise to greatly improve the economics over the traditional ABE fermentation process. Because of these developing technologies, it seems likely that the fermentative production of acetone, butanol, and ethanol will grow in industrial relevance in the near future.

Acetone-Butanol-Ethanol Fermentation Innovation Map



Concept Stage

Market and Competitive Analyses

The ABE Fermentation of sugar cane produces three main products: acetone, butanol and ethanol. Each of these compounds has its own market and demand.

Acetone

Acetone is used as a feedstock in the manufacture of methyl methacrylate (MMA) and bisphenol –A (BPA) whose derivatives are in turn used for making windows, skylights, signs, lighting fixtures, automotive parts, medical devices, and appliances. Acetone is also widely used as a solvent in many industrial processes, particularly in the pharmaceutical sector. The world acetone market is estimated to be 5.9m tonnes in 2010 and is expected to grow at 3% annually. The average selling price of acetone is about $1050/tonne with minimal variations depending on the region of the world (Hulsey, 2006).

Although the world market in MMA (acetone derivative) is expected to grow at 3-5%/year, much of it will not be reflected in future acetone demand because of the emergence of new non-acetone processes for the manufacture of MMA. For example some new processes use ethylene, methanol and carbon dioxide as feedstock while others are isobutylene based. Overall though, the demand and supply for acetone is expected to remain stable with no major increase in production capacity and no major change in demand for acetone as feedstock and solvent by industry.

Butanol

Butanol is an industrial commodity, with an average selling price of $3.75 per gallon and a 370 million gallon per year market that is growing at a current rate of 4% per annum (DuPont, 2010). Butanol is currently derived from petroleum and is used as an industrial solvent and in the manufacture of paint, resin, polymers and coatings.

Butanol also has the potential to be the next major biofuel. Although ethanol still makes up the biggest percentage of total global biofuel production, the production of biobutanol is expected to steadily rise in the near future. The key driver for this trend is that compared to ethanol, butanol is less corrosive, has a higher caloric value, is more nonpolar, so it mixes well with gasoline, and is less volatile. This similarity to gasoline means that it can be blended with gasoline up to a higher percentage level. Moreover, existing infrastructure for gasoline transportation, storage, and distribution (pipes, stations, pumps) can also be used for butanol without much modification because compared to ethanol, butanol is less corrosive and has less separation when in contact with water.

Growing consumer acceptance and name recognition for biobutanol, incentives to agriculture and industry, falling production costs, increasing prices and taxes for fossil fuels, and the desire for cleaner-burning sources of energy should drive an increase in the use of biobutanol as a fuel. In 2006, BP and DuPont announced a joint venture to develop and commercialize biobutanol biofuels that demonstrated that a 16% biobutanol blend in gasoline can be an effective automotive fuel. Another company, Cobalt Biofuels, based in Mountainview, California, recently raised $25 million in equity to research and commercialize biobutanol as an alternative to ethanol and biodiesel.

The main obstacle facing the adoption of butanol as a biofuel is competition from the existing use of ethanol and biodiesel in this role. The production cost of butanol is still higher than that of its biofuel counterparts. On the other hand, compared with gasoline, butanol has a lower energy density requiring a higher flow resulting in a lower fuel economy.

Ethanol

Ethanol's main use is as a motor fuel or fuel additive, with its largest industry being based in Brazil. Its use has become more popular in recent times due to its properties as a clean-burning, high-octane fuel that is produced from renewable sources, such as sugar cane and corn. Seen as an environmentally friendly fuel, ethanol reduces harmful tailpipe emissions of carbon monoxide and particulate matter, making it a popular choice with the recent rise in global awareness of pollution and harmful effects of fossil fuels on the planet. Finally, the economics of ethanol have been becoming more and more attractive as the price of gasoline has been increasing over the years. These factors have contributed both to public opinion as well as government policies encouraging the use and production of ethanol. With the increasing demand for renewable energy sources, the efficiency of ethanol as a fuel, and the increase in the price of gasoline, technologies for the production of ethanol need to be constantly evolving to make the production of ethanol cheaper and more efficient.

Ethanol is commonly used as an additive for gasoline, with mixtures ranging from 10% ethanol and 90% gasoline to 85% ethanol and 15% gasoline. In Brazil, flex-fuel motors can use 100% ethanol or 100% gasoline and any mixture in between. With the recent surge in demand for "green" products and energy sources, more and more people are switching to ethanol as a fuel along with, or instead of, unleaded gasoline. This has caused a tremendous increase in the demand, and therefore production, of ethanol worldwide. The leading producers of ethanol are the United States and Brazil, accounting together for 89% of the world production in 2007, when the US produced 6498.6 million gallons and Brazil produced 5019.2 million gallons (Market Research Analyst 2008). Although it is the world's largest producer, the US is still unable to meet its national demand, making Brazil the world's largest exporter of ethanol with about 90% of the global export market. Production in Brazil has been rising, going from 365,000 barrels per day (bbl/d) in 2207 up to 454,000 bbl/d in 2008, of which 86,000 bbl/d were exported. With production currently exceeding demand in the country, Brazil will be looking to export even more ethanol in the coming years.

The forecast for the global ethanol market is very promising, showing a great expectation for growth. By 2012, it is predicted that production in Brazil will increase by about 20% to 5,990 million gallons, and production in the US will go up about 43% to 8,838 million gallons (US Energy Information Administration, 2009). The current price for ethanol from Brazil is about 0.53 US $/L. Production in Brazil is regarded as the most efficient in the world, costing an average of 0.22 US$/L, while in the US it costs about 0.30 US$/L. The main difference between the two is the fact that Brazil uses mostly sugar cane while the US uses corn, which is a less efficient raw material for the process.

The main competitors for ethanol as a fuel are biobutanol, biodiesel and biogas. Biobutanol is butanol from biomass, and it performs in a similar way as gasoline in car engines. Currently, there are no vehicles that can run on 100% biobutanol and only some that run on mixtures of up to 10% biobutanol and 90% gasoline. DuPont and BP are currently the main producers of biobutanol. Biodiesel is created by reacting animal lipids or vegetable oil-based fat with an alcohol. It is used as a substitute for or blended with petrodiesel, usually used to fuel large trucks or buses. Biodiesel production in 2005 reached 3.8 million tons, about 85% of which came from the European Union, where diesel engines are far more common than in the Western Hemisphere. Biogas is gas produced by the biological breakdown of organic matter in the presence of oxygen, consisting usually of methane and carbon dioxide. These gases can be combusted in the presence of oxygen, so they can be used as fuel. It can be used for things such as cooking, waste disposal, and even vehicles. In Brazil, for example, it is becoming common practice to replace fuel tanks with a biogas tank that goes in the trunk of the car, and can be pumped up with new air at the gas station.

Fermentation Gases

The fermentation process produces a large amount of gaseous byproducts, with over 60% of the sugar weight going into CO_2 or H_2, in a mixture that is roughly 60% CO_2 and 40% H_2 by mole. In this process, these gases are sold by the price suggested by Qureshi and Blaschek (2001). Possible uses of this mixture include methanol synthesis (Fujita, Ito, and Takezawa, 1993), or membrane separation of these gases into a stream of pure CO_2 and H_2. The CO_2 can be used as an inert gas in chemical manufacturing processes, in liquid or solid form as a refrigerant, in the industrial synthesis of urea, in oil wells to decrease oil viscosity, in carbonated beverages, or to increase the yield of plant products grown in greenhouses (Universal Industrial Gases, Inc., 2008). Because it is present in such low concentrations in air, it is often recovered and sold from bioprocesses, such as this one (Universal Industrial Gases, Inc., 2008). Industrial applications of H_2 include use as an alternative fuel, to create a reducing environment, to remove oxygen from gaseous mixtures, or in the chemical synthesis of ammonia, methanol, hydrogen peroxide, polymers, and pharmaceuticals (Universal Industrial Gases, Inc., 2007).

Customer Requirements

Since this is a project tied with sustainability issues, it is imperative that the process be as energy efficient as possible. Water is also a big concern at the plant site, which will be a zero-discharge plant, so all process water is recycled within the plant. Water used with the sugar cane should contain less than 10 ppm organics as it will come in contact with sugar that will go into the market.

The butanol product must be at 99.5% purity, with less than 0.5% water and less than 0.01% ketones and aldehydes.

The acetone product must be at 99.5% purity, with less than 0.5% water and less than 0.001% evaporation residue.

The industry standard for ethanol is 99.5% purity by mass. At any purity lower than 97%, it is not miscible in gasoline and is therefore useless as a fuel additive. This presents a challenge, since there is an azeotrope at 94% purity that has to be dealt with in the process. The ethanol will be sold at US$2.50/gallon.

The fertilizer product is a solid product made with the older generations of organism that can no longer be used for fermentation. It is stored in a building before being shipped by truck or rail with 9% water and less than 10 ppm ABE.

Preliminary Process Synthesis

Traditionally, like many other industrial bioprocesses, ABE fermentation has been run in batch mode (Beesch, 1952). However, because *Clostridia* are inhibited by the solvent products of the fermentation, much recent research has focused on improving the practicality of continuous *in situ* product removal. Techniques that are being developed for this process include *in situ* gas stripping, liquid-liquid extraction, and pervaporation (Ezeji, Qureshi, and Blaschek, 2004). These continuous product removal options looked promising, but most experiments with them have been limited to a laboratory scale, and Ezeji, Qureshi, and Blaschek (2005) suggest that necessary substrate and product conditions to run these technologies on an industrial scale would be different than on a laboratory scale. Thus, we concluded that incorporating these technologies into our project without an experimental component would introduce too many uncertainties into the design and decided to design the process in traditional batch mode.

Although these technologies were not used in the design to continuously remove products from the fermenters, stripping and extraction are well-characterized as stand-alone separation units incorporated downstream of the fermentation section. In fact, complete product recovery by distillation, which was often used in traditional ABE fermentation (Beesch, 1952), is not economically competitive with butanol synthesis from petrochemicals, according to Billig (1998). To verify this claim, an ASPEN PLUS simulation was run that separated the chemicals entirely by distillation. This simulation used close to 700 GJ/hour of utilities, which was an unacceptably large amount given the sustainable nature of the project. Although this number could likely have been reduced by optimizing the distillation columns, we were able to reduce this number much more significantly by using other process units in place of the distillation column that removed water.

For the first step in the separation process, gas stripping was selected instead of liquid-liquid extraction for several reasons. Liquid-liquid extraction often uses oleyl alcohol to selectively remove ABE products (Shi, Zhang, Chen, and Mao, 2005); however, this compound was not in ASPEN PLUS, which was used in this project for all liquid-liquid equilibria calculations. Most other compounds considered for liquid-liquid extraction were either too hydrophobic to selectively extract acetone and ethanol from water or too hydrophilic to provide an efficient phase split with water. In addition, there is interest in using butanol as a food-grade extractant (Ezeji, Qureshi, and Blaschek, 2003), so adding an organic, and likely toxic, extractant to the butanol synthesis process would be undesirable. In contrast, gas stripping selectively removes the alcohols and solvents from water without introducing any new chemicals to the mixture, without requiring the addition of another distillation column to separate the extractant, as liquid-liquid extraction would. Thus, stripping was selected as the first unit operation in the separation section.

Initial work on the rest of the separation section focused on design heuristics presented in Seider, Seader, Lewin, and Widagdo (2009), as well as a description of the ABE process provided by Wu et al. (2007). The authors based their design on a sensitivity analysis that suggested the cheapest method of ABE separation was an initial stripping or extraction step to remove the bulk of the water, then a distillation column which separates butanol out in the bottoms, a distillation column that separates acetone out in the overhead stream, and then a final distillation column that separates ethanol and water to their azeotropic composition. We ended up using a very similar separation train, but industrial consultants suggested use of a decanter and molecular sieves in place of a distillation column to remove water. This ultimately decreased the energy use of the process, since one of the condensers and reboilers were eliminated, making it a more sustainable process.

Finally, extractive distillation and molecular sieves were considered for the removal of water beyond the azeotropic composition. Because all of the water needed to be purified and recycled and because of the lower energy demands of the molecular sieves compared to the extractive distillation column, molecular sieves were selected.

Assembly of Database

The thermodynamic and transport data were supplied by ASPEN PLUS.

The kinetics of fermentation and the product toxicity to the *Clostridium* culture are modeled by the following equation, which was supplied in the problem statement.

$$\frac{ds}{dt} = \left(\frac{ds}{dt}\right)_{max} \left(\frac{S}{5+S}\right)\left(\frac{7}{7+P}\right)$$

where $(ds/dt)_{max}$ = 1.1 g sugar / g DCW / hr in growth phase and 0.8 g sugar / g DCW / hr in production phase, S is the concentration of sugar in the reactor (g/L), and P is the concentration of ABE in the reactor (g/L).

The prices were also given in the problem statement as follows:

Cane juice = $0.05/lb sugar, as a 25% solution

Acetone = $3.00/gallon

Butanol = $4.00/gallon

Ethanol (denatured) = $2.50/gallon

Water = $3.00/thousand gallons (treated to be of adequate quality for the plant)

Unleaded gasoline = $2.00/gallon (wholesale)

Sewer = $5.00/thousand gallons (limit 10 ppm organics – boiler and cooling tower blowdown only)

Caustic (as a 50% solution, cost on a 100% basis) = $0.125/lb

Lime = $0.05/lb

Sulfuric acid (100% basis) = $0.06/lb

In addition, from the problem statement, for every 1000 pounds of sugar consumed, two pounds of sulfuric acid and one pound of lime must be added to the reactor to control the pH. Furthermore, when it is desired to maintain the organism in growth phase, 0.01 volumes of air/volume liquid/minute is required.

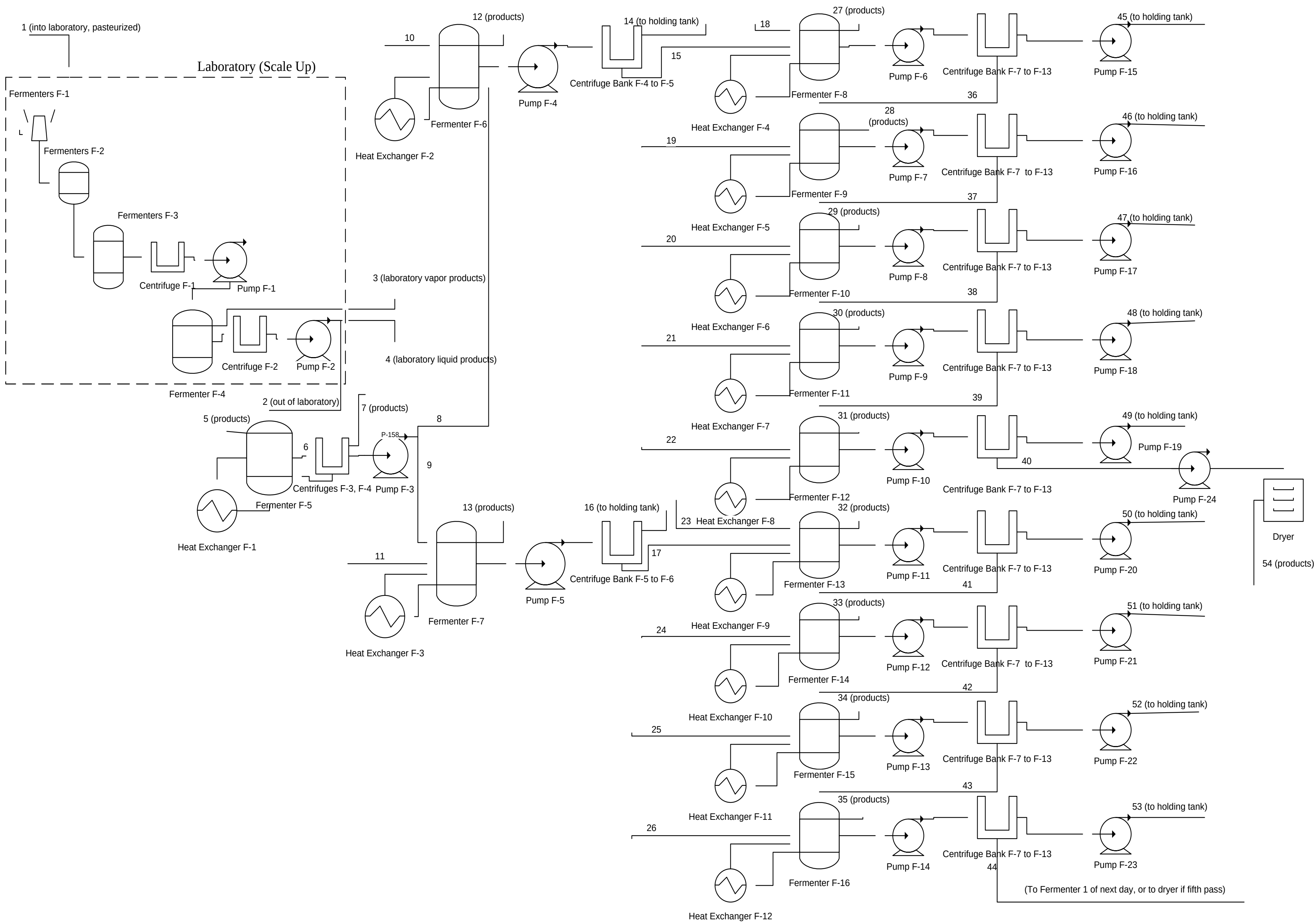
The price of ammonia refrigerant is given by ICIS Chemical Business (2006) as $385 per tonne. The price of fertilizer is about $47.95 per 50 lb bag (Planet Neutral). According to Qureshi and

Blaschek (2001), a reasonable net price for the sale of CO_2 and H_2, in the ratios produced by *Clostridia*, is $0.10/kg.

Bench-Scale Laboratory Work

This design project had no experimental component. However, as discussed in the Technology-Readiness Assessment section and the Preliminary Process Synthesis section, a laboratory component would be useful to further characterize continuous product removal and continuous concentrated sugar feed techniques to improve the economics of this process design.

