Butanol by Two Stage Fermentation

Design Project by:

Christina Chen

Amira Fawcett

Amy Posner

Tal Raviv

Presented To:

Professor Leonard A. Fabiano

Dr. Sean P. Holleran

April 14th, 2009

Department of Chemical and Biomolecular Engineering

University of Pennsylvania

School of Engineering and Applied Science

April 3rd, 2009
Dr. Sean P. Holleran
Professor Leonard K. Fabiano
University of Pennsylvania
School of Engineering and Applied Science
Department of Chemical and Biomolecular Engineering

Dear Professor Fabiano and Dr. Holleran,

Enclosed you will find our written report and solution to the design project proposed to us by Bruce Vrana, of DuPont, *Butanol by Two-Stage Fermentation*. The process involves the use of two strains of bacteria to convert glucose, derived from corn, into butanol fuel. The entire process is broken up into two main sections: the fermentation phase and the separations phase.

The fermentation phase begins with a series of continuous Fibrous Bed Bioreactors immobilized with *Clostridium tyrobutyricum*, performing acidogenesis by converting glucose into butyric acid. The butyric acid is then fed into another series of bioreactors where *Clostridium acetobutylicum* converts butyric acid into butanol via solventogenesis. The separations phase details a continuous separations stream to recover the butanol product at 99.5% purity and recycle as much of the raw materials and process water as possible.

The report outlines the necessary startup and investment costs required to implement the facility, as well as the potential profitability of the plant. The design requires 10 billion pounds of corn a year to produce 54.4 million gallons of butanol.

Financial analysis on the design yielded an NPV of $217 million at an interest rate of 15%. This corresponds to a 32.3% IRR, when butanol is $4.00 per gallon. Further analysis of these estimations are detailed inside.

Sincerely,

Christina Chen

Amira Fawcett

Amy Posner

Tal Raviv

Table of Contents

Table of Contents

Section I: Abstract

Abstract

Current techniques for producing butanol tend to have a low yield and form a large amount of other solvents, because there is only one stage for fermentation. Having one stage limits the type of bacteria that can be used, because the chosen bacteria must be able to both convert glucose to butyric acid, and then convert butyric acid to butanol. The only types of bacteria that can perform both these tasks also create a lot of other acids, which are turned to other solvents in the product stream. This is most prevalent in ABE fermentation, which creates significant amounts of acetone and ethanol along with the butanol. David Ramey of ButylFuel LLC, has created a distinct process that generates butanol, without significant amounts of acetone or ethanol, using a two-stage fermentation process. The first stage converts glucose to butyric acid through acidogenesis, while the second stage converts the butyric acid to butanol via solventogenesis. This process optimizes the efficiency and specific production of the desired solvent, butanol.

The purpose of this report is to scale-up Ramey's process and build a plant based on a two-stage fermentation procedure. The economical viability of producing 50 million gallons of butanol per year, at a purity of 99.5% from the plant will also been discussed. These results will allow the organization to determine the worth of licensing the technology from ButylFuel. Additionally, because this process will compete with many ethanol plants, it is necessary for the design to mirror a typical ethanol plant as much as possible. Because of this, aspects of the current production of ethanol were implemented in the design, including the Dry Grind process and the Dried Distillers Grain Drying process. These

implementations allow the process to be constructed from modified ethanol plants, rather than having to rebuild a new plant.

The fermentation phase of the design utilizes a series of fibrous bed reactors and two different strands of Clostridium bacteria for each stage. The product stream out of the second fermentation stage, containing butanol, is separated using a liquid-liquid extractor, and a series of distillation columns, to extract the butanol from water. Different separation options were researched, including pervaporation, decanters, and stripping. The liquid-liquid extractor with distillation columns was chosen in the end, because it was the simplest and most economical process for dealing with a product stream that was over 90% water. Also, a butanol/water azeotrope surfaces during the separations process that is efficiently dealt with by the extractor.

For the economic analysis, this report uses 50 million gallons per year producing ethanol plant as a comparison with the butanol process. The total capital investment for the ethanol plant is about $74.1 million with an investment rate of return (IRR) of 33.1%. This correlates to a total capital investment of $1.48/gallon of ethanol produced.

Since the design specifications involved the modification of an existing ethanol plant, it was assumed that some existing ethanol equipment would be integrated into the system. Specifically, the Dried Distiller's Grains (DDGS) dryer and the Dry Grind process are assumed to be installed and operational in year one. Additionally, it was assumed this equipment had been fully depreciated by the time of construction of the butanol plant.

The results of this report were based on 54.3 million gallons per year producing butanol plant. , For this design, a total capital investment of $219 million was determined. This is a substantial investment cost highlighted by the fact that the overall net present value (NPV) of the design, after 15 years, was found to be a negative $3.55 billion. The poor investment opportunity stems from the high cost of utilities needed to run the plant. Of the total annual costs, 94.5% is derived from the overall utility costs.

The profitability analysis and a review of current market conditions indicate that this investment should not be undertaken due to its high degree of unprofitability. Serious consideration of external factors and of the design itself must be taken before pursuing any investment. These factors, such as the price of corn, will be outlined more thoroughly at the end of the report.

Section II: Introduction

Introduction

Petroleum and natural gases are currently the main energy source used in the world, but the amount of viable fossil fuels are slowly depleting. Also, research is being done to produce more environmentally friendly fuels to combat green house gas emissions produced from petroleum. Scientists have turned to biofuels as an effective alternative to fossil fuels. Until now, bioethanol has been the primary biofuel, because it is economically favorable to produce and easy to manufacture. It is also a renewable fuel that is made from agricultural feedstock. However, biobutanol is proving to be much more advantageous than bioethanol.

Compared to bioethanol, biobutanol has more energy per gallon, thus more miles per gallon. Biobutanol has 110,000 BTUs per gallon, while bioethanol only has 84,000 BTUs per gallon. Butanol can also be blended with gasoline at much higher levels than bioethanol without any necessary engine alterations, because its physical attributes are more similar to gasoline. It has lower vapor pressure, which makes it safer to store, and handle. Because biobutanol is a better biofuel than ethanol, this project has designed a process to create this solvent. This new plant will hopefully replace ethanol plants, so it was mirrored as closely as possible to the ethanol process. The economic feasibility of the plant was investigated, and the cost was compared to that for producing ethanol[1].

[1] Ramey, David, and Shang-Tian Yang. Production of Butyric Acid and Butanol from Biomass. Tech. Morgantown: U.S. Department of Energy, 2004.

Traditionally, an Acetone-Butanol-Ethanol (ABE) process is used for biobutanol production. This one stage batch process uses a Clostridium strand (generally C. beijerinckii or C. Acetobutylicum) to produce a mixture of butanol, acetone and ethanol. First, fermentation produces a mixture of butyric, lactic and acetic acid. Later, the culture pH drops and butanol, acetone and ethanol are produced in a 6:3:1 by mass ratio respectively. The main problems with this process are the low conversion of glucose to biobutanol, and the large amount of undesired solvents produced. Also, the process is very complicated and difficult to control, so its use has dramatically declined since the 1950s. Now, butanol is mostly produced via petrochemical routes, which is not eco-friendly. This, however, is not an environmentally conscious method of making butanol[2].

David Ramey of Butyl Fuel, LLC, has created a process that uses two-stage anaerobic fermentation to produce green butanol with higher specificity and efficiency than the ABE process. Up until now, there have only been laboratory-scaled productions of butanol using this process. The objective of this project is to design a scaled up process that can produce at least 50 million gallons per year of butanol at 99.5% pure with less than 10ppm acetone. It will be a challenge to create a high yield of butanol. The separations process must also be able to handle a much larger amount of undesired solvents and water when the process is scaled up. Also, sterility will be a main concern for any stream entering and coming out of the fermenters. The process must be as energy efficient as possible, meaning energy use should be no more than 35,000 BTU per gallon of butanol.

[2] "BUTANOL Advances in Biofuels." The Light Party. 03 Apr. 2009 <http://www.lightparty.com/Energy/Butanol.html>.

The economics of the plant will be evaluated to determine if this plant is economically feasible and what the lowest price possible is to license the technology from ButylFuel. A profitability analysis will also allow the direct comparison to bioethanol plants.

In order to accurately compare the designed process to that of ethanol plants, the plant must draw as many parallels with the bioethanol plant as possible. A typical procedure for the production of bioethanol is dry grinding. In this process, corn is milled and mixed with water to form a slurry. Corn costs about $4.00 per bushel, and it takes 3.74 bushels of corn to create one gallon of butanol from this process. This slurry is passed through a liquefaction and saccharification stage to break up the starch into glucose to be sent for fermentation using different amylases and sulfuric acid. Any unfermented biomass out of the fermenter is separated out and dried to produce DDGS, an animal feed co-product, which can be sold at $150 per dry ton. Both of these processes are also used in the designs for the butanol plant with a different fermentation and separations sequence in the middle.[3]

The fermentation portion uses two series of fibrous bed reactors, one series converting the glucose to butyric acid, and the other converting the butyric acid to butanol product. The product stream from fermentation is sent to the separations train, which includes liquid-liquid extractors with dodecane as the solvent, and distillation columns. The plant will be built in the Midwest near the source of corn production, allowing closer access to raw feed material. It will operate for 330 days a year. Because there will be little access to water, it

[3] Kwiatkowski, Jason, and Andrew McAloon. "Modeling the process and costs of fuel ethanol production by the corn dry-grind process." Industrial Crops and Products 23 (2006): 288-96. 1 Feb. 2009 <http://www.elsevier.com/locate/indcrop>.

will be a near zero-discharge plant, meaning as much of the process water possible will be recycled within the plant.[4]

The main reason for the production of biobutanol is to have an eco-friendly fuel for the future. As seen above, butanol is made from a biological material, thereby reducing greenhouse gas emissions. Also when burned, it produces no SOx or NOx, making it very environmentally beneficial. Because butanol is an organically friendly solvent, the design of the plant should also be very safe for the environment. The fermentation off-gas contains hydrogen will be burned off and safely disposed of. The rest of the gas is CO_2, which can be collected and used as means to help grow the corn feed. Also, the DDGS sold as animal feed will be made safe for ingestion.

[4] Ramey, David E. Continuous Two Stage, Dual Path Anaerobic Fermentation of Butanol and Other Organic Solvents Using Two Different Strains of Bacteria. Environmental Energy, Inc., assignee. Patent 5753474. 1996.

Importance of the Study

The population of the world is slowly growing every day, and this is increasing the demand for energy and fuel. From 1980, the consumption of oil has risen about 1,830,000 barrels per year in the United States alone. In Asia, it has risen 5,480,000 barrels per year, as seen in Figure 1. From 2000-2030, the world primary energy demand is expected to double.

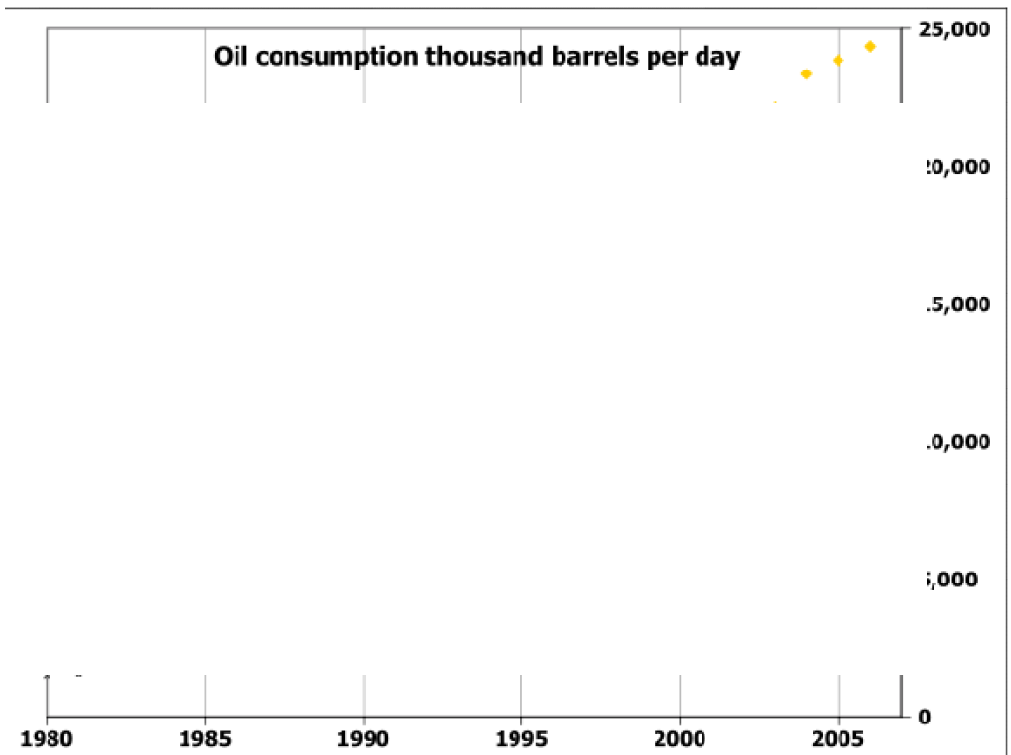


Figure 1: The oil consumption in thousands of barrels per day for the world shows a steady increase in demand for fuel in almost every single region of the world over the

With such a large demand in fuel, and the depletion of natural gases and petroleum, it is necessary to find renewable forms of energy. Biofuels are a great alternative form of energy. Present estimations show that the world oil production will peak sometime in the next 10-15 years. Also, the consumption of so much oil is causing harm to the environment through the large amount of CO_2 emissions. The CO_2 emission from a gallon of gasoline is 19.4 pounds. With 42 gallons per barrel, this is 171,000,000 pounds of CO_2 emitted per year in just the United States[5]. Biofuels will help to reduce the amount of green house gases. Even though CO_2 is emitted during fermentation and combustion of biofuels, it is cancelled out by the greater amount taken up by the plants used as raw feed material into biofuel producing plants. Also, because it is domestically produced, this will be beneficial to the internal economy by helping the agricultural markets and markets for other domestically make products well as creating new jobs[6].

The main obstacles the biofuels market faces are the cost of production, lack of favorable regulatory regimes, the cost of technology transfer, and scarcity of land available for growing biomass. Also, economic issues are correlated with the high prices and limited availability of organic feed products , selling prices of biofuel co-products such as DDGs, and the amount of energy used within the process.

Biobutanol has many advantages over other forms of biofuels. Because it is a four-carbon alcohol, it doubles the amount of carbon in ethanol, thus contains more BTUs per molecule. This translates to more miles driven per gallon. The specific energy per gallon of each biofuel is discussed in the conclusion. Butanol comes very close to the same fuel value as gasoline. Butanol can also be blended with fossil fuels at much higher levels than ethanol, because it is closer to gasoline physically than ethanol. It can eventually replace gasoline

---

[5] "Emission Facts: Average Carbon Dioxide Emissions Resulting from Gasoline and Diesel Fuel | US EPA." U.S. Environmental Protection Agency. 03 Apr. 2009 <http://www.epa.gov/otaq/climate/420f05001.htm#calculating>.

[6] Biofuel Guide - Ethanol and Biodiesel as alternative energy. 03 Apr. 2009 <http://biofuelguide.net>.

one to one without making modification to the engine, while ethanol can only replace 85% of the gasoline in a blend. There is also a less separation in water. Safety is always a key concern in the plant, and it is a big advantage that butanol is much safer than bioethanol. It has a Reid value of only 0.33 compared to 2.0 for ethanol, which means that it is much less volatile with less fear of explosion and no need for special blends during the summer months. Also, it is less corrosive than ethanol so safer to ship through existing pipelines.

Project Charter

Project Name	Economically Viable Production of Butanol by Two-Stage Fermentation
Project Champions	Bruce Vrana (Dupont), Leonard A. Fabiano
Project Leaders	Christina Chen, Amira Fawcett, Amy Posner, Tal Raviv
Specific Goals	- design process and plant to create 50,000,000 gallons per year of butanol – appropriately scale up from laboratory-scale findings from patent by David Ramey - 99.5% butanol purity in product stream with less than 10 ppm ketones - processing of unfermented biomass to be sold as animal feed – concentration of solvents must be below toxic level - sterilization of fermentation bacteria for any output streams - total conservation of water – recycled within plant minus purges - environmentally safe and efficient means of waste disposal – H2, purge streams, CO2, other solvents - economically sound process – approximately the same cost as ethanol, and financially feasible
Project Scope	In Scope: - Butanol by Fermentation of Corn - Process Similar to that of Ethanol – Dry Grind Out of Scope: - Butanol by Fermentation of Cellulose and Other Forms of Sugars - ABE Fermentation Process
Deliverables	- Business Opportunity Assessment - Technical Feasibility Assessment
Timeline	- Process Design within 4 months

Figure 2. Project Charter

Section III: Innovation Map

Technology-Readiness Assessment

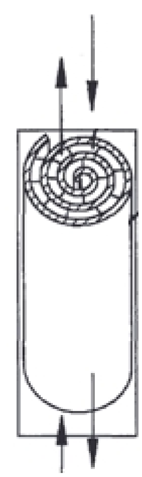
The biobutanol process mirrors many aspects of the traditional bioethanol process. However there are many technological innovations within the biobutanol process that optimize its production. Both processes involve the liquefaction and saccharification processes to produce glucose from corn, as well as the drying of unused solids to produce DDGS. The most striking differences between the butanol and ethanol processes involve fermenter design, continuous vs. batch operation, and the microbes used in the fermentation process. The benefits of the butanol process design are highlighted with a comparison of traditional ABE processes. .

The ABE process has been in use since the 1920s to produce acetone, butanol, and ethanol in a 6:3:1 ratio of butanol, acetone, and ethanol. This has traditionally been executed by fermenting a form of starch in a batch vessel using *Clostridium Acetobutylicum* as the microbe. With this design, the bacteria strain undergoes two growth periods: an acidogenesis phase followed by a solventogenesis phase. In the acidogenesis phase, the starch source is converted into acids, more specifically, butyric acid, acetic acid, and lactic acid. When these substrates are in a high enough concentration, the bacteria then transitions into its solventogenesis growth phase. Once this occurs, the acids in the vessel are converted into the three desired solvents: butanol, acetone, and ethanol. Using this method, a significant smaller amount of ancillary solvents (acetone and ethanol) is produced. Because this process is operated in batch, a limited amount of product will be produced since the bacteria growth is significantly inhibited at a solvent concentration as low as 13 g/L.

The technology used in this process differs from the traditional method in several respects. For example, this process is in continuous operation, which prevents solvent accumulation and cell inhibition. The concentration of solvents in the reactor never exceeds 6g/L, which is far from the bacteria's tolerance level. The Fibrous Bed Bioreactor—which will be discussed in depth below—also improves the productivity of the bacteria, leading to greater product yield. Also, two fermenters are used, one in only the acidogenesis phase, and one operating in only the solventogenesis phase.

The bacteria strain *Clostridium Tyrobutyricum* is used for the acidogenesis phase. This strain is unique because it will only convert glucose into the three acids, but will not convert any acids into solvents in the conditions of the fermenter. This particular bacteria also is highly selective in that it will convert the glucose into a greater amount of butyric acid compared to lactic acid and acetic acid. This is done by introducing a non-replicative integrational plasmid containing pta gene fragment into the bacteria using electroporation. This mutant is able to produce 15% more butyric acid and 14% less acetic acid. Conditions such as pH and glucose feed were optimized to sustain this behavior.

Figure 3. Schematic of Fibrous Matrix, showing flow channels through fibers.



The stream exiting the first fermenter is fed into the second solventogenesis fermenter, which utilizes *Clostridium Acetobutylicum* as the microbe. While this bacteria traditionally is used to produce both acids and solvents, under the specified conditions, it will only convert the existing acids into solvents. The presence of the acids in the feed stream will induce *Clostridium Acetobutylicum* to enter its solventogenesis growth phase. Therefore, it will not utilize any of the available glucose in the feed stream to produce more acids; its only activity will be converting the three acids into butanol, acetone, and ethanol. The butyric acid to glucose feed ratio as well as the pH will be optimized to sustain this behavior.

The Fibrous Bed Bioreactor (FBB) is another unique design in this process. Designed by ST Yang of The Ohio State University, it has many attributes that significantly enhance both the reaction yield and productivity. It is essentially an immobilized bed reactor, where the cells are immobilized on a roll of fibers, in this process, cotton fibers. It is a very unique design, with many advantages, in particular for a continuous fermentation. For example, it can run continuously for over one year without the need for maintenance or downstream processing. Figure 3 shows how the fibers are organized within the bioreactor[7]. The orientation of the fibers creates many channels which the reactor contents flow up through. Generally, 12% of all biomass are in the bulk fluid, 58% is weakly attached to the matrix, and the remaining 30% is strongly attached to the matrix.

The bioreactor has a very large void space (greater than 90%) as well as a large surface area, which allow for both a greater cell density (40-100 g/L) as well as a greater productivity, leading to much faster reactions. Because of the continuous fermentation

---

[7] Yang, Shang-Tian. Extractive Fermentation Using Convoluted Fibrous Bed Bioreactor. The Ohio State University Research Foundation, assignee. Patent 5,563,069. 1996.

operation, fresh medium is constantly being supplied, and solvents that normally inhibit cell growth are constantly being removed. These bioreactors are also self-renewing, meaning that the as one cell dies, more grow in its place because of the continuous supply of nutrients. Therefore no new cells need to be introduced into the bioreactors after the biomass adheres to the fibrous surface during the initial inoculation and immobilization period. This design also allows for dead cells to fall to the bottom of the vessel, while allowing gases that are produced during the process (CO_2 and H_2 in this process) to easily flow up through the channels, helping to mix the contents of the reactor.

Both ethanol and butanol processes produce CO_2 and H_2 during fermentation. However, the handling and disposal of the gases is quite different. In the ethanol process, a scrubber is used to separate out the CO_2 from the waste gas from the fermenter. A scrubber is optimal here, because there is a large amount of ethanol product in the gas stream that must be recovered as well as process water. In the butanol process, however, the temperature of the gas out the fermenter is not hot enough to have a significant amount of vaporized butanol or water, so the stream is mainly $CO_2$and H_2 in a 44:1 ratio by mass. Therefore, an additional scrubber is not needed to recover product, and the gas stream is fed to the thermal oxidizer in the DDGS drying process to burn off the excess H_2 and the trace amounts of solvents to get a pure CO_2 stream. The CO_2 can then be sold or safely discarded. Note that in the butanol process, these gases are not simply discarded as they are in the ethanol process. The effluent gases perform an important function in the process. This stream is recycled back into the fermenters in order to air-lift and mix the contents of the reactors, as well as to control the pressure of 5in water gauge to maintain anaerobic conditions.

Recovery of the butanol is traditionally done using distillation in the typical ABE process. Butanol has a higher boiling point than water, so this tends to have high energy demands and cost. Butanol and water also have an azeotrope that is unbreakable by distillation only. Other methods of separation were examined to overcome these problems. Gas stripping is the method used in the patent by David Ramey. Warm CO_2 pulls the butanol from the product stream. The butanol rich CO_2 stream is then pumped to an activated carbon adsorption bed to extract the butanol out. The adsorbed butanol is then stripped with another stream of warm CO_2, and the resulting stream is condensed to produce a nearly pure butanol stream. This process proved undesirable, because of it can't handle the large throughput from fermentation. Decanters were examined as a means of overcoming the azeotrope. Pervaporation was also considered as a potential separator. This process allows selective permeation of the butanol and potentially other solvents through the membrane while the water stays behind. Many butanol models use pervaporation with silicon rubber membranes or polypropylene membranes. The same problem as with stripping occurs here. Decanters use phase separations to create a water rich and butanol rich stream. However, the concentration of butanol out of fermentation is too low (around 1%) for phase separation to occur in the decanter. Liquid-liquid extraction separates components based on relative solubility in two different immiscible liquids. There is no added energy needed for the extractor to operate. Here, butanol can be extracted from mostly water into a solvent stream. The solvent used is dodecane. This method is favorable, because no azeotrope exists between dodecane and butanol. This allows direct distillation to be used to separate the butanol from the dodecane. Also, a smaller flow rate of dodecane is needed than of the original amount of water in the product stream out of fermentation. Now, the rest of the separations process has less mass to deal with, and this means less cost in buying and maintaining the distillation columns. Upon technical and economical analysis of the different separation technologies, the liquid-liquid extractor followed by distillation columns was the most efficient at separating out the butanol, and most economical.

Innovation Map

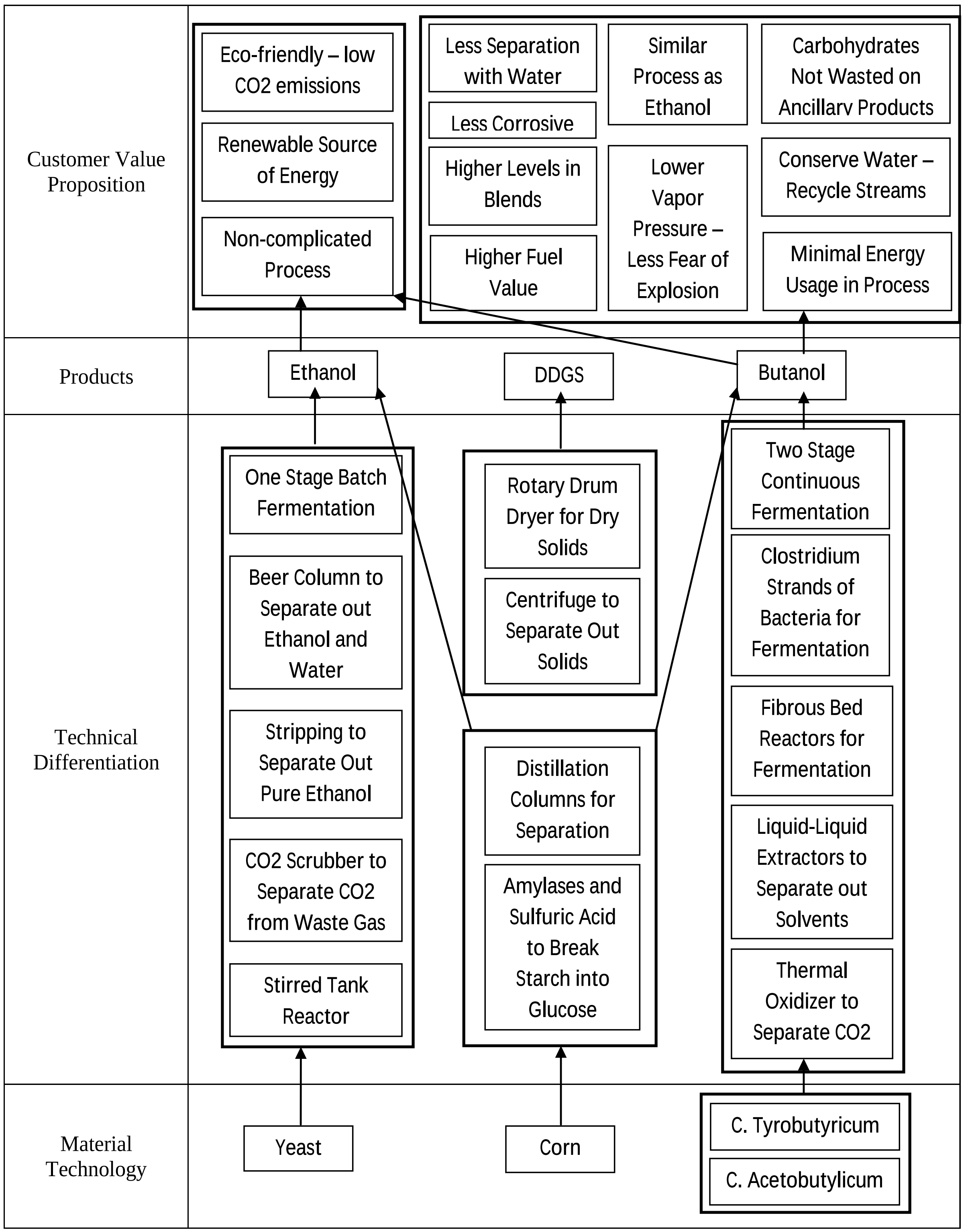


Figure 4. Innovation Map

Section IV: Concept Stage

Market Analysis

Market Outlook

Within the next thirty years, global energy demand is projected to double due to expanding population and developing economies. Of this demand, 14% is supposed to come from China. Almost 80% of the world's energy supply is currently derived from coal, gas, and oil. Between now and 2030, oil and gas will account for 60% of the world's increasing energy demand. The growing scarcity of fossil fuels will continue to push the price of these resources. In the coming decades, the inevitable uneven distribution of fossil fuels will push energy security into the spotlight as a critical economic and political issue. In fact, the recent party change within the United States is indicative of the shift towards making energy a top priority. President Obama's recently announced stimulus package is one of many measures that have been undertaken to injecting funds into the pursuance of alternative energy. Furthermore, in 2005, the US Energy Policy Act was passed, detailing a comprehensive legislation that included the Renewable Fuel Standards which was aimed at doubling the use of biofuels by 2012. While these new measures and projections of alternative energy growth seem optimistic, industry analysis suggests that biofuels will not be economically competitive in the short to medium term; however, biofuels do have the potential to capture 50% of total global fuel production in the next 50 years[8].

The biofuels market has been steadily increasing over the past few years. There are currently 10.3 billion gallons of biofuels sold per year throughout the world, amounting to

[8] Westwood, Gary. "The Biofuels Market Outlook"

roughly $27 billion. The production of biofuels is projected to grow to 87 billion gallons/year by the year 2020. Ethanol, the most prevalent biofuel, is expected to reach 27 billion gallons by 2014. Brazil and the US lead in the production of bioethanol, producing 65% of the world's ethanol, while the EU makes around 13%. The EU is also a leading producer of biodiesel, with approximately 85% of the world's biodiesel coming from the EU. This industry is projected to reach 2.9 billion gallons in 2014[9].

The biofuels market can be broken into two major components: biofuels and biomass. In 2005, this market accounts for almost 21% of the world's total renewable energy production. The United States, Brazil, and Germany are the present day world leaders in integrating ethanol, and biofuels, into their energy infrastructure.

Within the category of biofuels, many are familiar with the present day techniques of blending fuel ethanol with gasoline. Ethanol still leads the race in biofuel production, but another feedstock based product, butanol, is on the rise.

Unlike fossil fuels, butanol production is carbon neutral and, as a corn based product, is a renewable resource. There is no current competition in the market for a viable ethanol replacement using Butanol, minimizing barriers to entry that could be present from incumbents in the market. There are many key aspects of differentiation that butanol has over ethanol. For one, butanol can achieve higher levels in gasoline blends since it is less volatile. Additionally, butanol can integrate with existing gasoline infrastructure (such as gasoline pipes, stations, etc.) with minimal overhaul since it is less corrosive and has less

[9] Scott, A. et al. "Alternative Fuels: Rolling Out Next Generation Technologies." Chemical Week. New York, 27 December 2006, 168:45, 17-20.

separation when in contact with water, as compared to ethanol. Most importantly, butanol has the potential to outperform ethanol in CO_2 savings, with current fuel ethanol saving 30g CO_2/km.

Drivers

There are three main drivers for biofuel adoption: growing environmental concerns, energy security, and the rising cost of fossil fuels. Growing environmental concerns over CO_2 have pushed governments around the world to call for a reduction in the emission of greenhouse gases. The recent Bali Climate Conference, is one such example of governments meeting together to discuss and implement CO_2 emission reduction targets. Biofuels, like butanol, have a significant potential to reduce emissions of greenhouse gases as compared to fossil fuels.

Furthermore, increased pressure on energy security has forced governments and agencies to focus on mitigating the future threat rising costs of oil and gas pose. It is no surprise then that many biofuel projects, like butanol production, rest on the policy and incentives governments create to address this growing problem. Solutions such as financial incentives, crop subsidies, levies, excise duty exemptions, investment support, and tax incentives could all encourage adoption of biofuels into the market. The effectiveness of these implementations are exemplified in Sweden, Italy, and Spain. These countries provided general fiscal incentives for adoption of biofuels, and are moving forward in the assimilation of alternative energies.

One important factor in the future production of biofuels is the availability of arable land to farm for fuel. Countries with significant land and water resources tend to be leaders in the production of biofuels. Modern day examples of this fact include; Brazil, who specializes in ethanol derived from sugar cane, the U.S, who specializes in corn derived ethanol, and Germany, who utilize sugar beet to produce ethanol. In the long-term, India and China have the potential to be strong leaders in biofuel production.

Inhibitors

The largest obstacle facing the biofuels market is the cost of biofuel production. In order to become competitive with current fossil fuels, biofuel production must be cheaper or the same price as traditional gasoline production. With regard to butanol, for butanol to become a major player in the biofuel market, it must be the same as or less expensive to produce than ethanol. While butanol does have a higher energy value than ethanol, the current production techniques require substantially more money to implement than ethanol and do not generate any cost savings from switching fuels. Production costs take into consideration the process energy used, and the prices for feedstock and byproducts. These high production costs usually involve a degree of volatility as feedstock prices fluctuate.

The availability of land is an essential factor to the biofuels market. This driver is also a potential inhibitor as concern rises over the scarcity of land and sustainability concerns over large areas of agricultural land. This allocation of arable land for biomass production could also lead to an increased demand in the price of food crops, further hindering the acceptance of biofuels into existing infrastructures.

Overall, the adoption of biofuels is not barred by modern technology solutions, but by implementation. On the whole, securing investments and contracts is the biggest concern in launching the biofuel industry. The subsequent expansion is then heavily dependent on future government policies and frameworks to stimulate the market.

Competitors

Ethanol

Ethanol is the most commonly used liquid biofuel worldwide. Ethanol's attractiveness is derived from its ability to supplement gasoline at any percentage due to similar volatilities with gasoline. However, once 15% ethanol by volume has been reached, a vehicle's fuel system must be adapted, which is a costly procedure that most car manufacturers are unwilling to make.

Current production techniques of ethanol involve fermentation of sugars (such as sugar cane) or starch (such as corn). Brazil leads the world with the lowest commercial cost of ethanol production from sugar cane. In fact, Brazilian sugar cane ethanol is already in competition with production costs for diesel and gasoline, indicating that it is achievable for biofuels to become competitive with fossil fuel production costs. Industry analysis suggests that the cost of ethanol produced from corn is supposed to drop almost 20% between 2004 and 2010. Increasing advances in production technology will continue to drive down the cost of producing ethanol from sugar cane and starch. By 2010, Brazil and the U.S are expected to be the main suppliers of ethanol. At that time, European ethanol production from beet and wheat will become a 20 billion euro market; however, they will not meet the projected demands for the region.

Ethanol is not without its disadvantages. A huge problem across the industry is the issue associated with water mixing with ethanol and gasoline, causing a phase separation. When drawn in, the mixture can cause an engine to stall. Subsequently, these blends cannot be transported through existing gasoline pipelines since there is a high risk of moisture mixing along the way.

Biodiesel

Another direct competitor to butanol is biodiesel. Biodiesel fuels are derived from the esterification of oil or fat, making it one of the cheapest biofuels to produce. Current production techniques involve oilseeds like rapeseed and sunflower seeds, as well as vegetable waste, and animal fat. Like butanol, biodiesel involves low carbon emissions and could reduce carbon emissions by up to 40%. Furthermore, biodiesels do not involve costly modifications for integration with modern biodiesel engines and existing diesel infrastructure. There are some vehicle manufacturers who have begun to create vehicles that can run on pure biodiesel since the energy content is roughly 90% that of petrol diesel.

It is important to consider that, at this time, there is no market for high blends and pure biodiesel. Additionally, biodiesel attacks certain rubbers, elastomers, and paints, limiting its integration with all forms of fuel infrastructure. The cost of producing rapeseed biodiesel is much higher than the production costs of petrol or diesel, further deterring the integration of this biofuel. Substantial industry subsidies are required to make this fuel competitive in the near future.

Other Technologies

Two other technologies included in the industry are biofuels produced from solid waste and biogas.

Solid biofuels are made up of fuels produced from wood, charcoal, dried animal excrement, peat, and waste materials from crops. Since these are natural waste products from crops, there is no net release of CO_2. Additionally, utilizing crop waste mitigates concern over the use of arable land for fuel production instead of as a food source.

Gas biofuels, or biogas, is produced from the anaerobic digestion of organic material by microorganisms. The quality of biogas produced from these processes is similar to natural gas, allowing for easy integration with modern natural gas infrastructure. The main issue associated with this procedure is the large production of CO_2 and CH_4 which conflicts with greenhouse emission reduction goals.

Future Technologies

First generation biofuels are all threatened by advances in second generation biofuel production. Research into biofuels created from lingocellulosic, or non-food feedstock, suggest the potential for a more efficient and cleaner production process. A significant advantage advanced biofuels have over existing technologies, is that the net yields from perennial crops, grasses, and sugar cane have the potential to be much higher and can be grown on less valuable (arable) land. Additionally, the low cost of production and higher energy conversion makes the future of advanced biofuels more optimistic as compared to current methods.

Customer Requirements

Scope

The scope of the project was to determine if production of butanol via two strains of bacteria is economical in the short-term. Since current biofuel production is heavily based in ethanol, and the plant would be competing with over 100 fuel ethanol plants in the United States, the process design was to draw many parallels to the typical ethanol fuel production process.

With these constraints, the plant was charged with producing at the scale of 50,000,000 gallons of butanol per year, at 2009 prices of $4.00 per gallon. The process, as designed, offers a yield of approximately 54,000,000 gallons of butanol per year. Additionally, the dried distiller's grain solids (DDGS) that are produced during the corn mill process can be sold at $150 per dry ton.

Economic Considerations

After the initial investment requirements and profitability analysis was run on the process, the sensitivity to corn price was determined. In the past few years, the price of corn has fluctuated dramatically, pushing this variable to the top of the list in order of importance. The calculations and sensitivity analysis on this issue are detailed in the Financial Summary section of the report. Our process did not achieve an IRR at a butanol price of $4/gallon, and resulted in a negative NPV of $3.55 billion. .

