Ethanol from Sugar Beets:
A Process and Economic Analysis

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Abstract

The aim of this project was to design a process for producing bioethanol from sugar beets as a possible feedstock replacement for corn. Currently eighty-five percent of the energy used by the United States comes from non-renewable fossil fuels, which contribute to global climate change and are environmentally destructive (The Role of Renewable Energy Consumption, 2007). The process of converting corn to bioethanol is very energy intensive, as well as involving a staple food crop. Domestic sugar demand has decreased in recent years, leaving space for the growth of sugar beets for other purposes (SIC 2063, 2005).

A process was designed and modeled using commercially available chemical engineering design software and a cost analysis was performed. This process was compared to a process designed by the USDA that modeled the production of bioethanol from corn. Results showed that sugar beets are a superior feedstock for producing bioethanol compared to corn. For the same amount of bioethanol, sugar beets require a smaller acreage of land and fewer sugar beets, and potentially utilizes a more direct and inexpensive process with a smaller environmental cost. We recommend that further research into energy be conducted on the sugar beets to bioethanol process and that a pilot-scale plant be built to pave the way for future developments in bioethanol production.

Executive Summary

Fossil fuels are heavily relied on as a source of energy in the United States. Currently 85% of the energy used by the United States is from fossil fuels. Fossil fuels create pollution when they are burned, may contribute to climate change, and the supply of fossil fuels is limited (The Role of Renewable Energy Consumption, 2007). As a result considerable research is being done to find alternative fuels that can be used as a replacement for fossil fuels. Many types of clean alternative energy are currently being researched, such as solar, wind, and biofuels.

This project will focus on bioethanol as a promising alternative fuel. Bioethanol is made from plant material that is broken down and fermented by yeast. In Brazil, bioethanol from sugarcane is widely used as a fuel for automobiles. In fact, 45% of the fuel used for vehicles in Brazil is bioethanol from sugarcane (Rohter, 2006). Brazil has shown that this is a feasible and economical liquid fuel. However, due to the difference in climates, the United States cannot grow sugarcane in sufficient quantities to produce ethanol. Instead the United States currently uses corn as a feedstock for bioethanol production.

Corn is not an ideal feedstock for ethanol production. There are a number of reasons for this, but the primary concern is the net energy value (NEV) of corn. A net energy value is the ratio of the energy a fuel provides divided by energy required to produce that fuel. The net energy value of corn is highly debated, and has been reported to range from 0.79 to 1.3 (Shapouri et all, 2002, Pimentel and Patzek, 2005). Even if the more optimistic estimates are correct, 1.3 is a mediocre NEV at best. By comparison sugarcane has an NEV of at least eight (Rohter, 2006).

As both feedstocks currently used for ethanol production have serious flaws, this project set out to investigate another feedstock: sugar beets. Sugar beets contain sucrose that is used to produce table sugar (Cattanach et all, 1991). Sugar beets can be grown over much of the US (Jacobs, 2006), and could be converted to ethanol using a process similar to that used to convert sugarcane to ethanol. The goal of this project was to design and perform a cost analysis of a sugar beet to ethanol process using information known about the process to convert sugar cane to ethanol and information known about the process to convert sugar beets to sugar. This process was modeled using two commercially available Chemical Engineering design software packages: Aspen Plus and SuperPro Designer.

The USDA performed a similar simulation using SuperPro Designer for a corn to ethanol process. The file for this simulation was obtained, and compared to the sugar beet to ethanol process that was designed. Equipment, feedstock, and utility costs for each

simulation were compared. The sugar beet to ethanol equipment cost was found to be approximately $16 million while the corn to ethanol equipment cost approximately $18.5 million. Yearly utility costs were found to be approximately $23 million and $15 million for beets and corn, respectively. Feedstock costs per gallon were found to be $1.76 for corn and $1.66 for beets. Costs contributions associated with other process inlets, such as yeast, lime, and solvents, were found to be $0.10 per gallon for corn and $0.07 per gallon for sugar beets. The cost of utilities per gallon, such as cooling water, chilled water, and steam were found to be $0.38 for corn and $0.57 for sugar beets. The total cost for producing one gallon of ethanol from corn is $2.24, as opposed to $2.30 for sugar beets. This cost includes the feedstock cost in addition to all process and utility costs. This does not take into account the more expensive equipment costs for the corn to ethanol process, and may vary significantly with fluctuating feedstock costs if the purchase price per kilogram of corn or beets fluctuates. The utility cost was shown to be the deciding factor in cost analysis. The USDA used a molecular sieve adsorption tower in their process to separate ethanol and water. Using this separation mechanism would allow the utility costs to be significantly dropped and cause ethanol from sugar beets to cost less per gallon than ethanol from corn.