Process Flow Diagrams and Material Balances



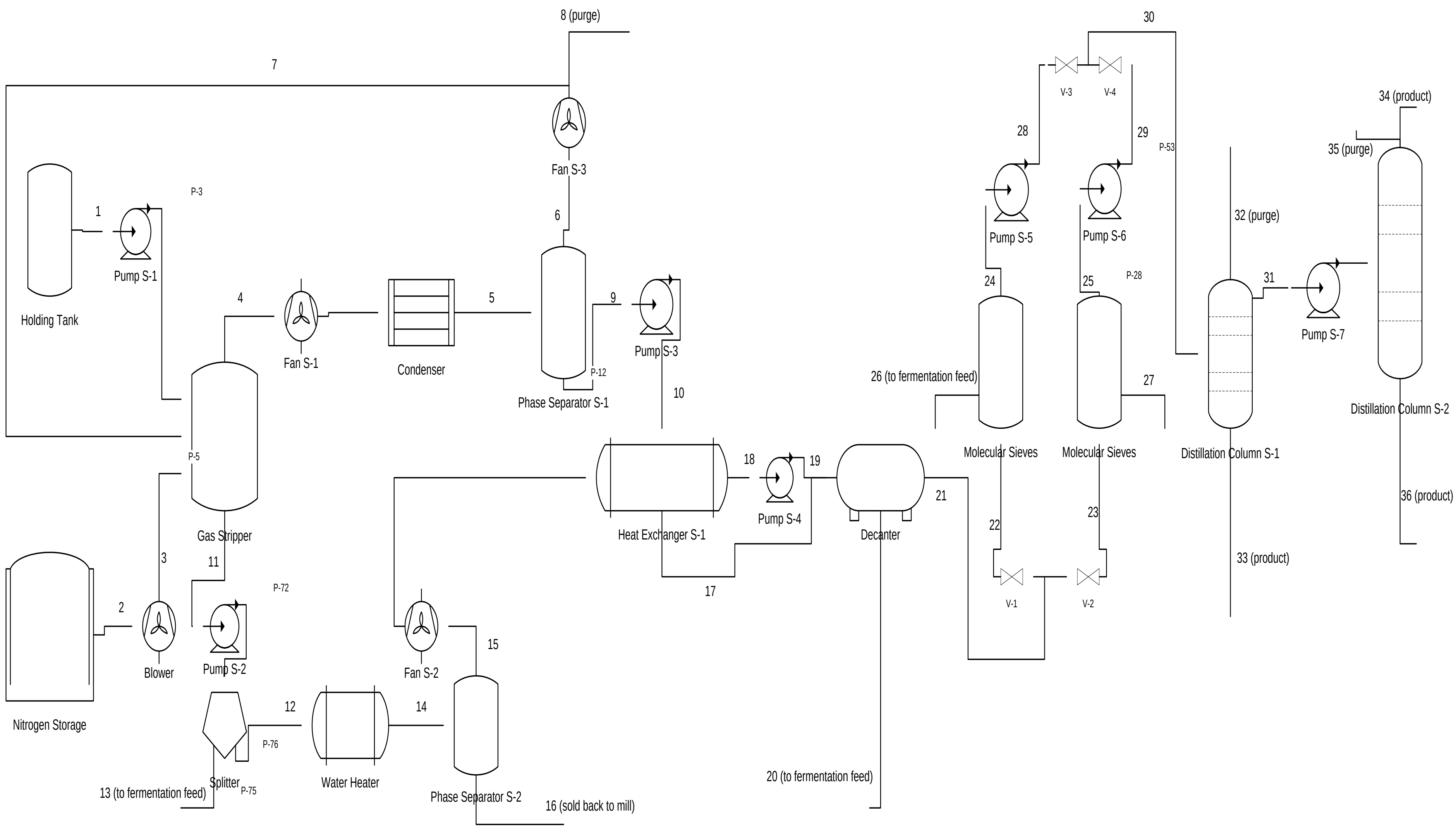
1 (into laboratory, pasteurized)
Laboratory (Scale Up)
Fermenters F-1
Fermenters F-2
Fermenters F-3
Centrifuge F-1
Pump F-1
Fermenter F-4
Centrifuge F-2
Pump F-2
3 (laboratory vapor products)
4 (laboratory liquid products)
2 (out of laboratory)
5 (products)
6
7 (products)
Centrifuges F-3, F-4
Pump F-3
P-158
Fermenter F-5
Heat Exchanger F-1
8
9
10
12 (products)
Fermenter F-6
Heat Exchanger F-2
Pump F-4
14 (to holding tank)
15
Centrifuge Bank F-4 to F-5
11
13 (products)
Fermenter F-7
Heat Exchanger F-3
Pump F-5
16 (to holding tank)
17
Centrifuge Bank F-5 to F-6
18
27 (products)
Fermenter F-8
Heat Exchanger F-4
Pump F-6
Centrifuge Bank F-7 to F-13
36
45 (to holding tank)
Pump F-15
19
28
(products)
Fermenter F-9
Heat Exchanger F-5
Pump F-7
Centrifuge Bank F-7 to F-13
37
46 (to holding tank)
Pump F-16
20
29 (products)
Fermenter F-10
Heat Exchanger F-6
Pump F-8
Centrifuge Bank F-7 to F-13
38
47 (to holding tank)
Pump F-17
21
30 (products)
Fermenter F-11
Heat Exchanger F-7
Pump F-9
Centrifuge Bank F-7 to F-13
39
48 (to holding tank)
Pump F-18
22
31 (products)
Fermenter F-12
Pump F-10
Centrifuge Bank F-7 to F-13
40
49 (to holding tank)
Pump F-19
Pump F-24
Dryer
54 (products)
23 Heat Exchanger F-8
32 (products)
Fermenter F-13
Heat Exchanger F-9
Pump F-11
Centrifuge Bank F-7 to F-13
41
50 (to holding tank)
Pump F-20
24
33 (products)
Fermenter F-14
Heat Exchanger F-10
Pump F-12
Centrifuge Bank F-7 to F-13
42
51 (to holding tank)
Pump F-21
25
34 (products)
Fermenter F-15
Heat Exchanger F-11
Pump F-13
Centrifuge Bank F-7 to F-13
43
52 (to holding tank)
Pump F-22
26
35 (products)
Fermenter F-16
Heat Exchanger F-12
Pump F-14
Centrifuge Bank F-7 to F-13
44
53 (to holding tank)
Pump F-23
(To Fermenter 1 of next day, or to dryer if fifth pass)

Stream Number	1	2	3	4	5	6	7	8*	9*	10	11	12	13	14
Temperature (F)	93	93	93	93	93	93	93	93	93	93	93	93	93	93
Pressure (psi)	19.7	19.7	19.7	19.7	19.7	19.7	19.7	19.7	19.7	19.7	19.7	19.7	19.7	19.7
Vapor Fraction	0.000	0.000	1.000	0.000	1.000	0.000	0.000	0.000	0.000	0.000	0.000	1.000	1.000	0.000
Mole Flow(lbmol/hr)	2420	2221	3	89	67	2216	2216	0	0	7535	7535	225	225	7515
Mass Flow(lb/hr)	46402	42479	68	1883	1733	40977	40709	134	134	144510	144510	5876	5876	138020
Comp. Flow(lb/hr)														
Acetone	15	14	0	11	0	288	288	0	0	47	47	0	0	978
Butanol	30	28	0	23	0	576	576	0	0	94	94	0	0	1954
Ethanol	5	4	0	4	0	96	96	0	0	15	15	0	0	327
Water	43125	39603	0	1565	0	39603	39603	0	0	134305	134305	0	0	134305
Nitrogen	6	6	0	0	6	0	0	0	0	19	19	19	19	0
Carbon Dioxide	1	1	66	0	1670	1	1	0	0	2	2	5666	5666	4
Hydrogen	0	0	2	0	57	0	0	0	0	0	0	191	191	0
Biomass	0	0	0	35	0	268	0	134	134	0	0	0	0	0
Sugar	3220	2823	0	245	0	145	145	0	0	10028	10028	0	0	452

Stream Number	15*	16	17*	18	19	20	21	22	23	24	25	26	27	28
Temperature (F)	93	93	93	93	93	93	93	93	93	93	93	93	93	93
Pressure (psi)	19.7	19.7	19.7	19.7	19.7	19.7	19.7	19.7	19.7	19.7	19.7	19.7	19.7	19.7
Vapor Fraction	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	1.000	1.000
Mole Flow (lbmol/hr)	0	7515	0	5686	5686	5686	5686	5686	5686	5686	5686	5686	173	173
Mass Flow (lb/hr)	967	138020	967	109044	109044	109044	109044	109044	109044	109044	109044	109044	4563	4563
Comp. Flow (lb/hr)														
Acetone	0	978	0	36	36	36	36	36	36	36	36	36	0	0
Butanol	0	1954	0	71	71	71	71	71	71	71	71	71	0	0
Ethanol	0	327	0	11	11	11	11	11	11	11	11	11	0	0
Water	0	134305	0	101344	101344	101344	101344	101344	101344	101344	101344	101344	0	0
Nitrogen	0	0	0	14	14	14	14	14	14	14	14	14	14	14
Carbon Dioxide	0	4	0	1	1	1	1	1	1	1	1	1	4405	4405
Hydrogen	0	0	0	0	0	0	0	0	0	0	0	0	144	144
Biomass	967	0	967	0	0	0	0	0	0	0	0	0	0	0
Sugar	0	452	0	7567	7567	7567	7567	7567	7567	7567	7567	7567	0	0

Stream Number	29	30	31	32	33	34	35	36*	37*	38*	39*	40*	41
Temperature (F)	93	93	93	93	93	93	93	93	93	93	93	93	93
Pressure (psi)	19.7	19.7	19.7	19.7	19.7	19.7	19.7	19.7	19.7	19.7	19.7	19.7	19.7
Vapor Fraction	1.000	1.000	1.000	1.000	1.000	1.000	1.000	0.000	0.000	0.000	0.000	0.000	0.000
Mole Flow(lbmol/hr)	173	173	173	173	173	173	173	0	0	0	0	0	0
Mass Flow(lb/hr)	4563	4563	4563	4563	4563	4563	4563	967	967	967	967	1533	967
Comp. Flow(lb/hr)													
Acetone	0	0	0	0	0	0	0	0	0	0	0	0	0
Butanol	0	0	0	0	0	0	0	0	0	0	0	0	0
Ethanol	0	0	0	0	0	0	0	0	0	0	0	0	0
Water	0	0	0	0	0	0	0	0	0	0	0	0	0
Nitrogen	14	14	14	14	14	14	14	0	0	0	0	0	0
Carbon Dioxide	4405	4405	4405	4405	4405	4405	4405	0	0	0	0	0	0
Hydrogen	144	144	144	144	144	144	144	0	0	0	0	0	0
Biomass	0	0	0	0	0	0	0	967	967	967	967	1533	967
Sugar	0	0	0	0	0	0	0	0	0	0	0	0	0

Stream Number	42*	43*	44*	45	46	47	48	49	50	51	52	53	54
Temperature (F)	93	93	93	93	93	93	93	93	93	93	93	93	93
Pressure (psi)	19.7	19.7	19.7	19.7	19.7	19.7	19.7	19.7	19.7	19.7	19.7	19.7	19.7
Vapor Fraction	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Mole Flow (lbmol/hr)	0	0	0	5676	5676	5676	5676	5676	5676	5676	5676	5676	5
Mass Flow (lb/hr)	967	967	967	104503	104503	104503	104503	104503	104503	104503	104503	104503	1063
Comp. Flow (lb/hr)													
Acetone	0	0	0	847	847	847	847	847	847	847	847	847	0
Butanol	0	0	0	1686	1686	1686	1686	1686	1686	1686	1686	1686	0
Ethanol	0	0	0	281	281	281	281	281	281	281	281	281	0
Water	0	0	0	101344	101344	101344	101344	101344	101344	101344	101344	101344	96
Nitrogen	0	0	0	0	0	0	0	0	0	0	0	0	0
Carbon Dioxide	0	0	0	4	4	4	4	4	4	4	4	4	0
Hydrogen	0	0	0	0	0	0	0	0	0	0	0	0	0
Biomass	967	967	967	0	0	0	0	0	0	0	0	0	967
Sugar	0	0	0	341	341	341	341	341	341	341	341	341	0


8 (purge)
30
7
V-3
V-4
34 (product)
28
29
P-53
35 (purge)
Fan S-3
P-3
Holding Tank
1
Pump S-1
6
32 (purge)
Pump S-5
Pump S-6
4
5
9
24
25
P-28
31
Fan S-1
Condenser
Pump S-3
Pump S-7
P-12
Phase Separator S-1
26 (to fermentation feed)
27
10
Distillation Column S-2
P-5
18
19
Molecular Sieves
Molecular Sieves
Distillation Column S-1
Heat Exchanger S-1
Pump S-4
21
36 (product)
22
23
Gas Stripper
Decanter
3
11
33 (product)
P-72
17
2
V-1
V-2
15
Blower
Pump S-2
Fan S-2
Nitrogen Storage
12
14
P-76
20 (to fermentation feed)
13 (to fermentation feed)
Splitter
P-75
Water Heater
Phase Separator S-2
16 (sold back to mill)

Stream Number	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18
Temperature (F)	93	77	142	92	41	41	31	41	41	41	82	82	82	217	217	217	102	213
Pressure (psi)	19.7	14.7	19.7	16.7	21.7	21.8	19.7	21.8	21.8	145	19.2	19.2	19.2	14.5	14.5	14.5	145	14.5
Vapor Fraction	0.007	1.000	1.000	1.000	0.958	1.000	1.000	1.000	0.000	0.000	0.000	0.000	0.000	0.019	1.000	0.000	0.011	0.010
Mole Flow(lbmol/hr)	68155	150	150	16622	16622	15635	15490	145	987	987	67173	14639	52534	14639	280	14360	987	280
Mass Flow(lb/hr)	1255320	4189	4189	485980	485980	444484	440373	4110	41496	41496	1213900	264555	949346	264555	5043	259512	41496	5043
Comp. Flow(lb/hr)																		
Acetone	9892	0	0	20705	20705	10914	10813	101	9791	9791	1	0	0	0	0	0	9791	0
Butanol	19718	0	0	20256	20256	543	538	5	19713	19713	1	0	0	0	0	0	19713	0
Ethanol	3289	0	0	4050	4050	842	834	8	3208	3208	73	16	57	16	13	3	3208	13
Water	1218260	0	0	9990	9990	1453	1439	13	8538	8538	1209710	263640	946065	263640	5029	258610	8538	5029
Nitrogen	0	4189	4189	428220	428220	427989	424032	3958	231	231	0	0	0	0	0	0	231	0
Carbon Dioxide	42	0	0	2758	2758	2742	2716	25	16	16	0	0	0	0	0	0	16	0
Hydrogen	0	0	0	1	1	1	1	0	0	0	0	0	0	0	0	0	0	0
Sugar	4120	0	0	0	0	0	0	0	0	0	4120	898	3222	898	0	898	0	0

Stream Number	19	20	21	22	23	24	25	26	27	28	29	30	31	32	33	34	35	36
Temperature (F)	213	168	168	168	168	169	169	169	169	169	169	169	90	90	275	90	90	202
Pressure (psi)	145	145	145	145	145	116	116	116	116	116	116	116	22	22	26	22	22	28
Vapor Fraction	0.010	0.000	0.000	0.000	0.000	0.000	0.000	0.120	0.120	0.000	0.000	0.000	0.000	1.000	0.000	0.000	1.000	0.000
Mole Flow (lbmol/hr)	280	665	600	300	300	251	251	50	50	251	251	501	235	2	264	167	0.03	68
Mass Flow (lb/hr)	5043	12011	34528	17264	17264	16219	16219	1046	1046	16219	16219	32437	12833	82	19522	9704	1.1	3128
Comp. Flow (lb/hr)																		
Acetone	0	1	9790	4895	4895	4846	4846	49	49	4846	4846	9692	9663	26	3	9655	0.5	8
Butanol	0	1	19712	9856	9856	9758	9758	99	99	9758	9758	19515	7	0	19507	0	0	7
Ethanol	13	20	3201	1601	1601	1585	1585	16	16	1585	1585	3169	3154	3	12	41	0	3113
Water	5029	11988	1579	790	790	2	2	788	788	2	2	3	3	0	0	3	0	0
Nitrogen	0	0	231	116	116	29	29	87	87	29	29	58	5	53	0	5	0.7	0
Carbon Dioxide	0	0	16	8	8	0	0	8	8	0	0	0	0	0	0	0	0	0
Hydrogen	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Sugar	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

Process Description

Process Description

The process is divided into two major sections: the fermentation section, where the sugar cane is converted to acetone, butanol, and ethanol by a novel strain of *Clostridium*¸ and the separation section, where acetone, butanol, and ethanol are separated to streams of the purity needed for sale.

The fermentation section consists of nine 500,000 gallon fermenters that produce most of the acetone, butanol, and ethanol. Cells for the fermentation section are supplied by two 500,000 tanks, in which the organism is exposed to slightly aerobic conditions, keeping it in growth phase. Each fermenter has a heat exchanger that removes the heat generated in the reaction at transfers it to cooling water. Furthermore, there is a train of centrifuges that recover the cells and recycle them to the other fermenters. The cells are used in five fermentation batches and then they are dried to 9% moisture and sold as fertilizer, keeping the time each cell is used under the 500 hours successfully used by Ezeji, Qureshi, and Blaschek (2005).

To avoid contamination, all of the incoming feed is sterilized in a pasteurizer consisting of two heat exchangers. First, the fresh feed is contacted with the recently sterilized feed, to cool the outgoing feed and pre-heat the incoming feed. The fresh feed is then heated to 212°F with low-pressure steam and held for five minutes before being considered sterile, when it is then contacted with fresh, cool incoming feed. In addition to sterilizing the feed, the process also has regular built-in cleaning time for the fermentation section, when the fermentation vessels are cleaned with caustic. An estimation for the amount of caustic required for this process is obtained by scaling up an estimate for the ethanol process by McAloon et al. (2000). With these procedures in place, according to the problem statement, the plant still suffers from, on average, one contaminated batch in one fermenter every 11 days. For this reason, a scale-up produces enough cells to seed the growth phase tanks every 11 days. This scale-up process consists of a test tube culture, a 600 mL culture vessel, a 3000 mL culture vessel, a 5000 gallon culture tank, and a 60,000 gallon tank, in accordance with that used in old industrial ABE fermentation processes (Beesch, 1952).

The liquid fermentation products are pumped from the fermenters into a holding tank so that the fermentation section can be run in batch mode but the separation can be run continuously. The products are first fed into a gas stripper, in which N_2 is fed into the bottom and strips the organics from the water. The bottoms water stream is then divided into two streams, one that is recycled back to the fermenters and the other that is sold back to the mill to keep the process zero-discharge. Since any water sold back to the mill must have less than 10 ppm organics, the water is first sent through a steam heat exchanger and a phase separator to concentrate the water stream. This phase separator produces an ethanol impurity stream that is fed into a decanter with the stripper condensate, consisting mostly of acetone, butanol, water, and ethanol. The decanter produces two liquid streams, one butanol-rich stream containing most of the acetone and ethanol and a water-rich stream containing small concentrations of organics. This water-rich stream is recycled back to the fermentation section to avoid the loss of product. The organic stream containing most of the products are sent to molecular sieves, which dehydrate the stream, eliminating the butanol-water and ethanol-water azeotrope and greatly simplifying the distillation sequence. The water stream is also recycled back to the fermenters and the organic stream is fed into two distillation columns, one that takes butanol off in the bottoms, at 99.9% purity, and one that separates acetone at 99.5% purity and ethanol at 99.5% purity. These are the necessary purity specifications for sale. The ethanol is denatured with 5% gasoline to prevent human consumption.

Energy Balance and Utility Requirements

Process Unit	Required Power
	Hp
Pump S-1	127.67
Pump S-2	2042
Pump S-3	12.6
Pump S-4	20
Pump S-5, S-6	1.55
Pump S-7	12.1
Blower	69
Fan S-1	3091
Fan S-2	1431
Fan S-3	165
TOTAL	6971.92

Process Unit	Required Power
	Hp
Pump F-1	Negligible
Pump F-2	Negligible
Pump F-3	0.36
Pumps F-4, F-5	3.02
Pumps F-6 - F-14	13.59
Pumps F-15 - F-23	13.59
Pump F-24	1.08
Centrifuge F-1	2
Centrifuge F-2	2
Centrifuges F-3, F-4	50
Centrifuges F-5, F-6	300
Centrifuges F-7 - F-12	900
Centrifuge F-13	150
TOTAL	1435.64

Process Unit	Heat Duty	50 psig Steam Req.	150 psig Steam Req.
	Btu/hr	lb/hr	lb/hr
Distillation Column 1	8982796.5		10386
Distillation Column 2	15115526	16472.4	
Water Heater	46,435,400	50916	
Pasteurizer	22,323,980	439966	
TOTAL		507354.4	10386

Process Unit	Required Cooling Duty	Cooling Water Requirement
	Btu/hr	lb/hr
Distillation Column 1	8158169.5	266621
Distillation Column 2	15115526	59300
TOTAL	23273695.5	325921

Process Unit	Heat Duty	Ammonia Req.	Chilled Water Req.
	Btu/hr	gal/hr	ton-day/hr
Condenser	1,930,972	836	
Fermenter Heat Ex. Growth Phase 1	299677		6.8
Fermenter Heat Ex. Growth Phase 2	4,994,616		114.4
Fermenter Heat Ex. Production Phase	1,555,200		514.8

Total BTU/hr IN from Utilities	Product Energy Potential (BUT/hr)	Difference
2.14E+06	3.36E+08	3.34E+08

The energy balance shows that the energy potential from the fuel products, ethanol and butanol, far exceeds the energy input required to make the process run. This is a very encouraging result, but a somewhat deceiving one, since we are not responsible for, and are not including, the utility requirements for harvesting and milling the sugar cane. This inclusion would have a large impact on the total energy in, and would decrease the difference between the energies we are "creating" and "consuming".

Equipment List and Unit Descriptions

Unit Descriptions: Fermentation

Fermenter F-1 – this fermenter holds the cells and juice in conditions suitable for growth phase of the cells for a period of 24 hours. Over the 24 hours the cells will grow and make product such that they use up all the sugar possible and is complete in the lab. It has the volume of a test tube with negligible cost.