Sterilization

A significant customer concern was the issue of sterility in the fermentation process. To mitigate this risk, sterile water was used for start up and recycled through the system. In order to ensure that no contaminant entered the process and generated undesired products, every stream entering a fermenter was sterilized to deactivate or kill any remaining organism. These sterilization procedures were implemented around the facility. Heat exchangers were placed in key positions to heat streams to 250°F to kill any microorganism remaining in the stream. These exchangers were placed before each fermenter series and in the recycle process water stream to ensure sterilization at every step. The product leaving the second fermenter is passed through a centrifuge to aid in the separations of solids from the butanol stream. Centrifugation removes virtually all biomass and exerts shear force currents on the living organism membranes. These solids are sent to the DDGS dryer where, at high temperatures, any trace organisms in the byproduct of feed are killed. During the separations process, the water stream enters a liquid-liquid extractor where it thoroughly interfaces with dodecane, a hydrophobic solvent, which disrupts cell membranes and deactivate all cell function. The exit water stream from the extractor remains 50°F hotter than the livable fermentation temperature for *Clostridia Acetobutylicum*, ensuring that any remaining bacteria is eliminated before recycling back to the first fermenter. In addition to these precautions, any buildup of biological contaminants in the water recycle is mitigated by the intermediate sterilization step between the two reactors. This sterilization reaches 250 F and ensures no circulation or buildup of foreign organisms.

Off-Gases

In the first fermentation series, hydrogen gas is produced as a byproduct. The process produces 3,700,000 lb per year of hydrogen gas, which represents only 2.3% of the gas stream. Since this amount is too small to serve as a source of heat, the gas stream containing carbon dioxide and hydrogen will be passed through the furnace included in the DDGS dryer system. This furnace will burn off any hydrogen in the stream, leaving a pure CO2 stream. The CO_2 collected can either be sold or released. The process produces 161,000,000 lb per year. The demand from soda companies for CO_2 isn't enough to encompass our CO_2 emissions, so the only possible method of disposing the gas is by releasing it. Currently, permits for the release of CO_2 are sold at $3 per ton, so that is an added $241,000 per year in permit fees. However, it is important to note that the butanol process as a whole is carbon-neutral, and the carbon dioxide being released is considered to be "green CO_2." The definition of what constitutes as "green off-gas" will be discussed further in Environmental Considerations.

Equipment Considerations

Since the process involves a large percentage of water, it was pivotal to construct the majority of the equipment out of stainless steel. As such, all six fermenters contain stainless steel interiors and piping, which can be maintenance every 15 years at $8.00 per square foot of stainless steel. Additionally, any heat exchanger involving the passing of a product stream (containing water and solvents) was specified as stainless steel to ensure no corrosion. All other equipment involving less than 1% of water was constructed out of carbon steel.

Energy Benchmarks

Current energy benchmark is about 35,000 BTU per gallon of product in fuel ethanol plants (Determined from the amount of heat and electricity needed by the process). An analysis of our heat and electrical requirements for the process resulted in an energy benchmark of

19,900 Btu per gallon of butanol produced. It would seem that the initial design of the process is exceeding current ethanol capabilities.

Water

In order to ensure a zero-discharge plant, all process water is recycled within the plant. The total amount of water needed at start-up is 6,810,000 gallons. This water will continually travel through the process via a recycle stream that occurs at the Liquid-Liquid Extractor stage. The recycled water is then used to dilute the incoming stream of glucose from the corn mill, which then re-enters the fermentation phase.

DDGS

The biomass solids from the drying process must be thoroughly cleaned to lower the butanol and acetone concentrations to below toxic levels, in order to be sold as animal feed. The butanol LD50 in rabbits is 3,400 mg per kg of rabbit. The main dangers are from prolonged exposure. In extreme cases this includes suppression of the central nervous system and even death. For acetone, ingestion of 200 mL has produced severe coma, hyperglycemia, and acetonuria in adult rabbits. This would approximate a dose of 2 to 3 milliliters per kilogram. This is not a concern in the ethanol plant, because ethanol is not as toxic as butanol and acetone during ingestion.

Section V: Process and Unit Descriptions
Process Overview

Process Description

The fermentation process involves two strains of bacteria converting glucose derived from corn into butanol in two phases: acidogenesis and solventogenesis. The first phase, acidogenesis, consists of three fibrous bed bioreactors, run in parallel, containing Clostridia tyrobutyricum. The Clostridium tyrobutyricum will take the glucose feed stream and convert the stream to butyric acid, with small concentrations of lactic acid and acetic acid. These product streams are then sent through a series of three heat exchangers that sterilize the streams at 250°F, and then cool back down to 98.6°F before entering the solventogenesis phase. In this phase, three fibrous bed bioreactors containing Clostridia acetobutylicum convert the acids into solvents. The product stream, containing butanol, acetone, and ethanol, are then pumped to a centrifuge where solids are removed for Dried Distillers Grain (DDGS) drying and liquids are sent to the separations process.

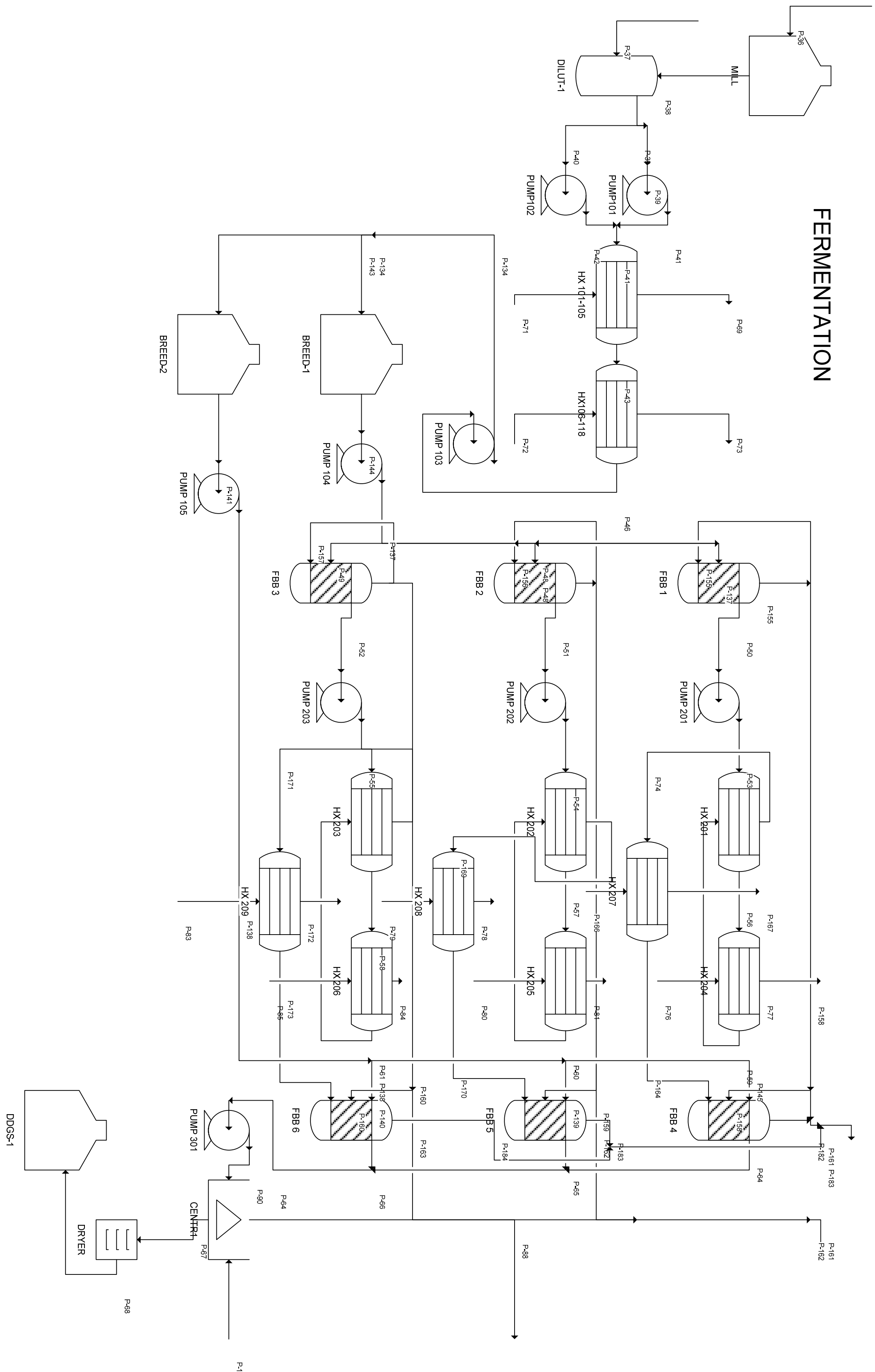
Separation of butanol from the reactor effluent is designed for a recovery of 96.5% of product. Solids are the first component to be removed by centrifugation, a step shared by the DDGS drying process. Next, liquid extraction is used to reduce the large stream size by transferring to a solvent. Water with trace amounts of all solvents is recycled from the extractor to the upstream fermentation process section, where it is used to dilute glucose produced by the dry-grind process.

Two distillation units in series (one parallel pair of columns and a final column) are subsequently used to remove butanol, ethanol, acetone, and water from the solvent, as well as separate butanol from these secondary products. No azeotropes are encountered except for the very last column, but 99.5% butanol is achieved. The solvent is recycled to the extraction, while residual products are either discarded or oxidized in the drying process.

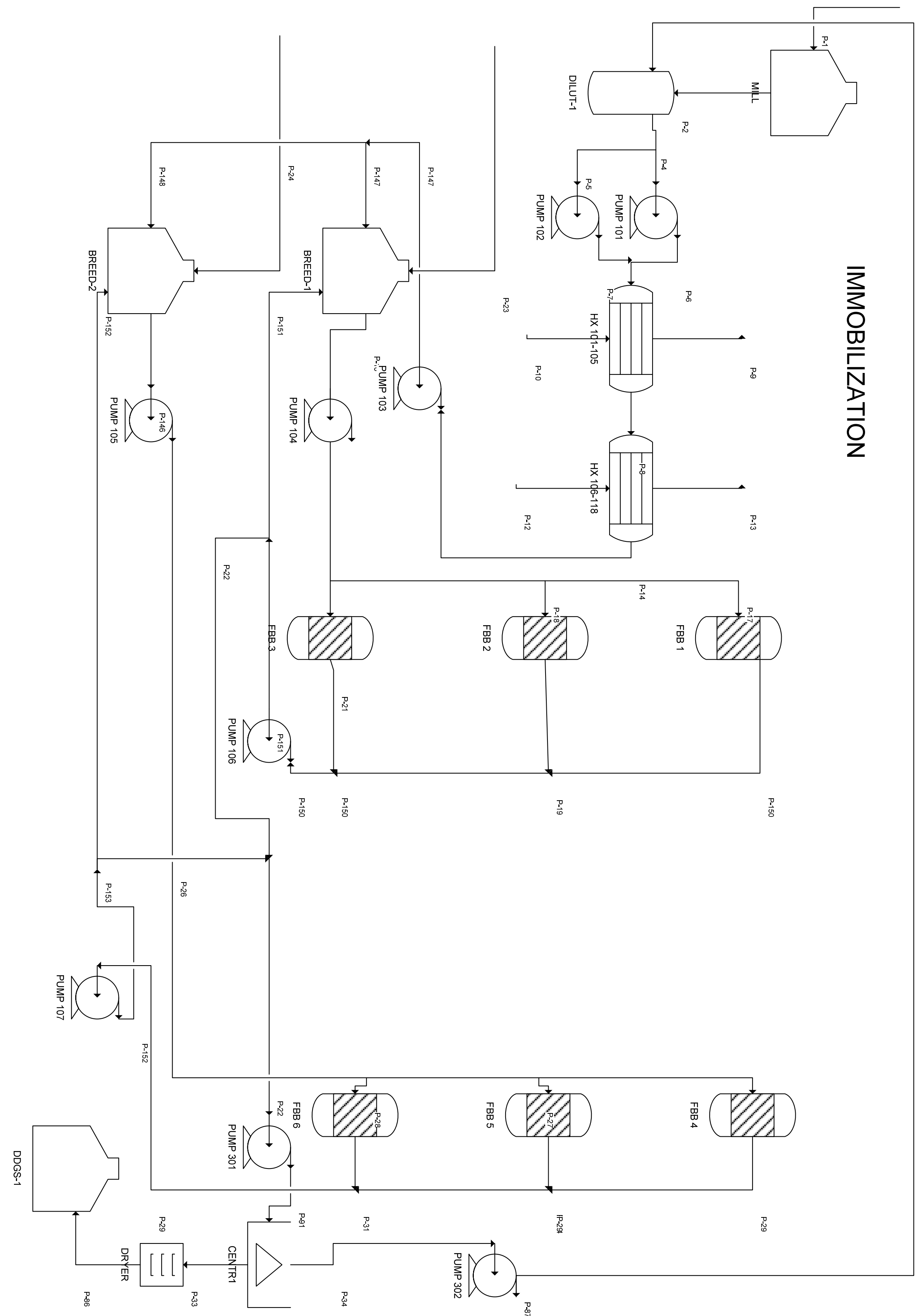
The beginning portion of the dry grind process to create bioethanol is used to generate the glucose needed for fermentation from corn feed. Corn is brought into the plant and stored in a holding tank until ready to use. The corn is milled down to remove husk, germ and shells to be easily broken down to simple sugars, and added to water to form a slurry. The corn then undergoes liquefaction to gelatinize the starch and hydrolyze it into dextrins. Further conversion of the sugars to glucose occurs in the saccharification stage. Sulfuric acid is used to reduce the pH so hydrolyzation is more efficient. The stream out of saccharification has a 65% by mass glucose, with the rest being water and leftover unfermented biomass solids, and it is ready to be sent to the fermenters and breeder tanks.

The product stream out of fermentation has contains unfermented solids that need to be separated out before the rest of the stream is sent to the separations train. Also, the solids can be sold as animal feed called DDGS for $150 per ton or $0.07 per lb, which is a significant source of revenue. The final portion of the dry grin process describes how the solids are processed. Centrifugation is used to separate the solids and liquids. The liquid stream is pumped to the liquid-liquid extractor, while the solids undergo drying in a rotary drum dryer to become DDGS.

Fermentation Process



Visio Design of Immobilization Process



Fermentation

Fermentation Overview

Each fermentation phase will have a separate breeder tank containing its respective bacteria and growth medium. These breeder tanks will be mostly active during the inoculation and immobilization phases. Inoculation occurs during the first two days of each production year, and immobilization follows for the next eight days. During immobilization and inoculation, the rest of the plant is shut down for cleaning and maintenance. For the next 320 days, the fermentation feed is passed through the breeder tanks.

Both fermentation phases will be fed a glucose solution from a large glucose holding tank. This tank will serve as a repository for glucose produced in the initial corn milling stage. Pumps will regulate the amount of glucose that enters each fermenter based on process specifications. Additionally, the holding tank will allow for the dilution of the glucose stream before it enters the fermentation process. The water stream recycled from the Liquid-Liquid Extractor in the Separations Process is used for all dilutions, as it feeds into the holding tank where it is mixed with the concentrated glucose.

At the end of the fermentation process approximately 8,400,000 pounds per hour of product will be entering the distillation stage. The product stream will be made up of 0.564% butanol, 0.0125% ethanol, 0.0250% acetone, and 1.10% biomass.

Fermentation Process Description

Heat Exchanger Unit HX 101

This shell and tube heat exchanger is used to preheat the unsterilized stream leaving the dry grind process with the sterilized stream exiting HX 102. Both sides of this unit are constructed from stainless steel, and it has an area of 25,680 ft^2 with a heat transfer coefficient of 74 btu/F-ft^2-hr. The cold stream enters the heat exchanger at 140°F and exits at 240°F. The hot, sterilized stream enters the heat exchanger at 250°F and exits at 130°F. There is a 10psi pressure drop across the exchanger.

Heat Exchanger Unit HX 102

This shell and tube heat exchanger is used to sterilize the preheated stream exiting HX 101. The product stream is on the tube side, which is constructed from stainless steel, and the steam is on the shell side, which is constructed from carbon steel. This unit has an area of 18,720 ft^2 with a heat transfer coefficient of 100 btu/F-ft2-hr. The cold stream enters the heat exchanger at 240°F and exits at 250°F. This stream is heated using 1,325,000 lb/hr of 50 psig steam which enters the heat exchanger at 281°F, and exits at 267°F. There is a 10 psi pressure drop across the exchanger.

Heat Exchanger Unit HX 103

This shell and tube heat exchanger is used to cool the sterilized corn slurry leaving HX 101 back down to 98.6°F, so it can be diluted with process water and fed into the fermenters. The product stream is on the shell side, which is constructed from stainless steel, and the cooling water is on the tube side, which is constructed from carbon steel. This unit has an area of 29,000ft2 with a heat transfer coefficient of 200 btu/F-ft2-hr. The hot stream enters the exchanger at 140°F and exits at 98.6°F. This stream is cooled using 1,042,000 lb/hr of cooling water that enters the exchanger at 90°F and exits at 120°F. There is a 10psi pressure drop across the exchanger.

Heat Exchanger Unit HX 104-115

This set of shell and tube heat exchangers is used to cool the water stream recycled from the liquid-liquid extractor (EXTRACT-1) from 111°F to 98.6°F so it can enter the breeder tanks (BREED 1-2) and both fermenter series (FBB 1-6) at the desired temperature after diluting the sterile corn slurry stream exiting HX 103. This is achieved with 4 exchangers in parallel each with three in series. The shell side holds the water stream and is constructed of stainless steel, while the tube side holds the cooling water and is constructed of carbon steel. During fermentation, the mass flow rate of the glucose stream is 8,100,000 lb/hr, and is cooled with 34,599,420 lb/hr cooling water that enters the exchanger at 80°F and exits at 112.8° F. The heat duty for this operation is 106,521,400 btu/hr. The total heat transfer area is 159840 ft^2 and the overall heat transfer coefficient is 476.6 btu/F-ft^2-hr.

Holding Tank Unit DILUT-1

There is one glucose dilution tank, constructed from stainless steel with a 1,980,000 gallon capacity. This tank has a height of 110 feet and a diameter of 55feet. It is in continuous operation as it dilutes the sterilized stream exiting the dry grind process at with recycled process water, both at 98.6°F. The glucose stream entering the holding tank has a concentration of 46.2 lb/ft^3, which is then diluted to the desired amount depending on the process phase. During cell inoculation and immobilization, the process calls for a glucose concentration of 3.74 lb/ft^3, whereas during fermentation, the process calls for a glucose concentration of 5.77 lb/ft^3. 10,400,000 lb of water are required to dilute the glucose for inoculation, whereas 120,000 lb/hr of water is necessary to dilute the required amount of glucose during immobilization, and 8,120,000 lb/hr of water are needed during the fermentation processes.

Inoculation/Immobilization

During cell inoculation and immobilization, the process calls for a glucose concentration of 3.74 lb/ft^3. Inoculation is essentially a batch process, lasting the first two days of the process year. A total of 10,400,000 lb of water are required to dilute the glucose for this process, bringing the total stream to 11,400,000 lb. After the stream is sterilized, a portion is fed to each of the two breeder tanks (BREED-1 and BREED-2).

During cell immobilization, 120,000 lb/hr of water is fed into the holding tank to provide fresh nutrients for the growing cells in the bioreactors. Once again, following dilution and sterilization, a portion of the stream is fed to each of the breeder tanks.

Fermentation

For the purposes of fermentation, a total of 8,120,000 lb/hr of water are required to dilute the incoming corn slurry. This stream is then sterilized, and fed to all six bioreactors. During the fermentation process, the glucose solution is fed into each Fermenter 1 (FBB 1-3) at a rate of 436,000 lb/hr. Glucose solution as well as the sterilized stream leaving Fermenter 1 (FBB 1-3) is fed into Fermenter 2 (FBB 4-6). The diluted glucose stream is fed into each Fermenter 2 at a rate of 2,630,000 lb/hr. Note that trace amounts of solvents will be present in the water recycled from the separations processes. These trace amounts will have no effect on either strain of bacteria or their fermentation performance.

Breeder Tanks

The function of the two breeder tanks is to grow each species of bacteria to the desired number prior to immobilization and fermentation. Following inoculation, each breeder tank is used to recycle nutrients to the fermenters during the immobilization period, and during fermentation it serves as a holding tank for the glucose and P2 medium feed streams into the fermenters.

There is one breeder tank for each bacteria species, one for *Clostridium Tyrobutyricum* (BREED-1) and one for *Clostridium Acetobutylicum* (BREED-2). Both operate at 98.6°F at all times.

Breeder Tank Unit BREED-1

This tank has a capacity of 264,000 gallons with height of 56.44ft and a diameter of 28.22ft. It is constructed from stainless steel and is equipped with an agitator to ensure complete mixing during inoculation and immobilization.

Inoculation/Immobilization

During inoculation, the suspended *Clostridium Tyrobutyricum* solution, P2 medium, and diluted glucose (at 60 g/L) are fed into BREED-1. The cells are then allowed to grow for a period of two days. Following these two days, one third of the contents of BREED-1 are pumped into each FBB 1-3. During the next eight days, the cells will continue to grow while immobilizing themselves on the fibrous matrix. Half of the exiting stream is purged, and a stream of equal mass of fresh nutrients (diluted glucose and P2 medium) is introduced into BREED-1 and again pumped through each FBB 1-3.

Fermentation

Following the first ten days of the process year, the breeder tank serves as a holding tank for the feed streams to the fermenters.

Breeder Tank Unit BREED-2

This tank has a capacity of 1,237,500gallons with a height of 94.45ft and a diameter of 47.22ft. It is constructed from stainless steel and is equipped with an agitator to ensure complete mixing during inoculation and immobilization.

Inoculation/Immobilization

During inoculation, the suspended *Clostridium Acetobutylicum* solution, P2 medium, and diluted glucose are fed into the breeder tank. The cells are then allowed to grow for a period of two days. Following these two days, one third of the contents of BREED-2 are pumped into each FBB 4-6. During the next eight days, the cells will continue to grow while immobilizing themselves on the fibrous matrix. Half of the exiting stream is purged, and a stream of equal mass of fresh nutrients (diluted glucose and P2 medium) is introduced into BREED-2 and again pumped through each FBB 4-6.

Fibrous Bed Bioreactors

Fibrous Bed Bioreactor Unit FBB 1-3

There are three Fibrous Bed Bioreactors (FBBs) for the acidogenesis phase of fermentation. Run in parallel, they are constructed from stainless steel and use cotton fibers as the immobilization matrix, and each has an 110,000 gallon capacity. *Clostridium Tyrobutyricum* are used for this stage of fermentation.

Inoculation/Immobilization

During Inoculation, the fermenters are not in use. At the conclusion of the first two days of the process year, the contents of BREED-1 are pumped into FBB 1-. All streams are at 98.6° F, and during immobilization, FBB1-3 each have a dilution rate of 0.02 hr^{-1}, which corresponds to a 48 hour residence time. This stream is sent through the fermenter and then recycled back into BREED-1, which is then sent back through FBB 1-3. Assuming that a portion of the nutrients were completely utilized in the bioreactor, half of the mass flow rate of the stream leaving FBB 1-3 (total of 22,6000 lb/hr) is purged and sent to the centrifuge (CENTR1), which will run at a much smaller rpm than during fermentation. It is assumed that this stream is mostly water and solids, with only trace amounts of P2 medium and glucose. The solids are then passed through the DDGS dryer while the liquid is recycled back to DILUT-1. The same mass that was purged from the process will be replaced by both the glucose solution from DILUT-1 and P2 medium. After eight days, a sufficient number of cells will have immobilized on the fibrous matrix, and the stream exiting the fermenter will no longer be recycled back to BREED-1.

Fermentation

The dilution rate for these fermenters is 0.6 hr^{-1}, which corresponds to a 1.67 hr residence time. As specified for ideal conditions of *Clostridium Tyrobutyricum*, each fermenter is kept at 98.6°F and at pH of5.4. A controller has been installed to monitor the pH, which is controlled by the addition of either NaOH or HCl.

As stated above, the diluted corn slurry is fed into each reactor at 438,000 lb/hr. P2 medium is also fed into the reactor at 5,300 lb/hr to supply nutrients to the bacteria. It is assumed that most of the P2 medium is consumed in the bioreactor, and only trace amounts leave these first three fermenters. It is also assumed that all of the glucose is either consumed by cells or converted into butyric acid.

Roughly 50% w/w of all glucose entering the fermenter is converted into butyric acid. Lactic and acetic acid are also produced, but in much smaller proportions; 0.100 w/w and 0.033 w/w, respectively. The gases CO_2 and H_2 are also products of this fermentation, with CO2 being produced in a 0.440 w/w proportion to glucose and H_2 being produced in 0.0053 w/w glucose. These gases mix the contents of the bioreactor as they flow upwards, and are then removed from the head space at the top of the reactor. A controller has been installed to maintain the pressure at 5in water gauge to keep the process anaerobic. This is accomplished by cycling enough of the effluent gas back through each reactor. The effluent gas produced in FBB 1-3 will also be sent into FBB 4-6 to control the pressure in an identical manner. This stream is then passed through the DDGS dryer, where H_2 is burned off and the CO_2 is released into the atmosphere.

Heat Exchanger Unit HX 201-HX 203

There are three shell and tube heat exchangers that act as preheaters for the sterilization of the product streams from the first stage of fermentation (FBB 1-3). Both the shell and tube size are constructed from carbon steel, and the unit has an area of 10,400 ft^2 with a heat transfer coefficient of 74 btu/F-ft^2-hr. The cold stream is the exit stream from FBB 1-3, which is at 98.6°F, and has a mass flow rate of 421,000 lb/hr. The hot stream is the sterilized FBB 1 product stream, which is exiting from units HX 204-206 at 250° F and also has a flow rate of 421,000 lb/hr. Since the mass flow rate and heat capacities of the two streams are identical, the temperature change for the hot and cold sides are also identical at 10° F. No utilities are required since a product stream is used to heat another. There is a 10 psi pressure drop across the heat exchanger.

Heat Exchanger Unit HX 204-206

There are three shell and tube heat exchangers that are used to sterilize the product from the first stage of fermentation before it can enter the second stage. The streams are considered sterilized once they reach a temperature of 250° F. For these exchangers, the shell side is constructed from carbon steel, the tube side is constructed from stainless steel, and each unit has an area of 2,580 ft^2 with a heat transfer coefficient of 100 btu/F-ft^2-hr. The cold stream enters the heat exchanger at 240° F, and exits 250° F. This stream is heated with 282,000 lb/hr of 50 psig steam, which enters the shell of the heat exchanger at 281° F, and exits at 267° F. There is a 10 psi pressure drop across the exchanger.

Heat Exchanger Unit HX 207-209

These three shell and tube heat exchangers cool the hot stream exiting the preheater to 98.6°F, which is the required feed temperature for all fermentation operations. Again, the shell side is constructed from carbon steel and the tube side is constructed from stainless steel. The heat exchanger has a heat transfer coefficient of 200 btu/F-ft^2-hr. The cooled sterilized stream enters tube side of the fermenter at 109° F, and exits at 98.6°F. Cooling is achieved with 410,000 lb/hr cooling water that enters the shell side at 90° F and exits at 108.6° F. There is a 10 psi pressure drop across the exchanger.

Fibrous Bed Bioreactor Unit FBB 4-FBB 6

The solventogenesis phase of fermentation also has three FBBs that run in parallel. These are also constructed from stainless steel and use cotton fibers as the immobilization matrix. These three reactors must be five times the size of the acidogenesis fermenters, and therefore each has a capacity of 550,000 gallons.

Inoculation/Immobilization

During inoculation, these fermenters are not in use. At the conclusion of the first two days of the process year, the contents of BREED-2 are pumped into the three solventogenesis fermenters. All streams are at 98.6°F, and during immobilization, FBB 4-6 each have a dilution rate of 0.02hr^{-1}, which corresponds to a 48 hour residence time. This stream is sent through the fermenter and then recycled back into BREED-2, which is then sent back through the Fermenter. Assuming that a portion of the nutrients were completely utilized in the bioreactor, half of the mass flow rate of the stream leaving FBB 4-6 (total of 105,000 lb/hr) is purged and combined with the purge stream from FBB 1-3 and is sent to the centrifuge (CENT 1). Again, it is assumed that this stream is mostly water and solids, with only trace amounts of P2 medium and glucose. The solids are then passed through the DDGS dryer while the liquid is recycled back to DILUT-1. The same mass that was purged from the process will be replaced by both the glucose solution from DILUT-1 and P2 medium. After eight days, a sufficient number of cells will have immobilized on the fibrous matrix, and the stream exiting the fermenter will no longer be recycled back to BREED-2.

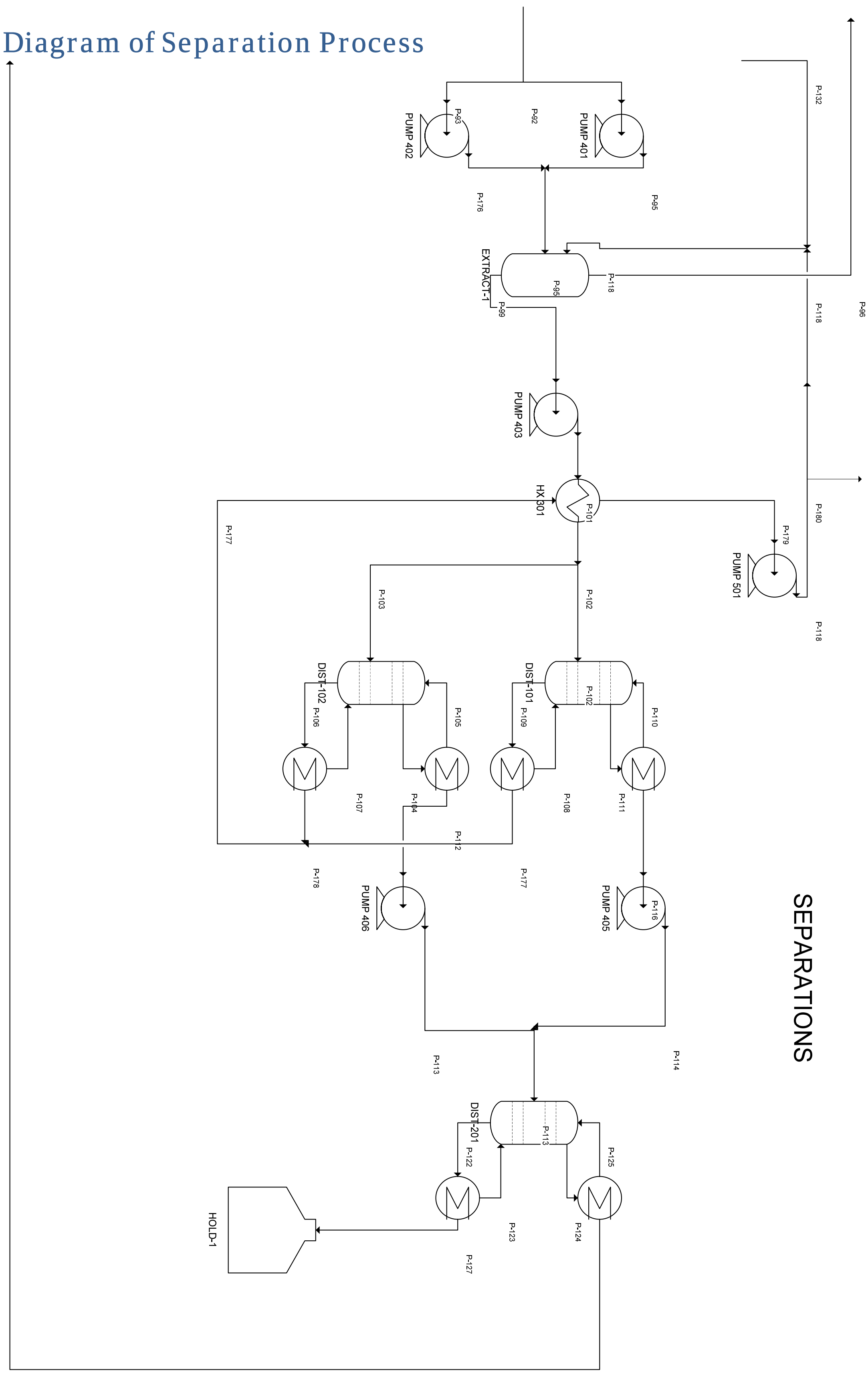
Fermentation

The dilution rate for these fermenters is $0.9 hr^{-1}$, which corresponds to a 1.11 hr residence time. For this reaction, *Clostridium Acetobutylicum* cells have a productivity of 4.6 g/L-hr. These fermenters are also kept at 98.6°F, but solventogenesis requires a more acidic pH, and it is therefore kept at 4.3. Just as in FBB 1-3, the pH is controlled using NaOH or HCl.

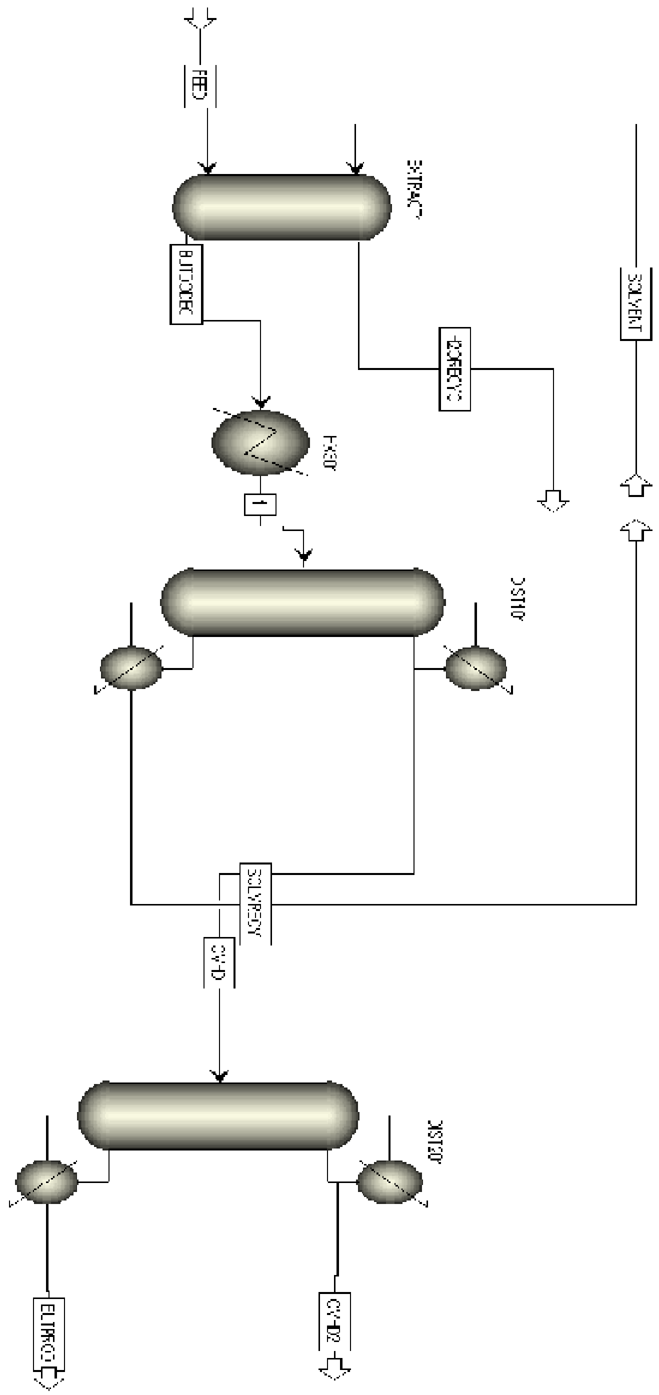
Three streams are fed into each solventogenesis fermenter: the sterilized product streams exiting FBB 1-3, fresh P2 medium at 53,000 lb/hr, and 2,630,000 lb/hr of the diluted glucose stream from DILUT-1. All streams enter FBB 4-6 at98.6°F. Again, it is assumed that only trace amounts of glucose and P2 medium exit the fermenter as most of the nutrients were utilized by the bacteria. The solventogenesis phase differs from the acidogenesis phase as the cells only utilize the glucose for nutrients, and it does not get converted into acids.

Separations Process

Visio Diagram of Separation Process



Aspen Simulation Flowsheet of Separations



Separations

Separations Process Description

Once solids leave the reactor effluent by centrifugation, the dissolved butanol must be removed from a mostly water solution. A large stream of 8,300,000 lb/hr, 99.3% of which is water, enters the separations process. In contrast from the traditional ethanol process, since mass fraction of butanol in the feed is less than 1% and amount of water from fermentation so high, liquid-liquid extraction is a more efficient method of initial separation. Therefore, the feed stream first encounters a liquid-liquid exchanger. The water stream exchanges nearly all its butanol with a one tenth relative mole flow of dodecane, which is far more soluble for butanol. The overhead water is recycled to the dry-grind portion of the process, where it is used to dilute glucose to levels acceptable by the reactors.

The butanol-rich dodecane continues to a heat exchanger, which raises its temperature from 123° F to 410° F before entering a series of distillation towers. The heat is removed from the bottoms of the first pair of distillation columns DIST-101 and DIST-102 at 483 F, in order to save utility costs. The bottoms of both DIST-101 and DIST-102 is nearly pure dodecane solvent, with trace amounts of water, acetone, ethanol, and butanol. It is recycled to the extractor via the heat exchanger.

The overhead of DIST-101 and DIST-102 is 94% butanol by mass. It proceeds to a final distillation column DIST-201 which raises the purity of butanol to 99.5% in the bottoms stream. The overhead stream of this column is a mixture of butanol, acetone, and ethanol, and nearly half water by mass, and is either to be discarded or can be dried and oxidized by integrating with the DDGS drying process.

The overall recovery of the process is approximately 96.5% butanol, with over half of this loss occurring in DIST-201, likely due to an azeotrope formed between the overhead aggregate of Acetone, Ethanol, Butanol, and 46% water, but quaternary data was not available to confirm this hypothesis.

Liquid-Liquid Extractor EXTRACT-1

Extraction follows solids removal to reduce the size of the stream containing Butanol. Dodecane was chosen as the solvent due to its high solubility and absence of azeotrope with butanol. The incoming feed from the drying process centrifuge contains 0.57 % butanol by mass and 99.4% water. This feed is 8,600,000 lb/hr, and only 6,810,000 lb/hr solvent stream is necessary to remove the butanol. This is 25% reduction in mass as well. Thus this unit operation allows for more efficient separation of butanol in further columns.

