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ABE FERMENTATION OF SUGAR IN BRAZIL

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ABE FERMENTATION OF SUGAR IN BRAZIL

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

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ABE Fermentation of Sugar Cane in Brazil

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ABE FERMENTATION OF SUGAR CANE IN BRAZIL

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Mohammad Zohaib

Advised by:
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
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 $3.36 \times 10^8$ BTU/hr and our utility inputs being only $2.14 \times 10^6$ 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.
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<th><strong>Project Charter</strong></th>
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<td><strong>Project Name</strong></td>
<td>Fermentation of Sugar Cane to Acetone, Butanol, and Ethanol</td>
</tr>
<tr>
<td><strong>Project Champions</strong></td>
<td>Mr. Bruce Vrana, Professor Leonard Fabiano, Professor Daniel Hammer</td>
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<tr>
<td><strong>Project Leaders</strong></td>
<td>Marcelo Mansur, Maclyn O’Donnell, Matthew Rehmann, Mohammad Zohaib</td>
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<tr>
<td><strong>Specific Goals</strong></td>
<td>Development of a profitable ABE fermentation process that has a net positive energy balance with a novel <em>Clostridium</em> strain</td>
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<td><strong>Project Scope</strong></td>
<td><strong>In-scope:</strong></td>
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<td>• 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</td>
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<td>• Profitable fermentation and separation to 99.5% acetone, 99.5% ethanol, and 99.5% butanol products</td>
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<td>• Net positive energy balance so that butanol and ethanol can be used as alternative fuels</td>
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<td></td>
<td>• Zero-discharge process design, so that all water is purified and recycled</td>
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<td>• Environmentally acceptable and economical disposal of large amounts of H₂ and CO₂ product</td>
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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

Customer-Value Proposition
- Lower cost solvents
- Lower cost solvent derivatives
- Improved energy efficiency
- Petrochemical conservation

Products
- Acetone
- Butanol
- Ethanol
- Fertilizer

Technical Differentiation
- Improved solvent yield
- Greater bacteriophage tolerance
- Decreased product inhibition
- Decreased stream volume
- Less energy for sterilization

Process/Manufacturing Technology
- Use of starchy feed
- Awareness of contamination issues
- In situ product recovery
- Use of concentrated substrate feed

Material Technology
- Clostridium acetobutylicum
- Clostridium saccharoacetobutylicum
- Clostridium beijerinckii
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\textsubscript{2} or H\textsubscript{2}, in a mixture that is roughly 60% CO\textsubscript{2} and 40% H\textsubscript{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\textsubscript{2} and H\textsubscript{2}. The CO\textsubscript{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\textsubscript{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.

\[
\frac{ds}{dt} = \left( \frac{ds}{dt} \right)_{\text{max}} \left( \frac{S}{5 + S} \right) \left( \frac{7}{7 + P} \right)
\]
where \((ds/dt)_{\text{max}} = 1.1 \text{ g sugar / g DCW / hr}\) in growth phase and \(0.8 \text{ g sugar / g DCW / hr}\) in production phase, \(S\) is the concentration of sugar in the reactor \((\text{g/L})\), and \(P\) is the concentration of ABE in the reactor \((\text{g/L})\).

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 CO₂ and H₂, 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
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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 N₂ 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
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<td>Pump F-24</td>
<td>1.08 hp</td>
</tr>
<tr>
<td>Centrifuge F-1</td>
<td>2 hp</td>
</tr>
<tr>
<td>Centrifuge F-2</td>
<td>2 hp</td>
</tr>
<tr>
<td>Centrifuges F-3, F-4</td>
<td>50 hp</td>
</tr>
<tr>
<td>Centrifuges F-5, F-6</td>
<td>300 hp</td>
</tr>
<tr>
<td>Centrifuges F-7 - F-12</td>
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</tr>
<tr>
<td>Centrifuge F-13</td>
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<td><strong>TOTAL</strong></td>
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### Process Unit Heat Duty 50 psig Steam Req. 150 psig Steam Req.

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<th>Heat Duty</th>
<th>50 psig Steam Req.</th>
<th>150 psig Steam Req.</th>
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<tbody>
<tr>
<td>Distillation Column 1</td>
<td>8982796.5</td>
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<td>Distillation Column 2</td>
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<td>Pasteurizer</td>
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<td><strong>TOTAL</strong></td>
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### Process Unit Required Cooling Duty Cooling Water Requirement

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<td>Distillation Column 1</td>
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<td>Distillation Column 2</td>
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<td><strong>TOTAL</strong></td>
<td>23273695.5</td>
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<thead>
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<tbody>
<tr>
<td>Condenser</td>
<td>1,930,972</td>
<td>836</td>
<td>6.8</td>
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<tr>
<td>Fermenter Heat Ex. Growth Phase 1</td>
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<td>114.4</td>
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<tr>
<td>Fermenter Heat Ex. Growth Phase 2</td>
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<td>Fermenter Heat Ex. Production Phase</td>
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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”.

<table>
<thead>
<tr>
<th>Total BTU/hr IN from Utilities</th>
<th>Product Energy Potential (BUT/hr)</th>
<th>Difference</th>
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<td>2.14E+06</td>
<td>3.36E+08</td>
<td>3.34E+08</td>
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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 °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°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°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 ft³/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 ft³/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 m³/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 ft³/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°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 ft³/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 ft³/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 ft³/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 ft³/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 ft³/hr of a 99.7% water stream from 82 °F to 217°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 ft³/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 ft³/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 ft³/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 ft³/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 ft³/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
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<th>Number Needed:</th>
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<td><strong>Bare Module Cost</strong></td>
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<tr>
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<tr>
<td>Butanol</td>
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<tr>
<td>Ethanol</td>
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<tr>
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<td>600 mL (26 containers)</td>
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Laboratory (Fermenters F-1 - F-4)
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<td>Process rate (kg sugar/day):</td>
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<td>Mass Fraction of each stream:</td>
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<td>Butanol</td>
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<td>Ethanol</td>
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<td>Water</td>
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<td>Nitrogen</td>
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<td>Sugar</td>
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<td>Carbon Dioxide</td>
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<tr>
<td>Acetone</td>
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<tr>
<td>Butanol</td>
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<tr>
<td>Ethanol</td>
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<tr>
<td>Water</td>
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<tr>
<td>Biomass</td>
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<td>Nitrogen</td>
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<td>Sugar</td>
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<td>Carbon Dioxide</td>
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<td>Hydrogen</td>
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<td><strong>Temperature (F):</strong></td>
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# Fermenters F-8 - F-16

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<tr>
<td>Butanol</td>
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<td>Hydrogen</td>
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**Centrifuge F-1**

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**Purpose:**

**Design Specs:**

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<tr>
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<td>3375</td>
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**Utilities:**

|                                |                |
|                                | 25 HP          |

**Price of Utilities:**

|                                | $5040/yr       |

**Comments and Figures:**

Price quote from US Centrifuge
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<tr>
<td>Centrifuge capacity (gpm):</td>
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</tr>
<tr>
<td>Average cells removed (kg/hr):</td>
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<tr>
<td>Material:</td>
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</tr>
<tr>
<td>Bowl Diameter (in):</td>
<td>30</td>
</tr>
<tr>
<td>Maximum G Force:</td>
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<td><strong>Utilities:</strong></td>
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<td>150 HP</td>
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<td><strong>Price of Utilities:</strong></td>
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<td>Maximum G Force:</td>
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<tr>
<td><strong>Mass Fraction of each stream:</strong></td>
<td></td>
</tr>
<tr>
<td>Acetone</td>
<td>0.000</td>
</tr>
<tr>
<td>Butanol</td>
<td>0.001</td>
</tr>
<tr>
<td>Ethanol</td>
<td>0.000</td>
</tr>
<tr>
<td>Water</td>
<td>0.929</td>
</tr>
<tr>
<td>Biomass</td>
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</tr>
<tr>
<td>Nitrogen</td>
<td>0.000</td>
</tr>
<tr>
<td>Sugar</td>
<td>0.069</td>
</tr>
<tr>
<td>Carbon Dioxide</td>
<td>0.000</td>
</tr>
<tr>
<td>Hydrogen</td>
<td>0.000</td>
</tr>
<tr>
<td><strong>Temperature:</strong></td>
<td></td>
</tr>
<tr>
<td></td>
<td>93.2°F</td>
</tr>
<tr>
<td><strong>Design Specs:</strong></td>
<td></td>
</tr>
<tr>
<td>Pump Efficiency:</td>
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<td>Volumetric Flow Rate (ft^3/hr):</td>
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<td>Type:</td>
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</tr>
<tr>
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</tr>
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<td>Stainless Steel</td>
</tr>
<tr>
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<td>5 psi</td>
</tr>
<tr>
<td><strong>Utilities:</strong></td>
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### Pump F-2

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**Materials in the Column**

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<th>Outlet</th>
</tr>
</thead>
<tbody>
<tr>
<td>Flow rate (ft(^3)/hr)</td>
<td>27.8</td>
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</table>

**Mass Fraction of each stream:**

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<thead>
<tr>
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<th>Inlet</th>
<th>Outlet</th>
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</thead>
<tbody>
<tr>
<td>Acetone</td>
<td>0.001</td>
<td>0.001</td>
</tr>
<tr>
<td>Butanol</td>
<td>0.001</td>
<td>0.001</td>
</tr>
<tr>
<td>Ethanol</td>
<td>0.000</td>
<td>0.000</td>
</tr>
<tr>
<td>Water</td>
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<td>0.926</td>
</tr>
<tr>
<td>Biomass</td>
<td>0.001</td>
<td>0.001</td>
</tr>
<tr>
<td>Nitrogen</td>
<td>0.000</td>
<td>0.000</td>
</tr>
<tr>
<td>Sugar</td>
<td>0.069</td>
<td>0.069</td>
</tr>
<tr>
<td>Carbon Dioxide</td>
<td>0.000</td>
<td>0.000</td>
</tr>
<tr>
<td>Hydrogen</td>
<td>0.000</td>
<td>0.000</td>
</tr>
<tr>
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**Design Specs:**

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<td>Type</td>
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</tr>
<tr>
<td>Motor Material</td>
<td>Stainless Steel</td>
</tr>
<tr>
<td>Pump Material</td>
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**Utilities:** negligible

**Price of Utilities:**

**Comments and Figures:**

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<td>0.007</td>
</tr>
<tr>
<td>Butanol</td>
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<td>0.013</td>
</tr>
<tr>
<td>Ethanol</td>
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<td>0.002</td>
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<tr>
<td>Water</td>
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<tr>
<td>Nitrogen</td>
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<td>0.039</td>
</tr>
<tr>
<td>Sugar</td>
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<td>0.001</td>
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<tr>
<td>Carbon Dioxide</td>
<td>0.006</td>
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<tr>
<td>Hydrogen</td>
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<td>0.003</td>
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<td>Volumetric Flow Rate (ft3/hr):</td>
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<td>Motor Material:</td>
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<td></td>
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<td>Pump Material:</td>
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<tr>
<td>Butanol</td>
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<tr>
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</tr>
<tr>
<td>Sugar</td>
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<td>0.003</td>
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<tr>
<td>Carbon Dioxide</td>
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<tr>
<td>Hydrogen</td>
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<tr>
<td>Volumetric Flow Rate (ft3/hr):</td>
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<tr>
<td>Head (ft):</td>
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<td></td>
</tr>
<tr>
<td>Type:</td>
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<td></td>
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<tr>
<td>Motor Material:</td>
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<td></td>
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<td>Pressure Change:</td>
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### Materials in the Column

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#### Mass Fraction of each stream:

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<th>Outlet</th>
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</thead>
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<td>0.008</td>
</tr>
<tr>
<td>Butanol</td>
<td>0.015</td>
<td>0.015</td>
</tr>
<tr>
<td>Ethanol</td>
<td>0.002</td>
<td>0.002</td>
</tr>
<tr>
<td>Water</td>
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<tr>
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<td>0.000</td>
</tr>
<tr>
<td>Sugar</td>
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<td>0.003</td>
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<tr>
<td>Carbon Dioxide</td>
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<td>0.040</td>
</tr>
<tr>
<td>Hydrogen</td>
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<td>0.001</td>
</tr>
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#### Temperature:

| Temperature: | 93.2°F | 93.2°F |

#### Design Specs:

| Pump Efficiency: | 0.7 |
| Volumetric Flow Rate (ft^3/hr): | 2783 |
| Head (ft): | 12 |
| Type: | Centrifugal Pump |
| Motor Material: | Stainless Steel |
| Pump Material: | Stainless Steel |
| Pressure Change: | 5 psi |