It was also found that corn and sugar beets require approximately the same amount of water and fertilizer per acre of land planted. However one acre of sugar beets will produce about 1930 kg ethanol while one acre of corn will produce only 1000 kg. This means that more energy from fossil fuels will be used to irrigate and fertilize corn per kg of ethanol produced. Whether ethanol is being produced from corn or sugar beets there are some unusable parts of the plant that are left over. In sugar beets the pulp and tops are a valuable food for sheep, cows, and other livestock. In corn the stover (stalks, leaves, husks, and cobs) are generally just tilled back into the earth (Kyle, 2010).

Overall sugar beets seem to be a very promising feedstock. However future research is still needed to determine several key pieces of information about the sugar beet to ethanol process. The NEV for sugar beets should be researched further; particularly the energy used for the utilities in the sugar beet to ethanol process, and the energy used to grow the sugar beets should be determined. Since the utility costs are the largest difference between the corn and ethanol processes, alternative methods of separation should be researched to lower utility costs in the sugar beets to ethanol process. If utility costs are lowered, the sugar beets to ethanol process will likely become more economical than the corn to ethanol process. Also lab scale followed by pilot scale version of this process should be assembled to further test the process' feasibility.

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1. Introduction

Over the past decade, global climate change has turned from a heavily debated to widely recognized and documented problem. Studies have shown that the rate of global warming has risen to 0.2 ± 0.05 degrees centigrade per year (Hansen et al., 2006), a number which is alarming due to the capacity of ecosystems to withstand changes in temperatures of within 2 – 4.5 degrees centigrade before becoming unstable (IPCC, 2007). Climate change is intensified by the release of certain chemicals into the atmosphere, known as greenhouse gases. These gases include carbon dioxide, methane, and nitrous oxide, and are produced in high quantities by the burning of fossil fuels, such as coal and petroleum, fuels that are the source of the majority of the world's energy demand and are non-renewable relative to the rate at which they are consumed. With the energy demand only growing higher, it is imperative to develop clean, renewable sources of energy.

One source of renewable energy is bioethanol. Bioethanol is ethanol produced from fermentation of sugars extracted from plant matter. In the United States, the most common sources of bioethanol are corn or wheat (EPA, 2010), but it can also be produced using sugar cane, or sugar beets. One problem with the current production of bioethanol in the United States is that it relies heavily on staple food crops, such as corn and wheat, thus creating a conflict between use as food versus fuel. Corn is also very energy-intensive to convert into ethanol, producing roughly the same amount of energy as was necessary for conversion, as opposed to crops such as sugarcane which produce six times more energy than they require to process and convert their sugars to ethanol. Whereas the corn process is fueled by fossil fuels, the leftover plant material from the sugar cane, known as bagasse, can be dried and burned as fuel, cutting down on the majority of the process' dependence on fossil fuels.

Though sugar cane can't be easily grown in the United States, sugar beets thrive in a wide variety of climates, and can be grown from Texas to as far north as the Dakotas, including places where other staple food crops can't be easily grown. Sugar beets are a food crop, but imported sugar and an increased reliance on corn syrup have decreased the demand for domestic sugar in the United States, so basing the production of bioethanol on this crop wouldn't result in the conflict between food versus fuel. Though the leftover pulp from the sugar beet process cannot be burned to recover fuel costs, it can be dried and fed to livestock.