The extractor is modeled as being 99% effective in transferring butanol from water to solvent stream and only 46% and 26% effective for ethanol and acetone, respectively.[10] The extractor employs three theoretical stages to reach the separation. These theoretical stages translate, with an efficiency of 0.15, to 21 real stages in the columns. This vertical, stainless steel column has a capacity of 135,000 ft^3/hr, diameter of 8.40 feet, a height of 54

[10] At a solvent temp of 220° F the butanol recovery is closer to 90%. The approximation used here relies on the recycling of any lost butanol to the reactors, and back to the separations by increasing feed concentration. Since the reactors and separations were modeled separately, the extractor unit is modeled as 99% recovery for simplicity of calculation. This is justified because the material balance would be the same in either case (no butanol is lost overall from the change) and the same amount of butanol proceeds to the rest of the separations process.

feet with 2 feet per stage, and a cross sectional area of 55 ft^2. The column operates at 30 psi and the temperature ranges between 123° F and 280° F from bottom to top.

The water leaving the extractor is nearly pure, with approximately 60-190 ppm maximum of solvents, levels which does not retard the first fermentation, nor inhibit product formation in the second fermentation. The dodecane solvent stream leaving the column at 6,900,000 lb/hr is modeled to contain 99% of the incoming butanol.

Heat Exchangers HX-301 to HX-324

In order to reduce the diameters of the first pair of distillation columns DIST-101 and 102, a series of heat exchangers pre-heats the butanol-rich dodecane stream from the extractor. Before entering the distillation as feed, the stream must be heated from 123° F to an optimal temperature of 410° F. Eight parallel lines of three heat exchangers in series accomplish this time. The heating fluid is simply the bottoms of the DIST-101 and 102 pair of columns, which is at nearly the same flow rate (nearly all dodecane is recovered by the column) and at 483° F. This stream reduces to a temperature of 220° F and recycles to the liquid-liquid extractor.

The heat exchangers are arranged as a network of eight parallel lines of three exchangers each in series. In total, the heat exchangers have 1800 tubes, each 30 ft long, and one pass per exchanger; they are made of stainless steel 304. The shells are 45 inches in diameter and 30 feet long. They are made of carbon steel. The overall heat transfer coefficient is modeled at 80 btu/h-ft^2-F, and the total area of all 24 exchangers together is 252,000 ft^2. The shell pressure drop is from 31 to 16 psi, while the tube pressure drop is from 30 to 28.7 psi

Examining the overall impact of the network, the shell side contains the dodecane to be recycled, starting at 483° F and dropping to 220° F. Conversely, the tube side contains the dodecane rich butanol feed to columns and rises from 123° F to 410° F. Thus, total Heat transferred amounts to 1,228,000,000 Btu/hr in the aggregate of the 24 heat exchangers, saving a great amount of utilities and operating cost.

Solvent recycle purge and makeup streams

Because of the large amount of water being recycled back to the dilution process, and dodecane recycled back to the liquid-liquid extractor, there is a major chance that contaminants could build up in the streams. To prevent this from occurring, a 1% purge stream for the water and 0.5% purge stream for the dodecane is included with each of the streams, and process water or dodecane is bought to replace that lost in the purge. Only a small purge is needed, because both the water stream out of the extractor and dodecane stream out of the first distillation column are 99.9% pure by mass. There is no reason to expect a large amount of build up. Also, each of the streams will eventually be heated enough to reach sterilization.

Particularly, the dodecane solvent recycle stream has a 1% purge stream to prevent buildup by any heavy contaminants, particularly trace solid particles. This is achieved with a flow splitter. The solvent stream also has a 1% makeup of fresh dodecane from stream. Both streams remove and replenish dodecane at a rate of 68,000 lb/hr and a temperature of 220° F.

Distillation Columns DIST-101 and DIST-102

The first pair of distillation columns removes the butanol from the dodecane stream and further shrinks the stream in which the dodecane is held. This column recovers dodecane as a bottoms product, reboiled at a temperature of 483 F, comprising nearly pure solvent. The overheads are a mixture of 94% butanol and the remainder solvents and water. Butanol is recovered at 99.2% split fraction.

Due to large liquid flows in the bottom stages nearing 7,000,000 lb/hr, two columns of 31 foot diameters and 64 ft height are used to handle the incoming feed. Only 13 theoretical stages were necessary to achieve this separation, translating to 26 real trays plus reboiler and condenser in the column. Sieve trays were used and modeled with the O'Connell correlation to calculate efficiency. The feed enters in real tray #9 and the columns produce 46,906 lb/hr butanol in total, at an overhead temperature of 251° F while 483° F dodecane exits the bottom. Operating pressure is 30 psi on average throughout the column. The reflux ratio is 2.3 and distillate rate is 49,800 lb/hr, only slightly more than the butanol product flow.

Each tower's reboiler and condenser cluster (three reboilers, two condensers per tower) transfer a heat duty of 1,600,000 Btu/hr and 2,100,000 Btu/hr, respectively. Constrained by a heat flux of 12,000 Btu/hr-ft^2 to prevent film boiling phenomena (and thus reduction of heat transfer properties), the reboiler area is 413 ft^2. Each condenser amounts to 260 ft^2 at a LMTD of 81° F. The reflux accumulators are horizontal, carbon steel vessels with a capacity of 155 ft^3, 9 feet long each, and a residence time of 5 minutes.

Reflux Pump DIST-PUMP 101-102

This is the centrifugal pump connected to distillation column DIST-101 and DIST-102. It is used to pump the reflux up all the way to the top tray. The pump is a single-stage, with a maximum shaft rpm of 3,600. The motor and the pump are made out of cast steel. The pump efficiency is 0.64 and the driver efficiency is 0.85, forming a 322 ft head. The mass flow rate is 102,000 lb/hr.

Reboiler Pump DIST-PUMP 103-104

This is the centrifugal pump connected to distillation column DIST-101 and DIST-102. The pump is a single-stage, with a maximum shaft rpm of 3,600. The motor and the pump are made out of cast steel. The pump efficiency is 0.88 and the driver efficiency is 0.74, forming a 413 ft head. The mass flow rate is 5,380,000 lb/hr.

Distillation Column DIST-201

The final distillation column further purifies the overhead of the first pair of columns into 99.5% product specification grade butanol, and a discarded overhead stream of mostly water mixed with residue solvents. Butanol is recovered as a bottoms product with ~120 ppm acetone, and is recovered at a 98% split fraction. This represents the maximum loss of butanol in the process.[11] The column produces 45,970 lbs/hr of butanol, which is 54.3 million gal/year at ~77° F.

Employing 15 theoretical stages and 22 actual trays, the tower amounts to a diameter of 3 feet and rises to a height of 45 feet. Carbon steel is used since water content is low. Top

[11] Adding stages or changing the reflux ratio did not help improve the fraction recovered, which indicates possible thermodynamic barriers. These could not be confirmed with ASPEN analysis and remain hypotheses.

temperature is 200° F while reaching 278° F at the bottoms. Operating pressure is 27 psi on average throughout.

Product specification of butanol will vary with application and thus the process is left flexible for updating based on market requirements. More is discussed in the preliminary process synthesis section of this report, but suffice it to say that the process is strategically left flexible since butanol is a novel fuel product and its application and resulting tolerances are not fully determined beyond the 99.5% purity requirement (i.e. what comprises the final 0.5%).

The reboiler and condenser clusters (also three reboilers, two condensers) each transfer a heat duty of 1,600,000 btu/hr and 2,100,000 btu/hr, respectively. Constrained by a heat flux of 12,000 btu/hr-ft^2 as well, each reboiler area is 413 ft^2. Each condenser amounts to 260 ft^2 at a LMTD of 81° F. The reflux accumulator is a horizontal, carbon steel vessel with a capacity of 46 ft^3, 6 feet long, and a residence time of 5 minutes.

Reflux Pump DIST-PUMP 201

This is the centrifugal pump connected to distillation column DIST-201. It is used to pump the reflux up all the way to the top tray. The pump is a single-stage, with a maximum shaft rpm of 3,600. The motor and the pump are made out of cast steel. The pump efficiency is 0.49 and the driver efficiency is 0.87, forming a 290 ft head. The mass flow rate is 30,400 lb/hr.

Reflux Pump DIST-PUMP 202

This is the centrifugal pump connected to distillation column DIST-202. It is used to pump the reflux up all the way to the top tray. The pump is a single-stage, with a maximum shaft rpm of 3,600. The motor and the pump are made out of cast steel. The pump efficiency is 0.57 and the driver efficiency is 0.86, forming a 331 ft head. The mass flow rate is 59,300 lb/hr.

Pump Unit PUMP401, PUMP402

This pump delivers the solids-free butanol stream from the centrifuge (CENTRI1) to the liquid-liquid extractor for separations (EXTRACT1). The mass flow rate is 3,720,000 lb/hr and the pressure increase is 33 psi. This is a radial, centrifugal, single-stage pump, with a 3,600 rpm shaft. The motor and pump are made out of stainless steel. The pump efficiency is 0.88 and motor efficiency is 0.92. The power consumption is 204 HP, and it requires 187 HP of brake power. It develops a head of 132 ft.

Pump Unit PUMP403, PUMP404

This pump delivers the dodecane stream carrying the butanol from the bottoms of the liquid-liquid extractor (EXTRACT1) to the first distillation column series (DIST1-2). The mass flow rate is 4,160,000 lb/hr and the pressure increase is 28 psi. This is a radial, centrifugal, single-stage pump, with a 3,600 rpm shaft. The motor and pump are made out of stainless steel. The pump efficiency is 0.88 and motor efficiency is 0.90. The power consumption is 77 HP, and it requires 70 HP of brake power. It develops a head of 153 ft.

Pump Unit PUMP405, PUMP406

This pump delivers the butanol rich stream from the overhead stream of the first set of distillation columns (DIST1-2) to the second fermenter (DIST3). The mass flow rate is 19,700 lb/hr and the pressure increase is 18 psi. This is a radial, centrifugal, single-stage pump, with a 3,600 rpm shaft. The motor and pump are made out of stainless steel. The pump efficiency is 0.42 and motor efficiency is 0.91. The power consumption is 158 HP, and it requires 145 HP of brake power. It develops a head of 122 ft.

Pump Unit PUMP501

This pump delivers the nearly pure dodecane stream from the bottoms of the first distillation series (DIST1-2) back to the liquid-liquid extractor as the solvent stream (EXTRACT1). The mass flow rate is 10,800,000 lb/hr and the pressure increase is 27 psi. This is a radial, centrifugal, single-stage pump, with a 3,600 rpm shaft. The motor and pump are made out of stainless steel. The pump efficiency is 0.89 and motor efficiency is 0.90. The power consumption is 77 HP, and it requires 69 HP of brake power. It develops a head of 110 ft.

Holding Tank Unit HOLD-1

This tank is used to store the butanol product stream from DIST 201 until it can be removed from the plant. The tank is constructed from stainless steel and has a 3,000,000 gallon capacity. The diameter of the tank is 63 ft and the height is 127 ft. A tank of these dimensions is able to hold at least two weeks of product. The butanol product stream flows into the tank at a mass flow rate of 46,200 lb/hr and is kept at ambient temperature and atmospheric pressure. The contents of the tank are liquid at this temperature, as the boiling point is 243° F. Butanol is also not corrosive and therefore will not compromise the integrity of the tank.

Holding Tank Unit DDGS-1

This tank is used to store DDGS that is recovered from the centrifuge (CENTR-1) and the dryer until it can be sold and removed from the plant. The tank is constructed from stainless steel and has a 4,110,000 gallon capacity. The diameter is 70 ft and the height is 141 ft. A tank of these dimensions is able to hold at least two weeks of product. DDGS is transferred into the holding tank at a mass flow rate of 97,600 lb/hr. The tank is held at atmospheric pressure and ambient temperature.

Process Description for Dry Grind to Form Glucose

The dry grinding of corn is very widely used in the formation of ethanol by fermentation. This process encompasses all the steps of the fermentation including the processing of the corn, fermentation by yeast, separations using a beer column and stripper, and processing of the DDGS using rotary motor dryers and a centrifuge. The first portion of the process for handling of the corn to the formation of glucose is very relevant for the butanol process. Corn is brought to the facility and held in a large storage tank. The tank is large enough to store a 12 day inventory of corn. About 11,410,000,000 lbs of corn is needed a year, which is about 203,000,000 bushels of corn. The corn goes through blowers and screens to dispose of loose husk and shells, before being sent to the hammer mill to be ground into smaller pieces and mixed with water to form a slurry. The slurry is sent to the liquefaction process, where ammonia and lime are added, and the starch is gelatinized using a steam injection heater and hydrolyzed with the amylases into dextrins. The pH is maintained at 6.5 for this process. The streams are held at 60 min at 190.4^{o} F for 60 minutes, and then cooked at 230^{o} F for 15 minutes. Afterward, the stream is sent to the saccharification process. Here, sulfuric acid is used to lower the pH to about 4.5, and the slurry is held for 5 hours. Glucoamylases are added to hydrolyze the dextrins to glucose at 140^{o} F, and the resulting glucose stream is pumped to the dilution tank. This process produces 940,233 lb/hr of glucose that will be fed to both fermenter series and both breeder tanks. Details on the economics and operations for this portion of the process can be found in Appendix D.

Process Description for DDGS Drying

The final section of the dry grind process is used to create the DDGS co-product. A series of centrifuges is used to pull the solids from the stream out of the fermentation process along with a small amount of water and trace amounts of solvents in the wet solids stream. The liquid stream out the centrifuge is sent directly to the separations process. The wet solids are sent along a conveyor belt to the rotary drum dryers. “These dryers feature single and multi-pass technology that moves materials through the drum in an air stream created by the dryer induced-draft fan. The multiple passes are mechanically interlocked to rotate at the same speed. As the drum rotates, the product is repeatedly shoved into the dryer hot gas.” (Onix Corporation) The hot air that is used to dry the solids collects a bit of water, and trace amounts of solvents, and this stream is sent to a thermal oxidizer to burn off the vaporized solvents. This thermal oxidizer is also used to burn off the H2 from the waste gas stream from the first series of fermenter. Out of the thermal oxidizer is air, small amounts of water vapor, and carbon dioxide to be released into the atmosphere. The solids out of the rotary drum dryer are sent through conveyor belts to a holding tank that stores a 2 week inventory. This process creates approximately 773,000,000 lbs per year of DDGS, which bring in about $51.8 million a year. The economic analysis and equipment for this part of the process can be found in Appendix D.

Specification Sheets

FERMENTER 1				
Identification:	Item		Fibrous Bed Bioreactor	
	Item No.		FBB 1-3	
	No. Required		3	
Cost of Bare Module:	$ 2,802,275.80			
Function:	Provide conditions necessary for the clostridium tryobutyricum to perform acidogenesis fermentation on corn slurry			
Operation:	Continuous			
Materials Handled:	*Inlet "Gluc dil"*	*Inlet "P2 Medium"*	*Outlet "Butyric Acid"*	*Outlet "Gases"*
Quantity (lb/hr)	436,526	5,035	420,875	20,686
Composition				
Glucose	0.0932	---	---	---
Butyric Acid	---	---	0.0453	---
Acetic Acid	---	---	0.0109	---
Lactic Acid	---	---	0.0036	---
CO2	---	---	---	0.9766
H2	---	---	---	0.0234
Butanol	---	---	---	---
Ethanol	---	---	---	---
Acetone	---	---	---	---
P2 Medium	---	1.0000	---	---
Water	0.8964	---	0.9294	---
Solids	0.0104	---	0.0108	---
Temperature (°F)	98.6	98.6	98.6	98.6
Design Data:	Volume: 110,000 gallons			
	Diameter: 21.08 ft			
	Height: 42.16 ft			
	Temperature: 98.6 °F			
	Material of Construction: Cotton fibers and Stainless Steel			
	Agitators: None			
	Void Space: 0.9			
Utilities:				
Controls:	Pressure			
	pH			
Tolerances:	None			
Comments and Drawings:	Sizing and costing calculations found in Appendix A			

FERMENTER 2				
Identification:	Item	Fibrous Bed Bioreactor		
	Item No.	FBB 4-6		
	No. Required	3		
Cost of Bare Module:	$ 4,620,430.82			
Function:	Provide conditions necessary for the clostridium acetobutylicum to perform solventogenesis fermentation			
Operation:	Continuous			
Materials Handled:	*Inlet "Butyric Acid"*	*Inlet "Gluc dil"*	*Inlet "P2 Medium"*	*Outlet "Butanol"*
Quantity (lb/hr)	421,025	2,632,797	31,965	2,808,830
Composition				
Glucose	---	0.09324	---	---
Butyric Acid	0.04532	---	---	---
Acetic Acid	0.01090	---	---	---
Lactic Acid	0.00363	---	---	---
CO2	---	---	---	---
H2	---	---	---	---
Butanol	---	---	---	0.005655
Ethanol	---	---	---	0.000125
Acetone	---	---	---	0.000250
P2 Medium	---	---	1.00000	---
Water	0.92937	0.89636	---	0.98257
Solids	0.01078	0.01040	---	0.01140
Temperature (°F)	98.6	98.6	98.6	98.6
Design Data:	Volume: 550,000 gallons			
	Diameter: 36.04 ft			
	Height: 72.08 ft			
	Temperature: 98.6 °F			
	Material of Construction: Cotton fibers and Stainless Steel			
	Agitators: None			
	Void Space: 0.9			
Utilities:				
Controls:	Pressure			
	pH			
Tolerances:	None			
Comments and Drawi	Sizing and costing calculations found in Appendix A			

BREEDER TANK 1

Identification:	Item	Breeder Tank
	Item No.	BREED-1
	No. Required	1
Cost of Bare Module:	$ 1,800,537.90	
Function:	Provide conditions to innoculate clostridium tyrobutyricum as well as serves as a vessel to pass corn slurry through during normal fermentation operation to Fermenter 1	
Operation:	Batch during innoculation and Continuous during immobilization and fermentation	

	During Innoculation			During Immobilization				
Materials Handled:	*Inlet "Cells"*	*Inlet "Gluc dil"*	*Inlet "P2 Medium"*	*Inlet "Gluc Dilution"*	*Inlet "P2 Medium"*	*Inlet "Recycle"*	*Outlet "Purge"*	*Outlet "Innoculum"*
Quantity (lb/hr)	216,793.76	1,951,143.80	16,854.39	22,582.68	195.00	22,582.68	22,582.68	45,165.36
Composition								
Glucose		0.0594	---	0.0594	---	trace	trace	trace
Clostridium Tyrobutyricum		---	---	---	---	---	---	---
P2 Medium		---	1.000	---	1.000	trace	trace	trace
Water		0.9342	---	0.9342	---	0.9936	0.9936	0.9936
Solids	---	0.0064		0.0064		0.0064	0.0064	0.0064
Temperature (°F)	98.6	98.6	98.6	98.6	98.6	98.6	98.6	98.6

Design Data:	Volume: 264,000 gallons
	Diameter: 28.22 ft
	Height: 56.44 ft
	Dilution Rate: 0.02 hr-1
	Temperature: 98.6 °F
	Material of Construction: Stainless Steel
	Agitators:
Utilities:	
Controls:	Temperature
	pH
Tolerances:	None
Comments and Drawings:	Sizing and costing calculations found in Appendix A

BREEDER TANK 2

Identification:	Item	Breeder Tank
	Item No.	BREED-2
	No. Required	1
Cost of Bare Module:	$ 3,352,193.82	
Function:	Provide conditions to innoculate clostridium acetobutylicum as well as serves as a vessel to pass corn slurry through during normal fermentation operation to Fermenter 2	
Operation:	Batch during innoculation and Continuous during immobilization and fermentation	

	During Inoculation			During Immobilization				
Materials Handled:	*Inlet "Cells"*	*Inlet "Gluc dil"*	*Inlet "P2 Medium"*	*Inlet "Gluc Dilution"*	*Inlet "P2 Medium"*	*Inlet "Recycle"*	*Outlet "Purge"*	*Outlet "Inoculum"*
Quantity (lb/hr)	1,016,220.64	9,145,985.75	79,005.00	105,856.32	914.00	105,856.32	105,856.32	211,712.64
Composition								
Glucose		0.0594	---	0.0594	---	trace	trace	trace
Clostridium Acetobutylicum		---	---	---	---	---	---	---
P2 Medium		---	1.000	---	1.000	trace	trace	trace
Water		0.9342	---	0.9342	---	0.994	0.994	0.994
Solids	---	0.0064		0.0064		0.006	0.006	0.006
Temperature (°F)	98.6	98.6	98.6	98.6	98.6	98.6	98.6	98.6

Design Data:	Volume: 1,237,500 gallons
	Diameter: 47.22 ft
	Height: 94.45 ft
	Dilution Rate: 0.02hr-1
	Temperature: 98.6 °F
	Material of Construction: Stainless Steel
	Agitators:
Utilities:	
Controls:	Temperature
	pH
Tolerances:	None
Comments and Drawings:	Sizing and costing calculations found in Appendix A

GLUCOSE HOLDING TANK									
Identification:	Item	Holding Tank							
	Item No.	DILUT-1							
	No. Required	1							
Cost of Bare Module:	$ 4,535,660.20								
Function:	Serves as holding tank to dilute glucose in the corn slurry coming from the corn mill for Innoculation/Immobilization as well as Fermentation								
Operation:	Continuous								
	During Innoculation*			During Immobilization**			During Fermentation***		
Materials Handled:	*Inlet "Corn Slurry"*	*Inlet "Water"*	*Outlet "Glucose Dil"*	*Inlet "Corn Slurry"*	Inlet "Water"	*Outlet "Glucose Dil"*	*Inlet "Corn Slurry"*	*Inlet "Water"*	*Outlet "Glucose Dil"*
Quantity (lb/hr)	1,042,241.94	10,364,407.83	11,406,649.78	12,062.99	119,958.42	132,021.41	1,008,481.48	8,118,595.50	9,127,076.98
Composition									
Glucose	0.650	---	0.0594	0.650	---	0.059	0.650	---	0.0718
Water + Other Inert Soluble Compounds	0.280	1.000	0.9342	0.280	1.000	0.934	0.280	1.000	0.9204
Solvents	---	trace	trace	---	trace	trace	---	trace	trace
Solids	0.070	---	0.0064	0.070	---	0.006	0.070	---	0.0077
Temperature (°F)	140	85	96.8	140	85	96.8	140	85	96.8
Design Data:	Volume: 1,980,000 gallons Diameter: 55.23 ft Height: 110.47 ft Temperature: 98.6 °F Material of Construction: Stainless Steel Agitators: None								
Utilities:									
Controls:	Tank Level/Incoming Flow Rates								
Tolerances:	None								
Comments and Drawings:	Sizing and costing calculations found in Appendix A								

* Quantity in lb, addition of glucose dilution is for complete 2 day period
** Immobilization is from day 2 - day 10

HEAT EXCHANGER

Identification:	Item	Shell and Tube Heat Exchanger		
	Item No.	HX 101		
	No. Required	1		
Cost of Bare Module:	$ 585,469.00			
Function:	Preheats stream exiting dry grind prior to dilution with sterilized stream exiting HX 102			
Operation:	Continuous			
Materials Handled:	*Inlet, Cold*	*Outlet, Cold*	*Inlet, Hot*	*Outlet, Hot*
Quantity (lb/hr)	1,042,000	1,042,000	1,042,000	1,042,000
Composition				
Glucose	0.650	0.650	0.650	0.650
Water + Other Inert Soluble Compounds	0.280	0.280	0.280	0.280
Solvents	---	---	---	---
Solids	0.070	0.070	0.070	0.070
Temperature (°F)	140.0	239.9	249.9	150.0
Design Data:	Heat Duty (BTU/hr):			189,000,000
	Heat Transfer Coefficient (BTU/F-ft2-hr):			74
	ΔT (°F)			99.8
	Heat Transfer Area (ft2):			25,687.00
	Material of Construction:	Shell:		
		Tube:		
Utilities:	None			
Comments and Drawings:	Sizing and costing calculations found in Appendix A			

HEAT EXCHANGER			
Identification:	Item	Shell and Tube Heat Exchanger	
	Item No.	HX 102	
	No. Required	1	
Cost of Bare Module:	$ 241,000.00		
Function:	Sterilized stream exiting HX 101 with 50psig steam		
Operation:	Continuous		
Materials Handled:	*Inlet, Cold*	*Outlet, Cold*	
Quantity (lb/hr)	1,042,000	1,042,000	
Composition			
Glucose	0.650	0.650	
Water + Other Inert Soluble Compounds	0.280	0.280	
Solvents	---	---	
Solids	0.070	0.070	
Temperature (°F)	239.9	249.9	
Design Data:	Heat Duty (BTU/hr):		18,720,000
	Heat Transfer Coefficient (BTU/F-ft2-hr):		100
	ΔTlm (°F)		17.43
	Heat Transfer Area (ft2):		10,741.00
	Heating Material		50psig steam at 281.03°F
	Material of Construction:	Shell:	Carbon Steel
		Tube:	Stainless Steel
Utilities:	Steam (50psig, 281.03F) at 1,325,167lb/hr		
Comments and Drawings:	Sizing and costing calculations found in Appendix A		

HEAT EXCHANGER			
Identification:	Item	Shell and Tube Heat Exchanger	
	Item No.	HX 103	
	No. Required	1	
Cost of Bare Module:	$ 687,000.00		
Function:	Cools hot stream exiting HX 101 with cooling water to 98.6°F		
Operation:	Continuous		
Materials Handled:	*Inlet, Hot*	*Outlet, Hot*	
Quantity (lb/hr)	1,042,000	1,042,000	
Composition			
Glucose	0.650	0.650	
Water + Other Inert Soluble Compounds	0.280	0.280	
Solvents	---	---	
Solids	0.070	0.070	
Temperature (°F)	150.0	98.6	
Design Data:	Heat Duty (BTU/hr):		101,000,000
	Heat Transfer Coefficient (BTU/F-ft2-hr):		200
	ΔTlm (°F)		17.13
	Heat Transfer Area (ft2):		29,638.00
	Cooling Material		Cooling water (90°-120°F)
	Material of Construction:	Shell:	Stainless Steel
		Tube:	Carbon Steel
Utilities:	Cooling water (90°F-120° F) at 1,042,000lb/hr		
Comments and Drawings:	Sizing and costing calculations found in Appendix A		

HEAT EXCHANGER			
Identification:	Item	Shell and Tube Heat Exchanger	
	Item No.	HX 104-115	
	No. Required	3 in series, 4 in parallel	
Cost of Bare Module:	$ 8,898,240.72		
Function:	Cools recycle water stream leaving EXTRACT-1		
Operation:	Continuous		
Materials Handled:	*Inlet, Hot*	*Outlet, Hot*	
Quantity (lb/hr)	8,118,596	8,118,596	
Composition			
Glucose	0.0718	0.0718	
Water + Other Inert Soluble Compounds	0.9204	0.9204	
Solvents	trace	trace	
Solids	0.0077	0.0077	
Temperature (°F)	111.0	98.6	
Design Data:	Heat Duty (BTU/hr):	106,521,400.00	
Heat Transfer Coefficient (BTU/F-ft2-hr):		477	
	ΔTlm (°F)	45.18	
Heat Transfer Area (ft2):		159,840.00	
	Heating Material:	Cooling water (80°F - 112°F)	
Material of Construction:			
	Shell:	Stainless Steel	
	Tube:	Carbon Steel	
Utilities:	Cooling water, 34,599,420lb/hr at 80°F - 112°F		
Comments and Drawin	Sizing and costing calculations found in Appendix A		

HEAT EXCHANGER				
Identification:	Item	Shell and Tube Heat Exchanger		
	Item No.	HX 201-203		
	No. Required	3		
Cost of Bare Module:	$ 739,814.93			
Function:	Preheats stream exiting FBB 1-3 with sterilized stream leaving second Heat Exchanger (HX 204-206)			
Operation:	Continuous			
Materials Handled:	*Inlet, Cold*	*Outlet, Cold*	*Inlet, Hot*	*Outlet, Hot*
Quantity (lb/hr)	420,874.90	420,874.90	420,874.90	420,874.90
Composition				
Glucose	---	---	---	---
Butyric Acid	0.0453	0.0453	0.0453	0.0453
Acetic Acid	0.0109	0.0109	0.0109	0.0109
Lactic Acid	0.0036	0.0036	0.0036	0.0036
P2 Medium	---	---	---	---
Water	0.9294	0.9294	0.9294	0.9294
Solids	0.0108	0.0108	0.0108	0.0108
Temperature (°F)	98.6	239.8	249.8	108.6
Design Data:	Heat Duty (BTU/hr):	0		
Heat Transfer Coefficient (BTU/F-ft2-hr):		74		
	ΔT (°F):	141.2		
Heat Transfer Area (ft2):		10,375.00		
Material of Construction:				
	Shell:	Stainless Steel		
	Tube:	Stainless Steel		
Utilities:	None			
Comments and Drawings:	Sizing and costing calculations found in Appendix A			

HEAT EXCHANGER				
Identification:	Item	Shell and Tube Heat Exchanger		
	Item No.	HX 204-206		
	No. Required	3		
Cost of Bare Module:	$ 247,584.73			
Function:	Sterilizes preheated stream leaving HX 201-203 with 50psig steam			
Operation:	Continuous			
Materials Handled:	*Inlet, Cold*	*Outlet, Cold*	*Inlet, Steam*	*Outlet, Steam*
Quantity (lb/hr)	420,874.90	420,874.90	282,201.00	282,201.00
Composition				
Glucose	---	---	---	---
Butyric Acid	0.0453	0.0453	---	---
Acetic Acid	0.0109	0.0109	---	---
Lactic Acid	0.0036	0.0036	---	---
P2 Medium	---	---	---	---
Water	0.9294	0.9294	1.0000	1.0000
Solids	0.0108	0.0108	---	---
Temperature (°F)	239.8	249.8	281.0	267.3
Design Data:	Heat Duty (BTU/hr):	7,561,596.80		
Heat Transfer Coefficient (BTU/F-ft2-hr):		100		
	ΔTlm (°F)	29.3		
Heat Transfer Area (ft2):		2,580.00		
	Heating Material:	50 psig steam (281.03°F - 267.26°F)		
Material of Construction:		Carbon Steel		
	Shell:	Stainless Steel		
	Tube:	Stainless Steel		
Utilities:	Steam (50psig, 281.03F) at 282,201lb/hr			
Comments and Drawi	Sizing and costing calculations found in Appendix A			

HEAT EXCHANGER				
Identification:	Item	Shell and Tube Heat Exchanger		
	Item No.	HX 207-209		
	No. Required	3		
Cost of Bare Module:	$ 329,610.07			
Function:	Cools hot stream exiting HX 201-203 to enter FBB 4-6 at required temperature of 98.6°F			
Operation:	Continuous			
Materials Handled:	*Inlet, Hot*	*Outlet, Hot*	*Inlet, Cooling Water*	*Outlet, Cooling*
Quantity (lb/hr)	420,874.90	420,874.90	409,508.00	409,508.00
Composition				
Glucose	---	---	---	---
Butyric Acid	0.0453	0.0453	---	---
Acetic Acid	0.0109	0.0109	---	---
Lactic Acid	0.0036	0.0036	---	---
CO2	---	---	---	---
H2	---	---	---	---
Butanol	---	---	---	---
Ethanol	---	---	---	---
Acetone	---	---	---	---
P2 Medium	---	---	---	---
Water	0.9294	0.9294	1.0000	1.0000
Solids	0.0108	0.0108	---	---
Temperature (°F)	108.6	98.6	90.0	108.6
Design Data:	Heat Duty (BTU/hr):	7,762,443.00		
	Heat Transfer Coefficient (BTU/F-ft2-hr):	200		
	ΔTlm (°F)	9.93		
	Heat Transfer Area (ft2):	3,907.00		
	Heating Material:	Cooling water (90°F - 120°F)		
	Material of Construction:	Carbon Steel		
	Shell:	Stainless Steel		
	Tube:	Stainless Steel		
Utilities:	Cooling water (90°F) at 409,508.00lb/hr			
Comments and Drawings	Sizing and costing calculations found in Appendix A			

BUTANOL HOLDING TANK				
Identification:	Item	Holding Tank		
	Item No.	HOLD-1		
	No. Required	1		
Cost of Bare Module:	$ 5,444,179.85			
Function:	Serve as a holding tank for the final butanol			
Operation:	Continuous			
Materials Handled:	*Inlet "Butanol"*			
Quantity (lb/hr)	46,199.32			
Composition				
Butanol	0.995			
Acetone	0.000123			
Ethanol	0.000574			
Water	0.001			
Dodecane	0.00324			
Temperature (°F)	85			
Design Data:	Volume: 3,000,000 gallons			
	Diameter: 63.44ft			
	Height: 126.88ft			
	Temperature: 85 °F			
	Material of Construction: Stainless Steel			
Utilities:	None			
Controls:	None			
Tolerances:	None			
Comments and Drawing	Sizing and costing calculations found in Appendix A			

DDGS HOLDING TANK				
Identification:	Item	Holding Tank		
	Item No.	DDGS-1		
	No. Required	1		
Cost of Bare Module:	$ 6,953,438.19			
Function:	Serve as a holding tank for DDGS prior to shipment. Holds two weeks of product at a time.			
Operation:	Continuous			
Materials Handled:	*Inlet "DDGS"*			
Quantity (lb/hr)	97,611.01			
Composition				
Ethanol	0.00005			
Solids	0.90998			
Water	0.08997			
Temperature (°F)	85			
Design Data:	Volume: 4,114,000 gallons			
	Diameter: 70.48ft			
	Height: 140.96ft			
	Temperature: 85 °F			
	Material of Construction: Stainless Steel			
Utilities:	None			
Controls:	None			
Tolerances:	None			
Comments and Drawings:	Sizing and costing calculations found in Appendix A			

DISTILLATION COLUMN				
Identification	Item:	Distillation Column		
	Item No.	DIST101-102		
	No. required:	2		
Cost of Bare Module:	$ 10,554,435	excluding reboiler and reflux pump		
Function:	To separate out most of the butanol from the dodecane stream out of EXTRACT1.			
Operation:	Continuous			
Materials Handled:	Feed from EXTRACT1	Bottoms	Overhead	
Quantity (lb/hr)	3,431,790	3,406,870	24,920	
Composition:				
Acetone	0.0001	0.0000	0.0111	
Butanol	0.0069	0.0000	0.9411	
Dodecane	0.9927	1.0000	0.0030	
Ethanol	0.0001	0.0000	0.0097	
Water	0.0003	0.0000	0.0351	
Temperature (°F)	126	484	251	
Design Data:	Theoretical Trays:	13	Molar Reflux Ratio:	2.28
	Real Trays:	26	Tray Spacing (ft):	2
	Tray Efficiency:	0.45	Headspace (ft):	4
	Tray Type:	Sieve	Sump Space (ft):	10
	Functional Height (ft):	64		
	Inside Diameter (ft):	31		
	Pressure:	30		
	Feed Stage:	5		
	Material:	Carbon Steel		
	Condenser	2 Needed		
	Temperature (°F):	251		
	Reflux Ratio	2.28		
	Overall Heat Transfer Coefficient (BTU/hr-ft^2-°F):	200		
	Area (ft^2):	767		
	Material:	Carbon Steel		
	Reboiler	3 Needed		
	Temperature (°F):	484		
	Area (ft^2):	5,780		
	Heat Flux (BTU/hr-ft^2):	12,000		
	Material:	Carbon Steel		
	Reflux Accumulator			
	Reflux Ratio:	2.28		
	Volume (ft^3/hr):	558		
	Diameter (ft):	5		
	Length (ft):	9		
	RA Material:	Carbon Steel		
Utilities:	6,552 kW cooling water and 60,983 kW steam			
Price of Utilities:	$12,477,696 per year			
Controls:	no			
Tolerances:	no			
Comments and Drawings:	Sizing and cost calculations in Appendix A.			

DISTILLATION COLUMN				
Identification	Item:	Distillation Column		
	Item No.	DIST201		
	No. required:	1		
Cost of Bare Module:	$ 539,378			
Function:	To further separate out the butanol to make a 99.5% pure product stream.			
Operation:	Continuous			
Materials Handled:	Feed from DIST101-102	Bottoms	Overhead	
Quantity (lb/hr)	49,840	46,199	3,641	
Composition:				
Acetone	0.0111	0.0001	0.1498	
Butanol	0.9411	0.9950	0.2577	
Dodecane	0.0030	0.0032	0.0000	
Ethanol	0.0097	0.0006	0.1254	
Water	0.0351	0.0011	0.4672	
Temperature (°F)	251.27	278.71	199.08	
Design Data:	Theoretical Trays:	13	Molar Reflux Ratio:	2.3
	Real Trays:	22	Tray Spacing (ft):	2
	Tray Efficiency:	0.46	Headspace (ft):	4
	Tray Type:	Sieve	Sump Space (ft):	10
	Functional Height (ft):	56		
	Inside Diameter (ft):	3		
	Pressure:	27		
	Feed Stage:	8		
	Material:	Carbon Steel		
	Condenser	2 Needed		
	Temperature (°F):	199		
	Reflux Ratio	2		
	Overall Heat Transfer Coefficient (BTU/hr-ft^2-°F):	200		
	Area (ft^2):	390		
	Material:	Carbon Steel		
	Reboiler	3 Needed		
	Temperature (°F):	279		
	Area (ft^2):	650		
	Heat Flux (BTU/hr-ft^2):	12,000		
	Material:	Carbon Steel		
	Reflux Accumulator			
	Reflux Ratio:	2.3		
	Volume (ft^3/hr):	72		
	Diameter (ft):	2		
	Length (ft):	5		
	RA Material:	Carbon Steel		
Utilities:	2,131 kW cooling water and 2,287 kW steam			
Price of Utilities:	$177,408 per year			
Controls:	no			
Tolerances:	no			
Comments and Drawings:	Sizing and cost calculations in Appendix A.			

REFLUX PUMP			
Identification:	Item	Centrifugal Pump	
	Item No.	DISTPUMP101-102	
	No. Required	2	
Cost of Bare Module:	$13,277		
Function:	To pump the reflux in distillation column DIST101 and DIST102		
Operation:	Continuous		
Materials Handled:			
Quantity:	24,922	lb/hr	
Composition:			
Water:	0.0351		
Ethanol:	0.0097		
Butanol:	0.9411		
Acetone:	0.0111		
Dodecane:	0.0030		
Temperature:	251	°F	
Feed Stream ID			
Exit Stream ID			
Design Data:	Pump Efficiency:	0.49	
	Driver Efficiency:	0.86	
	Volumetric Flow Rate (ft^3/hr):	558	
	Head Developed (ft-lbf/lb):	136	
	Single Stage	yes	
	Shaft:	stainless steel	
	Motor Material:	stainless steel	
	Pump Material:	stainless steel	
	Pressure Change	22	psi
Utilities:	18,624	kW-hr/yr	
Cost of Utilities:	$745	per yr	
Comments and Drawings: Sizing and cost calculations in A.			