#### Utilities:

| Utilities: | 1.51 HP |
| Price of Utilities: | $304/yr |

#### Comments and Figures:

Pumps F-6 - F-14
### Pumps F-15 - F-23

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<th>Outlet</th>
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<th><strong>Outlet</strong></th>
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</thead>
<tbody>
<tr>
<td>Acetone</td>
<td>0.008</td>
<td>0.008</td>
</tr>
<tr>
<td>Butanol</td>
<td>0.015</td>
<td>0.015</td>
</tr>
<tr>
<td>Ethanol</td>
<td>0.002</td>
<td>0.002</td>
</tr>
<tr>
<td>Water</td>
<td>0.930</td>
<td>0.930</td>
</tr>
<tr>
<td>Biomass</td>
<td>0.010</td>
<td>0.010</td>
</tr>
<tr>
<td>Nitrogen</td>
<td>0.000</td>
<td>0.000</td>
</tr>
<tr>
<td>Sugar</td>
<td>0.003</td>
<td>0.003</td>
</tr>
<tr>
<td>Carbon Dioxide</td>
<td>0.040</td>
<td>0.040</td>
</tr>
<tr>
<td>Hydrogen</td>
<td>0.001</td>
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| **Temperature:** | 93.2°F | 93.2°F |

<table>
<thead>
<tr>
<th><strong>Design Specs:</strong></th>
<th><strong>Pump Efficiency:</strong></th>
<th>0.7</th>
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</thead>
<tbody>
<tr>
<td>Volumetric Flow Rate (ft^3/hr):</td>
<td>2783</td>
<td></td>
</tr>
<tr>
<td>Head (ft):</td>
<td>12</td>
<td></td>
</tr>
<tr>
<td>Type:</td>
<td>Centrifugal Pump</td>
<td></td>
</tr>
<tr>
<td>Motor Material:</td>
<td>Stainless Steel</td>
<td></td>
</tr>
<tr>
<td>Pump Material:</td>
<td>Stainless Steel</td>
<td></td>
</tr>
<tr>
<td>Pressure Change:</td>
<td>5 psi</td>
<td></td>
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<table>
<thead>
<tr>
<th><strong>Utilities:</strong></th>
<th>1.51 HP</th>
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<td><strong>Price of Utilities:</strong></td>
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**Comments and Figures:**
## Pump F-24

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<td></td>
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<td><strong>Bare Module Cost</strong></td>
<td>$3,123.35</td>
<td></td>
</tr>
<tr>
<td><strong>Purpose:</strong></td>
<td>Pump solid suspension from centrifuge bank to dryer</td>
<td></td>
</tr>
<tr>
<td><strong>Materials in the Column:</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Flow rate (ft^3/hr):</td>
<td>53</td>
<td>53</td>
</tr>
<tr>
<td><strong>Mass Fraction of each stream:</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Acetone</td>
<td>0.000</td>
<td>0.000</td>
</tr>
<tr>
<td>Butanol</td>
<td>0.000</td>
<td>0.000</td>
</tr>
<tr>
<td>Ethanol</td>
<td>0.000</td>
<td>0.000</td>
</tr>
<tr>
<td>Water</td>
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<td>0.700</td>
</tr>
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<td>Biomass</td>
<td>0.300</td>
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<tr>
<td>Nitrogen</td>
<td>0.000</td>
<td>0.000</td>
</tr>
<tr>
<td>Sugar</td>
<td>0.000</td>
<td>0.000</td>
</tr>
<tr>
<td>Carbon Dioxide</td>
<td>0.000</td>
<td>0.000</td>
</tr>
<tr>
<td>Hydrogen</td>
<td>0.000</td>
<td>0.000</td>
</tr>
<tr>
<td><strong>Temperature:</strong></td>
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<td>93.2°F</td>
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<tr>
<td><strong>Design Specs:</strong></td>
<td>Pump Efficiency: 0.7</td>
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<tr>
<td></td>
<td>Volumetric Flow Rate (ft3/hr): 53</td>
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</tr>
<tr>
<td></td>
<td>Head (ft): 12</td>
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</tr>
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<td></td>
<td>Type: Centrifugal Pump</td>
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<tr>
<td></td>
<td>Motor Material: Stainless Steel</td>
<td></td>
</tr>
<tr>
<td></td>
<td>Pump Material: Stainless Steel</td>
<td></td>
</tr>
<tr>
<td></td>
<td>Pressure Change: 5 psi</td>
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<tr>
<td><strong>Utilities:</strong></td>
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## Distillation Column 1

<table>
<thead>
<tr>
<th>Bare Module Cost</th>
<th>$752,122</th>
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**Purpose:** To distill Butanol in the bottoms

### Materials in the Column:

<table>
<thead>
<tr>
<th></th>
<th>Feed</th>
<th>Bottoms</th>
<th>Overhead</th>
<th>N2 Purge</th>
</tr>
</thead>
<tbody>
<tr>
<td>Flow rate (ft^3/hr)</td>
<td>693</td>
<td>452</td>
<td>265</td>
<td>706</td>
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### Mass Fraction of each stream:

<table>
<thead>
<tr>
<th></th>
<th>Feed</th>
<th>Bottoms</th>
<th>Overhead</th>
<th>N2 Purge</th>
</tr>
</thead>
<tbody>
<tr>
<td>Acetone</td>
<td>0.229</td>
<td>146 PPM</td>
<td>0.753</td>
<td>0.385</td>
</tr>
<tr>
<td>Butanol</td>
<td>0.602</td>
<td>0.999</td>
<td>603 PPM</td>
<td>18 PPM</td>
</tr>
<tr>
<td>Ethanol</td>
<td>0.098</td>
<td>629 PPM</td>
<td>0.246</td>
<td>0.047</td>
</tr>
<tr>
<td>Water</td>
<td>243 PPM</td>
<td>612 PPM</td>
<td>471 PPM</td>
<td></td>
</tr>
<tr>
<td>Nitrogen</td>
<td>0.002</td>
<td>0</td>
<td>381 PPM</td>
<td>0.567</td>
</tr>
<tr>
<td>Sugar</td>
<td>0</td>
<td>0</td>
<td>0</td>
<td>0</td>
</tr>
<tr>
<td>Carbon Dioxide</td>
<td>5 PPM</td>
<td>0</td>
<td>9 PPM</td>
<td>555 PPM</td>
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<tr>
<td>Hydrogen</td>
<td>0</td>
<td>0</td>
<td>0</td>
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### Temperature:

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<tr>
<th></th>
<th>169°F</th>
<th>275°F</th>
<th>100°F</th>
<th>100°F</th>
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### Design Specs:

<table>
<thead>
<tr>
<th></th>
<th>Theoretical Trays:</th>
<th>15</th>
<th>Molar Reflux Ratio:</th>
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<tbody>
<tr>
<td>Real Trays:</td>
<td>22</td>
<td></td>
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<tr>
<td>Tray Efficiency:</td>
<td>0.7</td>
<td></td>
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<td></td>
</tr>
<tr>
<td>Tray Type:</td>
<td>Koch Flexitray</td>
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</tr>
<tr>
<td>Functional Height (ft):</td>
<td>41.5</td>
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</tr>
<tr>
<td>Inside Diameter (ft):</td>
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<tr>
<td>Pressure:</td>
<td>15.24</td>
<td></td>
<td></td>
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<tr>
<td>Feed Stage:</td>
<td>8</td>
<td></td>
<td></td>
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<tr>
<td>Material:</td>
<td>Carbon Steel</td>
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<td></td>
<td></td>
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<tr>
<td>Number of Man Holes:</td>
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<td></td>
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### Condenser

<table>
<thead>
<tr>
<th></th>
<th>Temperature (°F):</th>
<th>100</th>
</tr>
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<tbody>
<tr>
<td>Reflux Ratio:</td>
<td>1.1</td>
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</tr>
<tr>
<td>Overall Heat Transfer Coefficient (BTU/(hr °F ft^2)):</td>
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<tr>
<td>Area (ft^2):</td>
<td>3112.3</td>
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<tr>
<td>Material:</td>
<td>Carbon Steel Shell, Brass Tubes</td>
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### Reboiler

<table>
<thead>
<tr>
<th></th>
<th>Temperature (°F):</th>
<th>275.2</th>
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<tbody>
<tr>
<td>Area (ft^2):</td>
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<tr>
<td>Heat Flux (BTU/hr-ft^2):</td>
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<tr>
<td>Material:</td>
<td>Carbon Steel Shell and Tube, Kettle Vaporizer</td>
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### Reflux Accumulator

<table>
<thead>
<tr>
<th></th>
<th>Reflux Ratio:</th>
<th>1.1</th>
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</thead>
<tbody>
<tr>
<td>Volume (ft³/hr):</td>
<td>557</td>
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<tr>
<td>Diameter (ft):</td>
<td>2.5</td>
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</tr>
<tr>
<td>Length (ft):</td>
<td>7.4</td>
<td></td>
</tr>
<tr>
<td>RA Material:</td>
<td>Carbon Steel</td>
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### Utilities:

<table>
<thead>
<tr>
<th></th>
<th>Cooling water: 266621.4 lb/hr</th>
<th>Steam @ 150 psig: 10386 lb/hr</th>
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</thead>
</table>

### Price of Utilities:

<table>
<thead>
<tr>
<th></th>
<th>Cooling water: $12906 / yr</th>
<th>Steam @ 150 psig: $268,009 / yr</th>
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### Controls:

### Comments and Figures:
### Distillation Column 2

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<th>Bare Module Cost</th>
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<td><strong>Purpose:</strong></td>
<td>To distill Acetone in the overhead and Ethanol in the bottoms</td>
</tr>
<tr>
<td><strong>Materials in the Column:</strong></td>
<td>Feed</td>
</tr>
<tr>
<td>Flow rate (ft^3/hr)</td>
<td>265</td>
</tr>
<tr>
<td><strong>Mass Fraction of each stream:</strong></td>
<td></td>
</tr>
<tr>
<td>Acetone</td>
<td>0.753</td>
</tr>
<tr>
<td>Butanol</td>
<td>603 PPM</td>
</tr>
<tr>
<td>Ethanol</td>
<td>0.246</td>
</tr>
<tr>
<td>Water</td>
<td>612 PPM</td>
</tr>
<tr>
<td>Nitrogen</td>
<td>381 PPM</td>
</tr>
<tr>
<td>Sugar</td>
<td>0</td>
</tr>
<tr>
<td>Carbon Dioxide</td>
<td>9 PPM</td>
</tr>
<tr>
<td>Hydrogen</td>
<td>0</td>
</tr>
<tr>
<td><strong>Temperature:</strong></td>
<td>100.1°F</td>
</tr>
<tr>
<td><strong>Design Specs:</strong></td>
<td>Theoretical Trays: 25</td>
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<tr>
<td></td>
<td>Real Trays: 36</td>
</tr>
<tr>
<td></td>
<td>Tray Efficiency: 0.7</td>
</tr>
<tr>
<td></td>
<td>Tray Type: Koch Flexitray</td>
</tr>
<tr>
<td></td>
<td>Functional Height (ft): 64</td>
</tr>
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<td></td>
<td>Inside Diameter (ft): 5.31</td>
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<tr>
<td></td>
<td>Pressure: 18.04</td>
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<tr>
<td></td>
<td>Feed Stage: 13</td>
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<tr>
<td></td>
<td>Material: Carbon Steel</td>
</tr>
<tr>
<td></td>
<td>Number of Man Holes: 3</td>
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<tr>
<td><strong>Condenser</strong></td>
<td>Temperature (°F): 100</td>
</tr>
<tr>
<td></td>
<td>Reflux Ratio: 5</td>
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<td></td>
<td>Overall Heat Transfer Coefficient (BTU/(hr °F ft^2)): 100</td>
</tr>
<tr>
<td></td>
<td>Area (ft^2): 6659.4</td>
</tr>
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<td></td>
<td>Material: Carbon Steel Shell, Brass Tubes</td>
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<td><strong>Reboiler</strong></td>
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<td>Area (ft^2): 1253</td>
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<td></td>
<td>Heat Flux (BTU/hr-ft^2): 12000</td>
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<td>Material: Carbon Shell and Tube, Kettle Vaporizer</td>
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<td><strong>Reflux Accumulator</strong></td>
<td>Reflux Ratio: 5</td>
</tr>
<tr>
<td></td>
<td>Volume (ft^3/hr): 1207</td>
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<td></td>
<td>Diameter (ft): 2.26</td>
</tr>
<tr>
<td></td>
<td>Length (ft): 6.76</td>
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<tr>
<td></td>
<td>RA Material: Carbon Steel</td>
</tr>
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<td><strong>Utilities:</strong></td>
<td>Cooling Water: 493962 lb/hr</td>
</tr>
<tr>
<td><strong>Price of Utilities:</strong></td>
<td>Cooling Water: $23909.40 / yr</td>
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<td><strong>Comments and Figures:</strong></td>
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<tr>
<td>Pump S-1</td>
<td></td>
</tr>
<tr>
<td>----------</td>
<td></td>
</tr>
<tr>
<td><strong>Number needed:</strong></td>
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</tr>
<tr>
<td><strong>Bare Module Cost</strong></td>
<td>$165,000</td>
</tr>
<tr>
<td><strong>Purpose:</strong></td>
<td>Pump fluid from Holding Tank to Gas Stripper</td>
</tr>
<tr>
<td><strong>Materials in the Column:</strong></td>
<td></td>
</tr>
<tr>
<td><strong>Inlet</strong></td>
<td><strong>Outlet</strong></td>
</tr>
<tr>
<td><strong>Flow rate (ft^3/hr):</strong></td>
<td>150060</td>
</tr>
<tr>
<td><strong>Mass Fraction of each stream:</strong></td>
<td></td>
</tr>
<tr>
<td>Acetone</td>
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</tr>
<tr>
<td>Butanol</td>
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<td>Ethanol</td>
<td>0.003</td>
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<td>Water</td>
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<td>Nitrogen</td>
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<td>Sugar</td>
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<tr>
<td>Carbon Dioxide</td>
<td>33 PPM</td>
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<td><strong>Design Specs:</strong></td>
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</tr>
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<td>Pump Efficiency:</td>
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<td>Volumetric Flow Rate (ft^3/hr):</td>
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<td>Head (ft):</td>
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<td>Motor Material:</td>
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<tr>
<td>Pump Material:</td>
<td>Stainless Steel</td>
</tr>
<tr>
<td>Pressure Change:</td>
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<tr>
<td><strong>Utilities:</strong></td>
<td>127.67 HP</td>
</tr>
<tr>
<td><strong>Price of Utilities:</strong></td>
<td>$25738/yr</td>
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<tr>
<td><strong>Comments and Figures:</strong></td>
<td>Price quote obtained from Beijing Great Sources Technology Development Co, Ltd.</td>
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## Pump S-2

<table>
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<th>Pump ID</th>
<th>Type: Centrifugal Pump</th>
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<tbody>
<tr>
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<td>Number needed: 1</td>
</tr>
<tr>
<td>Bare Module Cost</td>
<td>$165,000</td>
</tr>
<tr>
<td>Purpose:</td>
<td>To pump water out of gas stripper bottom</td>
</tr>
</tbody>
</table>

### Materials in the Column:

<table>
<thead>
<tr>
<th></th>
<th>Inlet</th>
<th>Outlet</th>
</tr>
</thead>
<tbody>
<tr>
<td>Flow rate (ft$^3$/hr)</td>
<td>67173</td>
<td>67173</td>
</tr>
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</table>

### Mass Fraction of each stream:

<table>
<thead>
<tr>
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<th>Inlet</th>
<th>Outlet</th>
</tr>
</thead>
<tbody>
<tr>
<td>Acetone</td>
<td>509 PPB</td>
<td>509 PPB</td>
</tr>
<tr>
<td>Butanol</td>
<td>378 PPB</td>
<td>378 PPB</td>
</tr>
<tr>
<td>Ethanol</td>
<td>60 PPM</td>
<td>60 PPM</td>
</tr>
<tr>
<td>Water</td>
<td>0.997</td>
<td>0.997</td>
</tr>
<tr>
<td>Nitrogen</td>
<td>125 PPB</td>
<td>125 PPB</td>
</tr>
<tr>
<td>Sugar</td>
<td>0.003</td>
<td>0.003</td>
</tr>
<tr>
<td>Carbon Dioxide</td>
<td>125 PPB</td>
<td>125 PPB</td>
</tr>
<tr>
<td>Hydrogen</td>
<td>259 PPB</td>
<td>259 PPB</td>
</tr>
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</table>

### Temperature:

| Design Specs: | Pump Efficiency: 0.7 |
| Volumetric Flow Rate (ft$^3$/hr): | 67173 |
| Head (ft): | 12 |
| Type: | Centrifugal Pump |
| Motor Material: | Stainless Steel |
| Pump Material: | Stainless Steel |
| Pressure Change: | 5 psi |

### Utilities:

| Utilities: | 2042 HP |
| Price of Utilities: | $411667/yr |

### Comments and Figures:

Price quote obtained from Beijing Great Sources Technology Development Co, Ltd.
<table>
<thead>
<tr>
<th>Pump S-3</th>
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<tr>
<td><strong>Bare Module Cost</strong></td>
</tr>
<tr>
<td><strong>Purpose:</strong></td>
</tr>
<tr>
<td><strong>Materials in the Column:</strong></td>
</tr>
<tr>
<td>Flow rate (ft^3/hr)</td>
</tr>
<tr>
<td><strong>Mass Fraction of each stream:</strong></td>
</tr>
<tr>
<td>Acetone</td>
</tr>
<tr>
<td>Butanol</td>
</tr>
<tr>
<td>Ethanol</td>
</tr>
<tr>
<td>Water</td>
</tr>
<tr>
<td>Nitrogen</td>
</tr>
<tr>
<td>Sugar</td>
</tr>
<tr>
<td>Carbon Dioxide</td>
</tr>
<tr>
<td>Hydrogen</td>
</tr>
<tr>
<td><strong>Temperature:</strong></td>
</tr>
<tr>
<td><strong>Design Specs:</strong></td>
</tr>
<tr>
<td>Volumetric Flow Rate (ft^3/hr):</td>
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<tr>
<td>Head (ft):</td>
</tr>
<tr>
<td><strong>Type:</strong></td>
</tr>
<tr>
<td>Motor Material:</td>
</tr>
<tr>
<td>Pump Material:</td>
</tr>
<tr>
<td>Pressure Change:</td>
</tr>
<tr>
<td><strong>Utilities:</strong></td>
</tr>
<tr>
<td>Price of Utilities:</td>
</tr>
<tr>
<td><strong>Comments and Figures:</strong></td>
</tr>
<tr>
<td>Pump S-4</td>
</tr>
<tr>
<td>----------</td>
</tr>
<tr>
<td><strong>Pump ID</strong></td>
</tr>
<tr>
<td>Number needed:</td>
</tr>
<tr>
<td><strong>Bare Module Cost</strong></td>
</tr>
<tr>
<td><strong>Purpose:</strong></td>
</tr>
<tr>
<td><strong>Materials in the Column:</strong></td>
</tr>
<tr>
<td>Flow rate (ft^3/hr)</td>
</tr>
<tr>
<td><strong>Mass Fraction of each stream:</strong></td>
</tr>
<tr>
<td>Acetone</td>
</tr>
<tr>
<td>Butanol</td>
</tr>
<tr>
<td>Ethanol</td>
</tr>
<tr>
<td>Water</td>
</tr>
<tr>
<td>Nitrogen</td>
</tr>
<tr>
<td>Sugar</td>
</tr>
<tr>
<td>Carbon Dioxide</td>
</tr>
<tr>
<td>Hydrogen</td>
</tr>
<tr>
<td><strong>Temperature:</strong></td>
</tr>
<tr>
<td><strong>Design Specs:</strong></td>
</tr>
<tr>
<td>Volumetric Flow Rate (ft^3/hr):</td>
</tr>
<tr>
<td>Head (ft):</td>
</tr>
<tr>
<td>Type:</td>
</tr>
<tr>
<td>Motor Material:</td>
</tr>
<tr>
<td>Pump Material:</td>
</tr>
<tr>
<td>Pressure Change:</td>
</tr>
<tr>
<td><strong>Utilities:</strong></td>
</tr>
<tr>
<td><strong>Price of Utilities:</strong></td>
</tr>
<tr>
<td><strong>Comments and Figures:</strong></td>
</tr>
<tr>
<td>Pump ID</td>
</tr>
<tr>
<td>---------</td>
</tr>
<tr>
<td>Bare Module Cost</td>
</tr>
</tbody>
</table>

**Purpose:** Pump from Molecular Sieves to Distillation Tower S-1

**Materials in the Column:**

<table>
<thead>
<tr>
<th></th>
<th>Inlet</th>
<th>Outlet</th>
</tr>
</thead>
<tbody>
<tr>
<td>Flow rate (ft^3/hr)</td>
<td>698</td>
<td>698</td>
</tr>
</tbody>
</table>

**Mass Fraction of each stream:**

<table>
<thead>
<tr>
<th></th>
<th>Inlet</th>
<th>Outlet</th>
</tr>
</thead>
<tbody>
<tr>
<td>Acetone</td>
<td>0.299</td>
<td>0.299</td>
</tr>
<tr>
<td>Butanol</td>
<td>0.602</td>
<td>0.602</td>
</tr>
<tr>
<td>Ethanol</td>
<td>0.097</td>
<td>0.097</td>
</tr>
<tr>
<td>Water</td>
<td>521 PPM</td>
<td>521 PPM</td>
</tr>
<tr>
<td>Nitrogen</td>
<td>0.002</td>
<td>0.002</td>
</tr>
<tr>
<td>Sugar</td>
<td>0</td>
<td>0</td>
</tr>
<tr>
<td>Carbon Dioxide</td>
<td>5 PPM</td>
<td>5 PPM</td>
</tr>
<tr>
<td>Hydrogen</td>
<td>0</td>
<td>0</td>
</tr>
</tbody>
</table>

**Temperature:** 176°F

**Design Specs:**

<p>| | |</p>
<table>
<thead>
<tr>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>Pump Efficiency</td>
<td>0.7</td>
</tr>
<tr>
<td>Volumetric Flow Rate (ft^3/hr)</td>
<td>698</td>
</tr>
<tr>
<td>Head (ft)</td>
<td>46</td>
</tr>
<tr>
<td>Type:</td>
<td>Centrifugal Pump</td>
</tr>
<tr>
<td>Motor Material:</td>
<td>Carbon Steel</td>
</tr>
<tr>
<td>Pump Material:</td>
<td>Carbon Steel</td>
</tr>
<tr>
<td>Pressure Change:</td>
<td>15 psi</td>
</tr>
</tbody>
</table>

**Utilities:** 1.55 HP

**Price of Utilities:** $312/yr

**Comments and Figures:**

Pumps S-5, S-6
# Pump S-7

<table>
<thead>
<tr>
<th>Pump ID</th>
<th>Type: Centrifugal Pump</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Number needed: 1</td>
</tr>
<tr>
<td>Bare Module Cost</td>
<td>$46,261.00</td>
</tr>
<tr>
<td>Purpose:</td>
<td>Pump from Distillation Column S-1 to Dist. Col. S-2</td>
</tr>
</tbody>
</table>

### Materials in the Column:

<table>
<thead>
<tr>
<th>Fraction</th>
<th>Inlet</th>
<th>Outlet</th>
</tr>
</thead>
<tbody>
<tr>
<td>Flow rate (ft^3/hr)</td>
<td>260</td>
<td>260</td>
</tr>
</tbody>
</table>

### Mass Fraction of each stream:

<table>
<thead>
<tr>
<th>Substance</th>
<th>Inlet</th>
<th>Outlet</th>
</tr>
</thead>
<tbody>
<tr>
<td>Acetone</td>
<td>0.753</td>
<td>0.753</td>
</tr>
<tr>
<td>Butanol</td>
<td>445 PPM</td>
<td>445 PPM</td>
</tr>
<tr>
<td>Ethanol</td>
<td>0.245</td>
<td>0.245</td>
</tr>
<tr>
<td>Water</td>
<td>0.001</td>
<td>0.001</td>
</tr>
<tr>
<td>Nitrogen</td>
<td>425 PPM</td>
<td>425 PPM</td>
</tr>
<tr>
<td>Sugar</td>
<td>0</td>
<td>0</td>
</tr>
<tr>
<td>Carbon Dioxide</td>
<td>10 PPM</td>
<td>10 PPM</td>
</tr>
<tr>
<td>Hydrogen</td>
<td>0</td>
<td>0</td>
</tr>
<tr>
<td>Temperature</td>
<td>75°F</td>
<td>75°F</td>
</tr>
</tbody>
</table>