Fermenter F-2 – this fermenter holds the cells and juice in conditions suitable for growth phase of the cells for a period of 24 hours. Over the 24 hours the cells will grow and make product such that they use up all the sugar possible and is complete in the lab. It has a volume of 600 mL and priced at $136.

Fermenter F-3– this fermenter holds the cells and juice in conditions suitable for growth phase of the cells for a period of 24 hours. Over the 24 hours the cells will grow and make product such that they use up all the sugar possible and is complete in the lab. It has a volume of 3L and priced at $1676.

Centrifuge F-2 – separates the cells from the liquids that were in the fermenter F-3. It can process 100 GPD and priced at $1,015,000.

Pump F-1 – sends the cells from centrifuge F-1 to seed fermenter F-4. This pump causes no pressure change. It is a centrifugal pump made of stainless steel, handles a flow rate of 100 gallons per day.

Fermenter F-4– this fermenter holds the cells and juice in conditions suitable for growth phase of the cells for a period of 24 hours. Over the 24 hours the cells will grow and make product such that they use up all the sugar possible and is complete in the lab. It has a volume of 5000L and priced at $151109.

Centrifuge F-3 – separates the cells from the liquids that were in the fermenter F-4. It can process 5000 GPD and priced at $1,015,000.

Pump F-2 – sends the cells from centrifuge F-2 to seed fermenter F-5. This pump causes no pressure change. It is a centrifugal pump made of stainless steel, handles a flow rate of 5000 gal/day.

Fermenter F-5– this fermenter holds the cells and juice in conditions suitable for growth phase of the cells for a period of 24 hours. Over the 24 hours the cells will grow and make product such that they use up all the sugar possible. It has a volume of 60,000 gal and priced at $591,805

Heat Exchanger F-1– this unit takes hot liquid from fermenter F-5 and cools it before reinjecting it back into the same fermenter. This removes the heat created by the cells and keeps the reaction at 34 ^{o}C. A floating head heat exchanger made of stainless steel, this unit is priced at $194290

Centrifuges F-3, F-4 – separates the cells from the liquids that were in the fermenter F-5. It sends some of the cells to reseed fermenter F-5. It can process 120000 gal/day.

Pump F-3 – sends the cells from centrifuge F-3 to seed fermenter F-6 or F-7. This pump causes no net pressure change. It is a centrifugal pump made of stainless steel, handles a flow rate of 120,000 gal/day.

Fermenter F-6, F-7– these fermenters holds the cells and juice in conditions suitable for growth phase of the cells for a period of 24 hours. Over the 24 hours the cells will grow and make product such that they use up all the sugar possible and is complete in the lab. Each of these has a volume of 500,000 gal and priced at $1,767,075

Heat Exchanger F-2, F-3– this unit takes hot liquid from fermenter F-6, F-7 and cools it before reinjecting it back into the same fermenter. This removes the heat created by the cells and keeps the reaction at 34^{o}C. A floating head heat exchanger made of stainless steel, this unit is priced at $329,990

Pump F-4, F-5 – sends everything from fermenter F-6, F-7 to the centrifuge bank F-4 to F-5. This pump causes no net pressure change. It is a centrifugal pump made of stainless steel, handles a flow rate of 120,000 gal/day.

Centrifuge Bank F-5 to F-6 – separates the cells from the liquids and sends the cells to two of the fermenter in fermenter F-8 through F-16. It sends some of the cells to reseed fermenter F-6, F-7. Each can process the full volume of the fermenter.

Fermenter F-8 through F-16– these fermenters hold the cells and juice in conditions suitable for production phase of the cells for a period of 24 hours. Over the 24 hours the cells will make product such that they use up all the sugar possible and is complete in the lab. Each of these has a volume of 500,000 gal and priced at $1,767,075

Heat Exchanger F-4 through F-12– this unit takes hot liquid from fermenter F-8 through F-16 and cools it before reinjecting it back into the same fermenter. This removes the heat created by the cells and keeps the reaction at 34 ^{o}C. A floating head heat exchanger made of stainless steel, this unit is priced at $165,500

Pump F-6 through F-14– sends everything from fermenter F-8 through F-16 to the centrifuge bank F-6 to F-12. This pump causes no net pressure change. It is a centrifugal pump made of stainless steel, each handles a flow rate of 120,000 gal/day.

Centrifuge Bank F-7 to F-13 – separates the cells from the liquids and sends the cells to two of the fermenters in fermenter F-8 through F-16. One centrifuge of the bank will send cells that have been used five times to pump F-24.

Pump F-15 through F-23– sends the liquid phase from centrifuge bank F-6 to F-12 to the holding tank. This pump causes no pressure change. It is a centrifugal pump made of stainless steel, each handles a flow rate of 2783 ft^3/hr, priced at $45963.

Pump F-24– sends the five times used cells from Centrifuge Bank F-6 to F-12 to the dryer. This pump causes no pressure change. It is a centrifugal pump made of stainless steel, handles a flow rate of 53 ft^3/hr, priced at $3123.

Dryer– takes the cells provided by pump F-24 and dries out the cells so that they can be used as fertilizer and in the process kills the cells. It is made of stainless steel and is priced at $732,800.

Unit Descriptions: ABE Separation

Holding Tank – serves as a bridge between the fermenters and the separation process.

Pump S-1 – pumps the contents of the holding tank into the gas stripper. The average flow rate is of about 4200 m^3/hr of a solution comprised of the ABE products and 97% water by mass. The pump causes no pressure change, but compensates for the 5 psi pressure drop that is assumed to take place due to the piping. The pump is priced at $165,000, as quoted by supplier Beijing Great Sources Technology Development Co.

Nitrogen Storage – tank to hold the fresh Nitrogen feed to be used in the Gas Stripper. The fresh feed must flow at a rate of 1900 kg/hr, and it is contained at atmospheric pressure and ambient pressure.

Blower – sends the fresh nitrogen feed into the gas stripper. Compresses the nitrogen by 5 psi so the fresh feed pressure matches with the recycle stream pressure going into the stripper. Made of stainless steel, priced at $5,287.04.

Gas Stripper – The fermentation product is fed in the top while nitrogen enters the column from the bottom and strips the organics in vapor form to leave in the overhead stream, separating them from the water phase. The fermentation product comes in at 97% water, while the stripper overhead is at only 2.1% water by mass, allowing for a much easier and cheaper separation later on. Column has 8 sieve trays, a height of 23 ft, inside diameter of 18 ft, and is priced at $10,419,598.

Fan S-1 – sends the gas stripper overhead stream with nitrogen and organics into the condenser. Handles a total volume of 5,888,350 ft^3/hr and makes up the 5 psi pressure drop due to piping. This is a centrifugal backward curved fan made of fiberglass, priced at $73,354.

Condenser – a fixed head heat exchanger, refrigerates the stripper overhead stream to 41^{o}F so that the nitrogen phase can later be separated from the organics phase. Uses a refrigeration system with ammonia with a heat duty of 1,930,972 BTU/hr, handles a flow rate of 5,888,320 ft^3/hr, made of stainless steel and priced at $160,740.

Phase Separator S-1 – Separates the ABE products from the nitrogen gas. The inlet stream comes in at a flow rate of 3,930,000 ft^3/hr at 88% nitrogen by mass. The outlet gas stream takes almost all of the nitrogen, while the outlet liquid stream is at only 0.6% nitrogen, with nearly all the organics, at a flow rate of about 41,500 lb/hr. Separator is made of stainless steel, has a diameter of 2.85 ft, a height of 8.55 ft, and is priced at $74,267.

Fan S-2 – sends the recycled nitrogen stream from the phase separator to the gas stripper. Handles a flow rate of 3,823,120 ft^3/hr, centrifugal backward curved fan made of fiberglass, priced at $73,354.

Pump S-2 – pumps the water stream from the bottom of the stripper. Handles a flow rate of 67,173 ft^3/hr with 99.7% water by mass. A centrifugal pump made of stainless steel, priced at $165,000 as quoted by supplier Beijing Great Sources Technology Development Co.

Splitter – splits up the water stream into one stream to be sterilized and sent back to the mill, and another to be used to dilute the can juice for fermentation.

Water Heater – heats up the water stream to be going back to the mills so it can go through a successful phase splitting to get rid of the undesired organics. Heats up 4,266 ft^3/hr of a 99.7% water stream from 82 ^{o}F to 217^{o}F using 50,916 lb/he of saturated steam. This is where the plant sees one of its highest utility costs at $821,184/yr. Made of stainless steel, the heater is priced at $257,743.

Phase Separator S-2 – separates the impurities from the water stream so it can be sent back to the mills. The impurities stream leaves at 138,671 ft^3/hr at 98% water and 2% ethanol in the vapor phase. Separator is made of stainless steel with a diameter of 11 ft, a height of 33 ft, and is priced at $253,519.

Fan S-3 – sends the vapor impurities stream into the heat exchanger where it will come in contact with the organics stream from the stripper overhead. Makes up for the 5 psi pressure loss due to piping, centrifugal backward curved fan made of stainless steel, priced at $19,504.

Pump S-3 – Sends ABE liquid stream from phase separator to heat exchanger. Increases the pressure of the stream by 120 psi in order to facilitate decantation downstream. Centrifugal pump made of stainless steel, handles a flow rate of 767 ft^3/hr, priced at $7,694.

Heat Exchanger S-1 – the ABE stream from the stripper overhead comes in contact with the impurities stream separated from the water bottoms. The two streams leave as liquids ready to be decanted downstream. The ABE stream is at high pressure from Pump S-3 while the impurities stream will have its pressure increased by Pump S-4. A floating head heat exchanger made of carbon steel, this unit is priced at $62,125.

Pump S-4 – pumps the liquid impurities into the decanter and increases its pressure by 130 psi so it can be decanted more easily. Takes in a flow rate of 1465 ft^3/min, this unit is a centrifugal pump made of carbon steel, priced at $12,445.

Decanter – Separates organic-rich phase from water-rich phase to facilitate ABE separation in the distillation columns further downstream. The outgoing water stream is 99.8% water by mass while the organic stream is only 4.6% water. This unit is made of stainless steel, has a diameter of 3.67 ft, a height of 11.1 ft, and is priced at $93,350.

Molecular Sieves – further separate water from the organics, so that the distillation columns can remove the Butanol, Ethanol, and Acetone products at the desired purities of 99.5%. The sieves allow for normal distillation to be done instead of azeotropic, as they remove enough water that we will be past any azeotropes that would prevent us from getting the required purities. The outgoing organics stream has a flow rate of 698 ft^3/hr and is only 521 PPM water. There are two identical units that use 13x sieves, have a diameter of 14.2 ft, a height of 42.6 ft, and are priced at $2,188,202 each.

Pumps S-5, S-6 – These identical pumps connect the outgoing organic-rich streams from the molecular sieves to the first distillation column. These streams are at 30% acetone, 60% butanol, and about 10% ethanol. These pumps increase the pressure by 15 psi so as to facilitate the distillation. These centrifugal pumps are made of carbon steel and are priced at $3,884.

Distillation Column S-1 – this column separates the final butanol product at 99.9% purity from the acetone and ethanol product. The butanol leaves out of the bottoms at the desired purity while the acetone and ethanol streams will continue to the final distillation column to be separated. With 22 Koch Flexitrays, this carbon steel unit has an inside diameter of 4 ft, a height of 42 ft, and is priced at $752,122.

Pump S-7 – pumps the acetone-ethanol stream into the second distillation column where they are to be separated into the final products. With a flow rate of 260 ft^3/hr, this pump has a slight pressure increase of 12 psi to facilitate distillation, it is made of carbon steel and is priced at $4,349.

Distillation Column S-2 – this column separates the acetone from the ethanol. The overhead stream contains acetone at the desired 99.5% purity, and the bottoms contain the ethanol also at 99.5% purity by mass. It is also worthy to note that the condenser is a partial condenser due to the enormous cost in utilities that would be required to fully condense the overhead stream. We are choosing to lose about half a pound per hour of acetone product to save far more on utility costs. This unit has 36 Koch Flexitrays, it has an inside diameter of 5.31 ft, a functional height of 64 ft, it is made of carbon steel and is priced at $1,215,870.

Unit Specification Sheets

Laboratory (Fermenters F-1 - F-4)		
	Number Needed:	1
Bare Module Cost		$ 4,954.90
Purpose:	To provide cells for Fermenters F-6, F-7	
Materials in the Column:		
Process rate (kg sugar/day):		3504
Mass Fraction of each stream:	Inlet	Outlet
Acetone	0.000	0.001
Butanol	0.001	0.001
Ethanol	0.000	0.000
Water	0.929	0.926
Biomass	0.000	0.001
Nitrogen	0.000	0.000
Sugar	0.069	0.069
Carbon Dioxide	0.000	0.000
Hydrogen	0.000	0.000
Temperature (F):	34	34
Design Specs:	Volume:	5000 gal (One tank)
		3 L (127 containers)
		600 mL (26 containers)
		Test tubes
	Material of Construction:	Stainless Steel
Utilities:	.75 hp for agitation	
Price of Utilities:	$151/yr	
Comments and Figures:		

Fermenter F-5		
	Number Needed:	1
Bare Module Cost		$ 591,805.89
Purpose:	To provide cells for Fermenters F-6, F-7	
Materials in the Column:		
Process rate (kg sugar/day):		29214
Mass Fraction of each stream:	Inlet	Outlet
Acetone	0.000	0.007
Butanol	0.001	0.013
Ethanol	0.000	0.002
Water	0.932	0.928
Biomass	0.000	0.000
Nitrogen	0.000	0.039
Sugar	0.066	0.001
Carbon Dioxide	0.000	0.006
Hydrogen	0.000	0.003
Temperature (F):	34	34
Design Specs:	Volume (gal):	60000
	Material of Construction:	Stainless Steel
Utilities:	9 hp for agitation	
Price of Utilities:	$1814/yr	
Comments and Figures:		

Fermenters F-6, F-7		
	Number Needed:	2
Bare Module Cost		$1,767,075.77
Purpose:	To provide cells for Fermenters F-8 to F-16	
Materials in the Column:		
Process rate (kg sugar/day):		218400
Mass Fraction of each stream:	Inlet	Outlet
Acetone	0.000	0.007
Butanol	0.001	0.013
Ethanol	0.000	0.002
Water	0.929	0.927
Biomass	0.001	0.007
Nitrogen	0.000	0.000
Sugar	0.069	0.003
Carbon Dioxide	0.000	0.039
Hydrogen	0.000	0.001
Temperature (F):	34	34
Design Specs:	Volume (gal):	500000
	Material of Construction:	Stainless Steel
Utilities:	75 hp for agitation	
Price of Utilities:	$15120/yr	
Comments and Figures:		

Fermenters F-8- F-16		
	Number Needed:	9
Bare Module Cost		$ 1,767,075.77
Purpose:	To convert sugar cane to ABE	
Materials in the Column:		
Process rate (kg sugar/day):		82392
Mass Fraction of each stream:	Inlet	Outlet
Acetone	0.000	0.008
Butanol	0.001	0.015
Ethanol	0.000	0.002
Water	0.921	0.921
Biomass	0.008	0.010
Nitrogen	0.000	0.000
Sugar	0.069	0.003
Carbon Dioxide	0.000	0.040
Hydrogen	0.000	0.001
Temperature (F):	34	34
Design Specs:	Volume (gal):	500000
	Material of Construction:	Stainless Steel
Utilities:	75 hp for agitation	
Price of Utilities:	$15120/yr	
Comments and Figures:		

Centrifuge F-1		
	Number needed:	1
Bare Module Cost		$60,900.00
Purpose:		
Design Specs:	Centrifuge capacity (gpm):	20
	Average cells removed (kg/hr):	< 5
	Bowl Diameter (in):	not given
	Maximum G Force:	1750
Utilities:	2 HP	
Price of Utilities:	$403/yr	
Comments and Figures:	Price Quote From US Centrifuge	

Centrifuge F-2		
	Number needed:	1
Bare Module Cost		$60,900.00
Purpose:		
Design Specs:	Centrifuge capacity (gpm):	20
	Average cells removed (kg/hr):	35
	Bowl Diameter (in):	not given
	Maximum G Force:	1750
Utilities:	2 HP	
Price of Utilities:	$403/yr	
Comments and Figures:	Price Quote From US Centrifuge	

Centrifuge F-3 - F-4		
	Number needed:	2
Bare Module Cost		$304,500.00
Purpose:		
Design Specs:	Centrifuge capacity (gpm):	50
	Average cells removed (kg/hr):	134
	Material:	Stainless Steel
	Bowl Diameter (in):	48
	Maximum G Force:	3375
Utilities:	25 HP	
Price of Utilities:	$5040/yr	
Comments and Figures:	Price quote from US Centrifuge	

Centrifuges F-5 - F-6		
	Number needed:	2
Bare Module Cost		$1,015,000.00
Purpose:		
Design Specs:	Centrifuge capacity (gpm):	350
	Average cells removed (kg/hr):	967
	Material:	Stainless Steel
	Bowl Diameter (in):	30
	Maximum G Force:	2500
Utilities:	150 HP	
Price of Utilities:	$30240/yr	
Comments and Figures:	Price Quote From US Centrifuge	

Centrifuges F-7 - F-12		
	Number needed:	6
Bare Module Cost		$1,015,000.00
Purpose:		
Design Specs:	Centrifuge capacity (gpm):	350
	Average cells removed (kg/hr):	967
	Material:	Stainless Steel
	Bowl Diameter (in):	30
	Maximum G Force:	2500
Utilities:	150 HP	
Price of Utilities:	$30240/yr	
Comments and Figures:	Price Quote From US Centrifuge	

Centrifuge F-13		
	Number needed:	1
Bare Module Cost		$1,015,000.00
Purpose:		
Design Specs:	Centrifuge capacity (gpm):	350
	Average cells removed (kg/hr):	1533
	Material:	Stainless Steel
	Bowl Diameter (in):	30
	Maximum G Force:	2500
Utilities:	150 HP	
Price of Utilities:	$30240/yr	
Comments and Figures:	Price Quote From US Centrifuge	

Pump F-1		
	Number needed:	1
Bare Module Cost		$20,072.68
Purpose:	Pump fluid to Fermenter F-4	
Materials in the Colum	Inlet	Outlet
Flow rate (ft^3/hr):	0.56	0.56
Mass Fraction of each stream:		
Acetone	0.000	0.000
Butanol	0.001	0.001
Ethanol	0.000	0.000
Water	0.929	0.929
Biomass	0.000	0.000
Nitrogen	0.000	0.000
Sugar	0.069	0.069
Carbon Dioxide	0.000	0.000
Hydrogen	0.000	0.000
Temperature:	93.2°F	93.2°F
Design Specs:	Pump Efficiency:	0.7
	Volumetric Flow Rate (ft3/hr):	0.56
	Head (ft):	12
	Type:	Centrifugal Pump
	Motor Material:	Stainless Steel
	Pump Material:	Stainless Steel
	Pressure Change:	5 psi
Utilities:	negligible	
Price of Utilities:		
Comments and Figures:		

Pump F-2		
	Number needed:	1
Bare Module Cost		$40,269.14
Purpose:	Pump fluid to Fermenter F-5	
Materials in the Colum	Inlet	Outlet
Flow rate (ft^3/hr):	27.8	27.8
Mass Fraction of each stream:		
Acetone	0.001	0.001
Butanol	0.001	0.001
Ethanol	0.000	0.000
Water	0.926	0.926
Biomass	0.001	0.001
Nitrogen	0.000	0.000
Sugar	0.069	0.069
Carbon Dioxide	0.000	0.000
Hydrogen	0.000	0.000
Temperature:	93.2°F	93.2°F
Design Specs:	Pump Efficiency:	0.7
	Volumetric Flow Rate (ft3/hr):	27.8
	Head (ft):	12
	Type:	Centrifugal Pump
	Motor Material:	Stainless Steel
	Pump Material:	Stainless Steel
	Pressure Change:	5 psi
Utilities:	negligible	
Price of Utilities:		
Comments and Figures:		