REBOILER PUMP			
Identification:	Item	Centrifugal Pump	
	Item No.	DISTPUMP103-104	
	No. Required	2	
Cost of Bare Module:	$406,556		
Function:	To pump the reboil in distillation column DIST101 and DIST102		
Operation:	Continuous		
Materials Handled:			
Quantity:	3,406,870	lb/hr	
Composition:			
Water:	0		
Ethanol:	0		
Butanol:	0		
Acetone:	0		
Dodecane:	1		
Temperature:	484	°F	
Feed Stream ID			
Exit Stream ID			
Design Data:	Pump Efficiency:	0.88	
	Driver Efficiency:	0.93	
	Volumetric Flow Rate (ft^3/hr):	97,742	
	Head Developed (ft-lbf/lb):	156	
	Single Stage	yes	
	Shaft:	stainless steel	
	Motor Material:	stainless steel	
	Pump Material:	stainless steel	
	Pressure Change	22	psi
Utilities:	1,641,503	kW-hr/yr	
Cost of Utilities:	$64,580	per yr	
Comments and Drawings:	Sizing and cost calculations in Appendix A.		

REFLUX PUMP			
Identification:	Item	Centrifugal Pump	
	Item No.	DISTPUMP201	
	No. Required	1	
Cost of Bare Module:	$12,174		
Function:	To pump the reflux in distillation column DIST201.		
Operation:	Continuous		
Materials Handled:			
Quantity:	3,641	lb/hr	
Composition:			
Water:	0.4672		
Ethanol:	0.1254		
Butanol:	0.2577		
Acetone:	0.1498		
Dodecane:	0.0000		
Temperature:	199	°F	
Feed Stream ID			
Exit Stream ID			
Design Data:	Pump Efficiency:	0.15	
	Driver Efficiency:	0.84	
	Volumetric Flow Rate (ft^3/hr):	72	
	Head Developed (ft-lbf/lb):	119	
	Single Stage	yes	
	Shaft:	stainless steel	
	Motor Material:	stainless steel	
	Pump Material:	stainless steel	
	Pressure Change	22	psi
Utilities:	7,599	kW-hr/yr	
Cost of Utilities:	$304	per yr	
Comments and Drawings: Sizing and cost calculations in Appendix A.			

REBOILER PUMP			
Identification:	Item	Centrifugal Pump	
	Item No.	DISTPUMP202	
	No. Required	1	
Cost of Bare Module:	$22,072		
Function:	To pump the reflux in distillation column DIST201.		
Operation:	Continuous		
Materials Handled:			
Quantity:	46,199	lb/hr	
Composition:			
Water:	0.0011		
Ethanol:	0.0006		
Butanol:	0.9950		
Acetone:	0.0001		
Dodecane:	0.0032		
Temperature:	279	°F	
Feed Stream ID			
Exit Stream ID			
Design Data:	Pump Efficiency:	0.57	
	Driver Efficiency:	0.87	
	Volumetric Flow Rate (ft^3/hr):	1,074	
	Head Developed (ft-lbf/lb):	130	
	Single Stage	yes	
	Shaft:	stainless steel	
	Motor Material:	stainless steel	
	Pump Material:	stainless steel	
	Pressure Change	22	psi
Utilities:	28,235	kW-hr/yr	
Cost of Utilities:	$1,129	per yr	
Comments and Drawings:	Sizing and cost calculations in Appendix A.		

PUMP				
Identification:	Item	Centrifugal Pump		
	Item No.	Pump 101-102		
	No. Required	2		
Cost of Bare Module:	$153,844			
Function:	To pump the diluted glucose across the heat exchangers			
	before entering the fermenters.			
Operation:	Continuous			
Materials Handled:				
Quantity (lb/hr)	4,428,604			
Composition				
Water	0.9029			
Glucose	0.0874			
Solids	0.0097			
Temperature (°F)	243			
Feed Stream ID				
Exit Stream ID				
Design Data:	Pump Efficiency:	0.88		
	Driver Efficiency:	0.92		
	Volumetric Flow Rate (ft^3/hr):	74,283		
	Head Developed (ft-lbf/lb):	116		
	Single Stage	Yes		
	Shaft:	Stainless Steel 304		
	Motor Material:	Stainless Steel 304		
	Pump Material:	Stainless Steel 304		
	Pressure Change (psi):	48		
Utilities:	1,524,227	kW-hr/yr		
Cost of Utilities	$60,969	per yr		
Comments and Drawings:	Sizing and cost calculations in Appendix A.			

PUMP				
Identification:	Item	Centrifugal Pump		
	Item No.	Pump 103		
	No. Required	1		
Cost of Bare Module:	$145,032			
Function:	To pump the diluted glucose into the breeder tanks.			
Operation:	Continuous			
Materials Handled:				
Quantity	8,254,383	lb/hr		
Composition				
Butyric Acid	0.0057			
Acetic Acid	0.0001			
Lactic Acid	0.0003			
Water	0.9833			
Solids	0.0106			
Temperature	37	°C		
Feed Stream ID				
Exit Stream ID				
Design Data:	Pump Efficiency:	0.89		
	Driver Efficiency:	0.92		
	Volumetric Flow Rate (ft^3/hr):	148,567		
	Head Developed (ft-lbf/lb):	65		
	Single Stage	yes		
	Shaft:	stainless steel		
	Motor Material:	stainless steel		
	Pump Material:	stainless steel		
	Pressure Change	18	psi	
Utilities:	528,624	kW-hr/yr		
Cost of Utilities	$21,145	per yr		
Comments and Drawings:	Sizing and cost calculations in Appendix A.			

PUMP				
Identification:	Item	Centrifugal Pump		
	Item No.	Pump 104		
	No. Required	1		
Cost of Bare Module:	$38,353			
Function:	To pump the diluted glucose into the first stage fermenters.			
Operation:	Continuous			
Materials Handled:				
Quantity (lb/hr)	1,262,411			
Composition				
Water	0.9029			
Glucose	0.0874			
Solids	0.0097			
Temperature (°F)	99			
Feed Stream ID				
Exit Stream ID				
Design Data:	Pump Efficiency:	0.83		
	Driver Efficiency:	0.89		
	Volumetric Flow Rate (ft^3/hr):	21,175		
	Head Developed (ft-lbf/lb):	170		
	Single Stage	Yes		
	Shaft:	Stainless Steel 304		
	Motor Material:	Stainless Steel 304		
	Pump Material:	Stainless Steel 304		
	Pressure Change (psi):		18	psi
Utilities:	673,663	kW-hr/yr		
Cost of Utilities	$26,947	per yr		
Comments and Drawings:	Sizing and cost calculations in Appendix A.			

PUMP				
Identification:	Item	Centrifugal Pump		
	Item No.	Pump 105		
	No. Required	1		
Cost of Bare Module:	$127,725			
Function:	To pump the recycle from FBB4-6 to BREED2.			
Operation:	Continuous			
Materials Handled:				
Quantity (lb/hr)	7,613,874			
Composition				
Water	0.9029			
Glucose	0.0874			
Solids	0.0097			
Temperature (°F)	99			
Feed Stream ID				
Exit Stream ID				
Design Data:	Pump Efficiency:	0.89		
	Driver Efficiency:	0.92		
	Volumetric Flow Rate (ft^3/hr):	127,711		
	Head Developed (ft-lbf/lb):	260		
	Single Stage	Yes		
	Shaft:	Stainless Steel 304		
	Motor Material:	Stainless Steel 304		
	Pump Material:	Stainless Steel 304		
	Pressure Change		18	psi
Utilities:	5,825,278	kW-hr/yr		
Cost of Utilities	$233,011	per yr		
Comments and Drawings:	Sizing and cost calculations in Appendix A.			

PUMP				
Identification:	Item	Centrifugal Pump		
	Item No.	Pump 106		
	No. Required	1		
Cost of Bare Module:	$14,488			
Function:	To pump the recycle stream from FBB1-3 to BREED1.			
Operation:	Continuous			
Materials Handled:				
Quantity	40,837	lb/hr		
Composition				
Butyric Acid	0.0057			
Acetic Acid	0.0001			
Lactic Acid	0.0003			
Water	0.9833			
Solids	0.0106			
Temperature	37	°C		
Feed Stream ID				
Exit Stream ID				
Design Data:	Pump Efficiency:	0.52		
	Driver Efficiency:	0.82		
	Volumetric Flow Rate (ft^3/hr):	735		
	Head Developed (ft-lbf/lb):	65		
	Single Stage	yes		
	Shaft:	stainless steel		
	Motor Material:	stainless steel		
	Pump Material:	stainless steel		
	Pressure Change	18	psi	
Utilities:	328	kW-hr/yr		
Cost of Utilities	$13	per yr		
Comments and Drawings:	Sizing and cost calculations in Appendix A.			

PUMP				
Identification:	Item	Centrifugal Pump		
	Item No.	Pump 107		
	No. Required	1		
Cost of Bare Module:	$18,852			
Function:	To pump the product stream from FBB4-6 to the centrifuge.			
Operation:	Continuous			
Materials Handled:				
Quantity	191,460	lb/hr		
Composition				
Butyric Acid	0.0057			
Acetic Acid	0.0001			
Lactic Acid	0.0003			
Water	0.9833			
Solids	0.0106			
Temperature	37	°C		
Feed Stream ID				
Exit Stream ID				
Design Data:	Pump Efficiency:	0.70		
	Driver Efficiency:	0.85		
	Volumetric Flow Rate (ft^3/hr):	3,446	ft^3/hr	
	Head Developed (ft-lbf/lb):	65	ft-lbf/lb	
	Single Stage	yes		
	Shaft:	stainless steel		
	Motor Material:	stainless steel		
	Pump Material:	stainless steel		
	Pressure Change	18	psi	
Utilities:	1,152	kW-hr/yr		
Cost of Utilities	$46	per yr		
Comments and Drawings:	Sizing and cost calculations in Appendix A.			

PUMP				
Identification:	Item	Centrifugal Pump		
	Item No.	Pump 201-203		
	No. Required	3		
Cost of Bare Module:	$34,744			
Function:	To pump the diluted glucose into the first stage fermenters.			
Operation:	Continuous			
Materials Handled:				
Quantity (lb/hr)	420,784			
Composition				
Butyric Acid	0.045			
Acetic Acid	0.011			
Lactic Acid	0.004			
Water	0.930			
Solids	0.010			
Temperature (°F)	37			
Feed Stream ID				
Exit Stream ID				
Design Data:	Pump Efficiency:	0.76		
	Driver Efficiency:	0.89		
	Volumetric Flow Rate (ft^3/hr):	7,058		
	Head Developed (ft-lbf/lb):	212		
	Single Stage	Yes		
	Shaft:	Stainless Steel 304		
	Motor Material:	Stainless Steel 304		
	Pump Material:	Stainless Steel 304		
	Pressure Change		58	psi
Utilities:	305,374	kW-hr/yr		
Cost of Utilities	$12,215	per yr		
Comments and Drawings:	Sizing and cost calculations in Appendix A.			

PUMP				
Identification:	Item	Centrifugal Pump		
	Item No.	Pump 301		
	No. Required	1		
Cost of Bare Module:	$288,090			
Function:	To pump the diluted glucose into the first stage fermenters.			
Operation:	Continuous			
Materials Handled:				
Quantity (lb/hr)	8,272,106	lb/hr		
Composition				
Butyric Acid	0.0057			
Acetic Acid	0.0001			
Lactic Acid	0.0003			
Water	0.9833			
Solids	0.0106			
Temperature (°F)	37	°C		
Feed Stream ID				
Exit Stream ID				
Design Data:	Pump Efficiency:	0.89		
	Driver Efficiency:	0.93		
	Volumetric Flow Rate (ft^3/hr):	148,886		
	Head Developed (ft-lbf/lb):	194		
	Single Stage	yes		
	Shaft:	stainless steel		
	Motor Material:	stainless steel		
	Pump Material:	stainless steel		
	Pressure Change	68	psi	
Utilities:	4,723,783	kW-hr/yr		
Cost of Utilities	$188,951	per yr		
Comments and Drawings:	Sizing and cost calculations in Appendix A.			

PUMP				
Identification:	Item	Centrifugal Pump		
	Item No.	Pump 302		
	No. Required	1		
Cost of Bare Module:	$16,756			
Function:	To pump the diluted glucose into the first stage fermenters.			
Operation:	Continuous			
Materials Handled:				
Quantity	115,398	lb/hr		
Composition				
Butyric Acid	0.0057			
Acetic Acid	0.0001			
Lactic Acid	0.0003			
Water	0.9833			
Solids	0.0106			
Temperature	37	°C		
Feed Stream ID				
Exit Stream ID				
Design Data:	Pump Efficiency:	0.65		
	Driver Efficiency:	0.84		
	Volumetric Flow Rate (ft^3/hr):	2,077		
	Head Developed (ft-lbf/lb):	65		
	Single Stage	yes		
	Shaft:	stainless steel		
	Motor Material:	stainless steel		
	Pump Material:	stainless steel		
	Pressure Change	18	psi	
Utilities:	3,814	kW-hr/yr		
Cost of Utilities	$153	per yr		
Comments and Drawings:	Sizing and cost calculations in Appendix A.			

PUMP				
Identification:	Item	Centrifugal Pump		
	Item No.	Pump 401-402		
	No. Required	2		
Cost of Bare Module:	$113,910			
Function:	To raise the pressure of the feed before being fed into the liquid-liquid extractor.			
Operation:	Continuous			
Materials Handled:				
Quantity	414,660	lb/hr		
Composition				
Water	0.9939			
Ethanol	0.0001			
Butanol	0.0058			
Acetone	0.0003			
Temperature	99	°F		
Feed Stream ID				
Exit Stream ID				
Design Data:	Pump Efficiency:	0.88		
	Driver Efficiency:	0.92		
	Volumetric Flow Rate (ft^3/hr):	67,705		
	Head Developed (ft-lbf/lb):	132		
	Single Stage	yes		
	Shaft:	stainless steel		
	Motor Material:	stainless steel		
	Pump Material:	stainless steel		
	Pressure Change	33	psi	
Utilities:	163,176	kW-hr/yr		
Cost of Utilities	$6,527	per yr		
Comments and Drawings:	Sizing and cost calculations in Appendix A.			

PUMP				
Identification:	Item	Centrifugal Pump		
	Item No.	Pump 403-404		
	No. Required	2		
Cost of Bare Module:	$63,505			
Function:	To raise the pressure before being fed into the first distillation column.			
Operation:	Continuous			
Materials Handled:				
Quantity	3,431,800	lb/hr		
Composition				
Water	0.0003			
Ethanol	0.0001			
Butanol	0.0069			
Acetone	0.0001			
Dodecane	0.9926			
Temperature	410	°F		
Feed Stream ID				
Exit Stream ID				
Design Data:	Pump Efficiency:	0.88		
	Driver Efficiency:	0.90		
Volumetric Flow Rate (ft3/hr):		75,550		
Head Developed (ft-lbf/lb):		153		
	Single Stage	yes		
	Shaft:	stainless steel		
	Motor Material:	stainless steel		
	Pump Material:	stainless steel		
	Pressure Change:	28	psi	
Utilities:	1,555,197	kW-hr/yr		
Cost of Utilities	$62,208	per yr		
Comments and Drawings: Sizing and cost calculations in Appendix A.				

PUMP				
Identification:	Item	Centrifugal Pump		
	Item No.	Pump 405-406		
	No. Required	2		
Cost of Bare Module:	$56,030			
Function:	To raise the pressure before being fed into the second distillation column.			
Operation:	Continuous			
Materials Handled:				
Quantity	24,819	lb/hr		
Composition				
Water	0.0352			
Ethanol	0.0097			
Butanol	0.9410			
Acetone	0.0111			
Dodecane	0.0030			
Temperature	251	°F		
Feed Stream ID				
Exit Stream ID				
Design Data:	Pump Efficiency:	0.42		
	Driver Efficiency:	0.91		
Volumetric Flow Rate (ft3/hr):		556		
Head Developed (ft-lbf/lb):		122		
	Single Stage	yes		
	Shaft:	stainless steel		
	Motor Material:	stainless steel		
	Pump Material:	stainless steel		
	Pressure Change:	18	psi	
Utilities:	18,573	kW-hr/yr		
Cost of Utilities	$743	per yr		
Comments and Drawings:				

PUMP				
Identification:	Item	Centrifugal Pump		
	Item No.	Pump 501		
	No. Required	1		
Cost of Bare Module:	$104,462			
Function:	To raise the pressure of the solvent before being fed into the second distillation column.			
Operation:	Continuous			
Materials Handled:				
Quantity	6,813,900	lb/hr		
Composition				
Water	0.0000			
Ethanol	0.0000			
Butanol	0.0000			
Acetone	0.0000			
Dodecane	1.0000			
Temperature	483	°F		
Feed Stream ID				
Exit Stream ID				
Design Data:	Pump Efficiency:	0.89		
	Driver Efficiency:	0.90		
	Volumetric Flow Rate (ft3/hr):	195,480	ft^3/hr	
	Head Developed (ft-lbf/lb):	110	ft-lbf/lb	
	Single Stage	yes		
	Shaft:	stainless steel		
	Motor Material:	stainless steel		
	Pump Material:	stainless steel		
	Pressure Change	27	psi	
Utilities:	3,277,438	kW-hr/yr		
Cost of Utilities	$131,098	per yr		
Comments and Drawings:	Sizing and cost calculations in Appendix A.			

EXTRACTOR			
Identification:	Item	Liquid-Liquid extractor	
	Item No.	EXTRACT 101	
	No. Required	1	
Cost of Bare Module:	$312,419.72		
Function:	Separate most of the water out of the feed to the separations process.		
Operation:	Continuous		
Materials Handled:	FEED	SOLVENT	
Quantity (lb/hr)	8,290,000.00	6,810,000.00	
Composition			
Water	0.9940	0	
Ethanol	0.0001	0	
Butanol	0.0058	0	
Acetone	0.0003	0	
Dodecane	0.0000	1	
Feed Stage	20	1	
Temperature (°F)	98.60	43.00	
Feed Stream ID			
Exit Stream ID			
Design Data:	# Stages:	20	
	Diameter (ft);	8.40	
	Height (ft):	54.00	
	L/D Ratio:	6.5:1	
	Vol Flow Rate Feed (ft^3/hr):	135,457.52	
	Vol Flow Rate Solvent (ft3/hr):	195,128.94	
	Pressure (psi):	30	
	Top Stage Temp (°F):	280.00	
	Bot. Stage Temp (°F):	123.00	
	Material:	Stainless Steel	
Comments and Drawings:			

Start-up and Spare Equipment Needs

In order to startup the plant, all equipment needs to be cleaned and rinsed, to ensure that no ancillary material such as solids from the dry grind process or any biomass remain inside. At the beginning of the process year, all breeder tanks and fermenters are to be autoclaved at 250°F for one hour. The resulting wastewater would then be pumped to the centrifuge and the clean water would be recycled back to the glucose dilution tank.

Several pieces of spare equipment also need to be purchase to avoid process disruption in the event of operation difficulties. One spare pump for each major stream was incorporated into the total summary. Every stream that only had one pump in operation had a spare, while pumps in series had one spare for the whole set. These pumps will remain uninstalled since the installation of the pump will not significantly disrupt the continuous process.

Alternative Designs and Process Optimization

Introduction

Many options were weighed during preliminary process synthesis. To judge which of these options to adopt as part of the final design, consistent criteria were established. Beyond the standard criteria of safety, economic feasibility, simplicity, and others, criteria were established specific to the production of butanol from corn. Because the butanol process parallels the ethanol process, many Midwestern ethanol plants can be easily converted. Thus, retrofitting becomes a priority, and keeping the process as similar as possible in equipment type, resource consumption, and waste was a criteria. Raw materials costs had special meaning for Midwestern plants: a closed water process (minimal water replacement) was a firm constraint for all proposed designs. Additionally, ethanol plants sell DDGS byproducts, which are also produced by the corn fermentation for butanol. Reducing the need for replacement by keeping equipment simple and with minimal moving parts was – as shown further – overridden by the retrofication criteria. Bioreactor design considerations

Biobutanol is fermented by bacteria different than yeast for ethanol, and as a result require specialized conditions for continuous fermentation. To these ends, the literature unanimously recommended fibrous bed bioreactors (FBB) for both fermentation steps. The alternative is to host a continuous tank reactor with no substrate. This is a spiral wound matrix of fibers from one of several sources. FBB advantages are generally advised as tenfold increases in productivity and up to 1 year stable continuous operation. Bacteria grown on FBB are far more resilient to solids and can handle greater fractions of solids. These advantages result from the way cells grow and replenish on a high-surface-area fibrous surface; dead cells fall to the bottom and make room for new cells. The final advantage to FBB is its promising scalability to high volumes in pilot tests published by ButylFuel LLC.

Solids Removal

The ethanol process removes fermentation solids in two ways: first by concentration in the bottoms of a beer column, then by centrifugation of the final concentrated solids stream. The beer column is a baffle tray column that can handle solids to fractions up to 15%[12]. This would be ideal because it is not only already found on existing sites, but also produces DDGS solids feed that is completely free of solvents. Drying then only removes water and there is no doubt about product quality when selling to feedstock suppliers. Also there is no product loss in solids removal. Verifying that the feed and bottoms stream solid fractions were beneath the maximum levels (14-17%) this was the most straightforward and robust option. The beer column option was originally favored for reasons of engineering simplicity, cost efficiency, industry precedent, and market adaptability.

Next vacuum filtration was examined as an option because this would contain no moving parts. However, recommendations by our design consultants heavily weighted the factor of ethanol retrofitting. Centrifugation was chosen over vacuum filtration because these are

[12] Bruce Vrana, electronic correspondence, February 9, 2009.

already on-site in ethanol plant. Thus in the spirit of the ethanol industry we decided to use a centrifuge. Traditionally centrifuges are employed to remove liquid from beer column bottoms products, at around 14% solids by mass. The present process contains a far lower fraction of mass and thus provided confidence that the centrifuge alone would suffice even at lower rpm. Also, since solids removal is only 1% of the total reactor effluent by mass, the losses of solvent and products is 0.5% of 1%, a negligible amount when considering other tradeoffs.

Butanol removal from water stream

Originally, based on research literature it appeared that CO2 or N2 stripping may be the most efficient form of butanol removal. However it quickly became clear that this research was academic in nature and not suited for scale-up or the constraints of a reactor effluent with multiple solvents. Authorities in the field such as Nasib Qureshi at the Agricultural Research Service of the USDA advised that complete solids removal was necessary for efficient stripping of the fermentation broth. Additionally, our broth contained multiple solvents which would require distillation eventually.

The beer column proposal would also have assisted in separating butanol from the water, at least to the point of azeotropic composition. From there the process was modeled using a liquid phase separation decanter to break the azeotrope, and further distillation. This was our leading design for much of the process synthesis phase.

However, when the final material balances were calculated about the reactors, it was found that due to biological product inhibition effects, butanol was only 0.5% by mass in the yield. A much more efficient separation method is needed. Boiling off all the water during distillation was not an efficient form of separation due to very large stream sizes. Liquid-liquid extraction, although not found in ethanol plants, saved massive utility costs otherwise needed to boil off the water (which is more volatile than butanol). We were able to reduce the size of the stream containing butanol by 90% molar for distillation.

Solvent choice

A butanol-removing solvent was selected for a combination of an excellent solubility for butanol compared to water, with mediocre solubility at most for acetone and ethanol. Crucially, this solvent could *not form an azeotrope* with butanol, since distillation awaited downstream. After evaluating several hydrocarbons, it was determined that dodecane was the most optimal solvent for extraction. The drawback of dodecane is its very high boiling point (483 F) which requires either extremely high pressure steam or more likely, hot oil and a furnace mechanism. This is a decision normally left to local utility prices and safety tolerances.

Selection of dodecane was through trial and error, comparing both L-L extraction simulation results, and Txy diagrams as generated by the Aspen chemical properties database. Many hydrocarbons do form an azeotrope with butanol, but dodecane, due to its high molecular weight, will never have a partial pressure equal to that of butanol. Thus, dodecane does not form an azeotrope with butanol at any mole fraction and is fitting as an extracting solvent.

Distillation Feed Pre-heating

In order to reduce size of distillation columns DIST 101 and 102, particularly in the section below the feed of the column, pre-heating of the feed stream was suggested by Professor Leonard Fabiano. The butanol-rich dodecane left the extractor at 123° F and the next decision was what temperature it should be ideally heated. Trial and error showed that 410° F was an optimal temperature, based roughly on the tray sizing result given by Aspen. Although final calculations would be done manually, it was assumed that the diameters would correlate as feed stream changed and this was a credible optimization technique. We also verified it took that much load off of the reboiler, so there was no net energy cost even if we had used steam. However, there was a better idea for using heat integration.

Heat integration

Another suggestion by Professor Fabiano was to use the 483° F bottoms from the distillation columns to preheat the incoming butanol-rich dodecane. The two streams are nearly identical in flow rates and the temperature differences were sufficient for the task. The recycled bottoms solvent would also be cooled for liquid-liquid extraction. This modification saved significant steam utilities.

Waste stream handling

Discarding of the final column overhead, half-water half-solvent stream could be achieved by either landfill removal, or more resourcefully and environmentally consciously, by re-routing to the DDGS drying process. By mixing in with the liquids to be dried from the DDGS, irrecoverable water could be removed while unusable solvents oxidized to environmentally safe compounds. These were passed with the rest of the drying process vapors into the thermal oxidizer to break down the solvents.

Final Product Specifications

Product specification of Butanol will vary with application and thus the process is left flexible for updating based on market requirements. Butanol is a novel fuel product and its application tolerances are not fully determined beyond the 99.5% purity requirement (i.e. what comprises the final 0.5%). Product specification for fuel use will require testing, and different fuel consumers can handle the purity differently. Most in question is the acetone content, which has been reduced to the lowest concentration possible through distillation. However, even as fuel, the level of intended blending for the product decides what the final ppm of acetone will be in the fuel mixture and the current 120 ppm could be well tolerated. Further reduction of acetone would be achieved through a hydrogenation reaction.

Section VI: Operating Cost and Economic Analysis

Equipment Cost Estimates

Equipment cost estimates were determined by using Seider, Seader, and Lewin's book on Product Design and Process Synthesis Principles. Additionally, Ulrich Gaeld's textbook, A Guide to Chemical Engineering Process Design and Economics was utilized to determine aspects of the analysis. Cost estimates for the holding tanks were provided by Jason Noth of Natgun. Detailed tables and calculations are compiled in the Appendix for reference.

Spares	
PUMP 102-1	$ 290,619.49
PUMP 103-1	$ 145,031.83
PUMP 104-1	$ 38,352.93
PUMP 105-1	$ 127,725.28
PUMP 301-1	$ 288,090.17
PUMP 402-1	$ 113,910.49
PUMP 404-1	$ 63,505.13
PUMP 406-1	$ 56,029.88
PUMP 501-1	$ 104,461.67
DIST-PUMP 101-1 thru D	$ 56,937.47
DIST-PUMP-103-1 thru D	$ 960,231.62
DIST-PUMP-201-1	$ 16,831.06
DIST-PUMP-202-1	$ 18,957.81
DIST-RA 201-1	$ 33,623.35
Total	$ 2,314,308.17

Table 1. Spare Equipment Costs

Table 1 details the spares purchased at plant start-up. There is a spare for every major process stream and separations section. Additionally, there is a spare reflux accumulator for the distillation column. These spares will be installed on a need basis, since the installation will not significantly affect continuous operation.

Unit Name	C_P		F_{BM}	C_{BM}		Total C_{BM}	
Agitators							
A-1	$	46,086.32	2.03	$	93,555.23	$	93,555.23
A-2	$	111,175.36	2.03	$	225,685.98	$	225,685.98
Subtotal						$	319,241.21
Fermentors							
FBB-1 thru FBB-3	$	501,239.00	3.05	$	1,528,778.95	$	4,586,336.85
FBB-4 thru FBB-6	$	696,687.00	3.05	$	2,124,895.35	$	6,374,686.05
Subtotal						$	10,961,022.90
Columns							
DIST-101 thru DIST-102	$	2,430,437.81	4.16	$	10,110,621.29	$	20,221,242.58
DIST-201	$	71,728.60	4.16	$	298,390.98	$	298,390.98
EXTRACT-101	$	104,139.91	4.16	$	433,222.03	$	3,898,998.23
Subtotal						$	24,418,631.79
Holding Tanks							
BREED-1	$	590,340.29	3.05	$	1,800,537.88	$	1,800,537.88
BREED-2	$	1,099,079.94	3.05	$	3,352,193.82	$	3,352,193.82
DILUT-1	$	1,487,101.71	3.05	$	4,535,660.22	$	4,535,660.22
DDGS-1	$	2,279,815.80	3.05	$	6,953,438.19	$	6,953,438.19
HOLD-1	$	1,784,977.00	3.05	$	5,444,179.85	$	5,444,179.85
Subtotal						$	22,086,009.96
Pumps & Motors							
PUMP-101 thru PUMP-102	$	88,066.51	3.3	$	290,619.49	$	581,238.98
PUMP-103	$	43,949.04	3.3	$	145,031.83	$	145,031.83
PUMP-104	$	11,622.10	3.3	$	38,352.93	$	38,352.93
PUMP-105	$	38,704.63	3.3	$	127,725.28	$	127,725.28
PUMP-106	$	4,390.35	3.3	$	14,488.16	$	14,488.16
PUMP-107	$	5,712.82	3.3	$	18,852.31	$	18,852.31
PUMP-201 thru PUMP-203	$	10,528.59	3.3	$	34,744.35	$	104,233.04
PUMP-301	$	87,300.05	3.3	$	288,090.17	$	288,090.17
PUMP-302	$	5,077.51	3.3	$	16,755.78	$	16,755.78
PUMP-401 thru PUMP-402	$	34,518.33	3.3	$	113,910.49	$	227,820.98
PUMP-403 thru PUMP-404	$	19,243.98	3.3	$	63,505.13	$	127,010.27
PUMP-405 thru PUMP-406	$	16,978.75	3.3	$	56,029.88	$	112,059.75
PUMP-501	$	31,655.05	3.3	$	104,461.67	$	104,461.67
DIST-PUMP 101 thru DIST-PUMP 102	$	8,626.89	3.3	$	28,468.74	$	56,937.47
DIST-PUMP-103 thru DIST-PUMP 104	$	145,489.64	3.3	$	480,115.81	$	960,231.62
DIST-PUMP-201	$	5,100.32	3.3	$	16,831.06	$	16,831.06
DIST-PUMP-202	$	5,744.79	3.3	$	18,957.81	$	18,957.81
Subtotal						$	2,959,079.09

Heat Exchangers						
HX-101	$	585,469.60	3.17	$	1,855,938.63	$ 1,855,938.63
HX-102	$	241,024.39	3.17	$	764,047.32	$ 764,047.32
HX-103	$	687,111.32	3.17	$	2,178,142.88	$ 2,178,142.88
HX-104 thru HX-116	$	2,807,016.00	3.17	$	8,898,240.72	$ 8,898,240.72
HX-201 thru HX-203	$	233,380.10	3.17	$	739,814.92	$ 2,219,444.75
HX-204 thru HX-206	$	78,102.44	3.17	$	247,584.73	$ 742,754.20
HX-207 thru HX-209	$	103,977.94	3.17	$	329,610.07	$ 988,830.21
HX-301 thru HX-324	$	3,101,232.00	3.17	$	9,830,905.44	$ 9,830,905.44
Subtotal						$ 27,478,304.16
Reflux Accumulators						
DIST-RA 101 thru DIST-RA 102	$	18,112.99	3.05	$	55,244.62	$ 110,489.24
DIST-RA 201	$	11,024.05	3.05	$	33,623.35	$ 33,623.35
Subtotal						$ 144,112.59
Condensers and Reboilers						
DIST-COND 101 thru DIST-COND 102	$	50,905.58	3.17	$	161,370.69	$ 645,482.75
DIST-COND 201	$	40,972.86	3.17	$	129,883.97	$ 259,767.93
DIST-REBOIL 101 thru DIST-REBOIL 102	$	71,671.70	3.17	$	227,199.29	$ 1,363,195.73
DIST-REBOIL 202	$	24,441.54	3.17	$	77,479.68	$ 232,439.05
Subtotal						$ 2,500,885.47
Total						$ 90,867,287.16

Table 2. Equipment Costs

The Equipment costs listed in Table 2 include the base purchase cost, the bare module cost, and the total bare module cost for each equipment type. The bulk of the total equipment costs stems from the custom heat exchangers and distillation columns that will have to be built on site.

The designs for the heat exchangers were chosen using equations provided Product Design and Process Synthesis Principles and from discussions with Professor Fabiano. HX 104-116 and HX 301-324, are each a customized block of heat exchangers designed using ASPEN's Tasc+. The TEMA and specification sheets provided detail the requirements of all the heat exchangers.

Two equipment costs not listed are the prices for the DDGS and Dry Grind systems. Because this butanol plant will be built on existing ethanol plants, these equipment pieces will already be installed.

Utility Requirements

Unit Electrical Requirements					
Equipment	Power Req (Kw-hr/yr)	Price per kW-hr		Cost ($/yr)	
Agitators	120,924.19	$	0.04	$	4,836.97
Pumps	42,223,501.72	$	0.04	$	1,688,940.07
Total Cost per Year				$	1,693,777.04

Unit Cooling Water Requirements					
Equipment	CW Req (gal/hr)	Price per Gal		Cost ($/hr)	
Distillation Columns	208,015.35	$	0.00005	$	10.40
Heat Exchangers	4,426,169.21	$	0.00005	$	221.31
Total Cost per Hour				$	231.71

Additional Process Water					
Equipment	Gal/hr	Price per Gal		Cost ($/hr)	
Separations	988,115.25	$	0.00050	$	494.06
DDGS	1,080.43	$	0.00050	$	0.54
Total Cost per Hour				$	494.60

Unit Steam Requirements					
Equipment	LP Steam Req (lb/hr)	Price per lb		Cost ($/hr)	
Distillation Columns	597,834.04	$	0.0055	$	3,288.09
Heat Exchangers	2,171,770.86	$	0.0025	$	5,429.43
Total Cost per Hour				$	8,717.51

Additional Dodecane	L/year	Price per L		Cost ($/year)	
Replacement Dodecane (L/year)	4,373,853.94	$	1.16	$	5,091,165.99

Utility Costs Per Year	Cost $/hr		Cost $/yr	
Electricity		--	$	1,693,777.04
Cooling Water	$	231.71	$	1,807,331.98
Steam	$	8,717.51	$	73,227,120.71
Process Water	$	494.60	$	158,271.31
Dry Grind		--	$	752,706,000.00
DDGS		--	$	17,131,000.00
Dodecane Replacement		--	$	5,091,165.99
Total Utilities			$	851,814,667.02

Table 3. Total Utility Costs

Prices for each utility are given in Table 3 and were provided in Seider, Seader, and Lewin's book on Product Design and Process Synthesis Principles. The price for Dodecane was supplied by Arnie Sapuay from Alfa Chemicals.

Values for the DDGS and Dry Grind processes were provided from the SuperPro Analysis.

Income and Costs

Waste Removal Costs	lb/year	Price per ton	Cost $/yr
Carbon Dioxide	160,869,565.00	$ 3.00	$ 241,304.35

Table 4. Waste Removal

The amount of carbon dioxide produced is given in Table 4. In order to safely release this byproduct into the atmosphere it is necessary to sequester it, at $3.00/ton. This "green CO_2" can be released into the atmosphere (See Environmental Considerations for a more detailed analysis).

Raw Material	Amount	Price	Cost
Process Water (Gal)	6,810,000.00	$ 0.00050	$ 3,405.00
Dodecane (L)	1,422,978	$ 1.16	$ 1,656,346.39
Total Raw Materials			$ 1,659,751.39

Table 5. Raw Materials

Raw materials, given in Table 5, will be a one-time purchase at plant start-up. These two items will be continuously recycled throughout the system. Replacement purchases for water or solvent loss are included in the operating utilities per year.

Income per Year		
Fuel		
Total Butanol Produced (Gal/yr)		54,338,672.00
Price per Gal		$4.00
Income from Butanol	$	217,354,688.00
ByProduct		
Total DDGS Produced (lb/yr)		773079120
Price per lb	$	0.07
Income From DDGS	$	51,796,301.04
Total Income	$	269,150,989.04

Table 6. Income per Year

Electricity	
Total kW-hr/year	42,344,425.91
kW-hr/Gal Butanol	0.78
Cooling Water	
Total Gal/year	1,482,939,058.16
Gal/Gal Butanol	27.29
Process Water	
Total Gal/year	316,542,617.60
Gal/Gal Butanol	5.83
Steam	
Total lb LP/year	16,679,200,204.80
Total lb HP/year	4,591,365,427.20
lb LP/Gal Butanol	306.95
lb HP/Gal Butanol	84.50

Table 7. Ratio of Utilities to Butanol Produced

Utility Breakdown by Unit

Unit	Unit ID	kW-hr/yr	CW (gal/hr)	Steam (lb/hr)	Total Number	Total
Agitator	A-1	21,261.40			1	21,261.40
Agitator	A-2	99,662.79			1	99,662.79
Heat Exchanger	HX-102			1,325,167.35	1	1,325,167.35
Heat Exchanger	HX-103		125,095.42		1	125,095.42
Heat Exchanger	HX-104 thru HX-116		4,153,591.84		1	4,153,591.84
Heat Exchanger	HX-204 thru HX-206		LP 50psi	282,201.17	3	846,603.51
Heat Exchanger	HX-207 thru HX-209		49,160.65		3	147,481.95
Pump	PUMP-101-102	6,055,629.00			2	12,111,258.00
Pump	PUMP-103	528,623.71			1	528,623.71
Pump	PUMP-104	673,663.12			1	673,663.12
Pump	PUMP-105	5,825,278.14			1	5,825,278.14
Pump	PUMP-106	327.76			1	327.76
Pump	PUMP-107	1,152.01			1	1,152.01
Pump	PUMP-201-203	305,373.64			3	916,120.92
Pump	PUMP-301	4,723,783.32			1	4,723,783.32
Pump	PUMP-302	3,813.76			1	3,813.76
Pump	PUMP-401-402	163,176.06			2	326,352.12
Pump	PUMP-403-404	1,555,197.29			2	3,110,394.58
Pump	PUMP-405-406	18,572.96			2	37,145.92
Pump	PUMP-501	3,277,438.19			1	3,277,438.19
Pump	DIST-PUMP 101- 102	54,115.37			2	108,230.74
Pump	DIST-PUMP-103-104	5,234,537.13			2	10,469,074.26
Pump	DIST-PUMP-201	22,200.00			1	22,200.00
Pump	DIST-PUMP-202	88,645.18			1	88,645.18
Condensers	DIST-COND 101-102		89,460.75		2	178,921.50
Condensers	DIST-COND 201		29,093.85		1	29,093.85
Reboilers	DIST-REBOIL 101-102		HP	294,586.57	2	589,173.14
Reboilers	DIST-REBOIL 202		HP	8,660.90	1	8,660.90

Table 8. Utility Breakdown by Unit

Economic Analysis

Outlined below are the results of the profitability analysis conducted on the design. Depreciation was based on the 10 years MACRS schedule, which accounts for most major fuel producing plants. The plant was operated at 90% capacity. The equipment costs for the DDGS and Dry Grind were not included in these depreciation calculations. The assumption was made that the equipment in the ethanol plant being modified has already been fully depreciated.