### Design Specs:

<table>
<thead>
<tr>
<th>Specification</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>Pump Efficiency:</td>
<td>0.7</td>
</tr>
<tr>
<td>Volumetric Flow Rate (ft^3/hr):</td>
<td>260</td>
</tr>
<tr>
<td>Head (ft):</td>
<td>67</td>
</tr>
<tr>
<td>Type:</td>
<td>Centrifugal Pump</td>
</tr>
<tr>
<td>Motor Material:</td>
<td>Carbon Steel</td>
</tr>
<tr>
<td>Pump Material:</td>
<td>Carbon Steel</td>
</tr>
<tr>
<td>Pressure Change:</td>
<td>12 psi</td>
</tr>
</tbody>
</table>

### Utilities:

- 12.1 HP
- Price of Utilities: $2439/yr

### Comments and Figures:
**Stripper**

<table>
<thead>
<tr>
<th>Purpose:</th>
<th>To strip the organic solvents from the water using nitrogen gas</th>
</tr>
</thead>
</table>

### Materials in the Column:

<table>
<thead>
<tr>
<th>Feed Stage</th>
<th>Feed 1</th>
<th>Feed 2</th>
<th>Feed 3</th>
<th>Bottoms</th>
<th>Overhead</th>
</tr>
</thead>
<tbody>
<tr>
<td>Flow rate (ft^3/hr)</td>
<td>49,029</td>
<td>4,139,050</td>
<td>150,060</td>
<td>19,544</td>
<td>5.89E+06</td>
</tr>
<tr>
<td>Feed Stage</td>
<td>5</td>
<td>1</td>
<td>1</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Mass Fraction of each stream:</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Acetone</td>
<td>0</td>
<td>0.025</td>
<td>0.008</td>
<td>503 PPB</td>
<td>0.043</td>
</tr>
<tr>
<td>Butanol</td>
<td>0</td>
<td>0.001</td>
<td>0.016</td>
<td>373 PPB</td>
<td>0.042</td>
</tr>
<tr>
<td>Ethanol</td>
<td>0</td>
<td>0.002</td>
<td>0.003</td>
<td>60 PPM</td>
<td>0.008</td>
</tr>
<tr>
<td>Water</td>
<td>0</td>
<td>0.003</td>
<td>0.974</td>
<td>1</td>
<td>0.021</td>
</tr>
<tr>
<td>Nitrogen</td>
<td>1</td>
<td>0.963</td>
<td>0</td>
<td>124 PPB</td>
<td>0.881</td>
</tr>
<tr>
<td>Sugar</td>
<td>0</td>
<td>0</td>
<td>0</td>
<td>0</td>
<td>0</td>
</tr>
<tr>
<td>Carbon Dioxide</td>
<td>0</td>
<td>0.006</td>
<td>33 PPM</td>
<td>258 PPB</td>
<td>0.006</td>
</tr>
<tr>
<td>Hydrogen</td>
<td>0</td>
<td>2 PPM</td>
<td>7 PPB</td>
<td>0</td>
<td>2 PPM</td>
</tr>
<tr>
<td>Temperature:</td>
<td>142 °F</td>
<td>31 °F</td>
<td>93°F</td>
<td>82°F</td>
<td>31.9°F</td>
</tr>
</tbody>
</table>

### Design Specs:

<table>
<thead>
<tr>
<th>Theoretical Trays:</th>
<th>5</th>
<th>Tray Spacing (ft):</th>
<th>2</th>
</tr>
</thead>
<tbody>
<tr>
<td>Real Trays:</td>
<td>8</td>
<td>Headspace (ft):</td>
<td>3</td>
</tr>
<tr>
<td>Tray Efficiency:</td>
<td>0.7</td>
<td>Sump Space (ft):</td>
<td>4</td>
</tr>
<tr>
<td>Tray Type:</td>
<td>Sieve</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Functional Height (ft):</td>
<td>23</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Inside Diameter (ft):</td>
<td>18.17</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Pressure:</td>
<td>16.7</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Material:</td>
<td>Stainless Steel 304</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Number of Man Holes:</td>
<td>2</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

### Utilities:

| Utilities: | 0 |

### Price of Utilities:

| No utilities |

### Controls:

<p>| Comments and Figures: | |</p>
<table>
<thead>
<tr>
<th><strong>Condenser</strong></th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Number Needed:</strong></td>
</tr>
<tr>
<td><strong>Bare Module Cost</strong></td>
</tr>
<tr>
<td><strong>Purpose:</strong></td>
</tr>
<tr>
<td><strong>Materials in the Column:</strong></td>
</tr>
<tr>
<td><strong>Flow rate (ft³/hr)</strong></td>
</tr>
<tr>
<td><strong>Mass Fraction of each stream:</strong></td>
</tr>
<tr>
<td>Acetone</td>
</tr>
<tr>
<td>Butanol</td>
</tr>
<tr>
<td>Ethanol</td>
</tr>
<tr>
<td>Water</td>
</tr>
<tr>
<td>Nitrogen</td>
</tr>
<tr>
<td>Sugar</td>
</tr>
<tr>
<td>Carbon Dioxide</td>
</tr>
<tr>
<td>Hydrogen</td>
</tr>
<tr>
<td><strong>Temperature:</strong></td>
</tr>
<tr>
<td><strong>Design Specs:</strong></td>
</tr>
<tr>
<td></td>
</tr>
<tr>
<td></td>
</tr>
<tr>
<td></td>
</tr>
<tr>
<td></td>
</tr>
<tr>
<td></td>
</tr>
<tr>
<td></td>
</tr>
<tr>
<td></td>
</tr>
<tr>
<td><strong>Utilities:</strong></td>
</tr>
<tr>
<td><strong>Price of Utilities:</strong></td>
</tr>
<tr>
<td><strong>Comments and Figures:</strong></td>
</tr>
</tbody>
</table>
## Water Heater

<table>
<thead>
<tr>
<th>Number Needed:</th>
<th>1</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Bare Module Cost</strong></td>
<td>$257,743</td>
</tr>
<tr>
<td><strong>Purpose:</strong></td>
<td>Heat up water from Stripper for Phase splitting</td>
</tr>
<tr>
<td><strong>Materials in the Column:</strong></td>
<td></td>
</tr>
<tr>
<td>Inlet</td>
<td>Outlet</td>
</tr>
<tr>
<td>Flow rate (ft³/hr)</td>
<td>4,266</td>
</tr>
<tr>
<td><strong>Mass Fraction of each stream:</strong></td>
<td></td>
</tr>
<tr>
<td>Acetone</td>
<td>509 PPB</td>
</tr>
<tr>
<td>Butanol</td>
<td>378 PPB</td>
</tr>
<tr>
<td>Ethanol</td>
<td>60 PPM</td>
</tr>
<tr>
<td>Water</td>
<td>0.997</td>
</tr>
<tr>
<td>Nitrogen</td>
<td>125 PPB</td>
</tr>
<tr>
<td>Sugar</td>
<td>0.003</td>
</tr>
<tr>
<td>Carbon Dioxide</td>
<td>259 PPB</td>
</tr>
<tr>
<td>Hydrogen</td>
<td>0</td>
</tr>
<tr>
<td><strong>Temperature:</strong></td>
<td>82°F</td>
</tr>
<tr>
<td><strong>Design Specs:</strong></td>
<td>Heat Duty (BTU/hr):</td>
</tr>
<tr>
<td>Heats Transfer Coefficient (BTU/F-ft²-hr):</td>
<td>1000</td>
</tr>
<tr>
<td>ΔTlm (°F)</td>
<td>131.1</td>
</tr>
<tr>
<td>Heat Transfer Area (ft²):</td>
<td>338.7</td>
</tr>
<tr>
<td>Heating Material:</td>
<td>Steam</td>
</tr>
<tr>
<td>Type of Heat Exchanger:</td>
<td>Kettle Vaporizer</td>
</tr>
<tr>
<td>Shell:</td>
<td>Carbon Steel</td>
</tr>
<tr>
<td>Tube:</td>
<td>Stainless Steel</td>
</tr>
<tr>
<td><strong>Utilities:</strong></td>
<td>Saturated Steam @ 50 psig</td>
</tr>
<tr>
<td><strong>Price of Utilities:</strong></td>
<td>$821184/yr</td>
</tr>
<tr>
<td><strong>Comments and Figures:</strong></td>
<td></td>
</tr>
</tbody>
</table>
## Heat Exchanger S-1

<table>
<thead>
<tr>
<th>Number Needed:</th>
<th>1</th>
</tr>
</thead>
<tbody>
<tr>
<td>Bare Module Cost</td>
<td>$62,125</td>
</tr>
</tbody>
</table>

### Purpose:

**Materials in the Column:**

<table>
<thead>
<tr>
<th></th>
<th>Inlet 1</th>
<th>Inlet 2</th>
<th>Outlet 1</th>
<th>Outlet 2</th>
</tr>
</thead>
<tbody>
<tr>
<td>Flow rate (ft³/hr)</td>
<td>138,671</td>
<td>767</td>
<td>1465</td>
<td>3729</td>
</tr>
</tbody>
</table>

### Mass Fraction of each stream:

<table>
<thead>
<tr>
<th></th>
<th>Inlet 1</th>
<th>Inlet 2</th>
<th>Outlet 1</th>
<th>Outlet 2</th>
</tr>
</thead>
<tbody>
<tr>
<td>Acetone (PPM)</td>
<td>0.236</td>
<td>0.236</td>
<td>0.236</td>
<td>0.236</td>
</tr>
<tr>
<td>Butanol (PPM)</td>
<td>0.475</td>
<td>0.003</td>
<td>0.003</td>
<td>0.475</td>
</tr>
<tr>
<td>Ethanol</td>
<td>0.997</td>
<td>0.077</td>
<td>0.003</td>
<td>0.077</td>
</tr>
<tr>
<td>Water (PPM)</td>
<td>0.206</td>
<td>0.067</td>
<td>0.006</td>
<td>0.067</td>
</tr>
<tr>
<td>Nitrogen (PPM)</td>
<td>0.997</td>
<td>0.077</td>
<td>0.006</td>
<td>0.077</td>
</tr>
<tr>
<td>Sugar</td>
<td>0</td>
<td>0</td>
<td>0</td>
<td>0</td>
</tr>
<tr>
<td>Carbon Dioxide (PPM)</td>
<td>390 PPM</td>
<td>14 PPM</td>
<td>390 PPM</td>
<td>14 PPM</td>
</tr>
<tr>
<td>Hydrogen</td>
<td>0</td>
<td>0</td>
<td>0</td>
<td>0</td>
</tr>
<tr>
<td>Temperature (°F)</td>
<td>217°F</td>
<td>41°F</td>
<td>213°F</td>
<td>102°F</td>
</tr>
</tbody>
</table>

### Design Specs:

<p>| | |</p>
<table>
<thead>
<tr>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>Heat Duty (BTU/hr)</td>
<td>5136922</td>
</tr>
<tr>
<td>Heat Transfer Coefficient</td>
<td>150</td>
</tr>
<tr>
<td>(BTU/F-ft²-hr)</td>
<td></td>
</tr>
<tr>
<td>∆Tlm (°F)</td>
<td>141.24</td>
</tr>
<tr>
<td>Heat Transfer Area (ft²)</td>
<td>242.5</td>
</tr>
<tr>
<td>Heating Material:</td>
<td>Inlet 1 Stream</td>
</tr>
<tr>
<td>Type of Heat Exchanger:</td>
<td>Floating Head</td>
</tr>
<tr>
<td>Shell:</td>
<td>Carbon Steel</td>
</tr>
<tr>
<td>Tube:</td>
<td>Carbon Steel</td>
</tr>
</tbody>
</table>