Pump F-3		
	Number needed:	1
Bare Module Cost		$41,988.44
Purpose:	Pump fluid to Fermenters F-6, F-7	
Materials in the Colum	Inlet	Outlet
Flow rate (ft^3/hr):	668	668
Mass Fraction of each stream:		
Acetone	0.007	0.007
Butanol	0.013	0.013
Ethanol	0.002	0.002
Water	0.928	0.928
Biomass	0.000	0.000
Nitrogen	0.039	0.039
Sugar	0.001	0.001
Carbon Dioxide	0.006	0.006
Hydrogen	0.003	0.003
Temperature:	93.2°F	93.2°F
Design Specs:	Pump Efficiency:	0.7
	Volumetric Flow Rate (ft3/hr):	668
	Head (ft):	12
	Type:	Centrifugal Pump
	Motor Material:	Stainless Steel
	Pump Material:	Stainless Steel
	Pressure Change:	5 psi
Utilities:	0.36 HP	
Price of Utilities:	$73/yr	
Comments and Figures:		

Pumps F-4, F-5		
	Number needed:	2
Bare Module Cost		$45,952.65
Purpose:	Pump fluid to Fermenters F-8 - F-16	
Materials in the Colum	Inlet	Outlet
Flow rate (ft^3/hr):	2783	2783
Mass Fraction of each stream:		
Acetone	0.007	0.007
Butanol	0.013	0.013
Ethanol	0.002	0.002
Water	0.927	0.927
Biomass	0.007	0.007
Nitrogen	0.000	0.000
Sugar	0.003	0.003
Carbon Dioxide	0.039	0.039
Hydrogen	0.001	0.001
Temperature:	93.2°F	93.2°F
Design Specs:	Pump Efficiency:	0.7
	Volumetric Flow Rate (ft3/hr):	2783
	Head (ft):	12
	Type:	Centrifugal Pump
	Motor Material:	Stainless Steel
	Pump Material:	Stainless Steel
	Pressure Change:	5 psi
Utilities:	1.51 HP	
Price of Utilities:	$304/yr	
Comments and Figures:		

Pumps F-6 - F-14		
	Number needed:	9
Bare Module Cost		$45,952.65
Purpose:	Pump fluid in fermentation section	
Materials in the Colum	Inlet	Outlet
Flow rate (ft^3/hr):	2783	2783
Mass Fraction of each stream:		
Acetone	0.008	0.008
Butanol	0.015	0.015
Ethanol	0.002	0.002
Water	0.921	0.921
Biomass	0.010	0.010
Nitrogen	0.000	0.000
Sugar	0.003	0.003
Carbon Dioxide	0.040	0.040
Hydrogen	0.001	0.001
Temperature:	93.2°F	93.2°F
Design Specs:	Pump Efficiency:	0.7
	Volumetric Flow Rate (ft3/hr):	2783
	Head (ft):	12
	Type:	Centrifugal Pump
	Motor Material:	Stainless Steel
	Pump Material:	Stainless Steel
	Pressure Change:	5 psi
Utilities:	1.51 HP	
Price of Utilities:	$304/yr	
Comments and Figures:		

Pumps F-15 - F-23		
	Number needed:	9
Bare Module Cost		$45,952.65
Purpose:	Pump fluid in fermentation section	
Materials in the Colum	Inlet	Outlet
Flow rate (ft^3/hr):	2783	2783
Mass Fraction of each stream:		
Acetone	0.008	0.008
Butanol	0.015	0.015
Ethanol	0.002	0.002
Water	0.930	0.930
Biomass	0.010	0.010
Nitrogen	0.000	0.000
Sugar	0.003	0.003
Carbon Dioxide	0.040	0.040
Hydrogen	0.001	0.001
Temperature:	93.2°F	93.2°F
Design Specs:	Pump Efficiency:	0.7
	Volumetric Flow Rate (ft3/hr):	2783
	Head (ft):	12
	Type:	Centrifugal Pump
	Motor Material:	Stainless Steel
	Pump Material:	Stainless Steel
	Pressure Change:	5 psi
Utilities:	1.51 HP	
Price of Utilities:	$304/yr	
Comments and Figures:		

Pump F-24		
	Number needed:	1
Bare Module Cost		$3,123.35
Purpose:	Pump solid suspension from centrifuge bank to dryer	
Materials in the Colum	Inlet	Outlet
Flow rate (ft^3/hr):	53	53
Mass Fraction of each stream:		
Acetone	0.000	0.000
Butanol	0.000	0.000
Ethanol	0.000	0.000
Water	0.700	0.700
Biomass	0.300	0.300
Nitrogen	0.000	0.000
Sugar	0.000	0.000
Carbon Dioxide	0.000	0.000
Hydrogen	0.000	0.000
Temperature:	93.2°F	93.2°F
Design Specs:	Pump Efficiency:	0.7
	Volumetric Flow Rate (ft3/hr):	53
	Head (ft):	12
	Type:	Centrifugal Pump
	Motor Material:	Stainless Steel
	Pump Material:	Stainless Steel
	Pressure Change:	5 psi
Utilities:	1.08 HP	
Price of Utilities:	$217/yr	
Comments and Figures:		

Distillation Column 1

Bare Module Cost					$752,122	
Purpose:	To distill Butanol in the bottoms					
Materials in the Column:	Feed	Bottoms	Overhead	N2 Purge		
Flow rate (ft^3/hr)	693	452	265	706		
Mass Fraction of each stream:						
Acetone	0.229	146 PPM	0.753	0.385		
Butanol	0.602	0.999	603 PPM	18 PPM		
Ethanol	0.098	629 PPM	0.246	0.047		
Water	243 PPM	9 PPB	612 PPM	471 PPM		
Nitrogen	0.002	0	381 PPM	0.567		
Sugar	0	0	0	0		
Carbon Dioxide	5 PPM	0	9 PPM	555 PPM		
Hydrogen	0	0	0	0		
Temperature:	169°F	275°F	100°F	100°F		
Design Specs:	Theoretical Trays:	15			Molar Reflux Ratio:	1.1
	Real Trays:	22			Tray Spacing (ft):	1.5
	Tray Efficiency:	0.7			Headspace (ft):	4
	Tray Type:	Koch Flexitray			Sump Space (ft):	3
	Functional Height (ft):	41.5				
	Inside Diameter (ft):	4				
	Pressure:	15.24				
	Feed Stage:	8				
	Material:	Carbon Steel				
	Number of Man Holes:	2				
	Condenser					
	Temperature (°F):	100				
	Reflux Ratio:	1.1				
	Overall Heat Transfer Coefficient (BTU/(hr °F ft^2)):	100				
	Area (ft2):	3112.3				
	Material:	Carbon Steel Shell, Brass Tubes				
	Reboiler					
	Temperature (°F):	275.2				
	Area (ft2):	741.7				
	Heat Flux (BTU/hr-ft2):	12000				
	Material:	Carbon Steel Shell and Tube, Kettle Vaporizer				
	Reflux Accumulator					
	Reflux Ratio:	1.1				
	Volume (ft3/hr):	557				
	Diameter (ft):	2.5				
	Length (ft):	7.4				
	RA Material:	Carbon Steel				
Utilities:	Cooling water: 266621.4 lb/hr		Steam @ 150 psig: 10386 lb/hr			
Price of Utilities:	Cooling water: $12906 / yr		Steam @ 150 psig: $268,009 / yr			
Controls:						
Comments and Figures:						

Distillation Column 2						
Bare Module Cost					$1,215,870	
Purpose:	To distill Acetone in the overhead and Ethanol in the bottoms					
Materials in the Column:	Feed	Bottoms	Overhead	Acetone Purge		
Flow rate (ft^3/hr)	265	70	201	20		
Mass Fraction of each stream:						
Acetone	0.753	0.002	0.995	0.501		
Butanol	603 PPM	0.002	0	0		
Ethanol	0.246	0.996	0.005	0.001		
Water	612 PPM	473 PPB	808 PPM	790 PPM		
Nitrogen	381 PPM	0	429 PPM	0.497		
Sugar	0	0	0	0		
Carbon Dioxide	9 PPM	0	11 PPM	674 PPM		
Hydrogen	0	0	0	0		
Temperature:	100.1°F	201.9°F	100°F	100°F		
Design Specs:	Theoretical Trays:	25			Molar Reflux Ratio:	5
	Real Trays:	36			Tray Spacing (ft):	1.5
	Tray Efficiency:	0.7			Headspace (ft):	4
	Tray Type:	Koch Flexitray			Sump Space (ft):	3
	Functional Height (ft):	64				
	Inside Diameter (ft):	5.31				
	Pressure:	18.04				
	Feed Stage:	13				
	Material:	Carbon Steel				
	Number of Man Holes:	3				
	Condenser					
	Temperature (°F):	100				
	Reflux Ratio:	5				
	Overall Heat Transfer Coefficient (BTU/(hr °F ft^2)):	100				
	Area (ft2):	6659.4				
	Material:	Carbon Steel Shell, Brass Tubes				
	Reboiler					
	Temperature (°F):	201.9				
	Area (ft2):	1253				
	Heat Flux (BTU/hr-ft2):	12000				
	Material:	Carbon Shell and Tube, Kettle Vaporizer				
	Reflux Accumulator					
	Reflux Ratio:	5				
	Volume (ft3/hr):	1207				
	Diameter (ft):	2.26				
	Length (ft):	6.76				
	RA Material:	Carbon Steel				
Utilities:	Cooling Water : 493962 lb/hr		Steam @ 50 psig: 16472.40			
Price of Utilities:	Cooling Water : $23909.40 / yr		Steam @ 50 psig: $265,667.50 / yr			
Controls:						
Comments and Figures:						

Pump S-1		
	Number needed:	1
Bare Module Cost		$165,000
Purpose:	Pump fluid from Holding Tank to Gas Stripper	
Materials in the Column:	Inlet	Outlet
Flow rate (ft^3/hr)	150060	150060
Mass Fraction of each stream:		
Acetone	0.008	0.008
Butanol	0.016	0.016
Ethanol	0.003	0.003
Water	0.974	0.974
Nitrogen	0	0
Sugar		
Carbon Dioxide	33 PPM	33 PPM
Hydrogen	7 PPB	7 PPB
Temperature:	93.2°F	93.2°F
Design Specs:	Pump Efficiency:	0.7
	Volumetric Flow Rate (ft3/hr):	150060
	Head (ft):	99
	Type:	Centrifugal Pump
	Motor Material:	Stainless Steel
	Pump Material:	Stainless Steel
	Pressure Change:	5 psi
Utilities:	127.67 HP	
Price of Utilities:	$25738/yr	
Comments and Figures:	Price quote obtained from Beijing Great Sources	
	Technology Development Co, Ltd.	

Pump S-2		
Pump ID	Type:	Centrifugal Pump
	Number needed:	1
Bare Module Cost		$165,000
Purpose:	To pump water out of gas stripper bottom	
Materials in the Column:	Inlet	Outlet
Flow rate (ft^3/hr)	67173	67173
Mass Fraction of each stream:		
Acetone	509 PPB	509 PPB
Butanol	378 PPB	378 PPB
Ethanol	60 PPM	60 PPM
Water	0.997	0.997
Nitrogen	125 PPB	125 PPB
Sugar	0.003	0.003
Carbon Dioxide	125 PPB	125 PPB
Hydrogen	259 PPB	259 PPB
Temperature:		
Design Specs:	Pump Efficiency:	0.7
	Volumetric Flow Rate (ft3/hr):	67173
	Head (ft):	12
	Type:	Centrifugal Pump
	Motor Material:	Stainless Steel
	Pump Material:	Stainless Steel
	Pressure Change:	5 psi
Utilities:	2042 HP	
Price of Utilities:	$411667/yr	
Comments and Figures:	Price quote obtained from Beijing Great Sources	
	Technology Development Co, Ltd.	

Pump S-3		
Pump ID	Type:	Centrifugal Pump
	Number needed:	1
Bare Module Cost		$7,694.52
Purpose:	Pump from Phase Separator S-1 to Heat Exchanger	
Materials in the Column:	Inlet	Outlet
Flow rate (ft^3/hr)	767	767
Mass Fraction of each stream:		
Acetone	0.236	0.236
Butanol	0.475	0.475
Ethanol	0.077	0.077
Water	0.206	0.206
Nitrogen	0.006	0.006
Sugar	0	0
Carbon Dioxide	390 PPM	390 PPM
Hydrogen	0	0
Temperature:	41°F	41°F
Design Specs:	Pump Efficiency:	0.7
	Volumetric Flow Rate (ft3/hr):	767
	Head (ft):	295
	Type:	Centrifugal Pump
	Motor Material:	Stainless Steel
	Pump Material:	Stainless Steel
	Pressure Change:	120 psi
Utilities:	12.6 HP	
Price of Utilities:	$2540/yr	
Comments and Figures:		

Pump S-4		
Pump ID	Type:	Centrifugal Pump
	Number needed:	1
Bare Module Cost		$12,449.80
Purpose:	Pump liquid impurities from exchanger to decanter	
Materials in the Column:	Inlet	Outlet
Flow rate (ft^3/hr)	1465	1465
Mass Fraction of each stream:		
Acetone	27 PPM	27 PPM
Butanol	20 PPM	20 PPM
Ethanol	0.003	0.003
Water	0.997	0.997
Nitrogen	7 PPM	7 PPM
Sugar	0	0
Carbon Dioxide	14 PPM	14 PPM
Hydrogen	0	0
Temperature:	213°F	213°F
Design Specs:	Pump Efficiency:	0.7
	Volumetric Flow Rate (ft3/hr):	1465
	Head (ft):	312
	Type:	Centrifugal Pump
	Motor Material:	Stainless Steel
	Pump Material:	Stainless Steel
	Pressure Change:	130 psi
Utilities:	20 HP	
Price of Utilities:	$4032/yr	
Comments and Figures:		

Pumps S-5, S-6		
Pump ID	Type:	Centrifugal Pump
	Number needed:	2
Bare Module Cost		$3,844.51
Purpose:	Pump from Molecular Sieves to Distillation Tower S-1	
Materials in the Column:	Inlet	Outlet
Flow rate (ft^3/hr)	698	698
Mass Fraction of each stream:		
Acetone	0.299	0.299
Butanol	0.602	0.602
Ethanol	0.097	0.097
Water	521 PPM	521 PPM
Nitrogen	0.002	0.002
Sugar	0	0
Carbon Dioxide	5 PPM	5 PPM
Hydrogen	0	0
Temperature:	176°F	176°F
Design Specs:	Pump Efficiency:	0.7
	Volumetric Flow Rate (ft3/hr):	698
	Head (ft):	46
	Type:	Centrifugal Pump
	Motor Material:	Carbon Steel
	Pump Material:	Carbon Steel
	Pressure Change:	15 psi
Utilities:	1.55 HP	
Price of Utilities:	$312/yr	
Comments and Figures:		

Pump S-7		
Pump ID	Type:	Centrifugal Pump
	Number needed:	1
Bare Module Cost		$46,261.00
Purpose:	Pump from Distillation Column S-1 to Dist. Col. S-2	
Materials in the Column:	Inlet	Outlet
Flow rate (ft^3/hr)	260	260
Mass Fraction of each stream:		
Acetone	0.753	0.753
Butanol	445 PPM	445 PPM
Ethanol	0.245	0.245
Water	0.001	0.001
Nitrogen	425 PPM	425 PPM
Sugar	0	0
Carbon Dioxide	10 PPM	10 PPM
Hydrogen	0	0
Temperature:	75°F	75°F
Design Specs:	Pump Efficiency:	0.7
	Volumetric Flow Rate (ft3/hr):	260
	Head (ft):	67
	Type:	Centrifugal Pump
	Motor Material:	Carbon Steel
	Pump Material:	Carbon Steel
	Pressure Change:	12 psi
Utilities:	12.1 HP	
Price of Utilities:	$2439/yr	
Comments and Figures:		

Stripper

Bare Module Cost						$10,419,598
Purpose:	To strip the organic solvents from the water using nitrogen gas					
Materials in the Column:	Feed 1	Feed 2	Feed 3	Bottoms	Overhead	
Flow rate (ft^3/hr)	49,029	4,139,050	150,060	19,544	5.89E+06	
Feed Stage						
Mass Fraction of each stream:	5	1	1			
Acetone	0	0.025	0.008	503 PPB	0.043	
Butanol	0	0.001	0.016	373 PPB	0.042	
Ethanol	0	0.002	0.003	60 PPM	0.008	
Water	0	0.003	0.974	1	0.021	
Nitrogen	1	0.963	0	124 PPB	0.881	
Sugar	0	0	0	0	0	
Carbon Dioxide	0	0.006	33 PPM	258 PPB	0.006	
Hydrogen	0	2 PPM	7 PPB	0	2 PPM	
Temperature:	142 ?F	31 ?F	93?F	82?F	31.9?F	
Design Specs:	Theoretical Trays:	5			Tray Spacing (ft):	2
	Real Trays:	8			Headspace (ft):	3
	Tray Efficiency:	0.7			Sump Space (ft):	4
	Tray Type:	Sieve				
	Functional Height (ft):	23				
	Inside Diameter (ft):	18.17				
	Pressure:	16.7				
	Material:	Stainless Steel 304				
	Number of Man Holes:	2				
Utilities:	0					
Price of Utilities:	No utilities					
Controls:						
Comments and Figures:						

Condenser		
	Number Needed:	1
Bare Module Cost		$160,740
Purpose:	To refrigerate the stripper overhead stream	
Materials in the Column:	Inlet	Outlet
Flow rate (ft^3/hr)	5,888,320	3,928,370
Mass Fraction of each stream:		
Acetone	0.043	0.043
Butanol	0.042	0.042
Ethanol	0.008	0.008
Water	0.021	0.021
Nitrogen	0.881	0.881
Sugar	0	0
Carbon Dioxide	0.006	0.006
Hydrogen	2 PPM	2 PPM
Temperature:	92?F	41?F
Design Specs:	Heat Duty (BTU/hr):	1,930,972
	Heat Transfer Coefficient (BTU/F-ft2-hr):	100
	ΔTlm (°F)	49.1
	Heat Transfer Area (ft2):	393
	Heating Material:	Ammonia
	Type of Heat Exchanger:	Fixed Head
	Shell:	Stainless Steel
	Tube:	Stainless Steel
Utilities:	836 gal /hr of Ammonia	
Price of Utilities:	$103377 / yr	
Comments and Figures:		

Water Heater		
	Number Needed:	1
Bare Module Cost		$257,743
Purpose:	Heat up water from Stripper for Phase splitting	
Materials in the Column:	Inlet	Outlet
Flow rate (ft^3/hr)	4,266	143,202
Mass Fraction of each stream:		
Acetone	509 PPB	509 PPB
Butanol	378 PPB	378 PPB
Ethanol	60 PPM	60 PPM
Water	0.997	0.997
Nitrogen	125 PPB	125 PPB
Sugar	0.003	0.003
Carbon Dioxide	259 PPB	259 PPB
Hydrogen	0	0
Temperature:	82?F	217?F
Design Specs:	Heat Duty (BTU/hr):	46,435,400
	Heat Transfer Coefficient (BTU/F-ft2-hr):	1000
	ΔTlm (°F)	131.1
	Heat Transfer Area (ft2):	338.7
	Heating Material:	Steam
	Type of Heat Exchanger:	Kettle Vaporizer
	Shell:	Carbon Steel
	Tube:	Stainless Steel
Utilities:	Saturated Steam @ 50 psig	50916 lbs/hr
Price of Utilities:		$821184/yr
Comments and Figures:		

Heat Exchanger S-1				
	Number Needed:			1
Bare Module Cost				$62,125
Purpose:				
Materials in the Column:	Inlet 1	Inlet 2	Outlet 1	Outlet 2
Flow rate (ft^3/hr)	138,671	767	1465	3729
Mass Fraction of each stream:				
Acetone	27 PPM	0.236	27 PPM	0.236
Butanol	20 PPM	0.475	20 PPM	0.475
Ethanol	0.003	0.077	0.003	0.077
Water	0.997	0.206	0.997	0.206
Nitrogen	7 PPM	0.006	7 PPM	0.006
Sugar	0	0	0	0
Carbon Dioxide	14 PPM	390 PPM	14 PPM	390 PPM
Hydrogen	0	0	0	0
Temperature:	217?F	41?F	213?F	102?F
Design Specs:	Heat Duty (BTU/hr):	5136922		
	Heat Transfer Coefficient (BTU/F-ft2-hr):	150		
	ΔTlm (°F)	141.24		
	Heat Transfer Area (ft2):	242.5		
	Heating Material:	Inlet 1 Stream		
	Type of Heat Exchanger:	Floating Head		
	Shell:	Carbon Steel		
	Tube:	Carbon Steel		
Utilities:	No Utilities Used			
Price of Utilities:	0			
Comments and Figures:				