The following tables detail the results of the Product Design and Process Synthesis Principles profitability analysis. While many of the values were included the analysis, the cash flow table created was done by hand to accurately account for the assumptions made.

The total capital investment from the analysis was $217 million. Since the raw materials used in this process are a one-time purchase cost, and not a reoccurring fee, the total cost of raw materials, $1.65 million, was added to the total capital investment. The final value used in the cash flows was $219 million.

The cash flow summary outlines a plant-life of 15 years, with a one year build-up to full capacity. At the end of this 15 year study period, it was determined that the Net Present Value (interest rate of 15%) at the time would be roughly negative $3.55 billion, with an out of range IRR. It is clear from the investment analysis that this process is highly unprofitable and should not be pursued.

The negative NPV is reflected in the huge margin between total revenues and total costs. The total revenues for the plant, $271 million, include the sale of butanol at $4.00/gallon and DDGS at $150/bushel; however, this value is not enough to cover the costs of production. The total cost per year for this design is $897 million. The majority of this charge comes from the utility costs, in particular, from the Dry Grind process. An analysis of the operating costs for the Dry Grind indicate that 88.4% of the cost stems from the purchase of corn. In fact, the cost of corn accounts for 81.7% of the total cost for the process. A detailed analysis on the sensitivity of the process to the price of corn is outlined in the next section.

Investment Summary

April, 2009

Butanol Production Plant

			TOTAL
Bare Module Costs			
Fabricated Equipment			
Fibrous Bed Bioreactors	$10,961,000		
DIST 101-102	$22,126,000		
DIST 201	$575,200		
Heat Exchangers	$27,478,300		
Total Fabricated Equipment:	$61,140,500		
Process Machinery			
Agitators	$319,200		
All Non Distillation Pumps	$1,906,100		
Total Process Machinery:	$2,225,400		
Spares			
Pump Spares	$1,227,700		
Distillation Spares	$1,086,581		
Total Spares:	$2,314,300		
Storage			
Tanks	$22,086,000		
Total Storage:	$22,086,000		
Total Bare Module Costs:			$87,766,000
Direct Permanent Investment			
Cost of Site Preparation:	$4,388,300		
Cost of Service Facilities:	$4,388,300		
Allocated Costs for utility plants and related facilities:	$0		
Direct Permanent Investment:			$96,543,000
Total Depreciable Capital			
Cost of Contigencies and Contractor Fees:	$17,377,700		
Total Depreciable Capital:			$113,921,000
Total Permanent Investment			
Cost of Land:	$2,278,400		
Cost of Royalties:	$0		
Cost of Plant Start-Up:	$11,392,100		
Total Permanent Investment:			$127,592,000
Working Capital			
Inventory			
Butanol a 2,075,000 Gal		$8,299,000	
Process Water a 93,000 Gal		$0	
Dodecane Repla a 19,000 L		$22,500	
Total Inventory:		$8,321,600	
Accounts Receivable:		$9,879,800	
Cash Reservces:		$71,928,700	
Accounts Payable:		$150,400	
Total Working Capital:		$90,280,500	
TOTAL CAPITAL INVESTMENT			$217,872,500

Variable Cost Summary

April, 2009

Butanol Production Plant

		Per Gal Butanol		TOTAL
Raw Materials				
	Process Water	$0.00 per Gal of Butanol	$3,400	
	Dodecane Replacement	$0.03 per Gal of Butanol	$1,650,600	
	Total Raw Materials:	$0.03 per Gal of Butanol	$1,654,000	$1,654,000
Utilties				
	High Pressure Steam	$0.46 per Gal of Butanol	$25,253,900	
	Low Pressure Steam	$0.77 per Gal of Butanol	$41,698,100	
	Process Water	$0.00 per Gal of Butanol	$158,400	
	Cooling Water	$0.00 per Gal of Butanol	$74,100	
	Electricity	$0.03 per Gal of Butanol	$1,695,400	
	Dry Grind	$13.85 per Gal of Butanol	$752,693,700	
	DDGS	$0.32 per Gal of Butanol	$17,134,200	
	Dodecane Replacement	$0.09 per Gal of Butanol	$5,073,500	
	Total Raw Materials:	$15.53 per Gal of Butanol	$843,781,400	$845,435,400
Byproducts				
	Carbon Dioxide	$0.00 per Gal of Butanol	$241,300	
	DDGS	-$1.00 per Gal of Butanol	-$54,126,800	
	Total Byproducts:	-$0.99 per Gal of Butanol	-$53,885,400	$791,550,000
General Expenses				
	Selling / Transfer:	$0.12 per Gal of Butanol	$6,520,600	
	Direct Research:	$0.19 per Gal of Butanol	$10,433,000	
	Allocated Research:	$0.02 per Gal of Butanol	$1,086,800	
	Administrative Expense:	$0.08 per Gal of Butanol	$4,347,100	
	Management Incentives:	$0.05 per Gal of Butanol	$2,716,900	
	Total Byproducts:	$0.46 per Gal of Butanol	$25,104,500	$816,654,500
TOTAL		$15.03 per Gal of Butanol	$816,654,400	$816,654,400

Fixed Cost Summary

April, 2009

Butanol Production Plant

			TOTAL
Operations			
	Direct Wages and Benefits:	$3,432,000	
	Direct Salaries and Benefits:	$514,800	
	Operating Supplies and Services:	$205,920	
	Technical Assistance to Manufacturing:	$0	
	Control Laboratory:	$0	
	Total Operations:	$4,152,720	$4,152,720
Maintenance			
	Wages and Benefits:	$5,126,445	
	Salaries and Benefits:	$1,281,611	
	Materials and Services:	$5,126,445	
	Maintenance Overhead:	$256,322	
	Total Maintenance:	$11,790,823	$15,943,543
Operating Overhead			
	General Plant Overhead:	$735,195	
	Mechanical Department Services:	$248,517	
	Employee Relations Department:	$610,937	
	Business Services:	$766,259	
	Total Operating Overhead:	$2,360,908	$18,304,451
Property Insurance and Taxes			
	Total Property Insurance and Taxes:	$2,278,420	$20,582,871
TOTAL			$20,582,871

Year	Design Capacity	Revenues	Costs	BTCF	Depreciation Deduction Cost Basis	GDS	Deduction	Taxable Income	Taxes	ATCF	Present Value
0	0%	$ -	$ -	$ (1 $ (110,936,300)		---	---	---		$ (110,936,300)	$ (110,936,30
1	0%	$ -	$ -	$ (1 $ (111,850,200)		---	---	---		$ (111,850,200)	$ (97,261,04
2	45%	$ 122,161,602	$ (897,743,342)	$ (775,581,740)	$ 113,921,000	0.1000	$ 11,392,100	$ (786,973,840)	$ -	$ (775,581,740)	$ (586,451,222
3	90%	$ 244,323,204	$ (897,743,342)	$ (653,420,139)		0.1800	$ 20,505,780	$ (673,925,919)	$ -	$ (653,420,139)	$ (429,634,348
4	90%	$ 244,323,204	$ (897,743,342)	$ (653,420,139)		0.1440	$ 16,404,624	$ (669,824,763)	$ -	$ (653,420,139)	$ (373,595,085
5	90%	$ 244,323,204	$ (897,743,342)	$ (653,420,139)		0.1152	$ 13,123,699	$ (666,543,838)	$ -	$ (653,420,139)	$ (324,865,291
6	90%	$ 244,323,204	$ (897,743,342)	$ (653,420,139)		0.0922	$ 10,503,516	$ (663,923,655)	$ -	$ (653,420,139)	$ (282,491,558
7	90%	$ 244,323,204	$ (897,743,342)	$ (653,420,139)		0.0737	$ 8,395,978	$ (661,816,116)	$ -	$ (653,420,139)	$ (245,644,833
8	90%	$ 244,323,204	$ (897,743,342)	$ (653,420,139)		0.0655	$ 7,461,826	$ (660,881,964)	$ -	$ (653,420,139)	$ (213,604,202
9	90%	$ 244,323,204	$ (897,743,342)	$ (653,420,139)		0.0655	$ 7,461,826	$ (660,881,964)	$ -	$ (653,420,139)	$ (185,742,785
10	90%	$ 244,323,204	$ (897,743,342)	$ (653,420,139)		0.0656	$ 7,473,218	$ (660,893,356)	$ -	$ (653,420,139)	$ (161,515,465
11	90%	$ 244,323,204	$ (897,743,342)	$ (653,420,139)		0.0655	$ 7,461,826	$ (660,881,964)	$ -	$ (653,420,139)	$ (140,448,230
12	90%	$ 244,323,204	$ (897,743,342)	$ (653,420,139)		0.0328	$ 3,736,609	$ (657,156,747)	$ -	$ (653,420,139)	$ (122,128,896
13	90%	$ 244,323,204	$ (897,743,342)	$ (653,420,139)		0	$ -	$ (653,420,139)	$ -	$ (653,420,139)	$ (106,199,040
14	90%	$ 244,323,204	$ (897,743,342)	$ (653,420,139)		0	$ -	$ (653,420,139)	$ -	$ (653,420,139)	$ (92,346,991
15	90%	$ 244,323,204	$ (897,743,342)	$ (653,420,139)		0	$ -	$ (653,420,139)	$ -	$ (653,420,139)	$ (80,301,732

Revenues		
Fuel		
Total Butanol		54338672
Price per Gal		4
Total/year	$	217,354,688.00
DDGS		
Total DDGS		773079120
Price per lb	$	0.07
Total/year	$	54,115,538.40
Total Revenues	$	271,470,226.40

Total Capital Investment
$ 219,526,500.00

Expenses		
Utilities	$	851,814,667.02
CO2	$	241,304.35
General Expenses	$	25,104,500.00
Fixed Costs	$	20,582,871.00
Total Costs	$	897,743,342.37

NPV	$ (3,550,297,455.38)
IRR	Out of Range

Year 5		
Annual Sales:	$	244,323,203.76
Annual Costs:	$	(897,743,342.37)
Depreciation:	$	13,123,699.20
Income Tax:	$	-
Net Earnings:	$	(653,420,138.61)
Total Cap Inv:	$	219,526,500.00
ROI:		-3.98%

Sensitivity Analysis

Our greatest sensitivity lies with the price of corn and the total cost of utilities for the process. Another concern is the cost of ethanol. Detailed below is an analysis of the sensitivity of the design to the price of corn and ethanol.

Price of Corn

Table 9. Sensitivity to Corn Price

$/kg	$/bushel	Total Corn Cost	Total Costs
$ 0.200	$ 5.091	$ 1,060,987,551.80	$ 1,299,127,533.83
$ 0.180	$ 4.582	$ 954,888,796.62	$ 1,169,214,780.45
$ 0.160	$ 4.073	$ 848,790,041.44	$ 1,039,302,027.06
$ 0.140	$ 3.564	$ 742,691,286.26	$ 909,389,273.68
$ 0.120	$ 3.055	$ 636,592,531.08	$ 779,476,520.30
$ 0.100	$ 2.545	$ 530,493,775.90	$ 649,563,766.91
$ 0.080	$ 2.036	$ 424,395,020.72	$ 519,651,013.53
$ 0.060	$ 1.527	$ 318,296,265.54	$ 389,738,260.15
$ 0.040	$ 1.018	$ 212,197,510.36	$ 259,825,506.77
$ 0.038	$ 0.957	$ 199,536,897.69	$ 244,323,203.76
$ 0.024	$ 0.611	$ 127,318,506.22	$ 155,895,304.06
$ 0.023	$ 0.573	$ 119,361,099.58	$ 146,151,847.56
$ 0.020	$ 0.509	$ 106,098,755.18	$ 129,912,753.38
$ 0.016	$ 0.407	$ 84,879,004.14	$ 103,930,202.71
$ -	$ -	$ -	$ -
$ (0.020)	$ (0.509)	$ (106,098,755.18)	$ (129,912,753.38)
$ (0.040)	$ (1.018)	$ (212,197,510.36)	$ (259,825,506.77)
$ (0.060)	$ (1.527)	$ (318,296,265.54)	$ (389,738,260.15)
$ (0.080)	$ (2.036)	$ (424,395,020.72)	$ (519,651,013.53)
$ (0.100)	$ (2.545)	$ (530,493,775.90)	$ (649,563,766.91)
$ (0.120)	$ (3.055)	$ (636,592,531.08)	$ (779,476,520.30)
$ (0.140)	$ (3.564)	$ (742,691,286.26)	$ (909,389,273.68)
$ (0.160)	$ (4.073)	$ (848,790,041.44)	$ (1,039,302,027.06)
$ (0.180)	$ (4.582)	$ (954,888,796.62)	$ (1,169,214,780.45)
$ (0.200)	$ (5.091)	$ (1,060,987,551.80)	$ (1,299,127,533.83)

Year 5	Break Even
Annual Sales:	$ 244,323,203.76
Annual Costs:	$ (244,323,203.76)
Depreciation:	$ 13,123,699.20
Income Tax:	$ -
Net Earnings:	$ 4,855,768.70
Total Cap Inv:	$ 222,526,500.00
ROI:	2.18%
NPV	$ (280,755,597.67)
IRR	OUT OF RANGE

Year 5	12% IRR
Annual Sales:	$ 244,323,203.76
Annual Costs:	$ (103,930,202.71)
Depreciation:	$ 13,123,699.20
Income Tax:	$ 47,089,641.69
Net Earnings:	$ 93,303,359.37
Total Cap Inv:	$ 222,526,500.00
ROI:	41.93%
NPV	$ 198,797,443.98
IRR	12%

Year 5	Positive NPV
Annual Sales:	$ 244,323,203.76
Annual Costs:	$ (155,895,304.06)
Depreciation:	$ 13,123,699.20
Income Tax:	$ 27,862,554.19
Net Earnings:	$ 60,565,345.52
Total Cap Inv:	$ 222,526,500.00
ROI:	27.59%
NPV	$ 21,295,558.37
IRR	1%

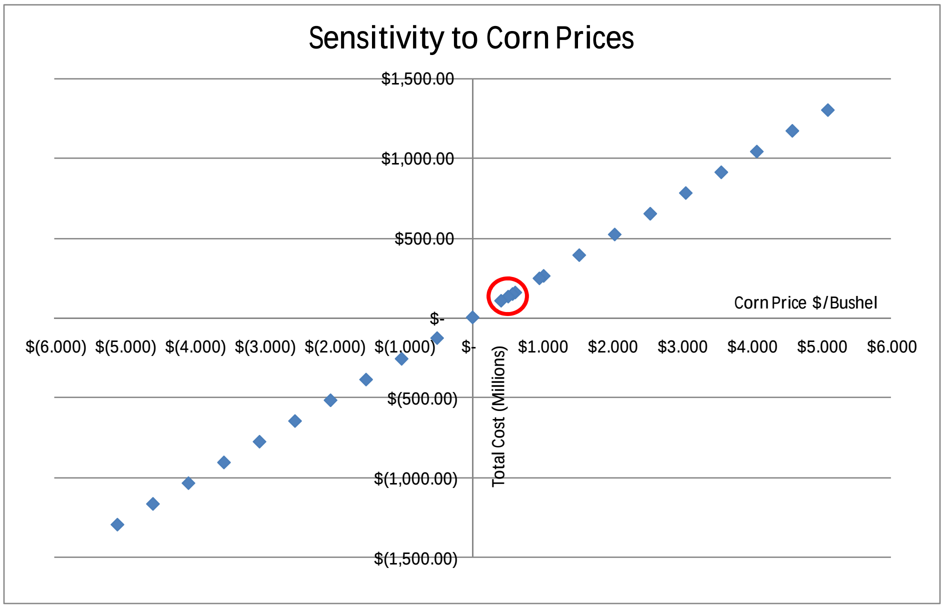


Figure 5. Sensitivity of Butanol Plant to Corn Prices

The cost of corn made up 81.7% of the total annual cost for the process. Since the price of corn significantly affects the overall profitability of the process, an analysis on the optimum price range was conducted. For a green-light on investment, the price of corn that results in a positive NPV needs to be found.

The circled area on Figure 5 refers to the value of corn price per bushel that results in a profitable investment. Three price points were determined, the breakeven point, the first point of positive NPV, and the point that achieved an IRR of 12% (See Table 9).

The breakeven price point was determined to be $0.957/bushel. At this point, the total costs equal the total revenues for the plant. It is important to note that the breakeven point is not the price at which to invest. The maximum price of corn that would allow for a profitable venture would be $0.611/bushel. Barring any other factors, if the price of corn is at or below this point, the process would be an excellent investment opportunity. Unfortunately, with the price of corn above these values, the recommendation is not to invest.

Sensitivity to DDGS and Butanol Price

	Break Even Price	Positive NPV	13% IRR	At Current DDGS Price	At Current Butanol Price
Butanol	$ 1.16	$ 4.80	$ 3.60	$ 26.00	$ 4.00
DDGS	$ 1.08	$ 1.60	$ 2.40	$ 0.07	$ 1.63
Total Revenue	$ 897,743,342	$ 1,497,752,218	$ 2,051,009,107	1466921010	1477473654
NPV	$ (191,137,151)	$ 12,951,513	$ 201,205,275	$ 2,460,743	$ 6,051,435
IRR	-6%	1%	13%	0%	0%

Table 10. DDGS and Butanol Price Sensitivity

Figure 6. DDGS Price Sensitivity

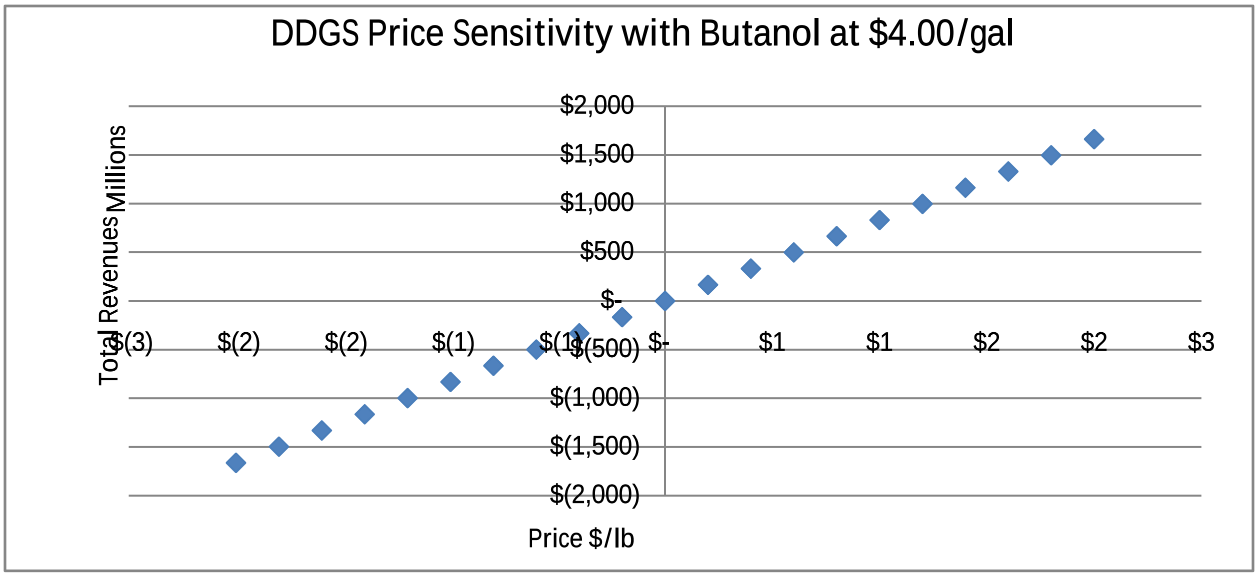


Figure 7. Butanol Price Sensitivity

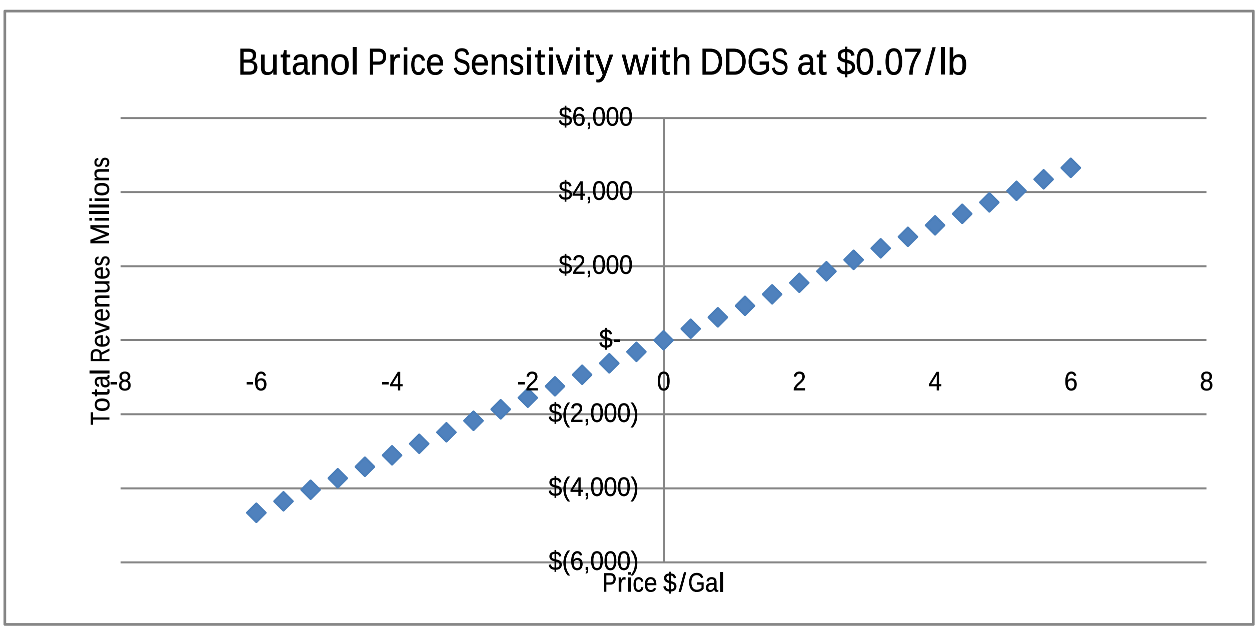
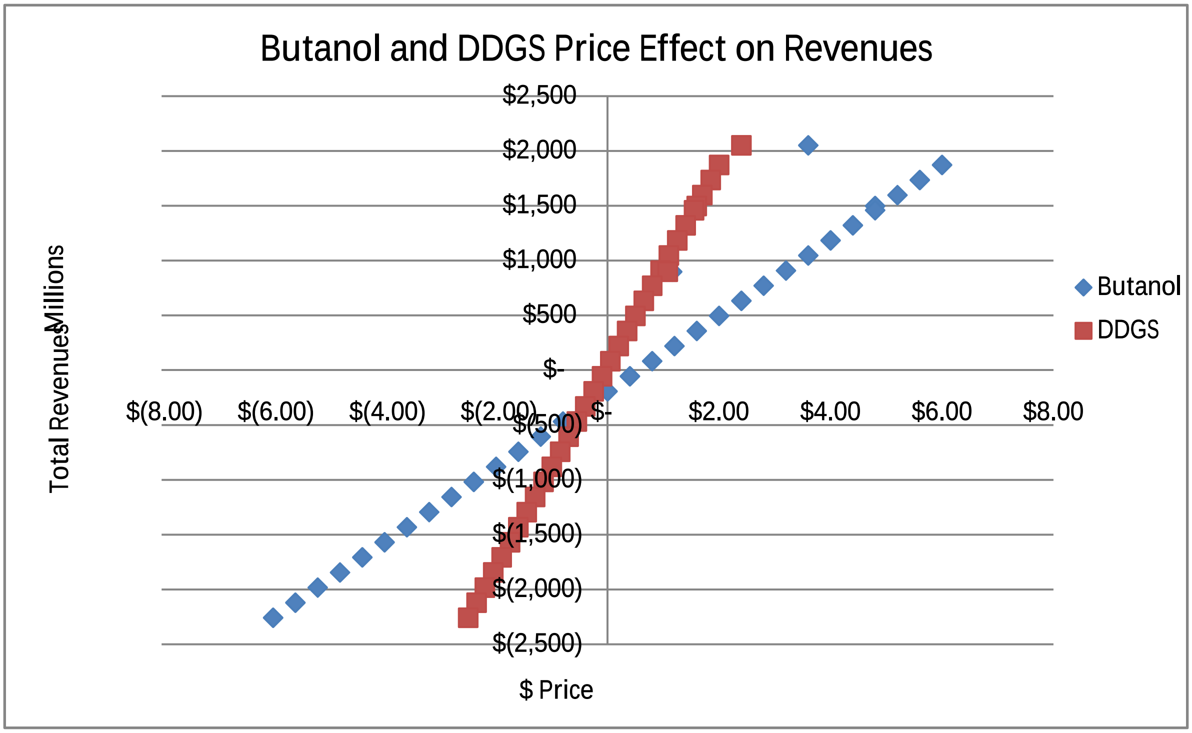


Figure 8. Butanol and DDGS Price Effect on Revenues



An analysis on the effect of varying prices for the products, DDGS and Butanol, was conducted to determine the sensitivity of the process to price fluctuations. As you can see from Figure 6, the process is very sensitive to fluctuations in the price of DDGS. This is due to the fact that DDGS makes up 93% of our revenues. Using 54.3 million gallons of butanol and 773 million pounds of DDGS as the set amount of products produced each year, the price of each commodity was varied to determine the effect on revenues and profitability. The price of butanol also affects the profitability as seen from Figure 7.

Table 10 refers to potential price points of each product that could generate positive investment values. Even if the cost of butanol were to increase in the next few years, since butanol only makes up a small fraction of the total revenues, the process is unlikely to see profitability unless the price of DDGS increases.

Sensitivity Analysis for Ethanol

In order to fully understand the gravity of the sensitivity analysis of the butanol plant, it was compared to that of ethanol. The biggest impact on the cost of producing ethanol is also from the price of the corn feed. Also, the fluctuations in selling price of ethanol vary dramatically throughout the years, and the revenue of the ethanol plant is dependent on this price. Both of these factors decrease the profitability of ethanol production as seen in the following chart, Figure 9[13]. However, the chart shows that the prices the ethanol plant is still profitable at a much higher price of corn. Even if ethanol is sold at only $1 per gallon (blue line), the plant breaks even if corn is bought at $1.97 per bushel. The price of corn must be a full dollar less in order for the butanol plant to be profitable. Also, the selling price of butanol is about 10 times higher than ethanol. For DDGS at $0.07 per lb, butanol would need to be sold at $26 per gallon, while ethanol can be sold at around $2.50.

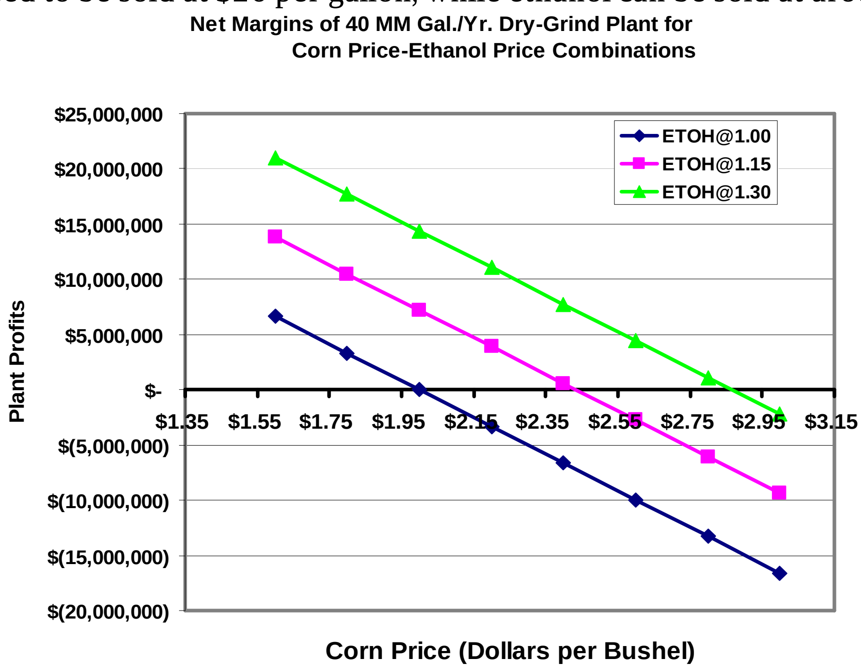


Figure 9: Sensitivity analysis of the plant profits with either rising corn prices or falling ethanol prices shows that the ethanol plant is more robust with either change.

[13] "Tracking ethanol profitability - Don Hofstrand January 2008." Iowa State University Extension. 03 Apr. 2009 <http://www.extension.iastate.edu/agdm/articles/hof/HofJan08.html>.

Another impact on the profitability of the ethanol plant is the starch content of the corn also has a big influence on the production cost. If the starch content is reduced from 59.5% to 55%, the ethanol production drops from 2.83gal/bushel to 2.62gal/bushel. For a 50 million gallon plant, that's a drop of 3.63 million gallons per year, and if ethanol is sold at $2/gallon, that's a loss of $7.27 million a year. Also, the DDGS will be affected by changes in composition of the corn feed. If there is less protein than the expected 8.3% protein mass in the feed, the DDGS will be less desirable, and the selling price will drop.

The major cost components of the ethanol plant are given in the Figure 10 below[14]. Also with most plants, the cost of the feed and the utilities make up a majority of the cost. The selling price of DDGS and any other co-products is a major factor reducing production costs.

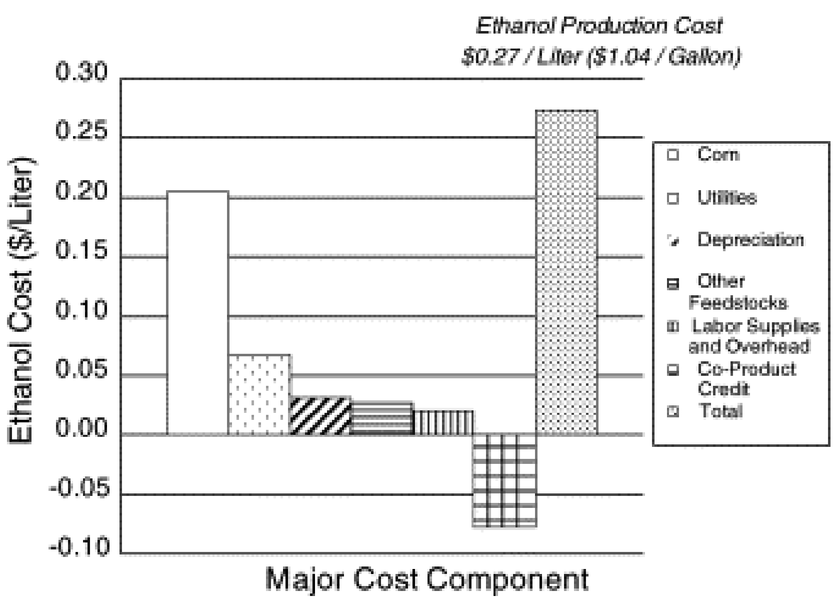


Figure 10: Major cost components that affect ethanol profitability.

[14] Campos, Edhilvia J., Hans P. Blaschek, and Nasib Qureshi. "Production of Acetone Butanol Ethanol from Degermed Corn Using Clostridium beijerinckii BA101." Applied Biotechnology and Biochemistry 99-100 (2002): 553-61.

Other Considerations

Feed

Corn. Poses no safety hazards

Dodecane. Dodecane is an industrial solvent that is hazardous in case of inhalation, slightly hazardous if ingested or if it comes in contact with skin. Severe over-exposure is fatal. Repeated exposure to chemical can result in targeted organ damage. Dodecane is combustible and is a fire hazard if in the presence of open flame, a spark, or extreme heat. Auto-ignition point is 398°F. Possible carcinogen based on laboratory animal data.

Bacteria

Clostridium Tyrobutyricum. C. Tyrobutyricum is benign to humans and animals.

Clostridium Acetobutylicum. C. Acetobutylicum is benign to humans and animals. It has been found in the human colon, but it is not known to be a part of normal human flora. This organism does not appear to be toxic to mammals, and would have to present in enormous quantities to pose any threat.

Products

Butanol. Butanol is very hazardous in case of skin contact, inhalation, and ingestion. It is also toxic at concentrations of 20g/L. Acute exposure can cause depression of the central nervous system. Holding tank must be tightly sealed and must be kept away from all possible sources of ignition.

DDGS. No health hazards, only contains dead biomass and solids from corn grind process.

Waste Disposal

CO_2. Released into the atmosphere, environmental implications discussed below.

Dodecane. A small fraction of the recycled dodecane is purged after each pass through the separation train. Since dodecane is a hazardous material, and must be handled as such. Disposal and storage regulations vary state-to-state, and the three options described below reflect hazardous waste regulations of Iowa. Note that all equipment that holds dodecane has the following safety features: explosion-proof electrical components, an automatic shutdown option, and pressure release valves. Extreme measures will be taken to ensure that dodecane is removed from any sources of ignition.

The first option is to store the waste on-site and ship the waste to a municipal landfill in appropriate containers. The fee for this is $40/ton plus transportation costs. With a 1.0% purge stream, this process will purge 262,000 tons of dodecane per year, which results in a cost of roughly $10,500,000 per year.

Another option is to store the waste on-site and ship the waste off-site to a waste treatment facility. The fee for this is $10/ton plus transportation costs, which comes to $2,620,000 per year. This option is much less expensive than the first option, although transportation costs will be slightly greater. It also eliminates a potential environmental hazard by disposing the dodecane in a waste treatment facility instead of a municipal landfill.

The third and most economically viable option would be to store and treat the solvent on-site and recycle the treated dodecane back into the system. This option would not only be more environmentally friendly than any sort of physical disposal, but it would also significantly reduce the cost of purchasing pure dodecane to feed into the extractor. The simplest way to treat hazardous waste on-site is distillation. SRS Engineering Corporation specialized in solvent recovery systems for industrial processes, and most systems obtain at least 99% purity. Assuming this degree of purity, only 683 lb/hr of dodecane would need to be purchased to replenish the extractor. The bottoms product is considered non-hazardous waste and can be disposed of in a municipal landfill for a minimal fee. A solvent recovery system can be purchased from SRS Engineering Corporation for roughly $3,000,000.[15]

Installing this system will significantly reduce the amount of fresh dodecane required per hour, since only 1% of the 1% purge is being lost. This results in a savings of over $5,000,000 per year. These savings in itself pays for the purchase and installation of the recovery system.

Water. All process water is either recycled or lost to the production of DDGS.

DDGS. Solids from corn grind as well as dead biomass is processed and sold as DDGS.

Safety and Health

The production of butanol involves few hazardous materials, yet great care needs to be taken when handling these materials, such as dodecane and butanol. All technicians must be sufficiently trained to take all necessary safety precautions. Also, all storage containers must be kept at the proper temperatures and pressures to minimize any combustion risks.

[15] Correspondence with Kevin Huisinga, Sales Manager at SRS Engineering Corporation. kevinh@srsengineering.com

Environmental Concerns

Because the first series of fermenters produce CO_2, there are environmental concerns about the release of the gas into the atmosphere. This process releases 78,000 tons CO_2 into the atmosphere each year, which is equivalent to the amount of CO_2 emitted by about 13,500 cars per year. However, this number is relatively low compared to the ethanol industry. An ethanol plant that produces 50 million gallons per year from glycerol releases 150,000 tons CO_2 each year, almost double that of this process.

While this process emits a significantly smaller amount of CO_2, any release of CO_2contributes to the already dangerously high level of Greenhouse Gases in the atmosphere. Still, all things considered, this process can reasonably be considered environmentally friendly, mainly due to the face that all CO_2produced is often referred to as "green CO_2." The concept of green CO_2comes from the idea that the release of this gas from a chemical process can mimic plants' natural uptake and release of CO_2through photosynthesis and respiration. Applied to this specific process, the corn fed into the dry grind process has, over its lifetime, absorbed some amount of CO_2, and any amount of CO_2released into the atmosphere because of its fermentation will ultimately be removed from the atmosphere by new corn crops that will be grown to produce more butanol. It is therefore believed that there would be no net change in the concentration of CO_2in the atmosphere due to this process. Although some of these claims cannot be confirmed, the concept of "green CO_2" basically proposes the idea that any CO_2 released from a biofuel process will be both on a smaller scale and less harmful to the environment than the combustion of fossil fuels.

As previously mentioned, the CO_2 produced in this process is also to both mix the contents of the fermenters and control the pressure in order to keep the system anaerobic. Utilizing the gas throughout the process not only makes the process more green, but also saves money since it eliminates the need for the purchase and installation of additional equipment or reagents that would normally be used to complete these tasks.

Key Comparison Statistics

Table 11: Comparison of resources consumed in both ethanol and butanol production processes. All values normalized to a btu of liquid fuel energy.

	Ethanol	Butanol
Equivalent of C02 released per 50 MM gallons	26,000 new cars	13,500 new cars
BTU per gallon fuel product	84,000	110,000
BTU of Fuel per BTU of process	2.40	5.53
BTU of Fuel per Dollar operating costs	43,299	6,667
BTU of Fuel per bushel of corn	240,000	29,333
C-to-C in-to-out	95%	8.00%
Reid value (Volatility)	2	0.33

This comparison of butanol and ethanol production processes, as seen from Table 11, normalizes values of both processes along btu - the measure of energy. While production of a btu of butanol is more carbon-friendly and produces a product with higher energy density, the true advantage is in operating energy invested to produce one btu of fuel. All other factors removed, butanol is ideal in that it produces more than twice as much fuel energy from the same amount of utilities.