### Utilities:
- No Utilities Used

### Price of Utilities:
- 0

### Comments and Figures:
- Heat Exchanger S-1
### Heat Exchanger F-1

<table>
<thead>
<tr>
<th></th>
<th>Inlet</th>
<th>Outlet</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Number Needed:</strong></td>
<td></td>
<td>1</td>
</tr>
<tr>
<td><strong>Bare Module Cost:</strong></td>
<td></td>
<td>$194,290</td>
</tr>
<tr>
<td><strong>Purpose:</strong></td>
<td></td>
<td>Remove heat from fermentation tank</td>
</tr>
<tr>
<td><strong>Materials in the Column:</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td><strong>Flow rate (ft^3/hr)</strong></td>
<td>657</td>
<td>657</td>
</tr>
<tr>
<td><strong>Mass Fraction of each stream:</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Acetone</td>
<td>0.007</td>
<td>0.007</td>
</tr>
<tr>
<td>Butanol</td>
<td>0.014</td>
<td>0.014</td>
</tr>
<tr>
<td>Ethanol</td>
<td>0.002</td>
<td>0.002</td>
</tr>
<tr>
<td>Water</td>
<td>0.973</td>
<td>0.973</td>
</tr>
<tr>
<td>Nitrogen</td>
<td>0</td>
<td>0</td>
</tr>
<tr>
<td>Sugar</td>
<td>0.003</td>
<td>0.003</td>
</tr>
<tr>
<td>Carbon Dioxide</td>
<td>0</td>
<td>0</td>
</tr>
<tr>
<td>Hydrogen</td>
<td>0</td>
<td>0</td>
</tr>
<tr>
<td><strong>Temperature:</strong></td>
<td>93°F</td>
<td>93°F</td>
</tr>
<tr>
<td><strong>Design Specs:</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Heat Duty (BTU/hr):</td>
<td></td>
<td>299,677</td>
</tr>
<tr>
<td>Heat Transfer Coefficient (BTU/F-ft2-hr):</td>
<td>100</td>
<td></td>
</tr>
<tr>
<td>ΔTlm (°F)</td>
<td></td>
<td>16.6</td>
</tr>
<tr>
<td>Heat Transfer Area (ft2):</td>
<td></td>
<td>180.53</td>
</tr>
<tr>
<td>Heating Material:</td>
<td></td>
<td>Chilled Water</td>
</tr>
<tr>
<td>Type of Heat Exchanger:</td>
<td></td>
<td>Shell and Tube</td>
</tr>
<tr>
<td>Shell:</td>
<td></td>
<td>Carbon Steel</td>
</tr>
<tr>
<td>Tube:</td>
<td></td>
<td>Stainless Steel</td>
</tr>
<tr>
<td><strong>Utilities:</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Chilled Water</td>
<td></td>
<td>6.8 ton-day/hr</td>
</tr>
<tr>
<td><strong>Price of Utilities:</strong></td>
<td></td>
<td>$44189 / yr</td>
</tr>
<tr>
<td><strong>Comments and Figures:</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td><strong>Heat Exchangers F-2 - F-3</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>-------------------------------</td>
<td></td>
<td></td>
</tr>
<tr>
<td><strong>Number Needed:</strong></td>
<td>2</td>
<td></td>
</tr>
<tr>
<td><strong>Bare Module Cost</strong></td>
<td>$329,990</td>
<td></td>
</tr>
<tr>
<td><strong>Purpose:</strong></td>
<td>Remove heat from fermentation tank</td>
<td></td>
</tr>
<tr>
<td><strong>Materials in the Column:</strong> &amp; <strong>Inlet</strong> &amp; <strong>Outlet</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Flow rate (ft³/hr)</td>
<td>2,226 &amp; 2,226</td>
<td></td>
</tr>
<tr>
<td><strong>Mass Fraction of each stream:</strong> &amp; inlet &amp; outlet</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Acetone</td>
<td>0.007 &amp; 0.007</td>
<td></td>
</tr>
<tr>
<td>Butanol</td>
<td>0.014 &amp; 0.014</td>
<td></td>
</tr>
<tr>
<td>Ethanol</td>
<td>0.002 &amp; 0.002</td>
<td></td>
</tr>
<tr>
<td>Water</td>
<td>0.973 &amp; 0.973</td>
<td></td>
</tr>
<tr>
<td>Nitrogen</td>
<td>0 &amp; 0</td>
<td></td>
</tr>
<tr>
<td>Sugar</td>
<td>0.003 &amp; 0.003</td>
<td></td>
</tr>
<tr>
<td>Carbon Dioxide</td>
<td>0 &amp; 0</td>
<td></td>
</tr>
<tr>
<td>Hydrogen</td>
<td>0 &amp; 0</td>
<td></td>
</tr>
<tr>
<td><strong>Temperature:</strong></td>
<td>93°F &amp; 93°F</td>
<td></td>
</tr>
<tr>
<td><strong>Design Specs:</strong></td>
<td>Heat Duty (BTU/hr): 2,497,308</td>
<td></td>
</tr>
<tr>
<td>Heat Transfer Coefficient (BTU/F·ft²·hr):</td>
<td>100</td>
<td></td>
</tr>
<tr>
<td>ΔTlm (°F)</td>
<td>16.6</td>
<td></td>
</tr>
<tr>
<td>Heat Transfer Area (ft²):</td>
<td>1504.4</td>
<td></td>
</tr>
<tr>
<td>Heating Material:</td>
<td>Chilled Water</td>
<td></td>
</tr>
<tr>
<td>Type of Heat Exchanger:</td>
<td>Shell and Tube</td>
<td></td>
</tr>
<tr>
<td>Shell:</td>
<td>Carbon Steel</td>
<td></td>
</tr>
<tr>
<td>Tube:</td>
<td>Stainless Steel</td>
<td></td>
</tr>
<tr>
<td><strong>Utilities:</strong></td>
<td>Chilled Water</td>
<td></td>
</tr>
<tr>
<td><strong>57.2 ton·day/hr</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td><strong>Price of Utilities:</strong></td>
<td>$368229 / yr</td>
<td></td>
</tr>
<tr>
<td><strong>Comments and Figures:</strong></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>
# 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:

<table>
<thead>
<tr>
<th>Flow rate (ft³/hr)</th>
<th>Inlet</th>
<th>Outlet</th>
</tr>
</thead>
<tbody>
<tr>
<td>1,685</td>
<td></td>
<td>1,685</td>
</tr>
</tbody>
</table>

### Mass Fraction of each stream:

<table>
<thead>
<tr>
<th></th>
<th>Inlet</th>
<th>Outlet</th>
</tr>
</thead>
<tbody>
<tr>
<td>Acetone</td>
<td>0.008</td>
<td>0.008</td>
</tr>
<tr>
<td>Butanol</td>
<td>0.016</td>
<td>0.016</td>
</tr>
<tr>
<td>Ethanol</td>
<td>0.002</td>
<td>0.002</td>
</tr>
<tr>
<td>Water</td>
<td>0.97</td>
<td>0.97</td>
</tr>
<tr>
<td>Nitrogen</td>
<td>0</td>
<td>0</td>
</tr>
<tr>
<td>Sugar</td>
<td>0.003</td>
<td>0.003</td>
</tr>
<tr>
<td>Carbon Dioxide</td>
<td>0</td>
<td>0</td>
</tr>
<tr>
<td>Hydrogen</td>
<td>0</td>
<td>0</td>
</tr>
<tr>
<td>Temperature:</td>
<td>93°F</td>
<td>93°F</td>
</tr>
</tbody>
</table>

### Design Specs:

<table>
<thead>
<tr>
<th>Heat Duty (BTU/hr):</th>
<th>172,800</th>
</tr>
</thead>
<tbody>
<tr>
<td>Heat Transfer Coefficient (BTU/F·ft²·hr):</td>
<td>100</td>
</tr>
<tr>
<td>ΔTlm (°F)</td>
<td>16.6</td>
</tr>
<tr>
<td>Heat Transfer Area (ft²):</td>
<td>104.1</td>
</tr>
<tr>
<td>Heating Material:</td>
<td>Chilled Water</td>
</tr>
<tr>
<td>Type of Heat Exchanger:</td>
<td>Shell and Tube</td>
</tr>
<tr>
<td>Shell:</td>
<td>Carbon Steel</td>
</tr>
<tr>
<td>Tube:</td>
<td>Stainless Steel</td>
</tr>
</tbody>
</table>

### Utilities:

| Chilled Water | 57.2 ton-day/hr |
| Chilled Water | $368229 / yr |

### Comments and Figures:
### Pasturizer Heat Exchanger

<table>
<thead>
<tr>
<th>Number Needed:</th>
<th>1</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Bare Module Cost</strong></td>
<td>$160,527</td>
</tr>
<tr>
<td><strong>Purpose:</strong></td>
<td>To use the pasteurized high-temperature stream to begin heating the incoming sugar cane juice</td>
</tr>
<tr>
<td><strong>Materials in the Column:</strong></td>
<td>Inlet 1</td>
</tr>
<tr>
<td>Flow rate (ft³/hr)</td>
<td>5,148</td>
</tr>
<tr>
<td><strong>Mass Fraction of each stream:</strong></td>
<td></td>
</tr>
<tr>
<td>Acetone</td>
<td>0</td>
</tr>
<tr>
<td>Butanol</td>
<td>0</td>
</tr>
<tr>
<td>Ethanol</td>
<td>0</td>
</tr>
<tr>
<td>Water</td>
<td>0.75</td>
</tr>
<tr>
<td>Nitrogen</td>
<td>0</td>
</tr>
<tr>
<td>Sugar</td>
<td>0.25</td>
</tr>
<tr>
<td>Carbon Dioxide</td>
<td>0</td>
</tr>
<tr>
<td>Hydrogen</td>
<td>0</td>
</tr>
<tr>
<td><strong>Temperature:</strong></td>
<td>77°F</td>
</tr>
<tr>
<td><strong>Design Specs:</strong></td>
<td></td>
</tr>
<tr>
<td>Heat Duty (BTU/hr):</td>
<td>192,439,332</td>
</tr>
<tr>
<td>Heat Transfer Coefficient (BTU/F-ft²-hr):</td>
<td>250</td>
</tr>
<tr>
<td>ΔTlm (°F):</td>
<td>73</td>
</tr>
<tr>
<td>Heat Transfer Area (ft²):</td>
<td>1,062</td>
</tr>
<tr>
<td>Heating Material:</td>
<td>Pasteurized Stream</td>
</tr>
<tr>
<td>Type of Heat Exchanger:</td>
<td>Floating Head</td>
</tr>
<tr>
<td>Shell:</td>
<td>Carbon Steel</td>
</tr>
<tr>
<td>Tube:</td>
<td>Carbon Steel</td>
</tr>
<tr>
<td><strong>Utilities:</strong></td>
<td>No Utilities Used</td>
</tr>
<tr>
<td><strong>Price of Utilities:</strong></td>
<td>0</td>
</tr>
<tr>
<td><strong>Comments and Figures:</strong></td>
<td></td>
</tr>
<tr>
<td><strong>Pasteurizer</strong></td>
<td><strong>Number Needed:</strong></td>
</tr>
<tr>
<td>----------------</td>
<td>--------------------</td>
</tr>
<tr>
<td><strong>Bare Module Cost</strong></td>
<td>$150,106</td>
</tr>
<tr>
<td><strong>Purpose:</strong></td>
<td>Pasteurizes incoming sugar cane juice</td>
</tr>
<tr>
<td><strong>Materials in the Column:</strong></td>
<td>Inlet</td>
</tr>
<tr>
<td><strong>Flow rate (ft^3/hr)</strong></td>
<td>5,377</td>
</tr>
<tr>
<td><strong>Mass Fraction of each stream:</strong></td>
<td></td>
</tr>
<tr>
<td>Acetone</td>
<td>0</td>
</tr>
<tr>
<td>Butanol</td>
<td>0</td>
</tr>
<tr>
<td>Ethanol</td>
<td>0</td>
</tr>
<tr>
<td>Water</td>
<td>0.75</td>
</tr>
<tr>
<td>Nitrogen</td>
<td>0</td>
</tr>
<tr>
<td>Sugar</td>
<td>0.25</td>
</tr>
<tr>
<td>Carbon Dioxide</td>
<td>0</td>
</tr>
<tr>
<td>Hydrogen</td>
<td>0</td>
</tr>
<tr>
<td><strong>Temperature:</strong></td>
<td>151°F</td>
</tr>
<tr>
<td><strong>Design Specs:</strong></td>
<td>Diameter(ft)</td>
</tr>
<tr>
<td></td>
<td>Length(ft)</td>
</tr>
<tr>
<td><strong>Material of Construction:</strong></td>
<td>Stainless Steel</td>
</tr>
<tr>
<td><strong>Heating Material:</strong></td>
<td>Steam</td>
</tr>
<tr>
<td><strong>Utilities:</strong></td>
<td>Saturated Steam @ 50 psig</td>
</tr>
<tr>
<td><strong>Price of Utilities:</strong></td>
<td>$7095770/yr</td>
</tr>
<tr>
<td><strong>Comments and Figures:</strong></td>
<td></td>
</tr>
</tbody>
</table>
## Decanter