Heat Exchanger F-1		
	Number Needed:	1
Bare Module Cost		$194,290
Purpose:	Remove heat from fermentation tank	
Materials in the Column:	Inlet	Outlet
Flow rate (ft^3/hr)	657	657
Mass Fraction of each stream:		
Acetone	0.007	0.007
Butanol	0.014	0.014
Ethanol	0.002	0.002
Water	0.973	0.973
Nitrogen	0	0
Sugar	0.003	0.003
Carbon Dioxide	0	0
Hydrogen	0	0
Temperature:	93?F	93?F
Design Specs:	Heat Duty (BTU/hr):	299,677
	Heat Transfer Coefficient (BTU/F-ft2-hr):	100
	ΔTlm (°F)	16.6
	Heat Transfer Area (ft2):	180.53
	Heating Material:	Chilled Water
	Type of Heat Exchanger:	Shell and Tube
	Shell:	Carbon Steel
	Tube:	Stainless Steel
Utilities:	Chilled Water	6.8 ton-day/hr
Price of Utilities:		$44189 / yr
Comments and Figures:		

Heat Exchangers F-2 - F-3		
	Number Needed:	2
Bare Module Cost		$329,990
Purpose:	Remove heat from fermentation tank	
Materials in the Column:	Inlet	Outlet
Flow rate (ft^3/hr)	2,226	2,226
Mass Fraction of each stream:		
Acetone	0.007	0.007
Butanol	0.014	0.014
Ethanol	0.002	0.002
Water	0.973	0.973
Nitrogen	0	0
Sugar	0.003	0.003
Carbon Dioxide	0	0
Hydrogen	0	0
Temperature:	93?F	93?F
Design Specs:	Heat Duty (BTU/hr):	2,497,308
	Heat Transfer Coefficient (BTU/F-ft2-hr):	100
	ΔTlm (°F)	16.6
	Heat Transfer Area (ft2):	1504.4
	Heating Material:	Chilled Water
	Type of Heat Exchanger:	Shell and Tube
	Shell:	Carbon Steel
	Tube:	Stainless Steel
Utilities:	Chilled Water	57.2 ton-day/hr
Price of Utilities:		$368229 / yr
Comments and Figures:		

Heat Exchangers F-4 - F-12		
	Number Needed:	9
Bare Module Cost		$165,500
Purpose:	Maintain 93 F temperature during fermentation	
Materials in the Column:	Inlet	Outlet
Flow rate (ft^3/hr)	1,685	1,685
Mass Fraction of each stream:		
Acetone	0.008	0.008
Butanol	0.016	0.016
Ethanol	0.002	0.002
Water	0.97	0.97
Nitrogen	0	0
Sugar	0.003	0.003
Carbon Dioxide	0	0
Hydrogen	0	0
Temperature:	93?F	93?F
Design Specs:	Heat Duty (BTU/hr):	172,800
	Heat Transfer Coefficient (BTU/F-ft2-hr):	100
	ΔTlm (°F)	16.6
	Heat Transfer Area (ft2):	104.1
	Heating Material:	Chilled Water
	Type of Heat Exchanger:	Shell and Tube
	Shell:	Carbon Steel
	Tube:	Stainless Steel
Utilities:	Chilled Water	57.2 ton-day/hr
Price of Utilities:		$368229 / yr
Comments and Figures:		

Pasteurizer Heat Exchanger				
	Number Needed:			1
Bare Module Cost				$160,527
Purpose:	To use the pasteurized high-temperature stream to begin heating the incoming sugar cane juice			
Materials in the Column:	Inlet 1	Inlet 2	Outlet 1	Outlet 2
Flow rate (ft^3/hr)	5,148	5586	5344	5377
Mass Fraction of each stream:				
Acetone	0	0	0	0
Butanol	0	0	0	0
Ethanol	0	0	0	0
Water	0.75	0.75	0.75	0.75
Nitrogen	0	0	0	0
Sugar	0.25	0.25	0.25	0.25
Carbon Dioxide	0	0	0	0
Hydrogen	0	0	0	0
Temperature:	77?F	212?F	41?F	151?F
Design Specs:	Heat Duty (BTU/hr):	19243933		
	Heat Transfer Coefficient (BTU/F-ft2-hr):	250		
	ΔTlm (°F)	73		
	Heat Transfer Area (ft2):	1062		
	Heating Material:	Pasteurized Stream		
	Type of Heat Exchanger:	Floating Head		
	Shell:	Carbon Steel		
	Tube:	Carbon Steel		
Utilities:	No Utilities Used			
Price of Utilities:	0			
Comments and Figures:				

Pasteurizer		
	Number Needed:	1
Bare Module Cost		$150,106
Purpose:	Pasteurizes incoming sugar cane juice	
Materials in the Column:	Inlet	Outlet
Flow rate (ft^3/hr)	5,377	5,586
Mass Fraction of each stream:		
Acetone	0	0
Butanol	0	0
Ethanol	0	0
Water	0.75	0.75
Nitrogen	0	0
Sugar	0.25	0.25
Carbon Dioxide	0	0
Hydrogen	0	0
Temperature:	151?F	212?F
Design Specs:	Diameter(ft)	7.46
	Length(ft)	22.40
	Material of Construction	Stainless Steel
	Heating Material:	Steam
Utilities:	Saturated Steam @ 50 psig	439966 lbs/hr
Price of Utilities:		$7095770/yr
Comments and Figures:		

Decanter				
			Number Needed:	1
Bare Module Cost				$97,391
Purpose:	To separate butanol-rich phase from water-rich phase			
Materials in the Column:	Inlet 1	Inlet 2	Outlet 1	Outlet 2
Flow rate (ft^3/hr)	1465	3729	204	732
Mass Fraction of each stream:				
Acetone	27 PPM	0.236	70 PPM	0.284
Butanol	20 PPM	0.475	92 PPM	0.571
Ethanol	0.003	0.077	0.002	0.093
Water	0.997	0.206	0.998	0.046
Nitrogen	7 PPM	0.006	15 PPM	0.007
Sugar	0	0	0	0
Carbon Dioxide	14 PPM	390 PPM	7 PPM	468 PPM
Hydrogen	0	0	0	0
Temperature:	77°F	102°F	168°F	168°F
Design Specs:				
			Diameter (ft):	3.8
			Height (ft):	11.4
			Material of Construction:	Stainless Steel
			Void Space:	0.5
Utilities:				0
Price of Utilities:				
Comments and Figures:				

Molecular Sieves			
	Number Needed:		2
Bare Module Cost			$2,188,202.84
Purpose:	To remove water from the organics stream		
Materials in the Column:	Inlet	Outlet 1	Outlet 2
Flow rate (ft^3/hr)	737	559	698
Mass Fraction of each stream:			
Acetone	0.283	0.045	0.299
Butanol	0.569	0.090	0.602
Ethanol	0.092	0.015	0.097
Water	0.049	0.764	521 PPM
Nitrogen	0.007	0.079	0.002
Sugar	0	0	0
Carbon Dioxide	467 PPM	0.007	5 PPM
Hydrogen	0	0	0
Temperature:	176°F	176°F	176°F
Design Specs:		Type of sieves:	13x
		Diameter (ft):	14.2
		Height (ft):	42.6
		Vessel Material:	Stainless Steel
		On Stream Time (hr):	8
		Price per lb:	$2
Utilities:			NA
Comments and Figures:			

Phase Separator S-1			
	Number Needed:	1	
Bare Module Cost			$74,267.73
Purpose:	To separate ABE and Nitrogen		
Materials in the Column:	Inlet	Outlet 1	Outlet 2
Flow rate (ft^3/hr)	3,930,000	3,860,000	767
Mass Fraction of each stream:			
Acetone	0.043	0.025	0.236
Butanol	0.042	0.001	0.475
Ethanol	0.008	0.002	0.077
Water	0.021	0.003	0.206
Nitrogen	0.881	0.963	0.006
Sugar	0	0	0
Carbon Dioxide	0.006	0.006	390 PPM
Hydrogen	2 PPM	2 PPM	0
Temperature:	41°F	41°F	41°F
Design Specs:			
	Diameter (ft):		2.85
	Height (ft):		8.55
	Material of Construction:	Stainless Steel	
	Void Space:		0.5
Utilities:			NA
Price of Utilities:			
Comments and Figures:			

Phase Separator S-2			
	Number Needed:	1	
Bare Module Cost			$149,690.00
Purpose:	To separate water and organic impurities		
Materials in the Column:	Inlet	Outlet 1	Outlet 2
Flow rate (ft^3/hr)	143,203	138,671	4,532
Mass Fraction of each stream:			
Acetone	509 PPB	27 PPM	2 PPB
Butanol	378 PPB	20 PPM	6 PPB
Ethanol	60 PPM	0.003	11 PPM
Water	0.997	0.997	0.997
Nitrogen	125 PPB	7 PPM	0
Sugar	0.003	0	0.003
Carbon Dioxide	258 PPB	8 PPM	0
Hydrogen	0	0	0
Temperature:	217°F	217°F	217°F
Design Specs:			
	Diameter (ft):		6.7
	Height (ft):		20.2
	Material of Construction:	Stainless Steel	
	Void Space:		0.5
Utilities:			NA
Price of Utilities:			
Comments and Figures:			

Blower		
Pump ID	Type:	Centrifugal Backward Curved
	Number needed:	1
Bare Module Cost		$5,287.04
Purpose:	Fan carrying fresh Nitrogen feed into Stripper	
Materials in the Column:	Inlet	Outlet
Flow rate (ft^3/hr)	58,597	49,029
Mass Fraction of each stream:		
Acetone	0	0
Butanol	0	0
Ethanol	0	0
Water	0	0
Nitrogen	1	1
Sugar	0	0
Carbon Dioxide	0	0
Hydrogen	0	0
Temperature:	77°F	142°F
Design Specs:	Fan Efficiency:	0.7
	Volumetric Flow Rate (ft3/hr):	58,597
	Head (in. H2O):	114
	Type:	Centrifugal Backward Curved
	Fan Material:	Stainless Steel
	Pressure Change:	5 psi
Utilities:	69 HP	
Price of Utilities:	$13910/yr	
Comments and Figures:		

Fan S-1		
Pump ID	Type:	Centrifugal Backward Curved
	Number needed:	1
Bare Module Cost		$73,354.30
Purpose:	Fan connecting stripper overhead to condenser	
Materials in the Column:	Inlet	Outlet
Flow rate (ft^3/hr)	5,888,350	5,297,080
Mass Fraction of each stream:		
Acetone	0.043	0.043
Butanol	0.042	0.042
Ethanol	0.008	0.008
Water	0.021	0.021
Nitrogen	0.881	0.881
Sugar	0	0
Carbon Dioxide	0.006	0.006
Hydrogen	2 PPM	2 PPM
Temperature:	92°F	126°F
Design Specs:	Fan Efficiency:	0.7
	Volumetric Flow Rate (ft3/hr):	5,888,350
	Head (in. H2O):	107
	Type:	Centrifugal Backward Curved
	Fan Material:	Fiberglass
	Pressure Change:	5 psi
Utilities:	3091 HP	
Price of Utilities:	$623146/yr	
Comments and Figures:		

Fan S-2		
Pump ID	Type:	Centrifugal Backward Curved
	Number needed:	1
Bare Module Cost		$36,280.30
Purpose:	Fan leading recycled Nitrogen into Stripper	
Materials in the Column:	Inlet	Outlet
Flow rate (ft^3/hr)	3,823,120	4,139,040
Mass Fraction of each stream:		
Acetone	0.025	0.025
Butanol	0.001	0.001
Ethanol	0.002	0.002
Water	0.003	0.003
Nitrogen	0.963	0.963
Sugar	0	0
Carbon Dioxide	0.006	0.006
Hydrogen	2 PPM	2 PPM
Temperature:	41°F	31°F
Design Specs:	Fan Efficiency:	0.7
	Volumetric Flow Rate (ft3/hr):	3,823,120
	Head (in. H2O):	77
	Type:	Centrifugal Backward Curved
	Fan Material:	Fiberglass
	Pressure Change:	-2 psi
Utilities:	1431 HP	
Price of Utilities:	$288490/yr	
Comments and Figures:		

Fan S-3		
Pump ID	Type:	Centrifugal Backward Curved
	Number needed:	1
Bare Module Cost		$19,504.90
Purpose:	Fan sending gas impurities to Heat Exchanger	
Materials in the Column:	Inlet	Outlet
Flow rate (ft^3/hr)	138,671	113,779
	Mass Fraction of each stream:	
Acetone	27 PPM	27 PPM
Butanol	20 PPM	20 PPM
Ethanol	0.003	0.003
Water	0.997	0.997
Nitrogen	7 PPM	7 PPM
Sugar	0	0
Carbon Dioxide	14 PPM	14 PPM
Hydrogen	0	0
Temperature:	217°F	289°F
Design Specs:	Fan Efficiency:	0.7
	Volumetric Flow Rate (ft3/hr):	1,010,530
	Head (in. H2O):	245
	Type:	Centrifugal Backward Curved
	Fan Material:	Stainless Steel
	Pressure Change:	5 psi
Utilities:	165 HP	
Price of Utilities:	$33264/yr	
Comments and Figures:		

Dryer		
	Number Needed:	1
Bare Module Cost		$732,799.51
Purpose:	Dry the cells down to 9% moisture so they can be sold as fertilizer	
Mass Flow (lb/hr):	2256	
Mass Fraction:		
Acetone	0	
Butanol	0	
Ethanol	0	
Water	1	
Nitrogen	0	
Sugar	0	
Carbon Dioxide	0	
Hydrogen	0	
Design Specs:	Heat Flux (lb/ hr/ft^2):	4.5
	Heat Transfer Area (ft^2):	501
	Material of Construction:	Stainless Steel
Utilities:	Low Pressure Steam 2715 lb/hr	
Price of Utilities:	$43787/yr	
Comments and Figures:		

Evaporator		
	Number Needed:	1
Bare Module Cost		$5,224,162.12
Purpose:	Purify the water being recycled to the mill	
Mass Flow (lb/hr):	258437	
Mass Fraction:		
Acetone	0	
Butanol	0	
Ethanol	0	
Water	1	
Nitrogen	0	
Sugar	0	
Carbon Dioxide	0	
Hydrogen	0	
Effect 1:	Temperature (F):	212
	Pressure (torr):	760
Effect 2:	Temperature (F):	152
	Pressure (torr):	211
Effect 3:	Temperature (F):	93
	Pressure (torr):	38
Design Specs:	Effect 1 Area (ft^2):	2459
	Effect 2 Area (ft^2):	6683
	Effect 3 Area (ft^2):	6929
	Material of Construction:	Stainless Steel
Utilities:	Low Pressure Steam 97396 lb/hr	
Price of Utilities:	$1,570,800 / yr	
Comments and Figures:		

Equipment Cost Summary

Unit Name	C_P	F_{BM}	C_{BM}	Total C_{BM}
Fermenters				
Fermenters F-1 - F-4			$4,954	$4,954
Fermenter F-5	$142,261	4.16	$591,806	$591,806
Fermenters F-6 - F-7	$424,778	4.16	$1,767,075	$3,534,150
Fermenters F-9 - F-19	$424,778	4.16	$1,767,075	$19,437,825
Subtotal				*$23,568,735*
Holding Tanks				
Holding Tank	$606,944	4.16	$2,524,886	$2,524,886
Subtotal				*$2,524,886*
Centrifuges				
Centrifuge F-1	$30,000	2.03	$60,900	$60,900
Centrifuge F-2	$30,000	2.03	$60,900	$60,900
Centrifuges F-3 - F-4	$150,000	2.03	$304,500	$609,000
Centrifuges F-5 - F-6	$500,000	2.03	$1,015,000	$2,030,000
Centrifuges F-7 - F-12	$500,000	2.03	$1,015,000	$6,090,000
Centrifuge F-13	$500,000	2.03	$1,015,000	$1,015,000
Subtotal				*$9,865,800*
Pressure Vessels				
Gas Stripper	$2,504,710	4.16	$10,419,598	$10,419,598
Phase Separator S-1	$24,349	3.05	$74,267	$74,267
Phase Separator S-2	$49,078	3.05	$149,690	$149,690
Decanter	$31,931	3.05	$97,391	$97,391
Molecular Sieves (2)	$526,010	4.16	$2,188,203	$4,376,410
Distillation Column S-1	$180,799	4.16	$752,122	$752,122
Distillation Column S-2	$292,276	4.16	$1,215,870	$1,215,870
Subtotal				*$17,085,348*
Heat Exchangers				
Heat Exchanger F-1	$61,290	3.17	$194,290	$194,290
Heat Exchangers F-2 - F-3	$104,098	3.17	$329,990	$659,980
Heat Exchangers F-4 - F-12	$52,208	3.17	$165,500	$1,489,500
Pasteurizer Heat Exchanger	$50,639	3.17	$160,527	$160,527
Pasteurizer	$49,215	3.05	$150,106	$150,106
Fertilizer Dryer	$339,694	2.06	$732,799	$732,799
Water Heater	$81,306	3.17	$257,743	$257,743
Condenser	$50,706	3.17	$160,740	$160,740
Heat Exchanger S-1	$19,597	3.17	$62,125	$62,125
Triple Effect Evaporator	$2,132,311	2.45	$5,224,162	$5,224,162
Subtotal				*$9,091,972*

Pumps & Motors				
Pump F-1	$6,083	3.3	$20,073	$20,073
Pump F-2	$12,203	3.3	$40,269	$40,269
Pump F-3	$12,724	3.3	$41,988	$41,988
Pumps F-4 - F-5	$13,925	3.3	$45,953	$91,906
Pumps F-6 - F-14	$13,925	3.3	$45,953	$413,577
Pumps F-15 - F-23	$13,925	3.3	$45,953	$413,577
Pump F-24	$946	3.3	$3,123	$3,123
Pump S-1	$50,000	3.3	$165,000	$165,000
Pump S-2	$50,000	3.3	$165,000	$165,000
Pump S-3	$2,331	3.3	$7,694	$7,694
Pump S-4	$3,773	3.3	$12,450	$12,450
Pump S-5, S-6	$1,165	3.3	$3,844	$7,688
Pump S-7	$14,018	3.3	$46,261	$46,261
Subtotal				*$1,428,606*
Fans				
Blower	$2,459	2.15	$5,287	$5,287
Fan S-1	$34,116	2.15	$73,350	$73,350
Fan S-2	$16,874	2.15	$36,280	$36,280
Fan S-3	$9,071	2.15	$19,504	$19,504
Subtotal				*$134,421*
Total				***$63,699,768***

Utility Requirements

Unit Electrical Requirements			
Equipment	Req kW-hr/hr	Price per kW-hr	Cost ($/yr)
Pumps	1688	$0.05	$453,729
Fans	3567	$0.05	$958,810
Agitators	626	$0.05	$168,269
Centrifuges	1166	$0.05	$313,421
Total			*$1,894,228*

Unit Cooling Water Requirements			
Equipment	CW gal/hr	Price per 1000 gal	Cost ($/yr)
Distillation Columns	91760	$0.075	$36,815
Total			*$36,815*

Unit Steam Requirements			
Equipment	Steam lb/hr	Price per 1000 lb	Cost ($/yr)
Pasteurizer 50 psig	439966	$3.00	$7,095,770
Distillation Columns 50 psig	16472	$3.00	$265,660
Distillation Columns 150 psig	10386	$4.80	$268,009
Heat Exchangers 50 psig	50916	$3.00	$821,173
Fertilizer Dryer 50 psig	2715	$3.00	$43,787
Triple Effect Evaporator 50 psig	97396	$3.00	$1,570,800
Total			*$10,065,199*

Unit Refrigerant Requirements			
Equipment	Ammonia gal / hr	Price per gal	Cost ($/yr)
Condenser	836	$0.023	$103,377
Equipment	Chilled water ton-day /hr	Price per ton-day	Cost ($/yr)
Fermentation Heat Exchangers	571.5	$1.20	$3,682,291
Total			*$3,785,668*