This is leaving out a comparison of costs and amount of energy per dollar invested in variable cost. Assuming that with more examination the process and utility optimization can be refined, a material and energy comparison is most salient.

Butanol's major disadvantage, therefore, is its clear material inefficiency. As seen in the cost of utilities for the dry grind process in Table3, production of butanol, in terms of raw materials alone, requires almost ten times more corn per btu than ethanol. If there is no issue of corn demand, or another source of glucose can be scaled to the needs of the process, butanol has a clear advantage over ethanol. However, in a world where corn is not only limited but a source of food, this is clearly an unwise choice of biofuel.

An important advantage of this butanol plant over ethanol is seen in the energy consumption versus energy produced. Current energy benchmark is about 35,000 BTU/gallon of product in fuel ethanol plants (Determined from the amount of heat and electricity needed by the process). At 84,000 BTU per gallon, this is 0.42 BTU consumed/BTU ethanol produced. An analysis of our heat and electrical requirements for the process resulted in an energy benchmark of 19,900 Btu/gallon of butanol produced. At 110,000 BTU/gallon butanol, this translates to 0.18 BTU consumed/BTU butanol produced. This is a 57% decrease in energy consumption from the ethanol plant.

As a fuel, the biggest advantage of butanol over ethanol is the greater amount of energy and therefore miles driven per gallon. Butanol has 110,000 BTUs per gallon, whereas ethanol only has 84,000 BTUs per gallon. That is an extra 30% of energy available per gallon. The energy per gallon of gasoline is 115,000 BTUs. A gallon of butanol gives 96% of the energy as a gallon of gasoline, while ethanol only produces 73% of gasoline's energy per gallon. With more energy per gallon, more miles per gallon can be driven using butanol than with ethanol. Butanol has been reported to get up to 24 highway miles per gallon, while ethanol has only reached about 16 miles per gallon, a 50% increase.

The relatively low yield of butanol indicates the inefficiency of this process.

Carbon Feed/Carbon Product Yield - Butanol

$$Carbon\ Atoms\ in\ Feed = 426482.6\frac{kg}{hr}*\frac{1mole}{0.180kg}*\frac{6.023E23mlcs}{1mole}*\frac{6C}{1mlc} = 8.56E30\ C/hr$$

$$Carbon\ Atoms\ in\ Product = 20850.9\frac{kg}{hr}*\frac{1mole}{0.074kg}*\frac{6.023E23mlcs}{1mole}*\frac{4C}{1mlc} = 6.79E29\ C/hr$$

$$Yield\colon \frac{6.79E29}{8.56E30} = 8\%$$

Carbon Feed/Carbon Product Yield - Ethanol

$$Carbon\ Atoms\ in\ Feed = 30252\frac{kg}{hr}*\frac{1mole}{0.180kg}*\frac{6.023E23mlcs}{1mole}*\frac{6C}{1mlc} = 6.07E29\ C/hr$$

$$Carbon\ Atoms\ in\ Product = 14688.84\frac{kg}{hr}*\frac{1mole}{0.046kg}*\frac{6.023E23mlcs}{1mole}*\frac{3C}{1mlc} = 5.77E29\ C/hr$$

$$Yield\colon \frac{5.77E29}{6.07E30} = 95\%$$

The great disparity in the amount of raw material that actually goes into making the product is seen when comparing the carbon yield from feed to product for butanol and ethanol. The amount of carbon in the glucose feed into the fermenters that becomes the butanol product is only 8%, while in the ethanol plant, 95% of the carbon is used in the ethanol plant.

Butanol production requires significantly more glucose than ethanol because the nutrient requirements for the bacteria strains used in this process are quite different from that of yeast, which are typically used for ethanol production.

For optimal growth conditions, a glucose feed supplemented with P2 medium is required. In the first series of fermenters, glucose is fed into the reactor at 92 g/L and approximately half of the glucose is converted into acids, while the other half is utilized as nutrients for *C.*

Tyrobutyricum. There are even greater glucose requirements for the second series of fermenters. The three solventogenesis fermenters are five times the size of the three acidogenesis fermenters, and it is necessary to have 15 times as much glucose in the feed stream as butyric acid. The cell density in the Fibrous Bed Bioreactors is also much larger than in traditional fermentation, and the number of viable cells in the reactor at any point in time remains relatively constant. A great deal of nutrients is required to sustain these microbes; however, the high cell density is also responsible for the remarkable speed of the fermentation process.

Ethanol plants traditionally use yeast to ferment the starch source, which do not require the addition of glucose, only enzymes. Because no additional glucose is needed for nutrients, it is virtually all eventually converted into solvents, which is not the case for butanol production.

Improvements

In theory, biofuels are a desirable alternative to fossil fuels. Not only do biofuels release fewer pollutants when used as gasoline, but increasing the use of domestically produced biofuels will greatly reduce the United States' dependence on foreign oil. Butanol in particular has an energy value closer to gasoline than ethanol, it can be blended into gasoline at high levels (up to 100%), and can be easily integrated into the current fuel infrastructure. However, the high operating costs are still an obstacle to the potential profitability of butanol production.

The large amount of water required for the process is a very significant operating cost. While little water is being wasted, a very large amount of energy is required to sterilize and subsequently cool the streams, an operation that must happen frequently throughout the process. The amount of water is fixed because the bacteria strains are very sensitive to solvent and nutrient concentrations. Cell growth is significantly inhibited by small concentrations of solvents, the highest concentration Clostridium Acetobutylicum can withstand is only 13g/L. Continuous operation and using the Fibrous Bed Bioreactor have certainly helped to offset the effects of this problem, but unfortunately do not reduce operating costs. In order for butanol production to become economically competitive with fossil fuels, however, further optimization is needed. Potential ways to reduce operating costs would be to genetically engineer bacteria strains that can sustain higher concentrations of solvents. If this is possible, less water will be needed, and therefore, less energy will be required to sterilize and subsequently cool the streams. Another consideration would be to find an alternative sterilization method. For example, sterilization via radiation may be a viable option in the future. Another major concern regarding biofuels production centers around the “food v. fuel” argument. There is constant debate regarding the risks associated with allocating farmland for the production of biofuels. There are concerns that the global food supply will begin to greatly diminish if more and more farmland is used towards the production of biofuels instead of for the supply of food.

If the U.S. decided to commit its entire corn and soybean production to biofuels would satisfy only 12% of the country’s entire demand for gasoline and 6% of the demand for diesel. With the scarcity of available land, it would be difficult to parcel out a significant portion of land that could support the biofuel industry. Because of the higher price of biofuel, farmers would be more inclined to cell their harvests to industry, thus reducing the overall supply of food, which would increase the price of food worldwide. Other countries, such as Brazil and Germany, found a solution to this problem by allocating land that would be used strictly to farm crops for the production of biofuel. Since this farmland supplied a sufficient amount of raw material, the biofuels industry caused minimal shifts in the price of food. It would be ideal for the United States to adopt this approach if it decided to invest significant resources into biofuels. Another solution to make biofuel production more sustainable would be to use a non-food source as the raw material. There are a range of materials that can be fermented into biofuel, for example, algae and solid biomass. Solid biomass materials consist of wood, charcoal, dried animal excrement, peat, waste materials from crops, etc. An additional benefit to using solid biomass is that there is no net release of CO_2. While these materials can be successfully used to produce biofuel, little has been done on a large, industrial scale.

Section VII: Conclusions and Recommendations

Our analysis of the process of producing butanol via a two-stage fermentation process has generated unprofitable projections about the economic viability and implementation of the facility. It is our recommendation that the process as a whole is not rushed into as an investment opportunity. Due to a variety of factors, we feel that this method of producing fuel butanol, would in the long-term not be economically practical nor sustainable.

One particular factor that significantly affected our financial analysis was the costing of utilities throughout the process. In order to achieve the correct concentration of glucose entering the fermenter, a large quantity of process water was required. Not only is process water costly, but the equipment requirements for handling such large stream flows were very expensive and too large to practically build. Most of our equipment will have to be manufactured on-site, and if it cannot be, high shipping costs will result. Additionally, the necessity of sterilization between the stages results in significant expenditures in purchasing steam and cooling water, not to mention enormous heat exchangers.

A major portion of our utilities stemmed from the cost of corn. This value represented 81.7% of the total annual costs for the process. At this time, the price per bushel of corn is too expensive to result in a profitable venture. Based on our sensitivity analysis, if the price of corn drops below $0.611/bushel, the venture would prove lucrative. However, as the market analysis indicated, there are multiple external factors that need to be considered before any investment considerations are made.

Additionally, it is important to note the dependence of DDGS on our profitability. If DDGS prices decrease, with no change it ethanol, it is highly unlikely that a process like ours can be implemented. Since DDGS makes up a huge fraction of our total revenues, considerable due diligence to the DDGS market must be taken before any investment can be made.

Furthermore, taking a look at the current economic landscape reveals a poor investment period. Despite the recent stimulus package, allocating $6 billion to alternative energy research, many venture capital and private equity firms are having difficulty generating funds for large investment ventures, such as this process. Even if funding could be secured, pessimistic views on the economy have led to low valuations, which could result in a significantly lower IRR then the 32.7% calculated in this analysis.

As discussed in the Customer Requirements, one major drawback to this process is the low yield of butanol relative to the amount of glucose added to the fermenter. Carbon recovery yield was only 5%. The majority of this loss of carbon results from the strain of bacteria used in the process. Clostridium only converts approximately 50% of the glucose entering the process, and the rest is consumed as cell nutrients.

Going forward, we recommend further research into the Clostridium strains being utilized in this process. Finding a way to mitigate the loss of carbon through the process could provide greater incentives to pursue this project. Additionally, a look into the necessity of sterilization should be conducted. Much of the operating costs could be reduced by finding alternative ways to sterilize the streams and ensure product purity.

Another area of interest would be finding an alternative carbon source. Rising cost of food crops will create a demand in the corn market, driving up the price of raw materials for this process. Since the design requires such significant amounts of corn, we feel the use of arable land for producing such a small percent yield of butanol to be unrealistic. Advances in biofuels have been investigating the use of biomass (crop waste) as a new carbon source.

Overall, more research should be conducted into the science behind this process. From an economic standpoint, until the production cost of butanol become competitive with current fossil fuel technologies, or existing ethanol production, this process will not be adapted by the mainstream market. Butanol as a fuel, has a multitude of advantages over existing fuels; however, at this point in time, we feel that the short-term is not the best time to pursue this project.

Section VIII: Acknowledgements

We would like to take the time to thank the following faculty and consultants for their great advice, expertise, and time during this project:

- Professor Fabiano for his never-ending assistance with Aspen, heat exchangers, pricing, sizing, and basically every part of the process.
- Dr. Holleran for his guidance during the meetings and asking the questions we didn't think of. Without him, we would be nowhere.
- Bruce Vrana for inspiring the problem statement for this project, and for his vast knowledge of the biofuels industry.
- Dr. Tiffany Rau for all of her assistance with the biological aspects of the project.
- Mr. Gary Sawyer for his valuable input during our weekly meetings, and his assistance in making tough decisions about our process synthesis.
- Mr. Stevie for always having something meaningful to say.
- Mr. David Kolesar and Mr. John Wismer for their insightful contributions.

Section IX: Bibliography

Breeze, Paul. "The Future of Global Biomass Power Generation." Business Insights. 2004.

Biofuel Guide - Ethanol and Biodiesel as alternative energy. 03 Apr. 2009 <http://biofuelguide.net>.

Bryner, Michelle. "BP and DuPont Expand Biofuels Pact." Chemical Week 25 Feb. 2008: 7.

"BUTANOL Advances in Biofuels." The Light Party. 03 Apr. 2009 <http://www.lightparty.com/Energy/Butanol.html>.

Campos, Edhilvia J., Hans P. Blaschek, and Nasib Qureshi. "Production of Acetone Butanol Ethanol from Degermed Corn Using Clostridium beijerinckii BA101." Applied Biotechnology and Biochemistry 99-100 (2002): 553-61.

Daugulis, Andrew J. and Finn Kollerup. "A Mathematical Model for Ethanol Production by Extractive Fermentation in a Continuous Stirred Tank Fermentor." Biotechnology and Bioengineering, Vol. XXVII, pp. 1335-1346 (1985) John Wiley & Sons, Inc.

Eastwood, Gary. The Biofuels Market Outlook. Publication. 2007. Business Insights.

"Emission Facts: Average Carbon Dioxide Emissions Resulting from Gasoline and Diesel Fuel | US EPA." U.S. Environmental Protection Agency. 03 Apr. 2009 <http://www.epa.gov/otaq/climate/ >.

Energy Information Administration - EIA - Official Energy Statistics from the U.S. Government. 03 Apr. 2009 <http://www.eia.doe.gov/ >.

"First U.S. Carbon-Dioxide Permits Sell for $3.07/Ton (Update3) -." Bloomberg.com. 03 Apr. 2009 <http://www.bloomberg.com/apps/news >.

Fischer, Curt R., Daniel Klein-Marcuschamer, and Gregory Stephanopoulos. "Selection and Optimization of microbial hosts for biofuels production." Metabolic Engineering 10 (2008): 295-304. Elsevier. <http://www.elsevier.com/locate/ymben>.

France. Interntational Energy Agency. 9 rue de la Federation. Biofuels for Transport: An International Perspective. 2004.

Green, Don W., and Robert H. Perry. Perry's Chemical Engineers' Handbook, Eighth Edition (Chemical Engineers Handbook). 8th ed. New York: McGraw-Hill Professional, 2007.

Groot, W.J., R.G.J.M Van der Lans, and K.Ch.A.M. Luyben. "Technologies for Butanol

Recovery Integrated with Fermentations." Process Biochemistry 27 (1992): 61-75.

Gottschal, J.C., and J.G. Morris. "The induction of acetone and butanol production in cultures of Clostridium acetobutylicum by elevated concentrations of acetate and butyrate." FEMS Microbiology Letters 12 (1981): 385-89.

Hill, Jason. "Environmental, economic, and energetic costs and benefits of biodiesel and ethanol biofuels." PNAS 103 (2006): 11206-1210.

Huang, Yu Liang, Likun Zhang, Zetang Wu, Chun Ming Cheung, and Shang-Tian Yang. "Production of carboxylic acids from hydrolyzed corn meal by immobilized cell fermentation in a fibrous-bed bioreactor." Bioresource Technology 82 (2002): 51-59.

Junker, B., and M. Lester. "A next generation, pilot-scale continuous sterilization system for fermentation media." Bioprocess Biosystem Engineering 28 (2006): 351-78.

Kister, Henry Z. Distillation operations. New York: McGraw-Hill, 1990.

Kwiatkowski, Jason, and Andrew McAloon. "Modeling the process and costs of fuel ethanol production by the corn dry-grind process." Industrial Crops and Products 23 (2006): 288-96. 1 Feb. 2009 <http://www.elsevier.com/locate/indcrop>. NOTE: Accompanying Superpro Designer process simulator found at http://www.intelligen.com/literature.shtml

Najafpour, Ghasem. Biochemical Engineering and Biotechnology. St. Louis: Elsevier Science, 2006.

"The Onix Corporation - Manufacturers of Rotary Drum Dryers, Rotary Dryers, Drum Dryers & Solid Fuel Burners." The Onix Corporation - Manufacturers of Solid Fuel Burners, Rotary Dryers, Sawdust Burners, Rotary Drum Dryers and Dryer Controls. 03 Apr. 2009 <http://www.theonixcorp.com/rotary.html>.

Papoutsakis, Eleftherios T. "Engineering solventogenic clostridia." Current Opinion in Biotechnology 19 (2008): 420-29. Science Direct. <http://www.sciencedirect.com>.

Park, Chang-Ho, and Qinghuang Geng. "Simultaneous Fermentation and Separation In the Ethanol and ABE Fermentation." Separation and Purification Methods 21 (1992): 127-74.

Perkis, D., W. Tyner, and R. Dale. "Economic analysis of a modified dry grind ethanol process with recycle of pretreated and enzymatically hydrolyzed distillers grains." Bioresource Technology 99 (2008): 5243-249. ScienceDirect. 26 Nov. 2008. 25 Feb. 2009 <http://www.sciencedirect.com/science>

Qureshi, Nasib, Xin-Liang Li, Stephen Hughes, Badal C. Saha, and Michael A. Cotta. "Butanol Production from Corn Fiber Xylan Using Clostridium acetobutylicum." Biotechnology Program 22 (2006): 673-80. EBSCO MegaFile.

Ramey, David, and Shang-Tian Yang. Production of Butyric Acid and Butanol from Biomass. Tech. Morgantown: U.S. Department of Energy, 2004.

Ramey, David E. Continuous Two Stage, Dual Path Anaerobic Fermentation of Butanol and Other Organic Solvents Using Two Different Strains of Bacteria. Environmental Energy, Inc., assignee. Patent 5753474. 1996.

RFA - Renewable Fuels Association. 03 Apr. 2009 <http://www.ethanolrfa.org>.

Schugerl, Karl. "Integrated processing of biotechnology products." Biotechnology Advances 18(2000): 581-99. Elsevier. <http://www.elsevier.com/>.

Seider, Warren D., J. D. Seader, and Daniel R. Lewin. Product and Process Design Principles Synthesis, Analysis, and Evaluation. New York: Wiley, 2003.

Shapouri, Hosein, James A. Duffield, and Michael S. Graboski. Estimating the Net Energy Balance of Corn Ethanol. Rep. no. 721. Washington, DC: U.S. Department of Agriculture, 1995.

Sneller, Todd, and Douglas Durante. "The Impact of Ethanol Production on Food, Feed, and Fuel." Ethanol Across America (2007). Business Source Premier.

Suszkiw, Jan. "Making Microbes Do More Work." Agricultural Research (October 2008): 9-9.

Tashiro, Yukihiro. "High Butanol Production by Clostridium saccharoperbutylacetonicum N1-4 in Fed-Batch Culture with pH-Stat Continuous Butyric Acid and Glucose Feeding Method." Journal of Bioscience and Bioengineering 98 (2004): 263-68.

Thompson, Kalee. "Beyond Ethanol." Popular Science Feb. 2008: 31-33.

"Tracking ethanol profitability - Don Hofstrand January 2008." Iowa State University Extension. 03 Apr. 2009 <http://www.extension.iastate.edu/agdm /articles/hof/HofJan08.html>.

Ulrich, Gael D. Chemical engineering process design and economics a practical guide. Durham, N.H: Process Pub., 2004.

United States. FDCH Regulatory Intelligence Database. DOE Announces Additional Steps in Developing Sustainable Biofuels Industry. 2008. Business Source Premier.

United States. The Clean Fuels Development Coalition, The Nebraska Ethanol Board. U.S.

Department of Agriculture. A Guide for Evaluating the Requirements of Ethanol Plants. 2007. EBSCO MegaFile.

Vijjeswarapu, Winston K. "Performance of a Cell Recycling Continuous Fermentation System for the Production of n-Butanol by Clostridium Acetobutylicum." Biotechnology and Bioengineering Symposium (1986): 471-78.

Wooley, Robert J., and Victoria Putsche. "Development of an ASPENPlus Physical Property Database for Biofuels Components." National Renewable Energy Laboratory (1996): 1-36.

Yang, Shang-Tian, and Ying Zhu. "Adaptation of Clostridium tyrobutyricum for Enhanced Tolerance to Butyric Acid in a Fibrous-Bed Bioreactor." Biotechnology Progress 19 (2003): 365-72.

Yang, Shang-Tian. Extractive Fermentation Using Convoluted Fibrous Bed Bioreactor. The Ohio State University Research Foundation, assignee. Patent 5,563,069. 1996.

Yang, Shang-Tian. Methods of Producing Butanol. Patent US20080248540. 2008.

Section X: Appendix

Appendix A: Example Calculations for Equipment Sizing and Costing

Appendix B: Spreadsheets for Pricing and Sizing of Units

Appendix C: Assembly of Database, Raw Materials, Material and Energy Balances, Block Diagrams for Fermentation

Appendix D: SuperPro Outputs

Appendix E: Aspen Outputs

Appendix F: Correspondences

Appendix G: MSDS

Appendix A

Product conversions

Fermenter 1, converting glucose to acid

Component	g/g glucose
Butyric Acid	0.469
Acetic Acid	0.0333
Lactic Acid	0.100
CO2	0.440
H	0.0106

Fermenter 2, converting acids to solvents

Component	g/g butyric acid
Butanol	0.830
Ethanol	.0183
Acetone	.0367

Fermenter 1 Sample calculations

3 reactors at 110,000 gallons each

Dilution rate = 0.6hr-1

$$Volume\ Fermenter\ 1 = \frac{1}{5} Volume\ Fermenter\ 2 = 110{,}000\ gallons$$

$$Working\ Volume\ Fermenter\ 1 = 0.75 * 110{,}000\ gallons = 88{,}000\ gallons = 333{,}116\ L$$

$$Total\ Flow\ Rate\ of\ Fermetner\ 1 = 333{,}116L * 0.6hr^{-1} = 199{,}870\frac{L}{hr}$$

$$Concentration\ of\ Butyric\ Acid\ in\ Fermenter\ 1\ Product\ Stream$$

$$\frac{\left(5.111\frac{g\ butanol}{L}\right)\left(1{,}405{,}334\ \frac{L}{hr\ Ferm\ 2}\right)}{\left(0.83\ \frac{g\ butanol}{g\ butyric\ acid}\right)\left(199{,}870\frac{L}{hr\ Ferm\ 1}\right)} = 43.30\ \frac{g\ butyric\ acid}{L}$$

$$Glucose\ Concentration\ in\ Ferm\ 1\ Feed = \frac{\left(43.40\ \frac{g\ butyric\ acid}{L}\right)}{\left(0.469\ \frac{g\ butyric\ acid}{g\ glucose}\right)} = 92.40\ \frac{g\ glucose}{L}$$

$Additional\ P2\ medium\ needed$

$$\left(199{,}870\ \frac{L}{hr}\right)\left(92.40\,\frac{g\ glucose}{L}\right)\left(0.130\,\frac{g\ P2\ medium}{g\ glucose}\right)\left(\frac{lb\ P2\ medium}{L}\right) = 5{,}300\ lb/hr$$

Fermenter 2 Sample Calculations

3 reactors at 550,000 gallons each

Dilution rate = 0.9hr-1

Productivity = 4.6g butanol/L-hr

Glucose/Butyric Acid in Fermenter 2 Feed = 15

$Working\ Volume = (0.75)(550{,}000\ gallons) = 412{,}500\ gallons = 1{,}561{,}482\ L$

$Mass\ flow\ rate\ of\ butanol\ out\ of\ Fermenter\ 2$

$$(1{,}561{,}482L)\left(4.6\,\frac{g}{L-hr}\right) = 7{,}182{,}819\,\frac{g\ butanol}{hr}$$

$Total\ Flow\ Rate\ of\ Fermenter\ 2 = (1{,}561{,}482L)(0.9hr^{-1}) = 1{,}405{,}334\,\frac{L}{hr}$

$Concentration\ of\ Butanol\ in\ Fermenter\ 2 = \dfrac{\left(7{,}182{,}819\,\frac{g}{hr}\right)}{\left(1{,}405{,}334\ \frac{L}{hr}\right)} = 5.111\ \dfrac{g\ butanol}{L}$

$Total\ nutrient\ flow\ rate = Total\ Fermetner\ 2\ Flow\ Rate -\ Flow\ rate\ from\ Fermenter\ 1$

$$\left(1{,}405{,}334\,\frac{L}{hr}\right) - \left(199{,}870\,\frac{L}{hr}\right) = 1205464\,\frac{L}{hr}$$

Additional P2 Nutrients needed

$$\left(1{,}205{,}464\,\frac{L}{hr}\right)\left(92.40\,\frac{g\ glucose}{L}\right)\left(0.130\,\frac{g\ P2\ medium}{g\ glucose}\right)\left(\frac{lb\ P2\ medium}{L}\right) = 31{,}965 lb/hr$$

Total Amount of Butanol Produced Per Year

$$\left(7{,}182{,}819\ \frac{g\ butanol}{hr}\right)(3\ Ferm\ 2)(24\ hr)(320\ days)(0.97\ recovery) = 54{,}300{,}000\ gal$$

Glucose Dilution

Water required for dilution

$$\frac{\left(741\,\frac{g\ glucose}{L}\right)\left(1{,}012{,}000\,\frac{lb}{hr}\ leaving\ dry\ grind\right)}{\left(92.40\ \frac{g\ glucose}{L}\right)} = 8{,}118{,}000\ \frac{lb\ water}{hr}$$

Pump Sizing and Pricing Calculations:

Sample Pump Calculations: Pump 401-402

Pump (shell):

$$Q = 67705\ \frac{ft^3}{hr} = 8{,}441{,}143\ \frac{gal}{min}$$

$$\rho = 61.246\,\frac{lb}{ft^3}$$

$$\Delta p = 33\,\frac{lb}{in^2}$$

$$Ws = \frac{Q\Delta p\rho * 144\,\frac{in^2}{ft^2}}{3600\,\frac{sec}{hr} * 550.22\,\frac{hp}{lbf - ft/s}} = 163.90HP$$

$$H = \frac{\Delta p}{\rho} = 78.29ft$$

$$S = QH^{0.5} = 74690.2\ gpm - ft^{0.5}$$

$$C_B = e^{9.2951 - 0.6019\ln(S) + 0.0519(\ln(S))^2} = \$8{,}743.03$$

$$F_T = 1$$

$$F_M = 2\ for\ stainless\ steel$$

$$C_{p,pump} = F_T F_M C_B = \$17{,}486.06$$

$$\eta_{pump} = -0.316 + 0.24015\ln(Q) - 0.0119(\ln(Q))^2 = 0.88$$

Motor:

$$P_B = \frac{Ws}{\eta_{pump}} = 187.29HP$$

$$\eta_{motor} = 0.8 + 0.0319\ln(P_B) - 0.00182\big(\ln(P_B)\big)^2 = 0.92$$

$$P_C = \frac{P_B}{\eta_{motor}} = 204.22HP$$

$$F_T = 1.80$$

$$C_{p,motor} = F_T e^{5.4866+0.13141 lnPc+0.053255(lnPc)^2+0.028628(lnPc)^3-0.0035549(lnPc)^4} = \$17{,}032.27$$

$$C_{p,total} = C_{p,pump} + C_{p,motor} = \$34{,}518.33$$

$$BMFactor = 3.30$$

$$C_{BM} = BMFactor * C_{p,total} = \$113{,}910.49$$

Calculating Utilities

Using PUMP101 as Example:

$$\Delta p = 33\frac{lb}{in^2}$$

$$\rho = 61.246\frac{lb}{ft^3}$$

$$m = 414{,}660\frac{lb}{hr}$$

$$H_{extractor} = 54ft$$

$$H = \frac{\Delta p * 144\frac{in^2}{ft^2}}{\rho} + H_{extractor} = 132.29ft$$

$$E = \frac{mH}{\eta_{pump}} = 31.66HP = 23.61kW$$

$$Plant\ Operations\ \left(\frac{hr}{yr}\right) = 320days * 24hrs * 0.9efficiency = 7128hr/yr$$

$$E_T = E * PlantOperations = 168275.31\ \frac{kW - hr}{yr}$$

$$C_{utilities} = E_T * \$0.04 = \$6{,}731.01$$

Liquid-Liquid Extractor Sizing and Pricing Calculations:

Sample Liquid-Liquid Extractor: Extract 101

Product Stream

$$m_p = 8{,}290{,}000\,\frac{lb}{hr}$$

$$\rho_p = 61.20\,\frac{lb}{ft^3}$$

$$Q_p = 135457.52\ \frac{ft^3}{hr} = 37.63\,\frac{ft^3}{sec}$$

$$V_p = 1.33\,\frac{ft}{sec}$$

$$A_p = \frac{Q_p}{V_p} = 28.29 ft^2$$

Solvent Stream

$$m_s = 6{,}810{,}000\,\frac{lb}{hr}$$

$$\rho_s = 34.90\,\frac{lb}{ft^3}$$

$$Q_s = 195{,}128.94\,\frac{ft^3}{hr} = 54.20\,\frac{ft^3}{sec}$$

$$V_s = 2.0\,\frac{ft}{sec}$$

$$A_s = \frac{Q_p}{V_p} = 27.10 ft^2$$

Total Extractor

$$A = A_p + A_s = 55.39 ft^2$$

$D = (\frac{4A}{\pi})^{0.5}$=8.40ft

TheoreticalStages = 3

η = 0.15

$$ActualStages = \frac{TheoreticalStages}{\eta} = 20 stages$$

$H_{stage} = 2ft$
$H = H_{stage} * ActualStages + 4ft\ top + 10ft\ bottom = 54ft$
$S = HD^{1.5} = 1{,}314.21$
$C_p = 250S^{0.84} = \$104{,}139.91$
$BMFactor = 3.0$

$$C_{BM} = BMFactor * C_p = \$312{,}419.72$$

Distillation Column Sizing and Pricing Calculations:

Example: DIST-101
From Aspen: Theoretical trays = 13

Calculate Tray Efficiencies:

$$\eta = 0.492(\bar{\mu} * \alpha)^{-0.245}$$

$\bar{\mu}$ = viscosity α = relative volatility

	Theo Stage	Butanol Volatility	Water Volatility	Relative Volatility	Viscosity Liquid from Top Stage	O'Connell Efficiency	1/Efficiency	# of Real stages
	CONDENSER	0.63	2.97		0.33			
	2	0.90	4.12	4.57	0.28	0.46	2.16	2.16
	3	1.26	7.34	5.82	0.26	0.45	2.24	4.41
	4	3.40	38.94	11.45	0.22	0.39	2.54	6.95
	5	6.62	74.06	11.19	0.19	0.41	2.46	9.40
FEED	6	6.92	74.47	10.76	0.19	0.41	2.42	11.82
	7	7.53	78.23	10.39	0.19	0.42	2.39	14.21
	8	8.82	85.43	9.68	0.18	0.43	2.33	16.54
	9	10.27	91.70	8.93	0.17	0.44	2.27	18.81
	10	11.03	94.17	8.54	0.17	0.45	2.23	21.05
	11	11.27	94.52	8.39	0.17	0.45	2.22	23.27
	12	11.32	94.21	8.32	0.17	0.45	2.21	25.48
	REBOILER	11.32	93.73		0.17			

Number of real stages determined by taking the inverse efficiency plus the number of real stages from the last theoretical stage.

Determine Column Parameters from ASPEN:

Headspace (ft)	4.00
Real Trays	26.00
Tray Spacing (ft)	2.00
Sump Space (ft)	10.00
Height (ft)	64.00

Height determined from sum of specifications.

Purchase Cost of Vessel:

$C_V = e^{7.0374+0.18255\ln(W)+0.02297(\ln(W)^2)} = \$267{,}021.00$

$W = \pi(D_i + T_s)(L + 0.8D_i)t_s\rho = 153{,}980\ lb$

$t_s = 0.44in$

$\rho = 490\ {}^{lb}/_{ft^3}$
$D_i = 31.0\,ft$
$L = 64\,ft$

Purchase Cost of Platforms and Ladders:

$$C_{PL} = 237.1(D)^{0.63313}(L)^{0.80161} = \$58{,}487.00$$

Base Cost of Trays:

$$C_{BT} = 369e^{0.1739(D)} = \$80{,}959.00$$

Cost of Trays:

$$C_T = F_{NT}F_{TT}F_{TM}C_{BT}N_T = \$2{,}104{,}931.00$$
$F_{NT} = Tray\ factor. Since\ N_T > 20$
$F_{TT} = Tray\ type = Sieve = 1.0$
$F_{TM} = Material\ Type = Carbon\ Steel = 1.0$

Purchase Cost of Column:

$$C_P = F_M C_V + C_{PL} + C_T = \$2{,}430{,}438.00$$
$F_M = Materials\ of\ Construction = Carbon\ Steel = 1.0$

Bare Module Cost of Column:

$$C_{BM} = F_{BM}C_P = \$10{,}110{,}621.00$$
$F_{BM} = Bare\ Module\ Factor = 4.16\ for\ Vertical\ Pressure\ Vessels$

Condenser Calculations DIST-COND-101

Assuming a fixed-head, shell-and-tube, carbon steel shell and 20ft long brass tubes.

Calculate Area of Condenser:

$$A = \frac{Q}{U\Delta T_{LM}} = 780ft^2$$
$$Q = 22{,}725{,}688\frac{Btu}{hr}(from\ ASPEN)$$

Since 2 condensers per column Q is split:

$$Q = 11{,}362{,}844\frac{Btu}{hr}$$
$$U = 200\ \frac{Btu}{hr\ ft^2\ F}\ from\ Table\ 13.5$$
$$\Delta T_{LM} = \frac{\Delta T_1 - \Delta T_2}{\ln(\Delta T_1/\Delta T_2)} = 145.70°F$$

Base Cost of Condenser:

$$C_B = e^{(11.0545 - 0.9228\ln(A) + 0.09861[\ln(A)^2])} = \$10{,}745.00$$

Purchase Cost of Condenser:

$$C_P = F_P F_M F_L C_B = \$51{,}231.00$$
$$F_P = Pressure\ Factor = 1.0$$
$$F_M = Material\ Factor = a + \left(\frac{A}{100}\right)^b = 4.77$$
$$a = 2.70\ b = 0.07\ for\ Stainless\ Steel$$
$$F_L = Tube\ Length\ Correlation = 1.0\ for\ 20ft$$

Bare Module Cost of Condenser:

$$C_{BM} = F_{BM} C_P = \$162{,}403.00$$
$$F_{BM} = Bare\ Module\ Factor = 3.17\ for\ Shell\ and\ Tube\ HX$$

Reboiler Calculations DIST-REBOIL-101

Assuming a kettle vaporizer with a carbon steel shell and tubes.

$$Q = 208{,}614{,}000\frac{Btu}{hr}(from\ ASPEN)$$

Since 3 condensers per column Q is split:

$$Q = 69{,}538{,}000\frac{Btu}{hr}$$
$$Heat\ Flux = 12{,}000\frac{Btu}{hr\ ft}\ Assuming\ this\ to\ avoid\ film\ boiling.$$
$$A = \frac{Q}{Heat\ Flux} = 5{,}795ft^2$$

Base Cost of Reboiler:

$$C_B = e^{(11.967 - 0.8709\ln(A) + 0.09005[\ln(A)^2])} = \$71{,}798.00$$

Purchase Cost of Reboiler:

$$C_P = F_P F_M F_L C_B = \$71{,}798.00$$
$$F_P = Pressure\ Factor = 1.0$$
$$F_M = Material\ Factor = a + \left(\frac{A}{100}\right)^b = 1.00$$
$$a = 0.00\ b = 0.00\ for\ Carbon\ Steel$$
$$F_L = Tube\ Length\ Correlation = 1.0\ for\ 20ft$$

Bare Module Cost of Reboiler:

$$C_{BM} = F_{BM} C_P = \$227{,}600.00$$
$$F_{BM} = Bare\ Module\ Factor = 3.17\ for\ Shell\ and\ Tube\ HX$$

Reflux Accumulator Calculations DIST-RA-101

Purchase Cost of Vessel:

$$C_V = e^{8.717+0.233\ln(W)+0.04333(\ln(W)^2)} = \$16{,}045.00$$

$$W = \pi(D_i + T_s)(L + 0.8D_i)t_s\rho = 3{,}369\ lb$$

$t_s = 0.44in$
$\rho = 490\ {}^{lb}/_{ft^3}$
$D_i = 4.62\ ft$
$L = 9.25\ ft$

Purchase Cost of Platforms and Ladders:

$$C_{PL} = 1580(D)^{0.20294} = \$2{,}156.00$$

Purchase Cost of Column:

$C_P = F_M C_V + C_{PL} = \$18{,}200.00$
$F_M = Materials\ of\ Construction = Carbon\ Steel = 1.0$

Bare Module Cost of Column:

$C_{BM} = F_{BM} C_P = \$55{,}512.00$
$F_{BM} = Bare\ Module\ Factor = 3.05\ for\ Horizontal\ Pressure\ Vessels$

Cost of Pumps and Motors DIST-PUMP-101 and DIST-PUMP-103
See Pump example for calculations

$C_{BM} = F_{BM} C_P = \$63{,}198.00$
$F_{BM} = Bare\ Module\ Factor = 3.30\ for\ Pumps\ and\ Drivers$

$C_{BM} = F_{BM} C_P = \$461{,}084.00$
$F_{BM} = Bare\ Module\ Factor = 3.30\ for\ Pumps\ and\ Drivers$

Total Bare Module Cost of Distillation Tower DIST-101

$$C_{BM\ DIST-101} = C_{BM\ Vessel} + C_{BM\ Cond} + C_{BM\ Reboil} + C_{BM\ Reflux} + C_{Pump} + C_{Pump} = \$11{,}080{,}420.00$$

Calculating Utilities

Using DIST-101 as Example:

Condenser DIST-COND-101 Cooling Water Requirements:

$$CW\ Flow\ Rate = \frac{Q}{\rho\Delta T C_p} = 90{,}940\ Gal/hr$$

$\rho = Density\ of\ Cooling\ Water = 8.33\ lb/Gal$

$C_p = Specific\ Heat\ of\ Water = 1.00\ {}^{Btu}/_{lb°F}$
$\Delta T = 120°F - 90°F = 30°F\ for\ Cooling\ Water$

$$CW\ Cost = \frac{\$0.05}{1{,}000\ Gal} Flow\ Rate = \$4.55/hr$$

Reboiler DIST-REBOIL-101 Steam Requirements:

$$Steam\ Flow\ Rate = \frac{Q}{\Delta H_{Vap}} = 295{,}340\ lb/hr$$

$$\Delta H_{Vap} = 706.35 \frac{Btu}{hr}\ at\ 717.58\ psig\ using\ Steam\ Tables$$

$$High\ Pressure\ Steam\ Cost = \frac{\$5.50}{1{,}000\ lb} Flow\ Rate = \$1{,}625/hr$$

Heat Exchanger Sizing and Pricing Calculations:

See sample calculations for condenser DIST-COND-101.

Holding Tank Sizing and Pricing Quote from Natgun Corporation:

Hi Amira,

First of all, concrete tanks are only used for liquids with a pH of 6.5 or greater. Acidic liquids will slowly strip the lime content of the concrete. For acidic liquids, welded steel tanks with steel floors are the best choice. The pricing for steel tanks is not that different from ours, so you can estimate the pricing of a steel tank as roughly the same as my pricing. If you need steel pricing try Dan Knight at Chicago Bridge and Iron (CB&I). If you are doing a life cycle cost as well, steel tanks need to be repainted every 15 years at a cost of $8 per square foot on the interior and exterior.