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

<table>
<thead>
<tr>
<th>Number Needed:</th>
<th>2</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Bare Module Cost</strong></td>
<td>$2,188,202.84</td>
</tr>
<tr>
<td><strong>Purpose:</strong></td>
<td>To remove water from the organics stream</td>
</tr>
<tr>
<td><strong>Materials in the Column:</strong></td>
<td><strong>Inlet</strong></td>
</tr>
<tr>
<td>Flow rate (ft^3/hr)</td>
<td>737</td>
</tr>
<tr>
<td><strong>Mass Fraction of each stream:</strong></td>
<td></td>
</tr>
<tr>
<td>Acetone</td>
<td>0.283</td>
</tr>
<tr>
<td>Butanol</td>
<td>0.569</td>
</tr>
<tr>
<td>Ethanol</td>
<td>0.092</td>
</tr>
<tr>
<td>Water</td>
<td>0.049</td>
</tr>
<tr>
<td>Nitrogen</td>
<td>0.007</td>
</tr>
<tr>
<td>Sugar</td>
<td>0</td>
</tr>
<tr>
<td>Carbon Dioxide</td>
<td>467 PPM</td>
</tr>
<tr>
<td>Hydrogen</td>
<td>0</td>
</tr>
<tr>
<td><strong>Temperature:</strong></td>
<td>176°F</td>
</tr>
<tr>
<td><strong>Design Specs:</strong></td>
<td><strong>Type of sieves:</strong></td>
</tr>
<tr>
<td></td>
<td><strong>Diameter (ft):</strong></td>
</tr>
<tr>
<td></td>
<td><strong>Height (ft):</strong></td>
</tr>
<tr>
<td></td>
<td><strong>Vessel Material:</strong></td>
</tr>
<tr>
<td></td>
<td><strong>On Stream Time (hr):</strong></td>
</tr>
<tr>
<td></td>
<td><strong>Price per lb:</strong></td>
</tr>
<tr>
<td><strong>Utilities:</strong></td>
<td>NA</td>
</tr>
<tr>
<td><strong>Comments and Figures:</strong></td>
<td></td>
</tr>
</tbody>
</table>
# Phase Separator S-1

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

| Number Needed: | 1 |
| Purpose: | To separate water and organic impurities |
| **Bare Module Cost** | $149,690.00 |
| **Materials in the Column:** |  |
| Flow rate (ft^3/hr) | Inlet | Outlet 1 | Outlet 2 |
| 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

<table>
<thead>
<tr>
<th>Pump ID</th>
<th>Type: Centrifugal Backward Curved</th>
<th>Number needed: 1</th>
</tr>
</thead>
<tbody>
<tr>
<td>Bare Module Cost</td>
<td>$5,287.04</td>
<td></td>
</tr>
<tr>
<td>Purpose:</td>
<td>Fan carrying fresh Nitrogen feed into Stripper</td>
<td></td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Materials in the Column:</th>
<th>Inlet</th>
<th>Outlet</th>
</tr>
</thead>
<tbody>
<tr>
<td>Flow rate (ft^3/hr)</td>
<td>58,597</td>
<td>49,029</td>
</tr>
</tbody>
</table>

**Mass Fraction of each stream:**

<table>
<thead>
<tr>
<th>Component</th>
<th>Inlet</th>
<th>Outlet</th>
</tr>
</thead>
<tbody>
<tr>
<td>Acetone</td>
<td>0</td>
<td>0</td>
</tr>
<tr>
<td>Butanol</td>
<td>0</td>
<td>0</td>
</tr>
<tr>
<td>Ethanol</td>
<td>0</td>
<td>0</td>
</tr>
<tr>
<td>Water</td>
<td>0</td>
<td>0</td>
</tr>
<tr>
<td>Nitrogen</td>
<td>1</td>
<td>1</td>
</tr>
<tr>
<td>Sugar</td>
<td>0</td>
<td>0</td>
</tr>
<tr>
<td>Carbon Dioxide</td>
<td>0</td>
<td>0</td>
</tr>
<tr>
<td>Hydrogen</td>
<td>0</td>
<td>0</td>
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**Temperature:**

- 77°F
- 142°F

**Design Specs:**

<table>
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<tr>
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<tr>
<td>Volumetric Flow Rate (ft^3/hr)</td>
<td>58,597</td>
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<tr>
<td>Head (in. H2O)</td>
<td>114</td>
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<tr>
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<tr>
<td>Pressure Change</td>
<td>5 psi</td>
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</table>

**Utilities:**

- 69 HP
- Price of Utilities: $13910/yr

**Comments and Figures:**

Blower Fan carrying fresh Nitrogen feed into Stripper
| **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 (ft³/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

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<td>Flow rate (ft(^3/)hr)</td>
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<td>Mass Fraction of each stream:</td>
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<td>Acetone</td>
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<td>Carbon Dioxide</td>
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<td>Fan sending gas impurities to Heat Exchanger</td>
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<td>Butanol</td>
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<td>Hydrogen</td>
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<td>Sugar</td>
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<td>Hydrogen</td>
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<td>Pressure (torr):</td>
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Equipment Cost Summary
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<th>$FBM$</th>
<th>$CBM$</th>
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<td>Fermenters F-1 - F-4</td>
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<td>$4,954</td>
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<td>Fermenters F-6 - F-7</td>
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<tr>
<td><strong>Centrifuges</strong></td>
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<tr>
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<td>$60,900</td>
<td>$60,900</td>
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<td><strong>Pressure Vessels</strong></td>
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### Pumps & Motors

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### Fans

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### Total

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Utility Requirements
### Unit Electrical Requirements

<table>
<thead>
<tr>
<th>Equipment</th>
<th>Req kW -hr/hr</th>
<th>Price per kW-hr</th>
<th>Cost ($/yr)</th>
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<tbody>
<tr>
<td>Pumps</td>
<td>1688</td>
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<td>Fans</td>
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<td>Agitators</td>
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<td>Centrifuges</td>
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### Unit Cooling Water Requirements

<table>
<thead>
<tr>
<th>Equipment</th>
<th>CW gal/hr</th>
<th>Price per 1000 gal</th>
<th>Cost ($/yr)</th>
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<tbody>
<tr>
<td>Distillation Columns</td>
<td>91760</td>
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### Unit Steam Requirements

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<tr>
<th>Equipment</th>
<th>Steam lb/hr</th>
<th>Price per 1000 lb</th>
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<tbody>
<tr>
<td>Pasteurizer 50 psig</td>
<td>439966</td>
<td>$3.00</td>
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<td>Distillation Columns 50 ps ig</td>
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<td>Distillation Columns 150 ps ig</td>
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<td>Heat Exchangers 50 psig</td>
<td>50916</td>
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<td>Fertilizer Dryer 50 ps ig</td>
<td>2715</td>
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<td>Triple Effect Evaporator 50 psig</td>
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<td><strong>Total</strong></td>
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### Unit Refrigerant Requirements

<table>
<thead>
<tr>
<th>Equipment</th>
<th>Ammonia gal / hr</th>
<th>Price per gal</th>
<th>Cost ($/yr)</th>
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<tbody>
<tr>
<td>Condenser</td>
<td>836</td>
<td>$0.023</td>
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<td>Equipment</td>
<td>Chilled water ton-day /hr</td>
<td>Price per ton-day</td>
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<td>Fermentation Heat Exchangers</td>
<td>571.5</td>
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### Unit Nitrogen Requirements

<table>
<thead>
<tr>
<th>Equipment</th>
<th>Nitrogen Ft^3/hr</th>
<th>Price per 100 ft^3</th>
<th>Cost($/yr)</th>
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<tr>
<td>Gas Stripper</td>
<td>58596</td>
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### Utility Costs per Year

<table>
<thead>
<tr>
<th>Item</th>
<th>Cost ($/yr)</th>
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<tbody>
<tr>
<td>Electrical</td>
<td>$1,894,228</td>
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<tr>
<td>Cooling Water</td>
<td>$36,815</td>
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<tr>
<td>Steam</td>
<td>$10,065,199</td>
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<tr>
<td>Refrigeration</td>
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<td>Nitrogen</td>
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<tr>
<td><strong>Total</strong></td>
<td><strong>$15,844,912</strong></td>
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</table>
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

<table>
<thead>
<tr>
<th>Year</th>
<th>Action</th>
<th>Distribution of Permanent Investment</th>
<th>Production Capacity</th>
<th>Depreciation 5 year MACRS</th>
<th>Product Price</th>
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<tr>
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<td>Construction</td>
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<td>Decanter Fabricated Equipment</td>
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<td>Fermentation Heat Exchangers Fabricated Equipment</td>
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<td>Fertilizer Dryer Fabricated Equipment</td>
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<td>Centrifuges Fabricated Equipment</td>
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<td><strong>Total</strong></td>
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### Raw Materials

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<th>Raw Material:</th>
<th>Unit:</th>
<th>Required Ratio:</th>
<th>Cost of Raw Material:</th>
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<tbody>
<tr>
<td>1 Sugar Cane Juice</td>
<td>lb</td>
<td>167.25 lb per gal of Ethanol</td>
<td>$0.050 per lb</td>
</tr>
<tr>
<td>2 Nitrogen</td>
<td>100 ft³</td>
<td>3.613 100 ft³ per gal of Ethanol</td>
<td>$0.02 per 100 ft³</td>
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<tr>
<td>3 Sulfuric Acid</td>
<td>lb</td>
<td>0.335 lb per gal of Ethanol</td>
<td>$0.06 per lb</td>
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<td>4 Lime</td>
<td>lb</td>
<td>0.1675 lb per gal of Ethanol</td>
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<td>5 Gasoline</td>
<td>gal</td>
<td>0.02 gal per gal of Ethanol</td>
<td>$2.00 per gal</td>
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<tr>
<td>6 Caustic</td>
<td>lb</td>
<td>0.75 lb per gal of Ethanol</td>
<td>$0.13 per lb</td>
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Total Weighted Average: $8.597 per gal of Ethanol

### Byproducts

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

### Utilities

<table>
<thead>
<tr>
<th>Utility:</th>
<th>Unit:</th>
<th>Required Ratio:</th>
<th>Utility Cost:</th>
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<tbody>
<tr>
<td>1 High Pressure Steam</td>
<td>1000 lb</td>
<td>0.01975 1000 lb per gal of Ethanol</td>
<td>$4.800 per 1000 lb</td>
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<td>2 Low Pressure Steam</td>
<td>1000 lb</td>
<td>1.17462 1000 lb per gal of Ethanol</td>
<td>$3.000 per 1000 lb</td>
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<tr>
<td>3 Cooling Water</td>
<td>1000 gal</td>
<td>0.17444 1000 gal per gal of Ethanol</td>
<td>$0.075 per 1000 gal</td>
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<tr>
<td>4 Chilled Water</td>
<td>ton-day</td>
<td>1.0865 ton-day per gal of Ethanol</td>
<td>$1.200 per ton-day</td>
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<tr>
<td>5 Electricity</td>
<td>kWh</td>
<td>15.6 kWh per gal of Ethanol</td>
<td>$0.050 per kWh</td>
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</table>

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

<table>
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<tr>
<th>Item</th>
<th>Percentage or Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>Cost of Site Preparations</td>
<td>5.00% of Total Bare Module Costs</td>
</tr>
<tr>
<td>Cost of Service Facilities</td>
<td>5.00% of Total Bare Module Costs</td>
</tr>
<tr>
<td>Allocated Costs for utility plants and related facilities</td>
<td>$0</td>
</tr>
<tr>
<td>Cost of Contingencies and Contractor Fees</td>
<td>18.00% of Direct Permanent Investment</td>
</tr>
<tr>
<td>Cost of Land</td>
<td>2.00% of Total Depreciable Capital</td>
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<tr>
<td>Cost of Royalties</td>
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<tr>
<td>Cost of Plant Start-Up</td>
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### Fixed Costs