Unit Nitrogen Requirements			
Equipment	Nitrogen Ft^3/hr	Price per 100 ft^3	Cost($/yr)
Gas Stripper	58596	$0.02	$63,002
Total			*$63,002*

Utility Costs per Year	Cost ($/yr)
Electrical	$1,894,228
Cooling Water	$36,815
Steam	$10,065,199
Refrigeration	$3,785,668
Nitrogen	$63,002
Total	*$15,844,912*

Economic Analysis

Economic Summary

General Information

Process Title:	ABE Fermentation of Sugar Cane in Brazil
Product:	Ethanol
Plant Site Location:	Brazil
Site Factor:	1.00
Operating Hours per Year:	5376
Operating Days Per Year:	224
Operating Factor:	0.6137

Product Information

This Process will Yield

526	gal of Ethanol per hour
12,624	gal of Ethanol per day
2,827,776	gal of Ethanol per year

Price $2.50 /gal

Chronology

Year	Action	Distribution of Permanent Investment	Production Capacity	Depreciation 5 year MACRS	Product Price
2008	Design		0.0%		
2009	Construction	100%	0.0%		
2010	Production	0%	45.0%	20.00%	$2.50
2011	Production	0%	67.5%	32.00%	$2.50
2012	Production	0%	90.0%	19.20%	$2.50
2013	Production		90.0%	11.52%	$2.50
2014	Production		90.0%	11.52%	$2.50
2015	Production		90.0%	5.76%	$2.50
2016	Production		90.0%		$2.50
2017	Production		90.0%		$2.50
2018	Production		90.0%		$2.50
2019	Production		90.0%		$2.50
2020	Production		90.0%		$2.50
2021	Production		90.0%		$2.50
2022	Production		90.0%		$2.50
2023	Production		90.0%		$2.50
2024	Production		90.0%		$2.50
2025	Production		90.0%		$2.50
2026	Production		90.0%		$2.50
2027	Production		90.0%		$2.50
2028	Production		90.0%		$2.50
2029	Production		90.0%		$2.50

Equipment Costs

Equipment Description		Bare Module Cost
Scale Up Fermentors	Fabricated Equipment	$4,130,909
Fermentors F-9 - F-19	Fabricated Equipment	$19,437,841
Holding Tank	Fabricated Equipment	$2,524,887
Triple Effect Evaporator	Fabricated Equipment	$5,224,162
Gas Stripper	Fabricated Equipment	$10,419,594
Phase Separator S-1	Fabricated Equipment	$74,264
Phase Separator S-2	Fabricated Equipment	$149,688
Decanter	Fabricated Equipment	$97,390
Molecular Sieves 1	Fabricated Equipment	$2,188,202
Molecular Sieves 2	Fabricated Equipment	$2,188,202
Distillation Column S-1	Fabricated Equipment	$752,124
Distillation Column S-2	Fabricated Equipment	$1,215,868
Fermentation Heat Exchangers	Fabricated Equipment	$2,343,771
Pastuerizer Heat Exchanger	Fabricated Equipment	$160,526
Pasteurizer	Fabricated Equipment	$97,390
Fertilizer Dryer	Fabricated Equipment	$699,770
Water Heater	Fabricated Equipment	$257,740
Condenser	Fabricated Equipment	$160,738
Heat Exchanger	Fabricated Equipment	$62,122
Separation Pumps	Fabricated Equipment	$404,092
Fermentation Pumps	Fabricated Equipment	$1,024,505
Blower	Fabricated Equipment	$5,287
Fan S-1	Fabricated Equipment	$73,349
Fan S-2	Fabricated Equipment	$36,279
Fan S-3	Fabricated Equipment	$19,503
Centrifuges	Fabricated Equipment	$9,865,800
Total		*$63,614,001*

Raw Materials

	Raw Material:	Unit:	Required Ratio:		Cost of Raw Material:	
1	Sugar Cane Juice	lb	167.25	lb per gal of Ethanol	$0.050	per lb
2	Nitrogen	100 ft^3	3.613	100 ft^3 per gal of Ethanol	$0.02	per 100 ft^3
3	Sulfuric Acid	lb	0.335	lb per gal of Ethanol	$0.06	per lb
4	Lime	lb	0.1675	lb per gal of Ethanol	$0.05	per lb
5	Gasoline	gal	0.02	gal per gal of Ethanol	$2.00	per gal
6	Caustic	lb	0.75	lb per gal of Ethanol	$0.13	per lb
	Total Weighted Average:				$8.597	per gal of Ethanol

Byproducts

	Byproduct:	Unit:	Ratio to Product		Byproduct Selling Price	
1	Acetone	gal	2.837	gal per gal of Ethanol	$3.000	per gal
2	Butanol	gal	6.435	gal per gal of Ethanol	$4.000	per gal
4	Water	1000 gal	0.0592	1000 gal per gal of Ethanol	$3.000	per 1000 gal
5	CO2/H2	kg	47.1	kg per gal of Ethanol	$0.100	per kg
6	Fertilizer	50 lb	0.0404	50 lb per gal of Ethanol	$47.950	per 50 lb
	Total Weighted Average:				$41.076	per gal of Ethanol

Utilities

	Utility:	Unit:	Required Ratio		Utility Cost	
1	High Pressure Steam	1000 lb	0.01975	1000 lb per gal of Ethanol	$4.800	per 1000 lb
2	Low Pressure Steam	1000 lb	1.17462	1000 lb per gal of Ethanol	$3.000	per 1000 lb
3	Cooling Water	1000 gal	0.17444	1000 gal per gal of Ethanol	$0.075	per 1000 gal
4	Chilled Water	ton-day	1.0865	ton-day per gal of Ethanol	$1.200	per ton-day
5	Electricity	kWh	15.6	kWh per gal of Ethanol	$0.050	per kWh
	Total Weighted Average:				$5.716	per gal of Ethanol

Variable Costs

General Expenses:

Selling / Transfer Expenses:	3.00%	of Sales
Direct Research:	4.80%	of Sales
Allocated Research:	0.50%	of Sales
Administrative Expense:	2.00%	of Sales
Management Incentive Compensation:	1.25%	of Sales

Working Capital

Accounts Receivable	⇨	30	Days
Cash Reserves (excluding Raw Materials)	⇨	30	Days
Accounts Payable	⇨	30	Days
Ethanol Inventory	⇨	4	Days
Raw Materials	⇨	2	Days

Total Permanent Investment

Cost of Site Preparations:	5.00%	of Total Bare Module Costs
Cost of Service Facilities:	5.00%	of Total Bare Module Costs
Allocated Costs for utility plants and related facilities:	$0	
Cost of Contingencies and Contractor Fees:	18.00%	of Direct Permanent Investment
Cost of Land:	2.00%	of Total Depreciable Capital
Cost of Royalties:	$0	
Cost of Plant Start-Up:	10.00%	of Total Depreciable Capital

Fixed Costs

Operations

Operators per Shift:	4	(assuming 5 shifts)
Direct Wages and Benefits:	$35	/operator hour
Direct Salaries and Benefits:	15%	of Direct Wages and Benefits
Operating Supplies and Services:	6%	of Direct Wages and Benefits
Technical Assistance to Manufacturing:	$0.00	per year, for each Operator per Shift
Control Laboratory:	$0.00	per year, for each Operator per Shift

Maintenance

Wages and Benefits:	4.50%	of Total Depreciable Capital
Salaries and Benefits:	25%	of Maintenance Wages and Benefits
Materials and Services:	100%	of Maintenance Wages and Benefits
Maintenance Overhead:	5%	of Maintenance Wages and Benefits

Operating Overhead

General Plant Overhead:	7.10%	of Maintenance and Operations Wages and Benefits
Mechanical Department Services:	2.40%	of Maintenance and Operations Wages and Benefits
Employee Relations Department:	5.90%	of Maintenance and Operations Wages and Benefits
Business Services:	7.40%	of Maintenance and Operations Wages and Benefits

Property Taxes and Insurance

Property Taxes and Insurance:	2%	of Total Depreciable Capital

Straight Line Depreciation

Direct Plant:	8.00%	of Total Depreciable Capital, less 1.18 times the Allocated Costs for Utility Plants and Related Facilities
Allocated Plant:	6.00%	of 1.18 times the Allocated Costs for Utility Plants and Related Facilities

Other Annual Expenses

Rental Fees (Office and Laboratory Space):	$0
Licensing Fees:	$0
Miscellaneous:	$0

Depletion Allowance

Annual Depletion Allowance:	$0

Variable Cost Summary

Variable Costs at 100% Capacity:

General Expenses

Selling / Transfer Expenses:	$	212,083
Direct Research:	$	339,333
Allocated Research:	$	35,347
Administrative Expense:	$	141,389
Management Incentive Compensation:	$	88,368
Total General Expenses	$	816,520

Raw Materials	$8.596985 per gal of Ethanol		$24,310,348
Byproducts	$41.075780 per gal of Ethanol		($116,153,105)
Utilities	$5.715543 per gal of Ethanol		$16,162,275
Total Variable Costs		$	(74,863,961)

Fixed Cost Summary

Operations		
Direct Wages and Benefits	$	1,456,000
Direct Salaries and Benefits	$	218,400
Operating Supplies and Services	$	87,360
Technical Assistance to Manufacturing	$	-
Control Laboratory	$	-
Total Operations	$	1,761,760
Maintenance		
Wages and Benefits	$	3,748,304
Salaries and Benefits	$	937,076
Materials and Services	$	3,748,304
Maintenance Overhead	$	187,415
Total Maintenance	$	8,621,099
Operating Overhead		
General Plant Overhead:	$	451,544
Mechanical Department Services:	$	152,635
Employee Relations Department:	$	375,227
Business Services:	$	470,624
Total Operating Overhead	$	1,450,030
Property Taxes and Insurance		
Property Taxes and Insurance:	$	1,665,913
Other Annual Expenses		
Rental Fees (Office and Laboratory Space):	$	-
Licensing Fees:	$	-
Miscellaneous:	$	-
Total Other Annual Expenses	$	-
Total Fixed Costs	$	13,498,802

Investment Summary

Bare Module Costs				
Fabricated Equipment	$	63,614,001		
Process Machinery	$	-		
Spares	$	-		
Storage	$	558,300		
Other Equipment	$	-		
Catalysts	$	-		
Computers, Software, Etc.	$	-		
Total Bare Module Costs:			$	64,172,301
Direct Permanent Investment				
Cost of Site Preparations:	$	3,208,615		
Cost of Service Facilities:	$	3,208,615		
Allocated Costs for utility plants and related facilities:	$	-		
Direct Permanent Investment			$	70,589,531
Total Depreciable Capital				
Cost of Contingencies & Contractor Fees	$	12,706,116		
Total Depreciable Capital			$	83,295,647
Total Permanent Investment				
Cost of Land:	$	1,665,913		
Cost of Royalties:	$	-		
Cost of Plant Start-Up:	$	8,329,565		
Total Permanent Investment - Unadjusted			$	93,291,124
Site Factor				1.00
Total Permanent Investment			$	93,291,124

Working Capital

		2009		2010		2011
Accounts Receivable	$	261,472	$	130,736	$	130,736
Cash Reserves	$	1,097,054	$	548,527	$	548,527
Accounts Payable	$	(1,496,933)	$	(748,466)	$	(748,466)
Ethanol Inventory	$	34,863	$	17,431	$	17,431
Raw Materials	$	59,943	$	29,972	$	29,972
Total	$	(43,600)	$	(21,800)	$	(21,800)
Present Value at 15%	*$*	*(37,913)*	*$*	*(16,484)*	*$*	*(14,334)*
Total Capital Investment			$	93,222,393		

Cash Flow Summary

Year	Percentage of Design Capacity	Product Unit Price	Sales	Capital Costs	Working Capital	Var Costs	Fixed Costs	Depreciation	Depletion Allowance	Taxible Income	Taxes	Net Earnings	Cash Flow	Cumulative Net Present Value at 15%
2008	0%		-	-	-	-	-	-	-	-	-	-	-	-
2009	0%		-	(93,291,100)	43,600	-	-	-	-	-	-	-	(93,247,500)	(81,084,800)
2010	45%	$2.50	3,181,200	-	21,800	33,688,800	(13,498,800)	(16,659,100)	-	6,712,100	(2,483,500)	4,228,600	20,909,600	(65,274,200)
2011	68%	$2.50	4,771,900	-	21,800	50,533,200	(13,498,800)	(26,654,600)	-	15,151,600	(5,606,100)	9,545,500	36,221,900	(41,457,700)
2012	90%	$2.50	6,362,500	-	-	67,377,600	(13,498,800)	(15,992,800)	-	44,248,500	(16,371,900)	27,876,600	43,869,300	(16,375,200)
2013	90%	$2.50	6,362,500	-	-	67,377,600	(13,498,800)	(9,595,700)	-	50,645,600	(18,738,900)	31,906,700	41,502,400	4,258,800
2014	90%	$2.50	6,362,500	-	-	67,377,600	(13,498,800)	(9,595,700)	-	50,645,600	(18,738,900)	31,906,700	41,502,400	22,201,400
2015	90%	$2.50	6,362,500	-	-	67,377,600	(13,498,800)	(4,797,800)	-	55,443,400	(20,514,100)	34,929,400	39,727,200	37,136,300
2016	90%	$2.50	6,362,500	-	-	67,377,600	(13,498,800)	-	-	60,241,300	(22,289,300)	37,952,000	37,952,000	49,542,900
2017	90%	$2.50	6,362,500	-	-	67,377,600	(13,498,800)	-	-	60,241,300	(22,289,300)	37,952,000	37,952,000	60,331,200
2018	90%	$2.50	6,362,500	-	-	67,377,600	(13,498,800)	-	-	60,241,300	(22,289,300)	37,952,000	37,952,000	69,712,400
2019	90%	$2.50	6,362,500	-	-	67,377,600	(13,498,800)	-	-	60,241,300	(22,289,300)	37,952,000	37,952,000	77,869,900
2020	90%	$2.50	6,362,500	-	-	67,377,600	(13,498,800)	-	-	60,241,300	(22,289,300)	37,952,000	37,952,000	84,963,400
2021	90%	$2.50	6,362,500	-	-	67,377,600	(13,498,800)	-	-	60,241,300	(22,289,300)	37,952,000	37,952,000	91,131,700
2022	90%	$2.50	6,362,500	-	-	67,377,600	(13,498,800)	-	-	60,241,300	(22,289,300)	37,952,000	37,952,000	96,495,400
2023	90%	$2.50	6,362,500	-	-	67,377,600	(13,498,800)	-	-	60,241,300	(22,289,300)	37,952,000	37,952,000	101,159,500
2024	90%	$2.50	6,362,500	-	-	67,377,600	(13,498,800)	-	-	60,241,300	(22,289,300)	37,952,000	37,952,000	105,215,200
2025	90%	$2.50	6,362,500	-	-	67,377,600	(13,498,800)	-	-	60,241,300	(22,289,300)	37,952,000	37,952,000	108,741,900
2026	90%	$2.50	6,362,500	-	-	67,377,600	(13,498,800)	-	-	60,241,300	(22,289,300)	37,952,000	37,952,000	111,808,600
2027	90%	$2.50	6,362,500	-	-	67,377,600	(13,498,800)	-	-	60,241,300	(22,289,300)	37,952,000	37,952,000	114,475,300
2028	90%	$2.50	6,362,500	-	-	67,377,600	(13,498,800)	-	-	60,241,300	(22,289,300)	37,952,000	37,952,000	116,794,200
2029	90%	$2.50	6,362,500	-	(87,200)	67,377,600	(13,498,800)	-	-	60,241,300	(22,289,300)	37,952,000	37,864,800	118,806,000

Profitability Measures

The Internal Rate of Return (IRR) for this project is 37.03%

The Net Present Value (NPV) of this project in 2008 is $ 118,806,000

ROI Analysis (Third Production Year)

Annual Sales	6,362,496
Annual Costs	53,878,763
Depreciation	(7,463,290)
Income Tax	(19,527,849)
Net Earnings	33,250,120
Total Capital Investment	93,203,924
ROI	35.67%

Sensitivity Analyses

Note: The Sensitivity Analyses section below takes quite a bit of memory to update each time a cell is changed; therefore, automatic calculations are turned off. After making your axis selections, press "F9" to recalculate the IRR values. (These two lines may be deleted before printing.)

	Vary Initial Value by +/-
x-axis	50%
y-axis	50%

Product Price

	Variable Costs										
	###########	###########	###########	###########	###########	############	###########	###########	###########	###########	###########
$1.25	16.16%	20.53%	24.56%	28.36%	31.99%	35.49%	38.89%	42.20%	45.44%	48.61%	51.73%
$1.50	16.58%	20.91%	24.92%	28.70%	32.32%	35.80%	39.19%	42.49%	45.72%	48.89%	52.00%
$1.75	17.00%	21.29%	25.27%	29.03%	32.64%	36.11%	39.49%	42.78%	46.00%	49.16%	52.27%
$2.00	17.41%	21.67%	25.62%	29.37%	32.95%	36.42%	39.79%	43.07%	46.29%	49.44%	52.54%
$2.25	17.82%	22.04%	25.97%	29.70%	33.27%	36.73%	40.08%	43.36%	46.57%	49.72%	52.81%
$2.50	18.23%	22.41%	26.32%	30.03%	33.59%	37.03%	40.38%	43.65%	46.85%	49.99%	53.08%
$2.75	18.63%	22.78%	26.66%	30.36%	33.90%	37.34%	40.68%	43.94%	47.13%	50.27%	53.35%
$3.00	19.03%	23.15%	27.01%	30.68%	34.22%	37.64%	40.97%	44.22%	47.41%	50.54%	53.62%
$3.25	19.42%	23.51%	27.35%	31.01%	34.53%	37.94%	41.26%	44.51%	47.69%	50.81%	53.89%
$3.50	19.81%	23.87%	27.69%	31.33%	34.84%	38.24%	41.56%	44.80%	47.97%	51.09%	54.15%
$3.75	20.19%	24.23%	28.03%	31.66%	35.15%	38.54%	41.85%	45.08%	48.25%	51.36%	54.42%

Sensitivity Analysis

Based on the assumptions in the problem statement, this plant seems to be a profitable venture, with an IRR of 37.03% and a net present value of $118,806,000 over the 20 year life of the plant. However, because the plant uses a sugar cane juice feedstock, and because biological feedstocks tend to fluctuate in price, it is desirable to know at what price of sugar cane juice the design becomes unprofitable.

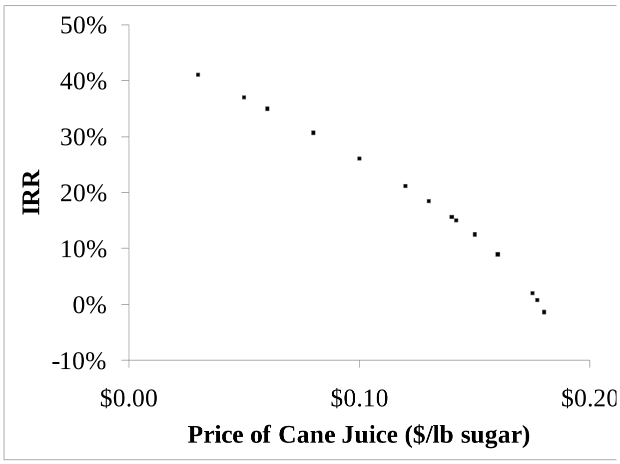


Figure 1. The sensitivity of the IRR of the process to the price of cane juice.

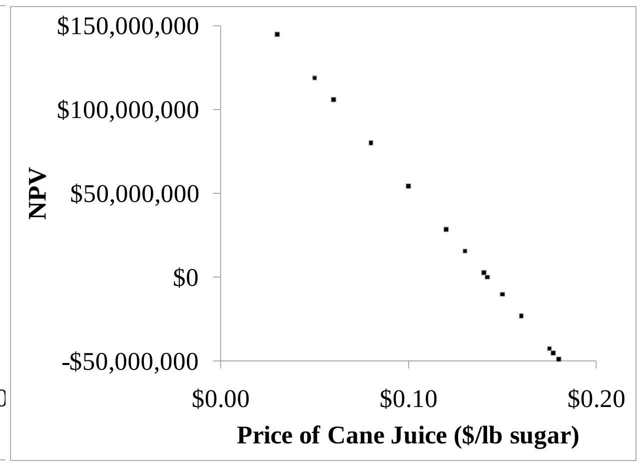


Figure 2. The sensitivity of the NPV of the process to the price of cane juice.