Natgun is a company that will construct the tank. Wall and dome panels are poured onsite and lifted into place with a crane, and the tank is put into compression by being wrapped with steel wires.

Construction of the tank takes about 3 months, you can construct multiple tanks at the same time if you have sufficient room. Each tank needs a 15 ft construction perimeter and an addition 100 square feet for temporary construction space for the wall and dome panels and a crane.

These are the prices (2009) for the tank construction:

520,000 gal tanks = $500,000 per tank

93,600 gal tanks = $275,000 per tank

265,000 gal tank = $380,000 per tank

1.25M gal tank = $800,000 per tank

As you see there is an economy of scale, on a per gallon basis. These prices are assuming that the tanks have a dome, if you do not need a dome (open-top) deduct 20%.

I would also suggest to add an additional $50,000 per tank for appurtenances(ladders, manways etc.), and an additional $100,000-$200,000 per tank for the earthwork.

Good luck on you project, I will attach some other info that may be helpful.

Thanks,

Jason North

Regional Manager
Natgun Corporation
10 Hearthstone Ct., Suite 2
Reading, PA 19606
Office: 610-370-2790
Fax: 610-370-2792
Mobile: 410-340-5217

Appendix B

Relevant Tables

Pumps and Motors within Distillation Columns				
	Reflux Pump and Motor		Reboiler Pump and Motor	
	DIST-PUMP 101 and 102	DIST-PUMP 201	DIST-PUMP 103 and 104	DIST-PUMP 202
Total Vol. Flow into Pump (gal/min):	232.06	68.85	12,186.37	134.38
Distillate Density (lb/ft3)	44.67	49.73	34.85	43.52
Pump Head (ft):	322.33	289.59	413.16	330.88
Size Factor (gpm-ft$^{1/2}$):	4,166.27	1,171.66	1,986,786.09	19,606.04
FM:	1.35	1.35	1.35	1.35
C_P of Pump:	$ 3,582.86	$ 2,789.13	$ 130,804.91	$ 6,100.60
Pump Efficiency:	0.64	0.49	0.88	0.57
Motor Efficiency:	0.85	0.87	0.74	0.86
Power Consumption:	186.60	70.72	8,176.78	118.52
C_P of Motor:	$ 15,568.10	$ 5,663.16	$ 8,917.51	$ 9,749.32
Total Purchase Cost:	$ 19,150.97	$ 8,452.29	$ 139,722.42	$15,849.92
C_{BM}:	$ 63,198.20	$ 27,892.54	$ 461,084.00	$52,304.73

Reflux Accumulators Sizing and Costing		
Distillation Tower	DIST-101 and DIST-102	DIST-201
Unit ID:	DIST-RA 101 and 102	DIST-RA 201
Number Required:	2	1
V Flow (ft3/hr):	560.43	165.91
Reflux Ratio:	2.32	2.33
Total Vol. Flow into Vessel (ft3/hr):	1,861.29	552.24
Residence Time (hr):	0.08	0.08
Vessel Vol. (ft3):	155.11	46.02
Diameter (ft):	4.62	3.08
Length (ft):	9.25	6.17
Thickness (in):	0.44	0.44
Density (lb/ft3):	490.00	490.00
Weight (lb):	3,369.46	1,504.72
C_V:	$ 16,045.09	$ 11,290.38
C_{PL}:	$ 2,155.73	$ 1,985.63
C_P:	$ 18,200.82	$ 13,276.01
C_{BM}:	$ 55,512.49	$ 40,491.83

Distilation Tower Sizing and Cost

	DIST-101 and DIST-102	DIST-201
Theoretical Trays:	13.00	15.00
Actual Trays:	26.00	22.00
Diameter (ft):	31.00	3.00
Pressure (psi):	30.00	20.00
Stress (psi):	15,000.00	15,000.00
Weld Efficiency:	0.85	0.85
Height (ft):	64.00	45.50
Density of Material (lb/ft^3):	490.00	490.00
Material Factor:	1.00	1.00
Thickness (in.):	0.31	0.31
Thickness Used (in.):	0.44	0.44
Weight (lbs):	153,980.23	8,266.46
F_{NT}:	1.00	1.00
F_{TT}:	1.00	1.00
F_{TM}:	1.00	1.00
C_{BT}:	$ 80,958.88	$ 627.15
C_T:	$ 2,104,930.87	$ 14,424.54
C_V:	$ 267,020.66	$ 38,285.51
C_{PL}:	$ 58,486.29	$ 10,248.24
C_P:	$ 2,430,437.81	$ 62,958.29
C_{BM}:	$ 10,110,621.31	$ 261,906.48

Condenser Areas and Costing for Each Tower

	Condenser (2 Per Tower)			\$/Unit	\$/Unit
Distillation Tower	Q (Btu/hr)	ΔT_{LM} (°F)	Area (ft^2)	C_P	C_{BM}
DIST-101 and DIST-102	11,362,844.00	145.70	779.86	\$ 51,231.10	\$ 162,402.59
DIST-201	2,103,671.65	80.95	259.89	\$ 37,573.48	\$ 119,107.92

**Q is representative of a 2 split stream

Reboiler Areas and Costing for Each Tower

	Reboiler (3 Per Tower)			\$/Unit	\$/Unit
Distillation Tower	Q (Btu/hr)	Heat Flux (Btu/hr-ft^2)	Area (ft^2)	C_P	C_{BM}
DIST-101 and DIST-102	69,538,000.00	12,000.00	5,794.83	\$ 71,798.16	\$ 227,600.17
DIST-201	1,649,071.32	12,000.00	412.27	\$ 21,763.45	\$ 68,990.13

**Q is representative of a 3 split stream

Pumps and Motors into Fermentation Process

	PUMP 101-102	PUMP 103	PUMP 104	PUMP 105	PUMP 106	PUMP 107	PUMP 201-203	PUMP 301	PUMP 302
Number Required:	2	1	1	1	1	1	3	1	
Total Vol. Flow into Pump (gal/min):	9,261.26	18,522.64	2,640.00	15,922.41	91.64	429.63	879.96	18,562.41	258.95
Distillate Density (lb/ft^3)	59.62	55.56	59.62	59.62	55.56	55.56	59.62	55.56	55.56
Pump Head (ft):	115.94	47.12	43.91	43.91	46.65	46.65	140.09	176.24	46.65
Size Factor (gpm-ft$^{1/2}$):	99,720.19	127,145.19	17,494.17	105,511.14	625.90	2,934.49	10,415.23	246,427.38	1,768.70
F_M:	2.00	2.00	2.00	2.00	2.00	2.00	2.00	2.00	2.00
C_P of Pump:	\$ 20,664.42	\$ 23,939.19	\$ 8,617.36	\$ 21,371.20	\$ 3,883.29	\$ 4,871.07	\$ 7,056.70	\$ 36,864.32	\$ 4,400.67
Pump Efficiency:	0.88	0.89	0.83	0.89	0.52	0.70	0.76	0.89	0.65
Motor Efficiency:	0.92	0.92	0.89	0.92	0.82	0.85	0.89	0.93	0.84
Power Consumption:	320.30	241.08	37.82	207.88	2.24	7.56	43.81	891.05	4.98
C_P of Motor:	\$ 25,954.90	\$20,009.85	\$ 3,004.74	\$ 17,333.44	\$507.06	\$841.75	\$ 3,471.89	\$ 50,435.73	\$676.84
Total Purchase Cost:	\$ 46,619.32	\$ 43,949.04	\$ 11,622.10	\$ 38,704.63	\$ 4,390.35	\$ 5,712.82	\$ 10,528.59	\$ 87,300.05	\$ 5,077.51
CBM:	\$ 153,843.76	\$ 145,031.82	\$ 38,352.94	\$ 127,725.29	\$ 14,488.15	\$ 18,852.31	\$ 34,744.33	\$288,090.15	\$ 16,755.77

Agitators	A-1	A-2
Volume of Tank (gal):	264,000.00	1,237,500.00
Horse Power (hP):	132.00	618.75
Power (kW):	98.43	461.40
CP:	\$ 46,086.32	\$ 111,175.36

Pumps and Motors into Separations Process				
	PUMP 401-402	PUMP 403-404	PUMP 405-406	PUMP 501
Number Required:	2	2	2	1
Total Vol. Flow into Pump (gal/min):	8,441.14	9,419.22	44.67	24,371.53
Distillate Density (lb/ft3)	61.25	45.42	45.42	34.89
Pump Head (ft):	78.29	88.77	57.70	109.79
Size Factor (gpm-ft$^{1/2}$):	74,690.29	88,745.32	339.35	255,360.91
FM:	2.00	1.35	2.00	1.35
C_P of Pump:	$ 17,486.06	$ 13,026.05	$ 3,801.63	$ 25,501.14
Pump Efficiency:	0.88	0.88	0.42	0.89
Motor Efficiency:	0.92	0.90	0.91	0.90
Power Consumption:	204.22	77.32	158.35	76.56
C_P of Motor:	$ 17,032.27	$ 6,217.93	$ 13,177.12	$ 6,153.91
Total Purchase Cost:	$ 34,518.33	$ 19,243.98	$ 16,978.75	$ 31,655.05
CBM:	$ 113,910.49	$ 63,505.13	$ 56,029.87	$ 104,461.65

Heat Exchangers between Fermentation			
Heat Exchangers	HX-201 thru HX-203	HX-204 thru HX-206	HX-207 thru HX-209
Number Required:	3	3	3
Mass Flow (lb/hr):	357,569.57	357,569.57	357,569.57
ΔT_{LM} (°C):	141.20	3.91	13.15
Q (Btu/hr):	108,407,055.89	7,561,596.88	7,762,443.32
Area (ft^2):	10,375.07	2,580.35	3,906.88
C_B:	$ 57,145.73	$ 19,745.27	$ 26,043.36
F_M:	3.98	3.98	3.95
C_P:	$ 233,380.10	$ 78,102.44	$ 103,977.94
C_{BM}:	$ 739,814.93	$ 247,584.75	$ 329,610.08

Tank Volumes and Pricing: Price = 0.4442*(Gallons) + 252377 + Extra Charges						
	Acutal L	Actual gal	Working L	Working gal	Extra Charges	$/Gal Tank
FERM1	416,395.30	110,000.00	333,116.24	88,000.00	200,000.00	501,239.0000
FERM2	2,081,976.48	550,000.00	1,561,482.36	412,500.00	200,000.00	696,687.0000
BREED1	999,348.71	310,588.24	999,348.71	264,000.00	200,000.00	590,340.2941
BREED2	4,684,447.09	1,455,882.35	4,684,447.09	1,237,500.00	200,000.00	1,099,079.9412
DILUT-1	7,495,115.34	2,329,411.76	7,495,115.34	1,980,000.00	200,000.00	1,487,101.7059
HOLD-1	11,363,636.36	3,000,000.00	9,659,090.91	2,550,000.00	200,000.00	1,784,977.0000

Heat Exchangers for Glucose Stream

Heat Exchangers	HX-101	HX-102	HX-103
Number Required:	1	1	1
Mass Flow (lb/hr):	1,042,000.00	1,042,000.00	1,042,000.00
ΔT_{LM} (°C):	99.80	17.43	17.13
Q (Btu/hr):	189,700,285	18,720,960	101,527,895
Area (ft^2):	25,687	10,742	29,638
C_B:	$ 140,245.09	$ 58,968.88	$ 164,009.38
F_M:	4.17	4.09	4.19
C_P:	$ 585,469.60	$ 241,024.39	$ 687,111.32
C_{BM}:	$ 1,855,938.62	$ 764,047.30	$ 2,178,142.90

Heat Exchanger Specification Sheet

HX 301-324

Size	44 /	360 in	Type	BEM	Hor	Connected in	8	parallel 3	series
Surf/unit(eff.)	251692.7 ft2		Shells/unit	24		Surf/shell (eff.)	10487.2		ft2

PERFORMANCE OF ONE UNIT

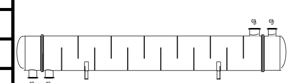
Fluid allocation		Shell Side		Tube Side	
Fluid name					
Fluid quantity, Total	lb/h	6809887		6859885	
Vapor (In/Out)	lb/h	0	0	0	0
Liquid	lb/h	6809887	6809887	6859885	6859885
Noncondensable	lb/h	0		0	
Temperature (In/Out)	F	483.6	220.26	123	410
Dew / Bubble point	F				
Density (Vap / Liq)	lb/ft3	/ 34.86	/ 42.94	/ 45.43	/ 37.37
Viscosity	cp	/ 0.17	/ 0.487	/ 0.918	/ 0.212
Molecular wt, Vap					
Molecular wt, NC					
Specific heat	BTU/(lb*F)	/ 0.788	/ 0.588	/ 0.521	/ 0.73
Thermal conductivity	BTU/(ft*h*F)	/ 0.048	/ 0.068	/ 0.075	/ 0.053
Latent heat	BTU/lb				
Pressure	psi	31.4	16.479	30	28.665
Velocity	ft/s	4.24		1.53	
Pressure drop, allow./calc.	psi	15	14.921	15	1.335
Fouling resist. (min)	ft2*h*F/BTU	0.002		0.0023	

Heat exchanged	122785700 BTU/h			MTD corrected	83.26	F	
Transfer rate, Service	58.59	Dirty	59.75	Clean	80.42	BTU/(h*ft2*F)	

CONSTRUCTION OF ONE SHELL

			Shell Side	Tube Side
Design/Test pressure		psi	70 / Code	70 / Code
Design temperature		F	670	590
Number passes per shell			1	1
Corrosion allowance		in	0.0625	0
Connections	In	in	16 / -	14 / -
Size/rating	Out		12 / -	12 / -
Nominal	Intermediate		12 / -	12 / -

Sketch



Tube No.	1801	OD 0.75	Tks- Avg 0.049	in	Length 30	ft	Pitch 0.9375	in
Tube type	Plain		Material	SS 304			Tube pattern	30
Shell	Carbon Steel	ID 44.4375	OD 45.3125	in	Shell cover	-		
Channel or bonnet		SS 304			Channel cover	-		
Tubesheet-stationary		SS 304			Tubesheet-floating	-		
Floating head cover		-			Impingement protection		None	
Baffle-crossing	Carbon Steel	Type	Single segment	Cut(%d) 35.38 H	Spacing: c/c		27	in
Baffle-long	-		Seal type				Inlet 29.4375	in
Supports-tube			U-bend		Type			
Bypass seal			Tube-tubesheet joint		Exp.			
Expansion joint	-		Type					
RhoV2-Inlet nozzle	934	Bundle entrance	777		Bundle exit		991	lb/(ft*s
Gaskets - Shell side	Compressed Fiber 1/16		Tube Side		Compressed Fiber 1/16			
Floating head	-							
Code requirements	ASME Code Sec VIII Div 1			TEMA class			B - chemical service	
Weight/Shell	32314.6		Filled with water	52607.9			Bundle 23136.2	lb
Remarks								

Heat Exchanger Specification Sheet

COOLER
HX 104-115

Size	58 /	288 in	Type	BEM	Hor	Connected in		4	parallel	3	series
Surf/unit(eff.)	159840.3	ft2	Shells/unit	12		Surf/shell (eff.)			13320		ft2

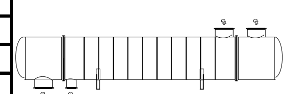
PERFORMANCE OF ONE UNIT

Fluid allocation		Shell Side		Tube Side	
Fluid name					
Fluid quantity, Total	lb/h	8118000		34599420	
Vapor (In/Out)	lb/h	73696	0	0	0
Liquid	lb/h	8118000	8118000	34599420	34599420
Noncondensable	lb/h	0		0	
Temperature (In/Out)	F	111	98.6	80	112.8
Dew / Bubble point	F	0	240.66		
Density (Vap / Liq)	lb/ft3	0.06 / 56.26	/ 61.32	/ 61.95	/ 60.84
Viscosity	cp	0.0132 / 0.238	/ 0.711	/ 0.88	/ 0.613
Molecular wt, Vap		18.02			
Molecular wt, NC					
Specific heat	BTU/(lb*F)	0.4528 / 1.115	/ 0.941	/ 0.924	/ 0.954
Thermal conductivity	BTU/(ft*h*F)	0.015 / 0.394	/ 0.36	/ 0.352	/ 0.365
Latent heat	BTU/lb	960			
Pressure	psi	25.236	10.612	65	55.376
Velocity	ft/s	28.87		5.93	
Pressure drop, allow./calc.	psi	15	14.624	15	9.624
Fouling resist. (min)	ft2*h*F/BTU	0.003		0.0035	

Heat exchanged	106521400	BTU/h			MTD corrected	59.39	F
Transfer rate, Service	112.21	Dirty	116.97		Clean	476.58	BTU/(h*ft2*F)

CONSTRUCTION OF ONE SHELL / **Sketch**

			Shell Side	Tube Side
Design/Test pressure		psi	60 / Code	100 / Code
Design temperature		F	420	420
Number passes per shell			1	1
Corrosion allowance		in	0	0
Connections	In	in	30 / -	30 / -
Size/rating	Out		16 / -	30 / -
Nominal	Intermediate		16 / -	30 / -



Tube No.	2874	OD	0.75	Tks- Avg	0.049	in	Length 24 ft	Pitch 0.9375	in
Tube type	Plain			Material	SS 304			Tube pattern	30
Shell	SS 304	ID	57.625	OD	58.125	in	Shell cover	-	
Channel or bonnet		SS 304					Channel cover	-	
Tubesheet-stationary		SS 304					Tubesheet-floating	-	
Floating head cover		-					Impingement protection	None	
Baffle-crossing	SS 304	Type	Single segment	Cut(%d)	40.14 V		Spacing: c/c	24.25	in
Baffle-long	-			Seal type				Inlet 33.0394	in
Supports-tube			U-bend				Type		
Bypass seal				Tube-tubesheet joint			Exp.		
Expansion joint	-			Type					
RhoV2-Inlet nozzle	2036	Bundle entrance		4447			Bundle exit	962	lb/(ft*s
Gaskets - Shell side	Compressed Fiber 1/16			Tube Side			Compressed Fiber 1/16		
Floating head	-								
Code requirements	ASME Code Sec VIII Div 1						TEMA class	B - chemical service	
Weight/Shell	40125.2		Filled with water		74761.2			Bundle 30876.5	lb
Remarks									

Appendix C:

Assembly of Database

Thermophysical	Ethanol	Butanol	Acetone	Lactic Acid	Acetic Acid	Water	Sulfuric Acid	Glucose
Free Energy (BTU/lb)	-1698.271	-1012.526	-1193.896	-2571.487	-2799.523	-5748.3	-3084.504	-2276.457
CP (BTU/lb R)	0.6178107	0.6344135	0.4848571	0.4862745	0.2373825	0.8888829	0.196928	0.4767455
Heat Vap (BTU/lb)	409.3473	321.9961	239.1389	406.7995	167.5739	1065.3	254.9304	411.1883
Enthalpy (BTU/lb)	-2615.068	-1928.17	-1847.732	-3279.338	-3275.878	-6853.814	-3485.664	-3020.998
Fugacity Coefficient	0.0620151	0.1518379	0.4963896	0.0200911	4.74E-16	2.39E-07	0.00522821	4.65E-09
Vap Pressure (psi)	0.9113714	2.231403	7.294917	0.2952588	6.96E-15	3.51E-06	0.0768334	6.83E-08
Density (lb/ft^3)	50.26743	51.40768	50.65375	77.48511	66.43549	62.54803	115.4577	74.26934
Entropy (BTU/lb R)	-1.845887	-1.843568	-1.31644	-1.425193	-0.9590982	-2.225852	-0.8076995	-1.499067
Internal Energy (BTU/lb)	-2615.122	-1928.223	-1847.786	-3279.373	-3275.919	-6853.857	-3485.688	-3021.035

Transport	Water	Ethanol	Acetone	N-But	Glucose	Lactic Acid	Acetic Acid	Sulfuric Acid
Thermal Conductivity (BTU-ft/hr-ft^2-R)	0.3595226	0.0952892	0.0897687	0.0869662	0.11837	0.1422357	0.0944077	0.1798242
Viscosity (cP)	0.7111731	0.8633581	0.2750804	1.890545	1.15E+23	131.6118	1.587255	52.31043
Surface Tension (dyne/cm)	70.43156	21.09028	21.56663	23.34319	21.73511	44.06618	29.24957	53.0978

Price	Water	Ethanol	Acetone	N-But	Corn	Lactic Acid	Acetic Acid	Sulfuric Acid
per 1000 gal	3	2500	15118	4000			34950	5189.3
per 1000 lbs					71.42857143	500		

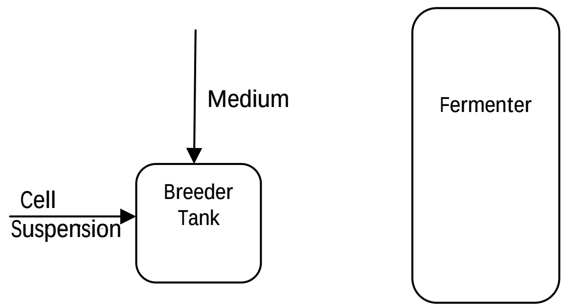
Energy (BTU)	Butanol	Ethanol	Gasoline
per gallon	110000	84000	115000

Kinetic Reactions			
1	C6H12O6 glucose	<-->	2 C3H6O3 lactic acid
2	C6H12O6 glucose	<-->	3 C2H4O2 acetic acid
3	(2/3) C6H12O6 glucose	<-->	C4H8O2 + O2 butyric acid
4	C3H6O3 lactic cid	<-->	C3H6O + O2 acetone
5	C2H4O2 + H2O acetic acid	<-->	C2H6O + O2 ethanol
6	C4H8O2 + H2O butyric acid	<-->	C4H10O + O2 butanol

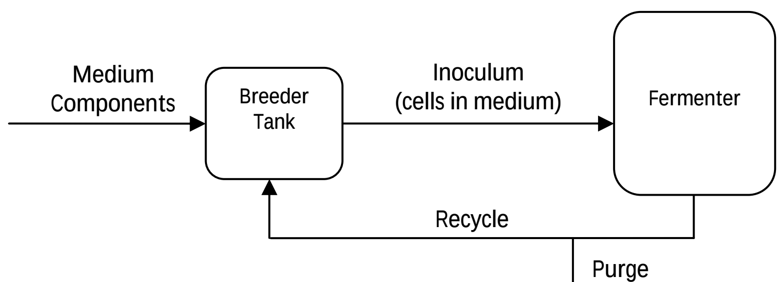
Material Balances for Fermentation Process								
Note: Actual F2, F3, F4, and F5 are three times value given below								
				MASS FLOWS				Mass Frac
m	F1	524417.77	L/hr	499335390.96	g/hr	1100847.00	lb/hr	
	xg,1	741.00	g/L	388593566.39	g/hr	856702.87	lb/hr	0.6500
	xw,1	131.28	g/L	68845564.64	g/hr	151778.61	lb/hr	0.2800
	xs,1	79.89	g/L	41896259.94	g/hr	92365.52	lb/hr	0.0700
dil	F2	199869.74	L/hr	198678092.69	g/hr	438010.57	lb/hr	
	FW	3682529.97	L/hr	3682529972.07	g/hr	8118595.50	lb/hr	
	F3	1205464.38	L/hr	1198277246.56	g/hr	2641751.28	lb/hr	
	xg,2	92.37	g/L	18461863.76	g/hr	40701.48	lb/hr	0.0929
	xw,2	891.71	g/L	178225760.97	g/hr	392920.86	lb/hr	0.8971
	xs,2	9.96	g/L	1990467.96	g/hr	4388.23	lb/hr	0.0100
	xw,w	1000.00	g/L	3682529972.07	g/hr	8118595.50	lb/hr	
	xg,3	92.37	g/L	111348115.81	g/hr	245480.78	lb/hr	0.0929
	xw,3	891.71	g/L	1074924120.86	g/hr	2369803.97	lb/hr	0.8971
	xs,3	9.96	g/L	12005009.88	g/hr	26466.54	lb/hr	0.0100
	Xn2							
FERM1	F4	199869.74	L/hr	190905481.96	g/hr	420874.89	lb/hr	
	xba,4	43.30	g/L	8653998.64	g/hr	19078.82	lb/hr	0.0453
	xaa,4	10.42	g/L	2082513.04	g/hr	4591.16	lb/hr	0.0109
	xla,4	3.47	g/L	694171.01	g/hr	1530.39	lb/hr	0.0036
	xw,4	888.00	g/L	177484331.31	g/hr	391286.29	lb/hr	0.9297
	xs,4	9.96	g/L	1990467.96	g/hr	4388.23	lb/hr	0.0104
FERM 2	F5	1405334.13	L/hr	1274062718.81	g/hr	2808829.78	lb/hr	
	xb,5	5.11	g/L	7182818.87	g/hr	15835.42	lb/hr	0.0056
	xe,5	0.11	g/L	158656.64	g/hr	349.78	lb/hr	0.0001
	xa,5	0.23	g/L	317313.28	g/hr	699.56	lb/hr	0.0002
	xw,5	891.18	g/L	1252408452.17	g/hr	2761090.26	lb/hr	0.9830
	xs,5	9.96	g/L	13995477.84	g/hr	30854.77	lb/hr	0.0110

Breeder Tanks

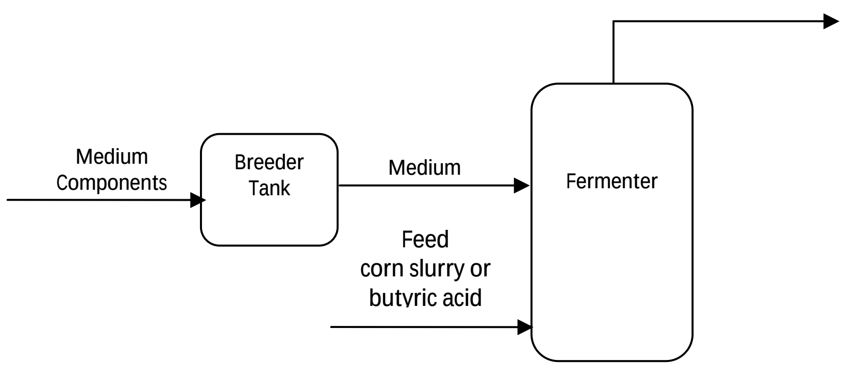
Day 0-2: Inoculate cells in growth medium



Day 2-10: Immobilize and grow cells on FBB while purging fully utilized nutrients



Day 10-330: Continuous fermentation

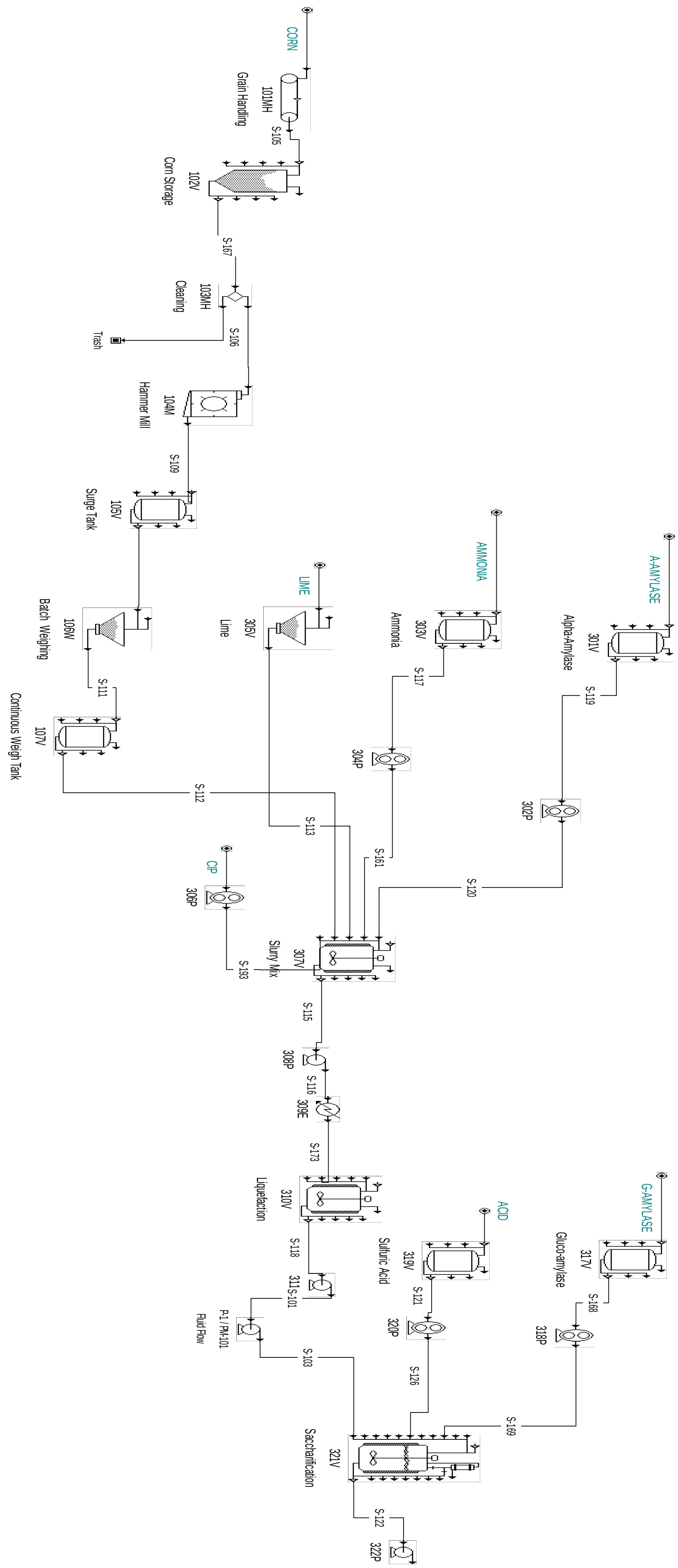


Day 330-365: Autoclave/clean Fermenter

Appendix D:

Dry Grind: Corn to

Glucose



Materials & Streams Report
for corn to glucose 4-13

April 13, 2009

1. OVERALL PROCESS DATA

Annual Operating Time	7,920.00h
Annual Throughput	0.00kg MP
Operating Days per Year	330.00

MP = Main Product = Undefined

2.1 STARTING MATERIAL REQUIREMENTS (per Section)

Section	Starting Material	Active Product	Amount Needed (kg	Molar Yield (%)	Mass Yield (%)	Gross Mass Yield (%)
Main Section	(none)	(none)	Unknow	Unknow	Unknow	Unknow
Grain Handling & Milling	(none)	(none)	Unknow	Unknow	Unknow	Unknow
Starch to Sugar	(none)	(none)	Unknow	Unknow	Unknow	Unknow
Fermentation	(none)	(none)	Unknow	Unknow	Unknow	Unknow
Ethanol Processing	(none)	(none)	Unknow	Unknow	Unknow	Unknow
CoProduct Processing	(none)	(none)	Unknow	Unknow	Unknow	Unknow
Common Suport	(none)	(none)	Unknow	Unknow	Unknow	Unknow

Sin = Section Starting Material, Aout = Section Active Product

2.2 BULK MATERIALS (Entire Process)

Material	kg/yr	kg/h	kg/kg MP
Corn	5,304,937,759	669,815.374	
Lime	718,270	90.691	
Liq. Ammonia	1,200,956	151.636	
Alpha-Amylase	298,452	37.683	
Glucoamylase	503,297	63.548	
Sulfuric Acid	993,746	125.473	
Caustic	17,700,150	2,234.867	
TOTAL	5,326,352,630	672,519.271	

2.3 BULK MATERIALS (per Section)

SECTIONS IN: Main Branch

Grain Handling & Milling

Material	kg/yr	kg/h	kg/kg MP
Corn	5,304,937,759	669,815.374	
TOTAL	5,304,937,759	669,815.374	

Starch to Sugar Conversion		
Material	**kg/yr**	**kg/h**
Lime	718,270	90.691
Liq. Ammonia	1,200,956	151.636
Alpha-Amylase	298,452	37.683
Glucoamylase	503,297	63.548
Sulfuric Acid	993,746	125.473
Caustic	17,700,150	2,234.867
TOTAL	21,414,871	2,703.898

2.4 BULK MATERIALS (per Material)

Corn			
Corn	**% Total**	**kg/yr**	**kg/h**
Grain Handling & Milling (Main Branch)			
101MH	100.00	5,304,937,759	669,815.374
TOTAL	100.00	5,304,937,759	669,815.374

Lime			
Lime	**% Total**	**kg/yr**	**kg/h**
Starch to Sugar Conversion (Main Branch)			
305V	100.00	718,270	90.691
TOTAL	100.00	718,270	90.691

Liq. Ammonia			
Liq. Ammonia	**% Total**	**kg/yr**	**kg/h**
Starch to Sugar Conversion (Main Branch)			
303V	100.00	1,200,956	151.636
TOTAL	100.00	1,200,956	151.636

Alpha-Amylase			
Alpha-Amylase	**% Total**	**kg/yr**	**kg/h**
Starch to Sugar Conversion (Main Branch)			
301V	100.00	298,452	37.683
TOTAL	100.00	298,452	37.683

Glucoamylase			
Glucoamylase	**% Total**	**kg/yr**	**kg/h**
Starch to Sugar Conversion (Main Branch)			
317V	100.00	503,297	63.548
TOTAL	100.00	503,297	63.548

Sulfuric Acid			
Sulfuric Acid	**% Total**	**kg/yr**	**kg/h**
Starch to Sugar Conversion (Main Branch)			
319V	100.00	993,746	125.473
TOTAL	100.00	993,746	125.473

Caustic			
Caustic	**% Total**	**kg/yr**	**kg/h**
Starch to Sugar Conversion (Main Branch)			
306P	100.00	17,700,150	2,234.867
TOTAL	100.00	17,700,150	2,234.867

2.5 BULK MATERIALS: SECTION TOTALS (kg/h)

Raw Material	Main Section	Grain Handling & Milling	Starch to Sugar Conversion
Corn	0.000	669,815.374	0.000
Lime	0.000	0.000	90.691
Liq. Ammonia	0.000	0.000	151.636
Alpha-Amylase	0.000	0.000	37.683
Glucoamylase	0.000	0.000	63.548
Sulfuric Acid	0.000	0.000	125.473
Caustic	0.000	0.000	2,234.867
TOTAL	0.000	669,815.374	2,703.898

Raw Material	Ethanol Processing	CoProduct Processing	Common Suport Systems
Corn	0.000	0.000	0.000
Lime	0.000	0.000	0.000
Liq. Ammonia	0.000	0.000	0.000
Alpha-Amylase	0.000	0.000	0.000
Glucoamylase	0.000	0.000	0.000
Sulfuric Acid	0.000	0.000	0.000
Caustic	0.000	0.000	0.000
TOTAL	0.000	0.000	0.000

2.6 BULK MATERIALS: SECTION TOTALS (kg/yr)

Raw Material	Main Section	Grain Handling & Milling	Starch to Sugar Conversion
Corn	0	5,304,937,759	0
Lime	0	0	718,270
Liq. Ammonia	0	0	1,200,956
Alpha-Amylase	0	0	298,452
Glucoamylase	0	0	503,297
Sulfuric Acid	0	0	993,746
Caustic	0	0	17,700,150
TOTAL	0	5,304,937,759	21,414,871

Raw Material	Ethanol Processing	CoProduct Processing	Common Suport Systems
Corn	0	0	0
Lime	0	0	0
Liq. Ammonia	0	0	0
Alpha-Amylase	0	0	0
Glucoamylase	0	0	0
Sulfuric Acid	0	0	0
Caustic	0	0	0
TOTAL	0	0	0

3. STREAM DETAILS

Stream Name	**ACID**	**S-121**	**S-126**	**G-AMYLASE**
Source	**INPUT**	**319V**	**320P**	**INPUT**
Destination	**319V**	**320P**	**321V**	**317V**
Stream Properties				
Activity (U/ml)	0.00	0.00	0.00	0.00
Temperature (°C)	21.00	21.00	21.27	21.00
Pressure (bar)	1.01	1.01	7.91	1.01
Density (g/L)	1,832.36	1,832.36	1,832.11	996.16
Component Flowrates (kg/h averaged)				
Sulfuric Acid	125.473	125.473	125.473	0.000
Water	0.000	0.000	0.000	63.548
TOTAL (kg/h)	125.473	125.473	125.473	63.548
TOTAL (L/h)	68.476	68.476	68.486	63.792

Stream Name	**S-168**	**S-169**	**CIP**	**S-193**
Source	**317V**	**318P**	**INPUT**	**306P**
Destination	**318P**	**321V**	**306P**	**307V**
Stream Properties				
Activity (U/ml)	0.00	0.00	0.00	0.00
Temperature (°C)	21.00	21.00	82.20	82.22
Pressure (bar)	1.01	1.01	1.01	3.29
Density (g/L)	996.16	996.16	980.70	980.70
Component Flowrates (kg/h averaged)				
Other Solids	0.000	0.000	111.743	111.743
Water	63.548	63.548	2,123.124	2,123.124
TOTAL (kg/h)	63.548	63.548	2,234.867	2,234.867
TOTAL (L/h)	63.792	63.792	2,278.838	2,278.859

Stream Name	**LIME**	**S-113**	**AMMONIA**	**S-117**
Source	**INPUT**	**305V**	**INPUT**	**303V**
Destination	**305V**	**307V**	**303V**	**304P**
Stream Properties				
Activity (U/ml)	0.00	0.00	0.00	0.00
Temperature (°C)	25.00	25.00	25.00	25.00
Pressure (bar)	1.01	1.01	1.01	1.01
Density (g/L)	1,173.66	1,173.66	1,173.66	1,173.66
Component Flowrates (kg/h averaged)				
Other Solids	90.691	90.691	151.636	151.636
TOTAL (kg/h)	90.691	90.691	151.636	151.636
TOTAL (L/h)	77.272	77.272	129.199	129.199

Stream Name	S-161	A-AMYLASE	S-119	S-120
Source	**304P**	**INPUT**	**301V**	**302P**
Destination	**307V**	**301V**	**302P**	**307V**
Stream Properties				
Activity (U/ml)	0.00	0.00	0.00	0.00
Temperature (°C)	25.13	25.00	25.00	25.08
Pressure (bar)	7.91	1.01	1.01	4.46
Density (g/L)	1,173.57	994.70	994.70	994.67
Component Flowrates (kg/h averaged)				
Other Solids	151.636	0.000	0.000	0.000
Water	0.000	37.683	37.683	37.683
TOTAL (kg/h)	151.636	37.683	37.683	37.683
TOTAL (L/h)	129.209	37.884	37.884	37.885