#### Operations

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<tr>
<th>Item</th>
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<td>Operators per Shift</td>
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<tr>
<td>Direct Wages and Benefits</td>
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<tr>
<td>Direct Salaries and Benefits</td>
<td>15% of Direct Wages and Benefits</td>
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<tr>
<td>Operating Supplies and Services</td>
<td>6% of Direct Wages and Benefits</td>
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<tr>
<td>Technical Assistance to Manufacturing</td>
<td>$0.00 per year, for each Operator per Shift</td>
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<tr>
<td>Control Laboratory</td>
<td>$0.00 per year, for each Operator per Shift</td>
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#### Maintenance

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<td>Wages and Benefits</td>
<td>4.50% of Total Depreciable Capital</td>
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<tr>
<td>Salaries and Benefits</td>
<td>25% of Maintenance Wages and Benefits</td>
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<tr>
<td>Materials and Services</td>
<td>100% of Maintenance Wages and Benefits</td>
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<tr>
<td>Maintenance Overhead</td>
<td>5% of Maintenance Wages and Benefits</td>
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#### Operating Overhead

<table>
<thead>
<tr>
<th>Item</th>
<th>Description</th>
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<tbody>
<tr>
<td>General Plant Overhead</td>
<td>7.10% of Maintenance and Operations Wages and Benefits</td>
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<tr>
<td>Mechanical Department Services</td>
<td>2.40% of Maintenance and Operations Wages and Benefits</td>
</tr>
<tr>
<td>Employee Relations Department</td>
<td>5.90% of Maintenance and Operations Wages and Benefits</td>
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<tr>
<td>Business Services</td>
<td>7.40% of Maintenance and Operations Wages and Benefits</td>
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</table>

#### Property Taxes and Insurance

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<th>Item</th>
<th>Description</th>
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</thead>
<tbody>
<tr>
<td>Property Taxes and Insurance</td>
<td>2% of Total Depreciable Capital</td>
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</table>

#### 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

<table>
<thead>
<tr>
<th>Item</th>
<th>Description</th>
</tr>
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<tbody>
<tr>
<td>Rental Fees (Office and Laboratory Space)</td>
<td>$0</td>
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<tr>
<td>Licensing Fees</td>
<td>$0</td>
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<tr>
<td>Miscellaneous</td>
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#### Depletion Allowance

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<tbody>
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<td>Annual Depletion Allowance</td>
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### Variable Cost Summary

**Variable Costs at 100% Capacity:**

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<th>General Expenses</th>
<th>Amount</th>
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<tbody>
<tr>
<td>Selling / Transfer Expenses:</td>
<td>$212,083</td>
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<tr>
<td>Direct Research:</td>
<td>$339,333</td>
</tr>
<tr>
<td>Allocated Research:</td>
<td>$35,347</td>
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<tr>
<td>Administrative Expense:</td>
<td>$141,389</td>
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<tr>
<td>Management Incentive Compensation:</td>
<td>$88,368</td>
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<td><strong>Total General Expenses</strong></td>
<td><strong>$816,520</strong></td>
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<table>
<thead>
<tr>
<th>Raw Materials</th>
<th>Amount</th>
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<tbody>
<tr>
<td>$8.596985 per gal of Ethanol</td>
<td>$24,310,348</td>
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</table>

<table>
<thead>
<tr>
<th>Byproducts</th>
<th>Amount</th>
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<td>$41.075780 per gal of Ethanol</td>
<td>$(116,153,105)</td>
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<table>
<thead>
<tr>
<th>Utilities</th>
<th>Amount</th>
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<tbody>
<tr>
<td>$5.715543 per gal of Ethanol</td>
<td>$16,162,275</td>
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</table>

**Total Variable Costs**  
$ (74,863,961)

### Fixed Cost Summary

<table>
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<tr>
<th>Operations</th>
<th>Amount</th>
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<tbody>
<tr>
<td>Direct Wages and Benefits:</td>
<td>$1,456,000</td>
</tr>
<tr>
<td>Direct Salaries and Benefits:</td>
<td>$218,400</td>
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<tr>
<td>Operating Supplies and Services:</td>
<td>$87,360</td>
</tr>
<tr>
<td>Technical Assistance to Manufacturing</td>
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</tr>
<tr>
<td>Control Laboratory</td>
<td>-</td>
</tr>
<tr>
<td><strong>Total Operations</strong></td>
<td><strong>$1,761,760</strong></td>
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<table>
<thead>
<tr>
<th>Maintenance</th>
<th>Amount</th>
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<tr>
<td>Wages and Benefits</td>
<td>$3,748,304</td>
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<tr>
<td>Salaries and Benefits</td>
<td>$937,076</td>
</tr>
<tr>
<td>Materials and Services</td>
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<td>Maintenance Overhead</td>
<td>$187,415</td>
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<td><strong>Total Maintenance</strong></td>
<td><strong>$8,621,099</strong></td>
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<thead>
<tr>
<th>Operating Overhead</th>
<th>Amount</th>
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<tr>
<td>General Plant Overhead:</td>
<td>$451,544</td>
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<tr>
<td>Mechanical Department Services:</td>
<td>$152,635</td>
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<td>Employee Relations Department:</td>
<td>$375,227</td>
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<td>Business Services:</td>
<td>$470,624</td>
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<td><strong>Total Operating Overhead</strong></td>
<td><strong>$1,450,030</strong></td>
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<table>
<thead>
<tr>
<th>Property Taxes and Insurance</th>
<th>Amount</th>
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</thead>
<tbody>
<tr>
<td>Property Taxes and Insurance:</td>
<td>$1,665,913</td>
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</table>

<table>
<thead>
<tr>
<th>Other Annual Expenses</th>
<th>Amount</th>
</tr>
</thead>
<tbody>
<tr>
<td>Rental Fees (Office and Laboratory Space):</td>
<td>-</td>
</tr>
<tr>
<td>Licensing Fees:</td>
<td>-</td>
</tr>
<tr>
<td>Miscellaneous:</td>
<td>-</td>
</tr>
<tr>
<td><strong>Total Other Annual Expenses</strong></td>
<td><strong>-</strong></td>
</tr>
</tbody>
</table>

**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

<table>
<thead>
<tr>
<th></th>
<th>2009</th>
<th>2010</th>
<th>2011</th>
</tr>
</thead>
<tbody>
<tr>
<td>Accounts Receivable</td>
<td>$261,472</td>
<td>$130,736</td>
<td>$130,736</td>
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<tr>
<td>Cash Reserves</td>
<td>$1,097,054</td>
<td>$548,527</td>
<td>$548,527</td>
</tr>
<tr>
<td>Accounts Payable</td>
<td>$(1,496,933)</td>
<td>$(748,466)</td>
<td>$(748,466)</td>
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<tr>
<td>Ethanol Inventory</td>
<td>$34,863</td>
<td>$17,431</td>
<td>$17,431</td>
</tr>
<tr>
<td>Raw Materials</td>
<td>$59,943</td>
<td>$29,972</td>
<td>$29,972</td>
</tr>
<tr>
<td><strong>Total</strong></td>
<td>$(43,600)</td>
<td>$(21,800)</td>
<td>$(21,800)</td>
</tr>
</tbody>
</table>

Present Value at 15%: $(37,913)  $(16,484)  $(14,334)

Total Capital Investment: $93,222,393
<table>
<thead>
<tr>
<th>Year</th>
<th>Percentage of Design Capacity</th>
<th>Product Unit Price</th>
<th>Sales</th>
<th>Capital Costs</th>
<th>Working Capital</th>
<th>Var Costs</th>
<th>Fixed Costs</th>
<th>Depreciation</th>
<th>Allowance</th>
<th>Taxible Income</th>
<th>Taxes</th>
<th>Net Earnings</th>
<th>Cash Flow</th>
<th>Cumulative Net Present Value at 15%</th>
</tr>
</thead>
<tbody>
<tr>
<td>2008</td>
<td>0%</td>
<td>-</td>
<td>-</td>
<td>(87,200)</td>
<td>-</td>
<td>-</td>
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<td>-</td>
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</tr>
<tr>
<td>2009</td>
<td>0%</td>
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<td>(93,291,100)</td>
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<tr>
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<td>48% $2.50</td>
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<td>(13,498,800)</td>
<td>(15,927,800)</td>
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<tr>
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<td>2012</td>
<td>90% $2.50</td>
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<td>67,377,600</td>
<td>(13,498,800)</td>
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<tr>
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<td>-</td>
<td>67,377,600</td>
<td>(13,498,800)</td>
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<tr>
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<td>67,377,600</td>
<td>(13,498,800)</td>
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<tr>
<td>2018</td>
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<td>67,377,600</td>
<td>(13,498,800)</td>
<td>-</td>
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<td>-</td>
<td>-</td>
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<tr>
<td>2019</td>
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<td>67,377,600</td>
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<tr>
<td>2020</td>
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<td>-</td>
<td>67,377,600</td>
<td>(13,498,800)</td>
<td>-</td>
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<tr>
<td>2021</td>
<td>90% $2.50</td>
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<td>-</td>
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<td>(13,498,800)</td>
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<td>(13,498,800)</td>
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<td>-</td>
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<td>(13,498,800)</td>
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<td>-</td>
<td>-</td>
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<tr>
<td>2025</td>
<td>90% $2.50</td>
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<td>-</td>
<td>67,377,600</td>
<td>(13,498,800)</td>
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<tr>
<td>2026</td>
<td>90% $2.50</td>
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<tr>
<td>2029</td>
<td>90% $2.50</td>
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<td>-</td>
<td>67,377,600</td>
<td>(13,498,800)</td>
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<td>-</td>
</tr>
</tbody>
</table>
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)

<p>| | |</p>
<table>
<thead>
<tr>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Annual Sales</strong></td>
<td>$6,362,496</td>
</tr>
<tr>
<td><strong>Annual Costs</strong></td>
<td>$53,878,763</td>
</tr>
<tr>
<td><strong>Depreciation</strong></td>
<td>($7,463,290)</td>
</tr>
<tr>
<td><strong>Income Tax</strong></td>
<td>($19,527,849)</td>
</tr>
<tr>
<td><strong>Net Earnings</strong></td>
<td>$33,250,120</td>
</tr>
<tr>
<td><strong>Total Capital Investment</strong></td>
<td>$93,203,924</td>
</tr>
<tr>
<td><strong>ROI</strong></td>
<td>35.67%</td>
</tr>
</tbody>
</table>

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.)