Figure 1 and Figure 2 above show how the IRR and NPV of the process change with the price of the sugar cane juice. The price of sugar cane at which the process breaks even in cost, giving an IRR of 15% and an NPV of $0, is $0.142/lb sugar. At this point, the revenues and costs for the plant over its lifetime are roughly equal, and no investment should be made in this plant if sugar cane juice gets above this price. Since the price of sugar cane is currently well below that level, this plant is a suggested investment. In fact, Qureshi and Blaschek (2001), in their evaluation of the ABE fermentation process, call an increase in feedstock cost by a factor of 2.5 the "worst-case scenario." Even in that case, if the price of sugar cane juice increased from

$0.05/lb sugar to $0.125/lb sugar, the plant would be profitable, generating an IRR of 19.84% and a net present value of $22 million.

Another uncertainty in the process design is the sale of the byproduct gases. Although CO_2 has several industrial uses, as discussed above, it is often considered an undesirable byproduct of ABE fermentation, and several industrial consultants suggested the plant might have difficulty finding a buyer for the gas. The price of $0.10/kg, as suggested by Qureshi and Blaschek (2001), was assumed in this design. However, if the plant could only sell the gases for $0.05/kg, the design would still be profitable, generating an IRR of 33.97% and a NPV of $100 million. In a worst-case scenario, the plant would be unable to find a buyer at any price for the CO_2/H_2, and the H_2 could be combusted for energy while the CO_2 is released into the atmosphere. This option would have undesirable environmental consequences, but, even if none of the energy from the H_2 is recovered to make steam, the design would be profitable, with an IRR of 30.83% and a NPV of $82 million.

Finally, when dealing with a bioprocess, the yield of useful products from batch to batch can be highly variable. If only 75% of the expected amount of ABE was actually obtained from the fermentation, the design would still be profitable, with an IRR of 27.25% and a NPV of $62 million. On the other hand, if genetic engineering improved the yield of ABE from 37% of the glucose weight to 40% of the glucose weight, the design would have a 40.07% IRR and a $137 million NPV. If 45% of the glucose weight was converted into ABE, the design would have an IRR of 44.91% and a NPV of $168 million. These values suggest the importance of research into the ABE process because, although it can currently be profitable, relatively modest yield increases can drastically improve the profitability of the process.

Conclusions and Recommendations

Conclusions and Recommendations

ABE Fermentation process using bacterial fermentation was phased out after World War II because cheaper ways of producing acetone, butanol, and ethanol from petroleum were found and employed. With new, cheaper separation technology, rising petroleum prices and energy requirements, and a growing need for reusable energy sources, there is promise again in using ABE fermentation to manufacture biofuels. Our design report shows that ABE fermentation can indeed serve as a cost-effective way of producing large quantities of butanol, acetone, and ethanol solvents/biofuels, providing a means of producing these chemicals without petrochemical consumption. With an IRR of 37.03%, ROI of 35.67%, and a NPV of $118,806,000 over the 20 year life of the plant, the design developed for this project is economically feasible and profitable. Therefore, a plant should be constructed and operated according to the design and specifications laid out in this report.

Although our design is economically profitable, there are a few areas where further modifications can be done to cut costs and generate new revenue streams. It seems likely that the largest cost reductions are to come from reducing the size of the process streams. Since the fermentation products are more than 97% water, pumping these streams and recovering the desired products becomes quite expensive. Moreover, further research could be carried out for improving the yield of acetone, butanol, and ethanol from the sugar, and improving the *Clostridium* strain's resistance to these solvents and the sugars could go a long way toward further improving the economics of this process.

A major revenue stream can come from finding buyers for the CO_2/H_2 off gas. Due to environmental considerations, the gas cannot be released into the atmosphere, providing further incentive to make beneficial use of the off gas. Thus we recommend carrying further research to come up with a cheaper way of separating and utilizing both the H_2 and CO_2 in off gas, either as fuel, or as feedstock for another chemical process. Another option would be to utilize the off gas in the gas stripper to separate the solvents from the water. Current quantity, composition, and purity requirements render the off gas unusable as the stripping gas in our design, since the solubility of CO_2 in water is greater than the solubility of N_2 in water. Therefore, we recommend conducting further research to determine optimal conditions where the off gas can be utilized as the stripping gas instead of nitrogen gas we had to use in our design. If such conditions were to be found, it would reduce the cost of the gas stripping process.

Additional Considerations

Start-Up

Before the plant can be started up all the parts and components need to be washed and sterilized. A onetime cost when the plant is first started will be the cost of the bulk of the nitrogen for the stripper and the water for diluting the cane juice. These are for the materials for the two recycle streams and can be saved to be used from year to year. At the startup of each year cycle the recycle will need to be pasteurized before it can be used again. Also, eleven days before the plant opens for the year the fermenter scale up from test tube will have to be started, so that when the first juice arrives a fermenter will be ready to seed. To do this the sugar for this process will have to be purchased and utilities for the scale up paid. Also due to the scale up's capacity it will take between three and four days for the plant to reach full capacity. The loss of juice from this will be mitigated by typically lower than harvests at the beginning of the growing season. Also, due to the reuse of cells there will not be fertilizer product until five days after the plant opens.

Safety

The production of ABE poses few major safety concerns. All the chemicals used are safe if they come in contact with the skin and can be inhaled at moderate concentrations. The main concern is that the ABE and the hydrogen produced are highly flammable. Since all fermentation vessels are kept at 5 psig, any leaks will not let air in. The striper is operated using inert nitrogen and the fermentation is preformed anaerobically, preventing oxygen from entering the process at those points. The nitrogen in the stripper will also contain trace amounts of an odorant (methanethiol) so leaks can easily be detected, since most of the ABE is in vapor form in this stage. The hydrogen will remain in a mixture with CO_2 and be carefully handled until sale. All staff will be trained in proper safety protocol and the tanks will be checked for leaks often.

Environmental Concerns

This process is one that is environmentally friendly. It takes carbon that is captured from the air by plants and uses it for fuel. Thus, ideally, when the fuel is used there should be no net carbon put back into the atmosphere. This is not quite correct since the plant gets carbon from a variety of sources and not all of that goes into the cane juice and not all of the carbon in the juice goes to ABE product, but it is fairly close. One thing that should be considered is if the ABE process makes more CO_2 than a similar process which makes only ethanol. The production of ethanol produces two moles of ethanol per mole of sugar, while the ABE makes only one mole of butanol or acetone per mole of sugar used. The processes for ethanol and for butanol make the same the same amount of CO_2 per mole sugar, but acetone makes one more per mol sugar used. This is shown below:

$$C_6H_{12}O_6 \rightarrow C_4H_{10}O\ (butanol) + 2CO_2 + H_2O$$

$$C_6H_{12}O_6 \rightarrow 2C_2H_5O\ (ethanol) + 2CO_2 + H_2$$

$$C_6H_{12}O_6 \rightarrow C_3H_6O\ (acetone) + 3CO_2 + 4H_2$$

These facts show that the ABE process should not have significantly more CO_2 off gas as a similar process making only ethanol since ABE makes ABE in about a 2:4:1 ratio respectively. However in regards to CO_2, ethanol fermentation is slightly more eco-friendly, even though it has a lower energy value. This also is, at best, an estimation since the sugar for these processes is cane juice, not glucose.

The fertilizer produced by this process is also better for than the environment than fertilizer that is more commonly used. This fertilizer is organism-based, instead of just loaded with free phosphate and nitrate, and will not lead to things like eutrophication. Even the water released by the plant is de-ionized and can be sold back to the cane mill.

Acknowledgements and References

Acknowledgements

Thank you to:

- Professor Leonard Fabiano, for his tremendous help in getting our ASPEN PLUS simulation to converge and equipment pricing
- Professor Daniel Hammer, for keeping us on track and providing us with advice about the biological aspects of the process
- Mr. Bruce Vrana, for writing the problem statement, answering numerous questions via email, and giving us advice with his knowledge about similar industrial processes
- All of the other industrial consultants, including, Mr. Adam Brostow, Mr. David Kolesar, Mr. John Wismer, Dr. William Retallick, and Dr. Tiffany Rau, for giving us advice during our weekly meetings

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Appendix

Appendix A: Sample Equipment Size and Cost Calculations

1. Yield of Products Per Glucose Molecule

A sample yield calculation is shown for acetone in the growth phase. The calculation is repeated for butanol, ethanol, carbon dioxide, and biomass in both growth phase and production phase according to the carbon mole balance given in the problem statement. Hydrogen production is given as 2 wt% of glucose consumed.

$$(100\ g\ sugar)\left(\frac{1\ mol\ sugar}{178\ g\ sugar}\right)\left(\frac{6\ mol\ C}{1\ mol\ sugar}\right)\left(\frac{0.149\ mol\ C\ in\ acetone}{1\ mol\ C}\right)\left(\frac{1\ mol\ acetone}{3\ mol\ C\ in\ acetone}\right)$$

$$\left(\frac{58.08\ g}{1\ mol}\right) = 9.72\ g\ acetone\ per\ 100\ g\ sugar$$

Total yield in mass percent of glucose consumed is:

Compound	*Growth Phase*	*Production Phase*
Acetone	9.72%	11.22%
Butanol	19.43%	22.36%
Ethanol	3.26%	3.73%
Carbon Dioxide	59.19%	61.12%
Hydrogen	2.00%	2.00%
Biomass	8.68%	0.87%

Note that the percentages sum to over 100% since some of the biomass mass is supplied by salts and nitrogen.

2. Production-Phase Design Calculations

Using the equation in the problem statement, to make the units match, the reciprocal of the cell concentration (X, g/L) must be included on the left side of the equation. The initial cell charge to the fermenter is designated as X_o (g/L). The initial sugar concentration is 67.5 g/L and the final sugar concentration is 4.05 g/L. In the production phase, using the mass yield values from above, the ABE concentration is calculated as a function of sugar remaining in the fermenter as:

$$P = (0.1122 + 0.2236 + 0.0373)\,(67.5 - S)$$

Similarly, the cell concentration in the fermenter is calculated as:

$$X = X_o + 0.0087(67.5 - S)$$

Therefore,

$$\left(\frac{dS}{dt}\right)\left(\frac{1}{X_o + 0.587 - 0.0087S}\right) = -0.8\,\left(\frac{S}{5+S}\right)\left(\frac{7}{32.18 - 0.373S}\right)$$

Integrating from S = 67.5 g/L to S = 4.05 g/L, and setting the fermentation time at ~24 hours:

$$X_o = 11.15\ {}^{g}\!/_{L}$$

$$t = 23.74\ h$$

The number of fermenters in use can be determined by considering the amount of sugar that must be consumed each day (each set of batches).

$$\left(\frac{30900\ kg}{1\ hour}\right)\left(\frac{24\ hours}{1\ day}\right)\left(\frac{1\ day}{1\ set\ of\ batches}\right)\left(\frac{1000\ g}{1\ kg}\right)\left(\frac{1\ L}{67.5\ g}\right)\left(\frac{0.264\ gal}{1\ L}\right)\left(\frac{1\ tank}{450000\ gal}\right)$$

$$= 6.5\ fermenters$$

Therefore, 7 fermenters are needed in use at any time. Because the fermenters are in use for 168 hours out of every 216 hours, 9 fermenters are needed.

3. Growth-Phase Design Calculations

To achieve an initial dry cell weight of 11.15 g/L, 1.90 x 10^7 g DCW are needed per fermenter. Since 9 fermenters must be changed every 216 hours, 9100 kg sugar/hr are needed to grow the cells before the final fermentation. This leaves 30900 kg sugar/hr for the production phase (see Appendix A.2). The growth phase kinetics are:

$$\left(\frac{dS}{dt}\right)\left(\frac{1}{X_o + 5.87 - 0.087S}\right) = -1.1\ \left(\frac{S}{5+S}\right)\left(\frac{7}{28.88 - 0.324S}\right)$$

Integrating from S = 67.5 g/L to S = 4.05 g/L, and setting the fermentation time at ~12 hours:

$$X_o = 11.75\ {}^{g}/_{L}$$

$$t = 12.03\ h$$

Thus, the cells go from 11.75 g/L to a final concentration of 22.78 g/L in the course of 24 hours. This means one 500000 gallon vessel, operating at 93% capacity, is required to provide the cells for the fermenters. In order to account for cleaning and down time, two will be purchased.

To account for contamination, new cells must be produced in a scale-up process. Since the problem statement says we have to deal with three days of contaminated products per year, our plant suffers from one contaminated fermentation every 11 days, on average. To replace the cells lost from contamination, the 60,000 gallon vessel must have 2.2×10^7 g DCW total every 11 days, or every 264 hours. This requires, on average, 1016 kg sugar per hour, which is accounted for by the 4.05 g/L not consumed in the fermentation section. In the case of an extreme situation, such as two contaminated fermentation batches in the same day, the cells in the fermenter will simply be reused for additional batches until the new cells are ready, since we are well below the 500 hours of cell use employed by Ezeji, Qureshi, and Blaschek (2005).

The final cell concentration in the 60,000 gallon vessel must be 96.8 g/L. Since 5.27 g/L cells can be grown per batch, due to substrate and product inhibition, 18 batches are required every 11 days, starting from an initial concentration of 4 g/L. The batch times are as follows: 25.4 hours, 14.4 hours, 10.2 hours, 7.83 hours, 6.38 hours, 5.38 hours, 4.66 hours, 4.10 hours, 3.67 hours, 3.31 hours, 3.02 hours, 2.78 hours, 2.57 hours, 2.39 hours, 2.24 hours, 2.10 hours, 1.98 hours, and 1.87 hours, leaving ample time for changing and cleaning the fermenter.

To feed the 60,000 gallon vessel, the final cell concentration in the 5000 gallon vessel needs to be 48 g/L. If the 5000 gallon vessel starts at 2 g/L, 38000 g DCW needs to come from the 3 liter vessels. If their final concentration is 100 g/L, 127 3 liter vessels are needed. If their initial concentration is 4 g/L, 26 600 mL vessels are needed. This can be fed by a test tube bank.

4. Sample Pump Price Calculation

Pump S-7

Flow rate from ASPEN PLUS = 260 ft^3/hr = 32.41 gpm

Pump will increase the pressure of the stream by 12 psi to facilitate distillation in the acetone-ethanol separating column,. And we also need to add half the height of the butanol column the stream is coming from.

$$\Delta P = \left(\frac{12\,psi}{8.338\,lb/ft^3}\right)\left(\frac{12\,in}{1\,ft}\right)^2 + 32\,ft = 66.98\,ft$$

From Seider, Seader, Lewin, and Widadgo (2009):

$$S = QH^{0.5} = (32.41\,gpm)(67)^{0.5} = 265.29$$

From Eq. 22.14, C_B = \$2,905.86. Multiply by F_T = 2 and F_M = 2 to account for stainless steel construction and the high process volume to get C_P = \$11,623.40. The horsepower requirement is:

$$P_C = \frac{QH\rho}{33000 * \eta} = \frac{(32.41\,gpm)(67\,ft)(48\,lb/ft^3)}{(33000)(0.7)(7.48\,gal/ft^3)} = 12.1HP$$

Using Eq.22.19, C_B = \$889.70. Multiplying by a factor of two to account for stainless steel construction, C_P = \$1779.40. The total pump cost is

$$C_{BM} = (\$11623.40 + \$1779.40)\left(\frac{523.6}{500}\right)(3.30) = \$46{,}261$$

5. Holding Tank Price Calculation

The price of a 1,250,000 gallon holding tank is as follows:

$$C_{BM} = 265(1.7)(1250000)^{0.51}\left(\frac{523.6}{500}\right)(4.16) = \$2{,}524{,}886.53$$

6. Decanter Price Calculation

From Eq. 22.61, calculate the design pressure at an operating pressure of 130.3 psig to be P_d = 190.7 psig. From the drum sizing sheet supplied by Professor Fabiano, the vessel length is 11.1 ft and the vessel diameter is 3.67 ft. The calculated wall thickness is:

$$t_P = \frac{P_d D_i}{2SE - 1.2P_d} = \frac{(190.7\,psi)(3.8\,ft)(12\ {}^{in}/_{ft})}{(2)(15000\,psi)(1) - (1.2)(190.7\,psi)} = 0.292\,in$$

The decanter weight is:

$$W = \pi(D_i + t_S)(L + 0.8D_i)t_S\rho_{carbon\ \ steel}$$

$$= \pi\left(3.8ft + (0.282\,in)\left(0.083\,{}^{ft}/_{in}\right)\right)\Big(11.4\,ft$$

$$+ (0.8)(3.67)\Big)(0.282\,in)\left(0.083\,{}^{ft}/_{in}\right)\left(490\ {}^{lb}/_{ft^3}\right) = 1974\,lb$$

From Eq. 22.53, C_v = \$16025.90. From Eq. 22.55, C_{PL} = \$3285.51. From Eq. 22.52, using a material factor of 1.7 to account for the stainless steel construction, C_P = \$30527.50. Then,

$$C_{BM} = (\$30527.50)(3.05)\left(\frac{523.6}{500}\right) = \$97391.90$$

7. Phase Separator Price Calcuation

From Seider, Seader, Lewin, and Widagdo (2009), use a design pressure of P_d = 10 psig for an operating pressure less than 10 psig. From the drum sizing sheet supplied by Professor Fabiano, the vessel length is 8.55 ft and the vessel diameter is 2.85 ft. The minimum recommended wall thickness for sufficient rigidity is given by Seider, Seader, Lewin, and Widagdo (2009) as 0.25 in. The phase separator weight is:

$$W = \pi(D_i + t_S)(L + 0.8D_i)t_S\rho_{carbo\ n\ steel}$$

$$= \pi\left(2.85ft + (0.25\ in)\left(0.083\ {}^{ft}/_{in}\right)\right)\big(8.55\ ft$$

$$+ (0.8)(2.85)\big)\ (0.25\ in)\left(0.083\ {}^{ft}/_{in}\right)\left(490\ {}^{lb}/_{ft^3}\right) = 993\ lb$$

From Eq. 22.53, C_v = \$12218.52. From Eq. 22.55, C_{PL} = \$2479.82. From Eq. 22.52, using a material factor of 1.7 to account for the stainless steel construction, C_P = \$23251.30. Then,

$$C_{BM} = (\$23251.30)(3.05)\left(\frac{523.6}{500}\right) = \$74267.73$$

From Seider, Seader, Lewin, and Widagdo (2009), use a design pressure of P_d = 10 psig for an operating pressure less than 10 psig. From the drum sizing sheet supplied by Professor Fabiano, the vessel length is 20.2 ft and the vessel diameter is 6.7 ft. The minimum recommended wall thickness for sufficient rigidity is given by Seider, Seader, Lewin, and Widagdo (2009) as 0.5 in. The decanter weight is:

$$W = \pi(D_i + t_S)(L + 0.8D_i)t_S\rho_{carbon\ \ steel}$$

$$= \pi\left(6.7ft + (0.5\ in)\left(0.083\ {}^{ft}/_{in}\right)\right)\big(20.2\ ft$$

$$+ (0.8)(6.7)\big)\ (0.25\ in)\left(0.083\ {}^{ft}/_{in}\right)\left(490\ {}^{lb}/_{ft^3}\right) = 5504\ lb$$

From Eq. 22.53, C_v = \$25865.00. From Eq. 22.55, C_{PL} = \$2949.57. From Eq. 22.52, using a material factor of 1.7 to account for the stainless steel construction, C_P = \$46920.50. Then,

$$C_{BM} = (\$46920.50)(3.05)\left(\frac{523.6}{500}\right) = \$149{,}690$$

8. Molecular Sieves Pricing Calculations

Professor Fabiano suggested the following simplifying assumptions for the molecular sieves design: "Size the two vessels for an 8 hour on stream time; L/D of around 3-4; use 13 X and use \$2/lb; when you cost the two vessels use a factor of 2 to include all the piping and auxiliary equipment; don't worry about the energy requirements." He also suggested that 10 lb of sieves adsorbs approximately 1 lb of water.