Stream Name	CORN	S-105	S-167	S-106
Source	**INPUT**	**101MH**	**102V**	**103MH**
Destination	**101MH**	**102V**	**103MH**	**104M**
Stream Properties				
Activity (U/ml)	0.00	0.00	0.00	0.00
Temperature (°C)	26.70	26.70	26.70	26.70
Pressure (bar)	1.01	1.01	1.01	1.01
Density (g/L)	1,335.34	1,335.34	1,335.34	1,335.34
Component Flowrates (kg/h averaged)				
Non-starch Poly	46,686.132	46,686.132	46,686.132	46,546.073
Oil	22,773.723	22,773.723	22,773.723	22,705.402
Other Solids	45,547.445	45,547.445	45,547.445	45,410.803
Protein - insol	33,021.898	33,021.898	33,021.898	32,922.832
Protein - solub	22,773.723	22,773.723	22,773.723	22,705.402
Starch	398,540.147	398,540.147	398,540.147	397,344.527
Water	100,472.306	100,472.306	100,472.306	100,170.889
TOTAL (kg/h)	669,815.374	669,815.374	669,815.374	667,805.927
TOTAL (L/h)	501,606.466	501,606.466	501,606.466	500,101.646

Stream Name	Trash	S-109	S-110	S-111
Source	103MH	104M	105V	106W
Destination	OUTPUT	105V	106W	107V
Stream Properties				
Activity (U/ml)	0.00	0.00	0.00	0.00
Temperature (°C)	26.70	26.70	26.70	26.70
Pressure (bar)	1.01	1.01	1.01	1.01
Density (g/L)	1,335.34	1,335.34	1,335.34	1,335.34
Component Flowrates (kg/h averaged)				
Non-starch Poly	140.058	46,546.073	46,546.073	46,546.073
Oil	68.321	22,705.402	22,705.402	22,705.402
Other Solids	136.642	45,410.803	45,410.803	45,410.803
Protein - insol	99.066	32,922.832	32,922.832	32,922.832
Protein - solub	68.321	22,705.402	22,705.402	22,705.402
Starch	1,195.620	397,344.527	397,344.527	397,344.527
Water	301.417	100,170.889	100,170.889	100,170.889
TOTAL (kg/h)	2,009.446	667,805.927	667,805.927	667,805.927
TOTAL (L/h)	1,504.819	500,101.646	500,101.646	500,101.646

Stream Name	S-112	S-115	S-116	S-173
Source	107V	307V	308P	309E
Destination	307V	308P	309E	310V
Stream Properties				
Activity (U/ml)	0.00	0.00	0.00	0.00
Temperature (°C)	26.70	27.81	27.88	87.80
Pressure (bar)	1.01	1.01	4.18	4.18
Density (g/L)	1,335.34	1,333.47	1,333.45	2,692.05
Component Flowrates (kg/h averaged)				
Non-starch Poly	46,546.073	46,546.073	46,546.073	46,546.073
Oil	22,705.402	22,705.402	22,705.402	22,705.402
Other Solids	45,410.803	45,764.873	45,764.873	45,764.873
Protein - insol	32,922.832	32,922.832	32,922.832	32,922.832
Protein - solub	22,705.402	22,705.402	22,705.402	22,705.402
Starch	397,344.527	397,344.527	397,344.527	397,344.527
Water	100,170.889	102,331.697	102,331.697	102,331.697
TOTAL (kg/h)	667,805.927	670,320.805	670,320.805	670,320.805
TOTAL (L/h)	500,101.646	502,687.860	502,695.208	249,000.571

Stream Name	S-118	S-101	S-103	S-122
Source	**310V**	**311P**	**P-1**	**321V**
Destination	**311P**	**P-1**	**321V**	**322P**
Stream Properties				
Activity (U/ml)	0.00	0.00	0.00	0.00
Temperature (°C)	88.32	88.37	88.74	60.00
Pressure (bar)	1.01	3.08	19.08	1.01
Density (g/L)	1,317.12	1,317.11	1,317.01	1,146.20
Component Flowrates (kg/h averaged)				
Glucose	0.000	0.000	0.000	437,078.980
Non-starch Poly	46,546.073	46,546.073	46,546.073	46,546.073
Oil	22,705.402	22,705.402	22,705.402	22,705.402
Other Solids	45,764.873	45,764.873	45,764.873	45,764.873
Protein - insol	32,922.832	32,922.832	32,922.832	32,922.832
Protein - solub	22,705.402	22,705.402	22,705.402	22,705.402
Starch	397,344.527	397,344.527	397,344.527	3,973.445
Sulfuric Acid	0.000	0.000	0.000	125.473
Water	102,331.697	102,331.697	102,331.697	58,687.346
TOTAL (kg/h)	670,320.805	670,320.805	670,320.805	670,509.825
TOTAL (L/h)	508,928.103	508,933.265	508,973.209	584,986.326

Stream Name	S-102
Source	**322P**
Destination	**OUTPUT**
Stream Properties	
Activity (U/ml)	0.00
Temperature (°C)	60.02
Pressure (bar)	2.25
Density (g/L)	1,146.18
Component Flowrates (kg/h averaged)	
Glucose	437,078.980
Non-starch Poly	46,546.073
Oil	22,705.402
Other Solids	45,764.873
Protein - insol	32,922.832
Protein - solub	22,705.402
Starch	3,973.445
Sulfuric Acid	125.473
Water	58,687.346
TOTAL (kg/h)	670,509.825
TOTAL (L/h)	584,993.456

4. OVERALL COMPONENT BALANCE (kg/h)

COMPONENT	IN	OUT	OUT-IN
Glucose	0.000	437,078.980	437,078.980
Non-starch Poly	46,686.132	46,686.132	0.000
Oil	22,773.723	22,773.723	0.000
Other Solids	45,901.515	45,901.515	0.000
Protein - insol	33,021.898	33,021.898	0.000
Protein - solub	22,773.723	22,773.723	0.000
Starch	398,540.147	5,169.066	- 393,371.082
Sulfuric Acid	125.473	125.473	0.000
Water	102,696.661	58,988.763	- 43,707.898
TOTAL	672,519.271	672,519.271	0.000

Economic Evaluation Report

April 13, 2009

for corn to glucose 4-13

1. EXECUTIVE SUMMARY (2007 prices)

Total Capital Investment	59576000.00	$
Capital Investment Charged to This Project	59576000.00	$
Operating Cost	752706000.00	$/yr
THE MAIN REVENUE STREAM HAS NOT BEEN IDENTIFIED. PRICING AND PRODUCTION/PROCESSING UNIT COST DATA HAVE NOT BEEN PRINTED		
Main Revenue	0.00	$/yr
Gross Margin	- 1.00	%
Return On Investment	- 1,253.44	%
Payback Time	- 1.00	years
IRR (After Taxes)	Out of search interval	(0-1000%)
NPV (at 5.0% Interest)	0.00	$
MT = Metric Ton (1000 kg)		

2. MAJOR EQUIPMENT SPECIFICATION AND FOB COST (2007 prices)

Quantity/ Standby/ Staggered	Name	Description	Unit Cost ($)	Cost ($)
1 / 0 / 0	101MH	Belt Conveyor	121000.00	121000.00
		Belt Length = 100.00 m		
1 / 0 / 0		Silo/Bin	14152000.00	14152000.00
		Vessel Volume = 267926.15 m3		
12 / 0 / 0	104M	Grinder	110000.00	1320000.00
		Size/Capacity = 55650.49 kg/h		
8 / 0 / 0	105V	Receiver Tank	46000.00	368000.00
		Vessel Volume = 138.92 m3		
10 / 0 / 0	106W	Hopper	63000.00	630000.00
		Vessel Volume = 145.85 m3		
8 / 0 / 0	107V	Receiver Tank	62000.00	496000.00
		Vessel Volume = 138.92 m3		
3 / 0 / 0	307V	Blending Tank	165000.00	495000.00
		Vessel Volume = 64.44 m3		

1 / 0 / 0	305V	Hopper	9000.00	9000.00
		Vessel Volume = 4.02 m3		
1 / 0 / 0	303V	Receiver Tank	28000.00	28000.00
		Vessel Volume = 14.36 m3		
1 / 0 / 0	301V	Receiver Tank	54000.00	54000.00
		Vessel Volume = 14.14 m3		
1 / 0 / 0	302P	Gear Pump	4000.00	4000.00
		Power = 0.20 kW		
1 / 0 / 0	310V	Blending Tank	226000.00	226000.00
		Vessel Volume = 249.00 m3		
1 / 0 / 0	321V	Stirred Reactor	222000.00	222000.00
		Vessel Volume = 188.56 m3		
1 / 0 / 0	317V	Receiver Tank	101000.00	101000.00
		Vessel Volume = 23.82 m3		
1 / 0 / 0	319V	Receiver Tank	23000.00	23000.00
		Vessel Volume = 25.56 m3		
1 / 0 / 0	304P	Gear Pump	4000.00	4000.00
		Power = 0.25 HP-E		
1 / 0 / 0	318P	Gear Pump	4000.00	4000.00
		Power = 0.25 HP-E		
1 / 0 / 0	320P	Gear Pump	4000.00	4000.00
		Power = 0.03 kW		
4 / 0 / 0	308P	Centrifugal Pump	25000.00	100000.00
		Power = 15.82 kW		
1 / 0 / 0	311P	Centrifugal Pump	15000.00	15000.00
		Power = 50.00 kW		
1 / 0 / 0	322P	Centrifugal Pump	15000.00	15000.00
		Power = 50.00 HP-E		
1 / 0 / 0	306P	Gear Pump	4000.00	4000.00
		Power = 5.00 HP-E		
2 / 0 / 0	309E	Heat Exchanger	39000.00	78000.00
		Heat Exchange Area = 93.69 m2		
1 / 0 / 0	103MH	Flow Splitter	303000.00	303000.00
		Size/Capacity = 669815.37 kg/h		
2 / 0 / 0	PM-101	Centrifugal Pump	131000.00	262000.00

Pump Power = 161.57 kW			
Unlisted Equipment			0.00
		TOTAL	19038000.00

3. DIRECT FIXED CAPITAL COST (DFC) SUMMARY (2007 prices in $)

Section Name	DFC ($)
Main Section	1048000.00
Grain Handling & Milling	52170000.00
Starch to Sugar Conversion	4158000.00
Fermentation	0.00
Ethanol Processing	0.00
CoProduct Processing	0.00
Common Suport Systems	2200000.00
Plant DFC	59576000.00

4. LABOR COST - PROCESS SUMMARY

Labor Type	Unit Cost ($/h)	Annual Amount (h)	Annual Cost ($)	%
Operator	0.00	0.00	0.00	0.00
Plant Operators	52.00	39600.00	2059200.00	100.00
TOTAL		39600.00	2059200.00	100.00

5. MATERIALS COST - PROCESS SUMMARY

Bulk Material	Unit Cost ($/kg)	Annual Amount (kg)	Annual Cost ($)	%
Corn	0.14	5304937759.00	730755176.00	99.66
Lime	0.09	718270.00	64644.00	0.01
Liq. Ammonia	0.22	1200956.00	264210.00	0.04
Alpha-Amylase	2.25	298452.00	671518.00	0.09
Glucoamylase	2.25	503297.00	1132419.00	0.15
Sulfuric Acid	0.11	993746.00	109312.00	0.01
Caustic	0.01	17700150.00	214526.00	0.03

TOTAL		5326352630.00	733211806.00	100.00

NOTE: Bulk material consumption amount includes material used as:
- Raw Material
- Cleaning Agent
- Heat Tranfer Agent (if utilities are included in the operating cost)

6. VARIOUS CONSUMABLES COST (2007 prices) - PROCESS SUMMARY

THE CONSUMABLES COST IS ZERO.

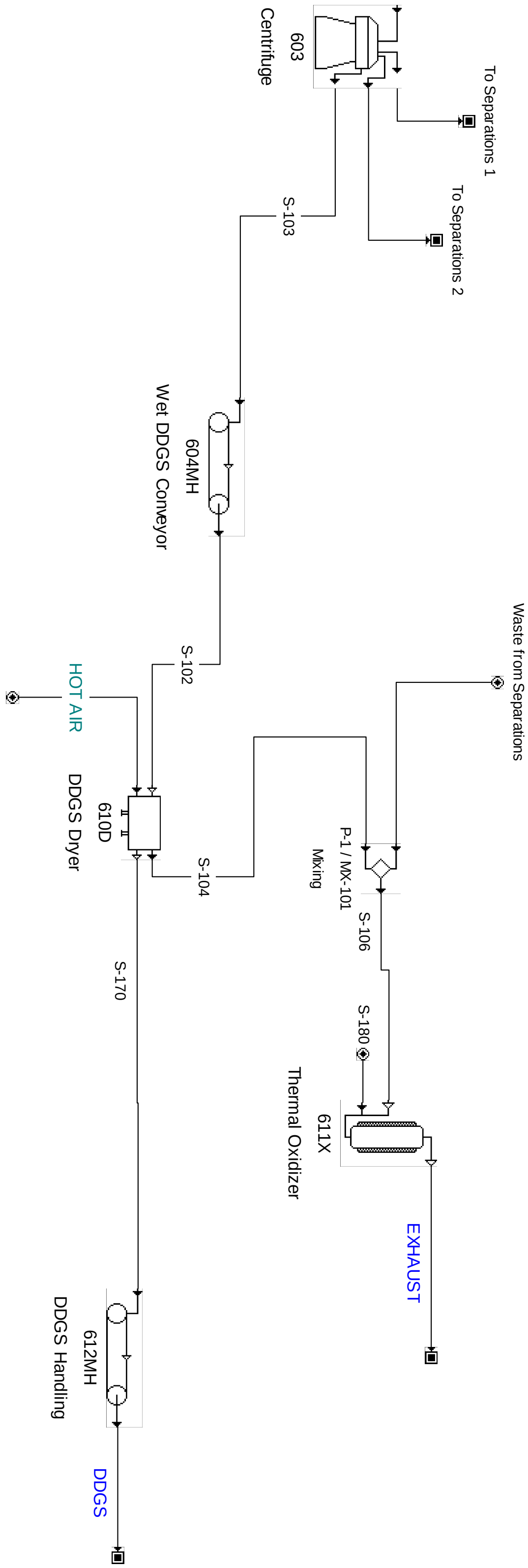
7. UTILITIES COST (2007 prices) - PROCESS SUMMARY

Utility	Annual Amount	Reference Units	Annual Cost ($)	%
Electricity	56511151.00	kWh	2825558.00	32.23
Steam	0.00	kg	0.00	0.00
Cooling Water	4554701699.00	kg	455470.00	5.20
Chilled Water	0.00	kg	0.00	0.00
CT Water	0.00	kg	0.00	0.00
Steam 50 PSI	256951478.00	kg	5485914.00	62.58
Steam (High P)	0.00	kg	0.00	0.00
TOTAL			8766942.00	100.00

8. ANNUAL OPERATING COST (2007 prices) - PROCESS SUMMARY

Cost Item	$	%
Raw Materials	733212000.00	97.41
Labor-Dependent	2059000.00	0.27
Facility-Dependent	8668000.00	1.15
Consumables	0.00	0.00
Utilities	8767000.00	1.16
Advertising/Selling	0.00	0.00
Running Royalties	0.00	0.00
Failed Product Disposal	0.00	0.00
TOTAL	752706000.00	100.00

DDGS Drying



To Separations 1
To Separations 2
603
Centrifuge
S-103
604MH
Wet DDGS Conveyor
S-102
HOT AIR
610D
DDGS Dryer
Waste from Separations
P-1 / MX-101
Mixing
S-104
S-106
S-180
611X
Thermal Oxidizer
EXHAUST
S-170
612MH
DDGS Handling
DDGS

Materials & Streams Report
for ddgs drying 4-13

April 13, 2009

1. OVERALL PROCESS DATA

Annual Operating Time	7,920.00h
Annual Throughput	0.00kg MP
Operating Days per Year	330.00

MP = Main Product = Undefined

2.1 STARTING MATERIAL REQUIREMENTS (per Section)

Section	Starting Material	Active Product	Amount Needed (kg Sin/kg	Molar Yield (%)	Mass Yield (%)	Gross Mass Yield (%)
Main Section	(none)	(none)	Unknow	Unknow	Unknown	Unknow
Grain Handling & Milling	(none)	(none)	Unknow	Unknow	Unknown	Unknow
Starch to Sugar Conversion	(none)	(none)	Unknow	Unknow	Unknown	Unknow
Fermentation	(none)	(none)	Unknow	Unknow	Unknown	Unknow
Ethanol Processing	(none)	(none)	Unknow	Unknow	Unknown	Unknow
CoProduct Processing	(none)	(none)	Unknow	Unknow	Unknown	Unknow
Common Suport Systems	(none)	(none)	Unknow	Unknow	Unknown	Unknow

Sin = Section Starting Material, Aout = Section Active Product

2.2 BULK MATERIALS (Entire Process)

Material	kg/yr	kg/h	kg/kg MP
Ethyl Alcohol	178,154,551	22,494.261	
Water	29,607,907,343	3,738,372.139	
Air	63,781,528	8,053.223	
Other Solids	319,096,121	40,289.914	
TOTAL	30,168,939,542	3,809,209.538	

2.3 BULK MATERIALS (per Section)

SECTIONS IN: Main Branch

Main Section

Material	kg/yr	kg/h	kg/kg MP
Ethyl Alcohol	6,969,230	879.953	
Water	6,110,971	771.587	
TOTAL	13,080,201	1,651.541	

CoProduct Processing			
Material	**kg/yr**	**kg/h**	**kg/kg MP**
Air	63,781,528	8,053.223	
Ethyl Alcohol	171,185,320	21,614.308	
Other Solids	319,096,121	40,289.914	
Water	29,601,796,372	3,737,600.552	
TOTAL	30,155,859,341	3,807,557.998	

2.4 BULK MATERIALS (per Material)

Ethyl Alcohol				
Ethyl Alcohol	**% Total**	**kg/yr**	**kg/h**	**kg/kg MP**
Main Section (Main Branch)				
P-1	3.91	6,969,230	879.953	
CoProduct Processing (Main Branch)				
603	96.09	171,185,320	21,614.308	
TOTAL	100.00	178,154,551	22,494.261	

Water				
Water	**% Total**	**kg/yr**	**kg/h**	**kg/kg MP**
Main Section (Main Branch)				
P-1	0.02	6,110,971	771.587	
CoProduct Processing (Main Branch)				
603	99.98	29,601,796,372	3,737,600.552	
TOTAL	100.00	29,607,907,343	3,738,372.139	

Air				
Air	**% Total**	**kg/yr**	**kg/h**	**kg/kg MP**
CoProduct Processing (Main Branch)				
610D	2.22	1,417,403	178.965	
611X	97.78	62,364,125	7,874.258	
TOTAL	100.00	63,781,528	8,053.223	

Other Solids				
Other Solids	**% Total**	**kg/yr**	**kg/h**	**kg/kg MP**
CoProduct Processing (Main Branch)				
603	100.00	319,096,121	40,289.914	
TOTAL	100.00	319,096,121	40,289.914	

2.5 BULK MATERIALS: SECTION TOTALS (kg/h)

Raw Material	Main Section	Grain Handling &	Starch to Sugar Conversion	Fermentation
Ethyl Alcohol	879.953	0.000	0.000	0.000
Water	771.587	0.000	0.000	0.000
Air	0.000	0.000	0.000	0.000
Other Solids	0.000	0.000	0.000	0.000
TOTAL	1,651.541	0.000	0.000	0.000

Raw Material	Ethanol Processing	CoProduct Processing	Common Suport
Ethyl Alcohol	0.000	21,614.308	0.000
Water	0.000	3,737,600.552	0.000
Air	0.000	8,053.223	0.000
Other Solids	0.000	40,289.914	0.000
TOTAL	0.000	3,807,557.998	0.000

2.6 BULK MATERIALS: SECTION TOTALS (kg/yr)

Raw Material	Main Section	Grain Handling &	Starch to Sugar Conversion	Fermentation
Ethyl Alcohol	6,969,230	0	0	0
Water	6,110,971	0	0	0
Air	0	0	0	0
Other Solids	0	0	0	0
TOTAL	13,080,201	0	0	0

Raw Material	Ethanol Processing	CoProduct Processing	Common Suport
Ethyl Alcohol	0	171,185,320	0
Water	0	29,601,796,372	0
Air	0	63,781,528	0
Other Solids	0	319,096,121	0
TOTAL	0	30,155,859,341	0

3. STREAM DETAILS

Stream Name	S-101	To Separations	To Separations	S-103
Source	INPUT	603	603	603
Destination	603	OUTPUT	OUTPUT	604MH
Stream Properties				
Activity (U/ml)	0.00	0.00	0.00	0.00
Temperature (°C)	25.00	25.01	25.01	25.01
Pressure (bar)	1.01	1.01	1.01	1.01
Density (g/L)	994.81	993.03	994.55	1,154.39
Component Flowrates (kg/h averaged)				
Ethyl Alcohol	21,614.308	21,398.165	213.982	2.161
Other Solids	40,289.914	0.000	0.000	40,289.914
Water	3,737,600.552	3,363,840.497	369,648.695	4,111.361
TOTAL (kg/h)	3,799,504.774	3,385,238.662	369,862.676	44,403.436
TOTAL (L/h)	3,819,330.486	3,408,995.825	371,890.949	38,464.801

Stream Name	S-102	HOT AIR	S-104	S-170
Source	604MH	INPUT	610D	610D
Destination	610D	610D	P-1	612MH
Stream Properties				
Activity (U/ml)	0.00	0.00	0.00	0.00
Temperature (°C)	25.01	104.00	70.00	70.00
Pressure (bar)	1.01	1.01	1.01	1.01
Density (g/L)	1,154.39	0.93	1.76	1,124.04
Component Flowrates (kg/h averaged)				
Ethyl Alcohol	2.161	0.000	0.067	2.094
Nitrogen	0.000	137.287	137.287	0.000
Other Solids	40,289.914	0.000	0.000	40,289.914
Oxygen	0.000	41.678	41.678	0.000
Water	4,111.361	0.000	127.765	3,983.596
TOTAL (kg/h)	44,403.436	178.965	306.797	44,275.604
TOTAL (L/h)	38,464.801	191,986.468	174,766.515	39,389.639

Stream Name	DDGS	Waste from Separations	S-106	S-180
Source	**612MH**	**INPUT**	**P-1**	**INPUT**
Destination	**OUTPUT**	**P-1**	**611X**	**611X**
Stream Properties				
Activity (U/ml)	0.00	0.00	0.00	0.00
Temperature (°C)	70.00	25.00	30.28	25.00
Pressure (bar)	1.01	1.01	1.01	1.01
Density (g/L)	1,124.04	871.35	12.52	1.18
Component Flowrates (kg/h averaged)				
Ethyl Alcohol	2.094	879.953	880.020	0.000
Nitrogen	0.000	0.000	137.287	6,040.484
Other Solids	40,289.914	0.000	0.000	0.000
Oxygen	0.000	0.000	41.678	1,833.774
Water	3,983.596	771.587	899.352	0.000
TOTAL (kg/h)	44,275.604	1,651.541	1,958.338	7,874.258
TOTAL (L/h)	39,389.639	1,895.386	156,453.658	6,677,790.886

Stream Name	EXHAUST
Source	**611X**
Destination	**OUTPUT**
Stream Properties	
Activity (U/ml)	0.00
Temperature (°C)	79.78
Pressure (bar)	1.01
Density (g/L)	1.31
Component Flowrates (kg/h averaged)	
Carb. Dioxide	1,681.342
Nitrogen	6,177.771
Oxygen	41.678
Water	1,931.996
TOTAL (kg/h)	9,832.787
TOTAL (L/h)	7,532,952.352

4. OVERALL COMPONENT BALANCE (kg/h)

COMPONENT	IN	OUT	OUT-IN
Carb. Dioxide	0.000	1,681.342	1,681.342
Ethyl Alcohol	22,494.261	21,614.241	- 880.020
Nitrogen	6,177.771	6,177.771	0.000
Other Solids	40,289.914	40,289.914	0.000
Oxygen	1,875.452	41.678	- 1,833.774
Water	3,738,372.139	3,739,404.783	1,032.644
TOTAL	3,809,209.538	3,809,209.729	0.191

Economic Evaluation Report

April 13, 2009

for ddgs drying 4-13

1. EXECUTIVE SUMMARY (2007 prices)

Total Capital Investment	92980000.00	$
Capital Investment Charged to This Project	92980000.00	$
Operating Cost	17131000.00	$/yr
THE MAIN REVENUE STREAM HAS NOT BEEN IDENTIFIED. PRICING AND PRODUCTION/PROCESSING UNIT COST DATA HAVE NOT BEEN PRINTED		
Main Revenue	0.00	$/yr
Other Revenues	6635.00	$/yr
Total Revenues	7000.00	$/yr
Gross Margin	- 258,098.83	%
Return On Investment	- 8.42	%
Payback Time	- 1.00	years
IRR (After Taxes)	Out of search interval	(0-1000%)
NPV (at 5.0% Interest)	- 155,907,000	$
MT = Metric Ton (1000 kg)		

2. MAJOR EQUIPMENT SPECIFICATION AND FOB COST (2007 prices)

Quantity/ Standby/ Staggered	Name	Description	Unit Cost ($)	Cost ($)
1 / 0 / 0	604MH	Belt Conveyor	56000.00	56000.00
		Belt Length = 100.00 m		
1 / 0 / 0	610D	Rotary Dryer	277000.00	277000.00
		Drying Area = 37.17 m2		
1 / 0 / 0	612MH	Belt Conveyor	123000.00	123000.00
		Belt Length = 100.00 m		
1 / 0 / 0	611X	Wet Air Oxidizer	204000.00	204000.00
		Vessel Volume = 1.07 m3		
32 / 0 / 0	603.00	Disk-Stack Centrifuge	925000.00	29600000.00
		Throughput = 1989.23 L/min		
1 / 0 / 0	MX-101	Mixer	0.00	0.00
		Size/Capacity = 1958.34 kg/h		
		Unlisted Equipment		0.00
			TOTAL	30260000.00

3. DIRECT FIXED CAPITAL COST (DFC) SUMMARY (2007 prices in $)

Section Name	DFC ($)

Main Section	0.00
Grain Handling & Milling	0.00
Starch to Sugar Conversion	0.00
Fermentation	0.00
Ethanol Processing	0.00
CoProduct Processing	90780000.00
Common Suport Systems	2200000.00
Plant DFC	92980000.00

4. LABOR COST - PROCESS SUMMARY

Labor Type	Unit Cost ($/h)	Annual Amount (h)	Annual Cost ($)	%
Operator	0.00	0.00	0.00	0.00
Plant Operators	52.00	39600.00	2059200.00	100.00
TOTAL		39600.00	2059200.00	100.00

5. MATERIALS COST - PROCESS SUMMARY

Bulk Material	Unit Cost ($/kg)	Annual Amount (kg)	Annual Cost ($)	%
Ethyl Alcohol	0.00	178154551.00	0.00	0.00
Water	0.00	29607907343.00	1302748.00	100.00
Air	0.00	63781528.00	0.00	0.00
Other Solids	0.00	319096121.00	0.00	0.00
TOTAL		30168939542.00	1302748.00	100.00

NOTE: Bulk material consumption amount includes material used as:
- Raw Material
- Cleaning Agent
- Heat Tranfer Agent (if utilities are included in the operating cost)

6. VARIOUS CONSUMABLES COST (2007 prices) - PROCESS SUMMARY

THE CONSUMABLES COST IS ZERO.

7. UTILITIES COST (2007 prices) - PROCESS SUMMARY

Utility	Annual Amount	Reference Units	Annual Cost ($)	%
Electricity	4372295.00	kWh	218615.00	91.10
Steam	0.00	kg	0.00	0.00
Cooling Water	0.00	kg	0.00	0.00
Chilled Water	0.00	kg	0.00	0.00
Natural Gas	60746.00	kg	21364.00	8.90
CT Water	0.00	kg	0.00	0.00
TOTAL			239978.00	100.00

8. ANNUAL OPERATING COST (2007 prices) -

PROCESS SUMMARY

Cost Item	$	%
Raw Materials	1303000.00	7.60
Labor-Dependent	2059000.00	12.02
Facility-Dependent	13529000.00	78.97
Consumables	0.00	0.00
Utilities	240000.00	1.40
Advertising/Selling	0.00	0.00
Running Royalties	0.00	0.00
Failed Product Disposal	0.00	0.00
TOTAL	17131000.00	100.00

Appendix E:

Aspen Stream Summary

STREAM:		1	7OVHD	8BTMS	B1BTMS	B1OVHD	FEED	OVHD	SOLVENT	SOLVRECY
From		HX301	DIST3	DIST3	EXTRACT1	EXTRACT1		DIST1		DIST1
To		DIST1			HX301		EXTRACT1	DIST3	EXTRACT1	
Substream: MIXED										
Phase:		Liquid	Liquid	Liquid	Liquid	Liquid	Liquid	Liquid	Liquid	Liquid
Component Mole Flow										
WATER	LBMOL/HR	97.16	94.42	2.73	97.16	457427	457524	97.16	0	0
ETHANOL	LBMOL/HR	10.48	9.91	0.58	10.48	12.28	22.76	10.48	0	0
ACETONE	LBMOL/HR	9.49	9.39	0.1	9.49	26.77	36.25	9.49	0	0
N-BUT-01	LBMOL/HR	637.6	12.66	620.16	637.6	5.27	642.87	632.82	0	4.78
SOLIDS	LBMOL/HR	0	0	0	0	0	0	0	0	0
N-DOD-01	LBMOL/HR	40000	0	0.88	40000	0	0	0.88	40000	39999.12
Component Mole Fraction										
WATER		0	0.75	0	0	1	1	0.13	0	0
ETHANOL		0	0.08	0	0	0	0	0.01	0	0
ACETONE		0	0.07	0	0	0	0	0.01	0	0
N-BUT-01		0.02	0.1	0.99	0.02	0	0	0.84	0	0
SOLIDS		0	0	0	0	0	0	0	0	0
N-DOD-01		0.98	0	0	0.98	0	0	0	1	1

STREAM:		1	7OVHD	8BTMS	B1BTMS	B1OVHD	FEED	OVHD	SOLVENT	SOLVRECY
Component Mass Flow										
WATER	LB/HR	1750.3	1701.1	49.25	1750.31	8240670	8E+06	1750.3	0	0
ETHANOL	LB/HR	482.99	456.45	26.53	482.99	565.62	1048.6	482.99	0	0
ACETONE	LB/HR	551.05	545.36	5.68	551.05	1554.55	2105.6	551.05	0	0
N-BUT-01	LB/HR	47260	938.15	45968	47260.5	390.94	47651	46906	0	353.99
SOLIDS	LB/HR	0	0	0	0	0	0	0	0	0
N-DOD-01	LB/HR	7E+06	0	149.53	6813537	0.6	0	149.53	6813538	6813387
Component Mass Fraction										
WATER		0	0.47	0	0	1	0.99	0.04	0	0
ETHANOL		0	0.13	0	0	0	0	0.01	0	0
ACETONE		0	0.15	0	0	0	0	0.01	0	0
N-BUT-01		0.01	0.26	0.99	0.01	0	0.01	0.94	0	0
SOLIDS		0	0	0	0	0	0	0	0	0
N-DOD-01		0.99	0	0	0.99	0	0	0	1	1

Appendix F

Tal Raviv <ravivt@seas.upenn.edu> **Sun, Feb 8, 2009 at 11:29 AM**

To: bruce.m.vrana@usa.dupont.com

Cc: Christina Chen <chencl@seas.upenn.edu>, Amira Fawcett <amirafawcett@gmail.com>, Amy Posner <posneram@seas.upenn.edu>

Hi Bruce,

I hope you are having a great weekend.
After our meeting this week, the team has been investigating separations alternatives for post-fermentation of butanol. I have found a few mentions of solids entering distillation columns, and it seems to be the way ethanol is indeed separated. We are now looking for specific processes that do this explicitly and that I can cite to show it is a legitimate method of separation and what equipment terminology it is (models that won't clog reboiler, etc.)

The other option on the table is to hydrocyclone the mash, then strip the remaining liquid with N2 or CO2 to remove solvents only. But the purged mash that's separated in the hydrocyclone will have both reactants and products in there, unfortunately. So for now the team is preferring the distillation route. Do you have any suggestions on companies to contact, or downstream separation processes examples we should seek out?

Thank you very much,
Tal Raviv

Bruce M Vrana <Bruce.M.Vrana@usa.dupont.com> **Mon, Feb 9, 2009 at 2:00 PM**

To: ravivt@seas.upenn.edu

Cc: Amira Fawcett <amirafawcett@gmail.com>, Christina Chen <chencl@seas.upenn.edu>, Amy Posner <posneram@seas.upenn.edu>, talsraviv@gmail.com

The dry grind corn to ethanol process in the U.S. is the process you want to compare to. But very little is published about the distillation column design. The USDA worked on a flowsheet that I think I gave you the reference to, and published a Superpro Designer flowsheet for. They previously had an Aspen Plus model, although I don't think they published it. But we've worked with their Aspen Model, and it has about 11.5% solids in the feed to the beer column. Newer ethanol plants run at even higher solids, probably as high as 14%. You could cite that as a "private communication" from me, if you want, in your report. If someone questions you during the presentation, your answer is that you're running the same solids loading as over 100 corn to ethanol plants in this country, using the same type of column internals.

Solids are best handled with baffle trays. These are briefly described in the APV Distillation Handbook, which can be found, among other places, at http://www.research.umbc.edu/~dfrey1/ench445/apv_distill.pdf . I would assume 40% tray efficiency to be on the safe side. You can assume that your costing correlations for trays will be conservative for the cost of baffle trays, which are mechanically simpler.

Let me know if you have any other questions.

Bruce

chencl@seas.upenn.edu <chencl@seas.upenn.edu> **Mon, Feb 9, 2009 at 11:25 PM**

To: Bruce M Vrana <Bruce.M.Vrana@usa.dupont.com>

Cc: ravivt@seas.upenn.edu, Amira Fawcett <amirafawcett@gmail.com>, Amy Posner <posneram@seas.upenn.edu>, talsraviv@gmail.com

Dear Bruce,

I just had a quick question on the corn refinery process. I was reading up on the wet milling process which extracts out starch specifically which is the ideal feed to the fermenters. Do you know if this process or the dry grind process would be preferred? Also, I was confused whether you meant we should model after the dry grind ethanol process or the dry milling process. From my reading it seemed like these two either produced ethanol or just degermed corn, not the starch that we need for the feeds to the fermenters. Do you have more information about this process? Thank you!

Christina Chen

Bruce M Vrana <Bruce.M.Vrana@usa.dupont.com> **Tue, Feb 10, 2009 at 10:59 AM**

To: chencl <chencl@seas.upenn.edu>

Cc: Amira Fawcett <amirafawcett@gmail.com>, Amy Posner <posneram@seas.upenn.edu>, ravivt@seas.upenn.edu, talsraviv@gmail.com

Christina,

The wet mill process does produce starch and then sugar which would be an ideal feedstock. And in fact, some ethanol is made that way in the U.S. But wet mills are very capital intensive and thus need to be built at much bigger scale to be economical, and nobody is building new ones for that reason. The vast majority of new ethanol capacity being built today is the corn dry grind process, and that's what I intended you to use. You obviously will only need to use the front end of the process, through making the sugars - the milling, cooking and saccharification steps will look very much like an ethanol plant. I think the references I gave you in the problem statement should have enough information for you to design the front end of your butanol process, but feel free to ask more questions once you get into that part of the design.

Bruce

Bruce M Vrana <Bruce.M.Vrana@usa.dupont.com> **Tue, Feb 10, 2009 at 5:00 PM**

To: chencl <chencl@seas.upenn.edu>

Cc: Amira Fawcett <amirafawcett@gmail.com>, Amy Posner <posneram@seas.upenn.edu>, ravivt@seas.upenn.edu, talsraviv@gmail.com

Christina,

My earlier note to Tal, that I think you were cc-ed on, gave 11.5% solids as a safe reference point, as that is what is practiced in most corn ethanol plants. Some go somewhat higher, perhaps up to 14% solids. I'm not aware of any good references on handling solids in distillation - since most folks avoid it very carefully.

One other reference might be helpful, if Towne library has it (which I doubt it does, but perhaps they could borrow a copy from interlibrary loan and then suggest they order a copy for themselves, since fuel alcohol is a topic of increasing interest). It's called "The Alcohol Textbook". Published by Nottingham University Press, written by Alltech Inc., lead editor K. A. Jacques. 4th edition was in 2003 and is now out of print (ISBN 1-897676-13-1), but I understand there is a 5th edition coming out soon.

Just looking, The Alcohol Handbook mentions "disc and donut" trays as being used in beer columns. I think that is basically the same as the baffle trays I mentioned earlier.

Bruce

---

chencl@seas.upenn.edu <chencl@seas.upenn.edu> **Tue, Feb 10, 2009 at 5:11 PM**

To: Bruce M Vrana <Bruce.M.Vrana@usa.dupont.com>

Cc: Amira Fawcett <amirafawcett@gmail.com>, Amy Posner <posneram@seas.upenn.edu>, ravivt@seas.upenn.edu, talsraviv@gmail.com

Hi Bruce,

I saw that the 11.5% was the feed concentration to the beer column, but in our meeting today, the consultants told us to find the max percentage of the bottoms product out the column as well because that could effect bubble point calculations and things like that. I was just wondering if you knew that offhand. I will definitely look up the Alcohol Textbook for more information. Thank you!

Christina

Bruce M Vrana <Bruce.M.Vrana@usa.dupont.com> **Tue, Feb 10, 2009 at 5:32 PM**

To: chencl <chencl@seas.upenn.edu>

Cc: Amira Fawcett <amirafawcett@gmail.com>, Amy Posner <posneram@seas.upenn.edu>, ravivt@seas.upenn.edu, talsraviv@gmail.com

Christina,

Sorry, I guess I misunderstood your question. But the answer is not much different, since most of the feed winds up in the bottoms. Our look at the USDA flowsheet shows about 14%   solids in the bottoms stream practiced widely in the industry. Some of the newer plants might be as much as 17% solids in the bottoms.

Bruce