<table>
<thead>
<tr>
<th>Variable Costs</th>
<th>$1.25</th>
<th>16.16%</th>
<th>20.53%</th>
<th>24.56%</th>
<th>28.36%</th>
<th>31.59%</th>
<th>35.49%</th>
<th>38.89%</th>
<th>42.20%</th>
<th>45.44%</th>
<th>48.61%</th>
<th>51.73%</th>
</tr>
</thead>
<tbody>
<tr>
<td>Product Price</td>
<td>$1.50</td>
<td>16.58%</td>
<td>20.91%</td>
<td>24.92%</td>
<td>28.70%</td>
<td>32.32%</td>
<td>35.80%</td>
<td>39.19%</td>
<td>42.49%</td>
<td>45.72%</td>
<td>48.99%</td>
<td>52.00%</td>
</tr>
<tr>
<td></td>
<td>$1.75</td>
<td>17.00%</td>
<td>21.29%</td>
<td>25.27%</td>
<td>29.03%</td>
<td>32.64%</td>
<td>36.11%</td>
<td>39.49%</td>
<td>42.78%</td>
<td>46.00%</td>
<td>49.16%</td>
<td>52.27%</td>
</tr>
<tr>
<td></td>
<td>$2.00</td>
<td>17.41%</td>
<td>21.67%</td>
<td>25.62%</td>
<td>29.37%</td>
<td>32.95%</td>
<td>36.42%</td>
<td>39.79%</td>
<td>43.07%</td>
<td>46.29%</td>
<td>49.44%</td>
<td>52.54%</td>
</tr>
<tr>
<td></td>
<td>$2.25</td>
<td>17.82%</td>
<td>22.04%</td>
<td>25.97%</td>
<td>29.70%</td>
<td>33.27%</td>
<td>36.73%</td>
<td>40.08%</td>
<td>43.36%</td>
<td>46.57%</td>
<td>49.72%</td>
<td>52.81%</td>
</tr>
<tr>
<td></td>
<td>$2.50</td>
<td>18.23%</td>
<td>22.41%</td>
<td>26.32%</td>
<td>30.03%</td>
<td>33.59%</td>
<td>37.03%</td>
<td>40.38%</td>
<td>43.65%</td>
<td>46.85%</td>
<td>49.99%</td>
<td>53.08%</td>
</tr>
<tr>
<td></td>
<td>$2.75</td>
<td>18.63%</td>
<td>22.78%</td>
<td>26.66%</td>
<td>30.36%</td>
<td>33.90%</td>
<td>37.38%</td>
<td>40.68%</td>
<td>43.94%</td>
<td>47.13%</td>
<td>50.27%</td>
<td>53.35%</td>
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<td></td>
<td>$3.00</td>
<td>19.03%</td>
<td>23.15%</td>
<td>27.01%</td>
<td>30.68%</td>
<td>34.22%</td>
<td>37.64%</td>
<td>40.97%</td>
<td>44.22%</td>
<td>47.41%</td>
<td>50.54%</td>
<td>53.62%</td>
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<tr>
<td></td>
<td>$3.25</td>
<td>19.42%</td>
<td>23.51%</td>
<td>27.36%</td>
<td>31.01%</td>
<td>34.53%</td>
<td>37.94%</td>
<td>41.26%</td>
<td>44.51%</td>
<td>47.69%</td>
<td>50.82%</td>
<td>53.89%</td>
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<tr>
<td></td>
<td>$3.50</td>
<td>19.81%</td>
<td>23.87%</td>
<td>27.69%</td>
<td>31.33%</td>
<td>34.84%</td>
<td>38.24%</td>
<td>41.56%</td>
<td>44.80%</td>
<td>47.97%</td>
<td>51.09%</td>
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<tr>
<td></td>
<td>$3.75</td>
<td>20.19%</td>
<td>24.23%</td>
<td>28.03%</td>
<td>31.66%</td>
<td>35.15%</td>
<td>38.54%</td>
<td>41.86%</td>
<td>45.09%</td>
<td>48.25%</td>
<td>51.36%</td>
<td>54.42%</td>
</tr>
</tbody>
</table>
**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](image1.png)  ![Figure 2](image2.png)

**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 CO₂/H₂ 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 H₂ and CO₂ 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 CO₂ in water is greater than the solubility of N₂ 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 CO₂ 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 CO₂ 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 CO₂ per mole sugar, but acetone makes one more per mol sugar used. This is shown below:

\[
C_6H_{12}O_6 \rightarrow C_4H_{10}O (\text{butanol}) + 2CO_2 + H_2O
\]
\[
C_6H_{12}O_6 \rightarrow 2C_2H_5O (\text{ethanol}) + 2CO_2 + H_2
\]
\[
C_6H_{12}O_6 \rightarrow C_3H_6O (\text{acetone}) + 3CO_2 + 4H_2
\]

These facts show that the ABE process should not have significantly more CO₂ 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 CO₂, 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
References


Al-Aidaroos, Salma, Nicholas Bass, Brian Downey, and Jonathan Ziegler. 2009. Offshore LNG production.


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.

\[
\left(100 \text{ g sugar}\right) \left(\frac{1 \text{ mol sugar}}{178 \text{ g sugar}}\right) \left(\frac{6 \text{ mol C}}{1 \text{ mol sugar}}\right) \left(\frac{0.149 \text{ mol C in acetone}}{1 \text{ mol C}}\right) \left(\frac{1 \text{ mol acetone}}{3 \text{ mol C in acetone}}\right) \left(\frac{58.08 \text{ g}}{1 \text{ mol}}\right) = 9.72 \text{ g acetone per 100 g sugar}
\]

Total yield in mass percent of glucose consumed is:

<table>
<thead>
<tr>
<th>Compound</th>
<th>Growth Phase</th>
<th>Production Phase</th>
</tr>
</thead>
<tbody>
<tr>
<td>Acetone</td>
<td>9.72%</td>
<td>11.22%</td>
</tr>
<tr>
<td>Butanol</td>
<td>19.43%</td>
<td>22.36%</td>
</tr>
<tr>
<td>Ethanol</td>
<td>3.26%</td>
<td>3.73%</td>
</tr>
<tr>
<td>Carbon Dioxide</td>
<td>59.19%</td>
<td>61.12%</td>
</tr>
<tr>
<td>Hydrogen</td>
<td>2.00%</td>
<td>2.00%</td>
</tr>
<tr>
<td>Biomass</td>
<td>8.68%</td>
<td>0.87%</td>
</tr>
</tbody>
</table>

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 (\(X, \text{ g/L}\)) must be included on the left side of the equation. The initial cell charge to the fermenter is designated as \(X_0, \text{ (g/L)}\). 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 \( \sim 24 \) 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, \( 1.90 \times 10^7 \) 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 \( S = 67.5 \ \text{g/L} \) to \( S = 4.05 \ \text{g/L} \), and setting the fermentation time at \( \sim 12 \ \text{hours} \):

\[
X_o = 11.75 \ \text{g/L}
\]

\[
t = 12.03 \ \text{h}
\]

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 \( 2.2 \times 10^7 \ \text{g DCW} \) 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 ft³/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.

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

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

\[
S = QH^{0.5} = (32.41 \text{ gpm})(67)^{0.5} = 265.29
\]

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:

\[
P_C = \frac{QH\rho}{33000 \times \eta} = \frac{(32.41 \text{ gpm})(67 \text{ ft}) (48 \text{ lb/ft}^3)}{(33000)(0.7)(7.48 \text{ gal/ft}^3)} = 12.1 \text{ HP}
\]

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:

\[
t_p = \frac{P_d D_i}{2SE - 1.2P_d} = \frac{(190.7 \text{ psi})(3.8 \text{ ft})(12 \text{ in/ft})}{(2)(15000 \text{ psi})(1) - (1.2)(190.7 \text{ psi})} = 0.292 \text{ in}
\]

The decanter weight is:

\[
W = \pi (D_i + t_s)(L + 0.8D_i)t_s \rho_{\text{carbon steel}}
\]

\[
= \pi \left( 3.8 \text{ ft} + (0.282 \text{ in}) \left( \frac{0.083 \text{ ft}}{\text{in}} \right) \right) (11.4 \text{ ft})
\]

\[
+ (0.8)(3.67)(0.282 \text{ in}) \left( \frac{0.083 \text{ ft}}{\text{in}} \right) \left( 490 \frac{\text{lb}}{\text{ft}^3} \right) = 1974 \text{ lb}
\]

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 Calculation

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:

\[ W = \pi(D_i + t_s)(L + 0.8D_i)t_s\rho_{carbon\ steel} \]

\[ = \pi \left( 2.85\text{ft} + (0.25\text{in}) \left(0.083\text{ft/ in}\right) \right)(8.55\text{ft}) \]

\[ + (0.8)(2.85)) (0.25 \text{ in}) \left(0.083\text{ft/ in}\right) \left(490 \text{ lb/ft}^3\right) = 993 \text{ lb} \]

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 \( P_d = 10 \text{ psig} \) 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:

\[ W = \pi(D_i + t_s)(L + 0.8D_i)t_s\rho_{carbon\ steel} \]

\[ = \pi \left( 6.7\text{ft} + (0.5\text{ in}) \left(0.083\text{ft/ in}\right) \right)(20.2\text{ft}) \]

\[ + (0.8)(6.7)) (0.25 \text{ in}) \left(0.083\text{ft/ in}\right) \left(490 \text{ lb/ft}^3\right) = 5504 \text{ lb} \]

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,
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:

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

The density of molecular sieves 13X is obtained from Hengye USA (2008). Using a void space of 0.5, the vessel volume is 6756 ft³. 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:

\[
Number of trays \, N_T = \frac{Number of stages}{Tray Efficiency} = \frac{15}{0.7} \approx 22 \text{ trays}
\]

\[
L = (N^T \times \text{Tray spacing}) + \text{sump space} + \text{head space}
\]

\[
L = 22 \times \frac{18}{12} + 4 + 3 + 3 + 3 - \frac{18}{12} = 41.5 \text{ ft}
\]

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

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

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

Calculating the wall thickness:

\[
t_p = \frac{P_d D_t}{2SE - 1.2P_d} = \frac{(15.24 \text{ psi})(4 \text{ ft})(12 \text{ in/ft})}{(2)(15000 \text{ psi})(1.2)(15.24 \text{ psi})} = 0.0287 \text{ in}
\]
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 $t_s$ for $D_i = 4$ ft is:

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

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

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

$$= \pi \left( 4\text{ft} + (0.4375\text{ in})(0.083\text{ ft/in}) \right) (41.5\text{ ft})$$

$$+ (0.8)(4)(0.4375\text{ in})(0.083\text{ ft/in})(490\text{ lb/ft}^3) = 10121.2\text{ lb}$$

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:

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

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

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

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

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:

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

**Condenser (Fixed Head Heat Exchanger):**

\[T_{H,in} = 172.7^\circ F \quad T_{H,out} = 100^\circ F \quad T_{C,in} = 90^\circ F \quad T_{C,out} = 120^\circ F\]

\[\Delta T_{LM} = 25.7^\circ F\]

Heat Duty, \(Q = -7998642.8\) Btu/hr

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

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:

\[F_M = a + \left( \frac{A}{100} \right)^b = 1.08 + \left( \frac{3112.31}{100} \right)^{0.05} = 2.27\]

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

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

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

Total Bare Module Cost:

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

**Reboiler (Kettle Vaporizer):**

Heat Duty, \(Q = 8900486.18\) Btu/hr
Heat Flux = 12000 Btu/hr-ft²

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

Calculating the purchase cost:

\[
C_P = F_P F_M F_L C_B
\]

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

\[
F_M = F_P = F_L = 1,
\]

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

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

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

Total Bare Module Cost:

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

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:

\[
t_s = 0.25 + 0.125 = 0.375 \text{ in}
\]

Using this value to calculate the weight of the horizontal vessel:
\[ W = \pi (D_t + t_s) (L + 0.8D_t) t_s \rho_{carbon\ steel} \]

\[ = \pi \left( 2.5 ft + (0.375 \text{ in}) \left( 0.083 \text{ ft/in} \right) \right) (7.42 ft) \]

\[ + (0.8)(2.5) (0.375 \text{ in}) \left( 0.083 \text{ ft/in} \right) (490 \text{ lb/ft}^3) = 660 \text{ lb} \]

Total purchase cost \( C_p \):

\[ C_p = F_M C_V + C_{PL} \]

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

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

\[ C_{PL} = 2005(D_t)^{0.20294} = 2005(2.5)^{0.20294} = 2307.83 \]

\[ C_p = F_M C_V + C_{PL} = (1)(10602.4) + 2307.83 = 13504.1 \]

Total Bare Module Cost:

\[ C_{BM} = F_{BM} C_p \left( \frac{523}{500} \right) = (3.05)(13504.1) \left( \frac{523}{500} \right) = \$41187.53 \]

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 Widagdo (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/ft², meaning a heat transfer area of 501 ft² is required. The price of a dryer this size is:

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

\[ C_{BM} = (\frac{523.6}{500})(2.06) = $732799.51 \]
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<th>N2FEED</th>
<th>STRPRCYL</th>
<th>ABE1</th>
<th>H2OMILL</th>
<th>ACETONE</th>
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Appendix C: Equipment Quote Sheets

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

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**Vertical Axial Flow Pump (ZL)**

**Supplier Details**

Beijing Great Sources Technology Development Co., Ltd.

- [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 Valve, ... [more]
- **Online Showroom:** 786 products

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**Product Details**

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

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**Product Description**

- **Place of Origin:** Shanghai China (Mainland)
- **Theory:** Centrifugal Pump
- **Power:** Electric
- **Material:** Cast iron
- **Capacity:** 0.03-300m³/h
- **Total head:** 2-25m
- **Motor rating (22kW):** 0.5-220kW
- **Size:** 600-1600mm
- **Speed:** 245-980r/min

**ZL pump series are single-stage, vertical axial flow pump with the liquid flowing in axial direction. The pump is used to transfer clean water or other liquid similar with water. The maximum temperature of the liquid can be 80°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, stock water level control and other conservation projects.**