With these specifications, the needed molecular sieve volume is:

$$\left(1690\ \frac{lb\ water}{hr}\right)(8\ hr)\left(10\ \frac{lb\ sieves}{lb\ water}\right)\left(1.5625\ \frac{mL}{g\ sieves}\right)\left(3.53\ x\ 10^{-5}\frac{ft^3}{mL}\right)\left(453\ \frac{g}{lb}\right) = 3378\ ft^3$$

The density of molecular sieves 13X is obtained from Hengye USA (2008). Using a void space of 0.5, the vessel volume is 6756 ft^3. An L/D ratio of 3 corresponds to a diameter of 14.2 ft and a length of 42.6 ft. Use a design pressure of 10 psig. The minimum wall thickness is 0.563 in. The weight of the vessel is given from Eq. 22.59 (same equation as Appendix A.6 and A.7) to be 55299 lb. The price of an empty vertical vessel of this weight is given from Eq. 22.54 to be C_v = \$126258.27. The price for platforms and ladders, from Eq. 22.56, is C_{PL} = \$36511.77. From Eq. 22.52, using a material factor of 1.7 to account for stainless steel construction, C_P = \$251150.83. Thus,

$$C_{BM} = (\$251150.83)(2)(4.16)\left(\frac{523.6}{500}\right) = \$2{,}188{,}202.84$$

Note that the molecular sieve assembly requires two of these vessels, so the above cost must be multiplied by two.

9. Fermenter Pricing Calculation

Aden et al. (2002) suggests an agitation power consumption of 0.15 hp/1000 gal. Using this figure, each 500,000 gallon fermenter needs an agitator that can provide 75 hp. The approximate cost for this agitator is

$$C_P = (3620)(75)^{0.57} = \$42412.46$$

For a 500,000 gallon vessel, the cost is

$$C_P = 265(1.7)(500000)^{0.51} = \$363219.79$$

The total purchase cost is $405,632.25. Then,

$$C_{BM} = (\$405632.25)\left(\frac{523.6}{500}\right)(4.16) = \$1{,}767{,}075.77$$

10. Distillation Tower and Heat Exchanger Pricing Calculation

Tower:

Calculating the length L of the Tower:

$$Number\ of\ trays\ N_T = \frac{Number\ of\ stages}{Tray\ Efficiency} = \frac{15}{0.7} \cong 22\ trays$$

$$L = (N^T * Tray\ spacing) + \ sump\ space + head\ space$$

$$L = 22 * \frac{18}{12} + 4 + 3 + 3 + 3 - \frac{18}{12} = 41.5\ ft$$

As the operating P_o is 12 which is greater than 10, the internal design gauge pressure is calculated from the following equation:

$$P_d = \exp\{0.60608 + \ 0.91615[\ln(Po)] + \ 0.0015655[\ln(Po)]^2$$

$$P_d = \exp\{0.60608 + \ 0.91615[\ln(12)] + \ 0.0015655[\ln(12)]^2 = 15.24 \text{ psia}$$

Calculating the wall thickness:

$$t_P = \frac{P_d D_i}{2SE - 1.2P_d} = \frac{(15.24\,psi)(4\,ft)(12\ {}^{in}/_{ft})}{(2)(15000\,psi)(1) - (1.2)(15.24\,psi)} = 0.0287\ in$$

The minimum recommended wall thickness for sufficient rigidity is given by Seider, Seader, Lewin, and Widagdo (2009) as 0.25 in. From the table on Page 575 wall thickness ts for D_i = 4ft is:

$$t_s = \frac{5}{16} + 0.125 = 0.4375$$

From length and wall thickness, the weight W of the tower is calculate:

$$W = \pi(D_i + t_S)(L + 0.8D_i)t_S\rho_{carbon\ \ steel}$$

$$= \pi\left(4ft + (0.4375\ in)\left(0.083\,{}^{ft}/_{in}\right)\right)\big(41.5\ ft$$

$$+ (0.8)(4)\big)\ (0.4375\ in)\left(0.083\,{}^{ft}/_{in}\right)\left(490\ {}^{lb}/_{ft^3}\right) = 10121.2\ lb$$

Using the weight W of the tower to calculate the vessel cost:

$$C_V = \exp\{7.2756 + 0.18255[\ln(W)] + 0.02297[\ln(W)]^2$$

$$C_V = \exp\{7.2756 + 0.18255[\ln(10121.2)] + 0.02297[\ln(10121.6)]^2 = \$54874.23$$

Cost of platforms and ladders:

$$C_{PL} = 300.9(D_i)^{0.63316}(L)^{0.80161}$$

$$C_{PL} = 300.9(4)^{0.63316}(41.5)^{0.80161} = \$14343.86$$

Cost of trays:

$$C_T = N_T F_{NT} F_{TT} F_{TM} C_{BT}$$

NT = 22, $F_{NT} = F_{TT} = F_{TM} = 1$,

$$C_{BT} = 468\exp(0.1739D_i) = 468\exp(0.1739 * 4) = 938.8$$

$$C_T = 22 * 1 * 1 * 1 * 938.3 = \$20642.6$$

Purchase cost of tower:

$$C_P = F_M C_V + C_{PL} + C_{BT} = (1)(54874.23) + 14343.86 + 20642.6 = \$89860.7$$

Total Bare Module Cost:

$$C_{BM} = F_{BM} C_P \left(\frac{523}{500}\right) = (4.16)(89860.7)\left(\frac{523}{500}\right) = \$\mathbf{391015.72}$$

Condenser (Fixed Head Heat Exchanger):

$T_{H,in}$ = 172.7ºF $T_{H,out}$ = 100ºF $T_{C,in}$ = 90ºF $T_{C,out}$ = 120ºF

ΔT_{LM} = 25.7ºF

Heat Duty, Q = -7998642.8 Btu/hr

$$Area\ of\ heat\ exchange, A = \frac{Q}{(U)(\Delta T_{LM})} = \frac{7998642.8}{(100)(25.7)} = 3112.31\ ft^2$$

Calculating the purchase cost:

$$C_P = F_P F_M F_L C_B$$

$F_P = F_L = 1$,

For carbon steel shell and brass tubes a = 1.08 and b = 0.05 in the following formula:

$$F_M = a + (\frac{A}{100})^b = 1.08 + (\frac{3112.31}{100})^{0.05} = 2.27$$

$$C_B = \exp\{11.2927 - 0.9228[\ln(A)] + 0.09861[\ln(A)]^2$$

$$= \exp\{11.2927 - 0.9228[\ln(3112.31)] + 0.09861[\ln(3112.31)]^2 = \$28276$$

$$C_P = (1)(2.27)(1)(64187) = \$64187$$

Total Bare Module Cost:

$$C_{BM} = F_{BM} C_P \left(\frac{523}{500}\right) = (3.17)(64187)\left(\frac{523}{500}\right) = \$\mathbf{212831}$$

Reboiler (Kettle Vaporizer):

Heat Duty, Q = 8900486.18 Btu/hr

Heat Flux = 12000 Btu/hr-ft^2

$$Area\ of\ heat\ exchange, A\ =\ \frac{Q}{heat\ flux} = \frac{7998642.8}{12000} = 741.71\ ft^2$$

Calculating the purchase cost:

$$C_P = F_P F_M F_L C_B$$

For a carbon-steel/carbon-steel kettle vaporizer:

$F_M = F_P = F_L = 1$,

$$C_B = \exp\{12.2052 - 0.8709[\ln(A)] + 0.09005[\ln(A)]^2$$

$$= \exp\{12.2052 - 0.8709[\ln(741.71)] + 0.09005[\ln 741.71]^2 = \$32296$$

$$C_P = (1)(1)(1)(32296) = \$32296$$

Total Bare Module Cost:

$$C_{BM} = F_{BM} C_P \left(\frac{523}{500}\right) = (3.17)(32296)\left(\frac{523}{500}\right) = \$\mathbf{107087}$$

Reflux Accumulator (Horizontal Pressure Vessel):

From the drum sizing sheet supplied by Professor Fabiano, the vessel length is 7.42 ft and the vessel diameter is 2.5 ft. The minimum recommended wall thickness for sufficient rigidity is given by Seider, Seader, Lewin, and Widagdo (2009) as 0.25 in. Adding a 0.125 corrosion allowance give the wall thickness:

$$t_s = 0.25 + 0.125 = 0.375\ in$$

Using this value to calculate the weight of the horizontal vessel:

$$W = \pi(D_i + t_S)(L + 0.8D_i)t_S\rho_{carbon\ \ steel}$$

$$= \pi\left(2.5ft + (0.375\ in)\left(0.083\ {}^{ft}/_{in}\right)\right)\left(7.42\ ft + (0.8)(2.5)\right)(0.375\ in)\left(0.083\ {}^{ft}/_{in}\right)\left(490\ {}^{lb}/_{ft^3}\right) = 660\ lb$$

Total purchase cost C_P:

$$C_P = F_M C_V + C_{PL}$$

$$C_V = \exp\{8.9552 - 0.233[\ln(W)] + 0.04333[\ln(W)]^2$$

$$C_V = \exp\{8.9552 - 0.233[\ln(660)] + 0.04333[\ln(660)]^2 = \$10602.4$$

$$C_{PL} = 2005(D_i)^{0.20294} = 2005(2.5)^{0.20294} = \$2307.83$$

$$C_P = F_M C_V + C_{PL} = (1)(10602.4) + 2307.83 = \$13504.1$$

Total Bare Module Cost:

$$C_{BM} = F_{BM} C_P \left(\frac{523}{500}\right) = (3.05)(13504.1)\left(\frac{523}{500}\right) = \$\mathbf{41187.53}$$

Total cost of distillation column:

C = 391016 + 212831 + 107087 + 41187.53 = $752122

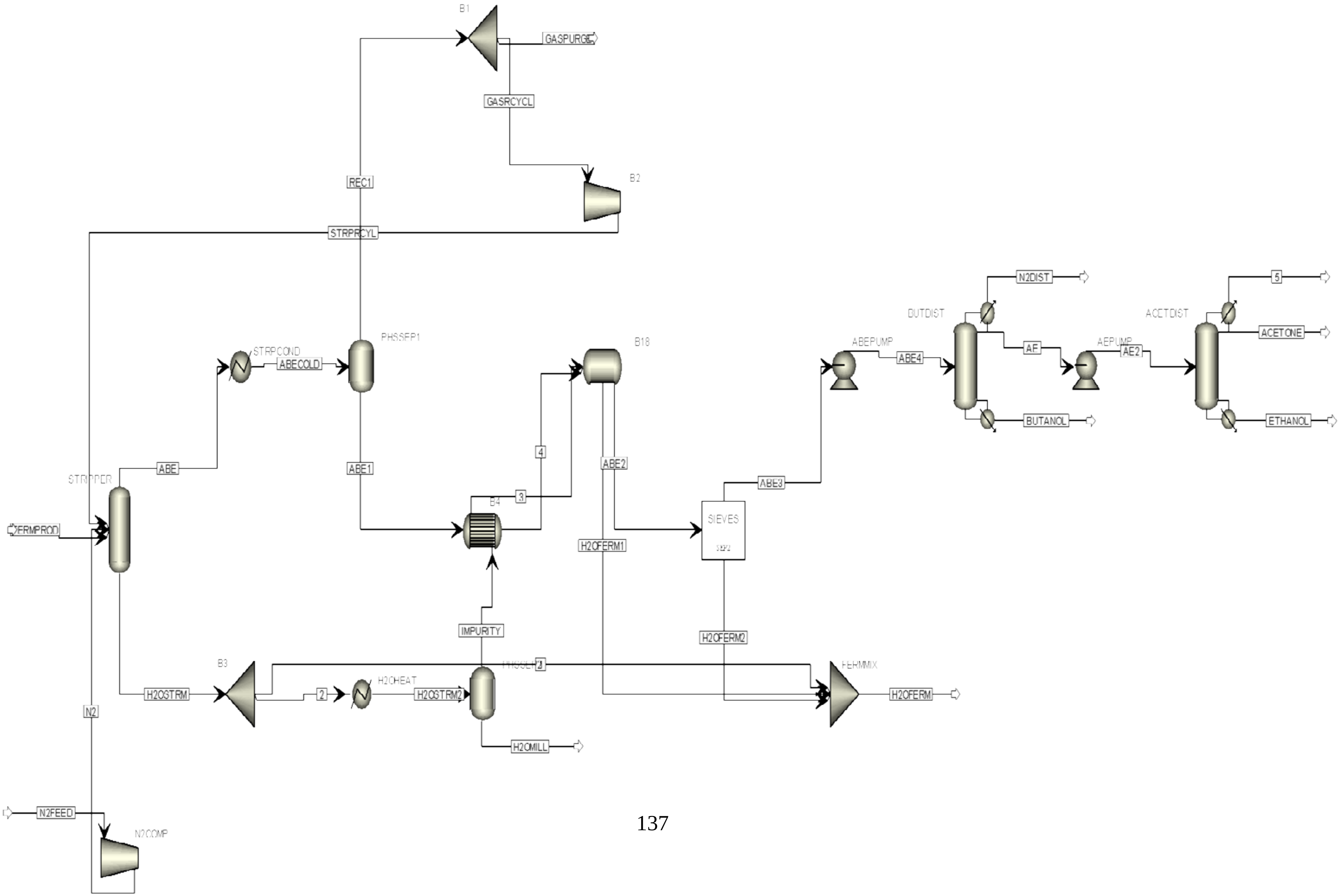
11. Dryer Pricing Calculation

For a drum dryer, the evaporation rate, from Seider, Seader, Lewin, and Widagdo (2009) ranges from 3-6 lb/hr/ft^2. The dryer must handle 967 lb DCW per hour, which is roughly 70% water (Muurahainen, 1998). This means 2256 lb water must be evaporated per hour, requiring a heat duty of approximately 2,188,320 Btu/hr of latent heat. The latent heat of vaporization of low pressure steam, from Seider, Seader, Lewin, and Widadgo (2009), is 806 btu/lb, meaning approximately 2715 lb steam per hour is needed. Using the median heat flux value in the range given above gives 4.5 lb/ hr/ft^2, meaning a heat transfer area of 501 ft^2 is required. The price of a dryer this size is:

$$C_P = (32000)(501)^{0.38} = \$339694.34$$

$$C_{BM} = (\$339694.34)\left(\frac{523.6}{500}\right)(2.06) = \$732799.51$$

Appendix B: ASPEN PLUS Simulation Results



				bh				
Stream ID		FERMPROD	N2FEED	STRPRCYL	ABE1	H2OMILL	ACETONE	BUTANOL	ETHANOL
Temperature	F	93.2	77.0	31.1	41.0	216.5	90.0	275.2	201.9
Pressure	psia	19.696	14.696	19.696	21.756	14.504	22.000	25.950	27.450
Vapor Frac		0.007	1.000	1.000	0.000	0.000	0.000	0.000	0.000
Mole Flow	lbmol/hr	68155.006	149.527	15490.313	986.681	14360.150	167.763	263.481	67.764
Mass Flow	lb/hr	1.25532E+6	4188.783	440373.543	41496.395	259512.014	9710.215	19521.934	3126.048
Volume Flow	cuft/hr	150008.618	58596.832	4.13904E+6	767.261	4531.645	199.479	452.482	70.313
Enthalpy	MMkcal/hr	-2128.809	> -0.001	-11.047	-31.374	-439.678	-4.494	-8.608	-1.972
Mass Flow	lb/hr								
WATER		1.21826E+6		1439.086	8537.564	258611.234	7.859	< 0.001	0.001
N-BUT-01		19718.145		537.978	19712.654	0.002	trace	19507.232	7.297
ETHAN-01		3289.297		833.922	3208.318	2.779	42.490	11.952	3111.614
ACETO-01		9892.142		10813.428	9790.596	< 0.001	9655.219	2.750	7.145
NITRO-01			4188.783	424031.714	231.061	trace	4.522		
CARBO-01		41.872		2716.419	16.172	< 0.001	0.115	trace	trace
HYDRO-01		0.009		0.935	trace	trace	trace	trace	
DEXTR-01		4120.440				897.999		trace	
Mass Frac									
WATER		0.970		0.003	0.206	0.997	809 PPM	9 PPB	452 PPB
N-BUT-01		0.016		0.001	0.475	6 PPB	trace	0.999	0.002
ETHAN-01		0.003		0.002	0.077	11 PPM	0.004	612 PPM	0.995
ACETO-01		0.008		0.025	0.236	2 PPB	0.994	141 PPM	0.002
NITRO-01			1.000	0.963	0.006	trace	466 PPM		
CARBO-01		33 PPM		0.006	390 PPM	trace	12 PPM	trace	trace
HYDRO-01		7 PPB		2 PPM	trace	trace	trace	trace	
DEXTR-01		0.003				0.003		trace	
Mole Flow	lbmol/hr								
WATER		67623.581		79.881	473.907	14355.105	0.436	trace	< 0.001
N-BUT-01		266.020		7.258	265.946	< 0.001	trace	263.175	0.098
ETHAN-01		71.399		18.102	69.642	0.060	0.922	0.259	67.542
ACETO-01		170.319		186.181	168.571	trace	166.240	0.047	0.123
NITRO-01			149.527	15136.703	8.248	trace	0.161		
CARBO-01		0.951		61.724	0.367	trace	0.003	trace	trace
HYDRO-01		0.004		0.464	trace	trace	trace	trace	
DEXTR-01		22.871				4.985		trace	
Mole Frac									
WATER		0.992		0.005	0.480	1.000	0.003	36 PPB	1 PPM
N-BUT-01		0.004		469 PPM	0.270	1 PPB	trace	0.999	0.001
ETHAN-01		0.001		0.001	0.071	4 PPM	0.005	985 PPM	0.997
ACETO-01		0.002		0.012	0.171	trace	0.991	180 PPM	0.002
NITRO-01			1.000	0.977	0.008	trace	962 PPM		
CARBO-01		14 PPM		0.004	372 PPM	trace	16 PPM	trace	trace
HYDRO-01		64 PPB		30 PPM	trace	trace	trace	trace	
DEXTR-01		336 PPM				347 PPM		trace	

Appendix C: Equipment Quote Sheets

Equipment from Mr. Benjamin Xu of Beijing Great Sources Technology Development Co. Ltd. (2010)

vertical axial flow pump (ZL) Add Product to Favorites

See larger image: vertical axial flow pump (ZL)

SHARE

Place of Origin:	Shanghai China (Mainland)
Brand Name:	Fashion
Model Number:	ZL
Theory:	Centrifugal Pump
Structure:	Single-stage Pump
Usage:	Water
Power:	Electric
Standard or Nonstandard:	Standard
FOB Price:	US $10,000 - 50,000
Port:	Shanghai
Payment Terms:	L/C,T/T
Minimum Order Quantity:	1 Set/Sets
Supply Ability:	5000 Set/Sets per Year
Package:	Suitable for Long Distance Ocean Transportation Packing
Delivery Time:	at least 30 days other wise there are stock

Offline Contact Now

Supplier Details

Beijing Great Sources Technology Development Co., Ltd.

[Beijing, China (Mainland)]

Gold Supplier [2nd Year]

Verified Supplier

- Business Type:
  Manufacturer, Trading Company, Buying Office, Agent, Distributor/Wholesaler
- Product/Services:
  Motor,Valve,API Valve,Butterfly Valve,Gate Valve,Ball Valve,Check ...
- Online Showroom: 736 products

other similar products from this supplier

Contact Details

View this Supplier's Website »

Add Company to Favorites

Product Details | Company Profile

Place of Origin: Shanghai China (Mainland)	Brand Name: Fashion	Model Number: ZL
Theory: Centrifugal Pump	Structure: Single-stage Pump	Usage: Water
Power: Electric	Standard or Nonstandard: Standard	material: cast iron

capacity:1020-39350m3/h
total head:2-25m
motor rating:28-1900kw
size:500-1600mm
speed:245-980r/min

ZL pump series are single-stage, vertical axial flow pump with the liquid flowing in axial direction. The pump are used to transfer clean water or other liquid fimiliar with water. The highest temperature of the liquid can be 50^0 C. Zl pumps have large capacity and low head, which is widely used in agriculture irrigation, civil water supply and drainage, water circulation in industrial power station, dock water level control and other conservancy projects.