Currently, there are no plants in place producing bioethanol from sugar beets. A process was developed for the extraction and conversion of ethanol from sugar beets, using as a reference existing processes for producing sugar from sugar beets and ethanol from sugar cane. Aspen Plus and SuperPro Designer, computer modeling software designed for chemical processes, were used to simulate the process and develop a working model. The cost of building and operating the sugar beet to ethanol plant was determined using SuperPro Designer, CAPCOST, and commercial vendors. The process was then compared to a process designed by the USDA for producing bioethanol from corn, which was detailed in the report "Modeling the Process and Costs of Fuel Ethanol Production by the Corn Dry-Grind Process" (Kwiatkowski et. al., 2006).

2. Background

2.1 - Global Warming and the Need for Clean, Renewable Energy Sources

As the world enters the twenty-first century, global climate change has become a widely recognized and heavily documented problem. Studies have shown that the current rate of global warming has remained constant at 0.2 ± 0.05 degrees centigrade per decade since approximately 1975, having increased from a slower but more fluctuating rate (Hansen et al., 2006). This might seem like a relatively small change. However, recent studies conducted by the International Panel for Climate Change (IPCC) have found that it is likely that the earth's climates and ecosystems can only withstand changes in temperatures of within 2 – 4.5 degrees centigrade (IPCC, 2007).

Some climate change is natural, as the earth progresses through cycles of heating and cooling. Changes in temperature of Pacific Ocean spanning 1.35 million years have been estimated by a collaborative of the National Aeronautics and Space Administration, Goddard Institute for Space Studies, the Columbia University Earth Institute, and the Sigma Space Partners, Inc. Their research concludes that excluding human impact, the earth would currently be entering a cooling phase, which it is not. This agrees with the Synthesis Report generated in 2007 by the IPCC, which states that, considering only natural forces such as solar energy, the world would currently be cooling, not warming.

Global warming is already having concrete impacts on the environment, as exemplified by the melting of the permafrost along the coast of Alaska. As documented by the Alaska Climate Research Center of the University of Fairbanks, temperatures in Alaska have risen significantly since 1977, sometimes by as much as 1.9 degrees centigrade above the norm (Temperature Change in Alaska, 2009). Many indigenous people living along the coast have been forced to relocate as their towns and villages experience flooding, erosion, and the loss of homes as the ground melts into the Ocean (Ansari, 2009). As global warming continues, it could lead to further flooding and coastal erosion, as well as droughts, food and water shortages, and salinated groundwater due to sea level rise (IPCC, 2007).

Humanity has exacerbated global warming. The IPCC concluded in their Synthesis Report that it is very unlikely that global warming is due to natural causes alone (IPCC, 2007). Using data from ice core samples, the IPCC showed how levels of various greenhouse gases - carbon dioxide, methane, and nitrous oxide – have increased in the atmosphere over the past 10,000 years. The levels of greenhouse gases began to rise steadily after the Industrial Revolution in the mid 1700's, with a steep increase in the mid to late 1900's into the present. This was a period of time in which the economies of the industrialized countries were becoming increasingly dependent on fossil fuels such as coal and petroleum.

Over time, the energy demands of the world have grown as world populations increased and more countries began industrializing, leading to increased amounts of greenhouse gases being released. According to the U.S. Energy Information Administration, the global energy demand in 2006 was 472 quadrillion Btu, and is expected to rise 44% by 2015 (International Energy Outlook 2009, 2009). 100 quadrillion Btu's of the global energy demand were used in the United States (Guey-Lee, 2007).

With such a high demand for energy and the need to scale back on emissions of greenhouse gases to slow further increases in global temperatures, it is highly important to research and implement alternative forms of energy that do not rely on fossil fuels.

2.2 – Different Types of Non-Renewable Fossil Fuels

Fossil fuels are formed when organic material is subjected to heat and pressure over millions of years. Though all fossil fuels form and collect underground, the means through which they are harvested and used vary.

Figure 2.1 shows the relative consumption of coal, petroleum, and natural gas (methane), compared with nuclear power and renewable energy sources such as solar and hydroelectric (The Role of Renewable Energy Consumption, 2007). As can be seen, fossil fuels still provide 84% of the energy consumed by the United States. Besides producing a large amount of greenhouse gases, these energy sources are non-renewable and are being consumed at an unsustainable rate. At the current rates of consumption, it is estimated that the world's supply of coal, petroleum, and natural gas will eventually be depleted, though estimates on how soon this will be vary (IPCC, 2007).

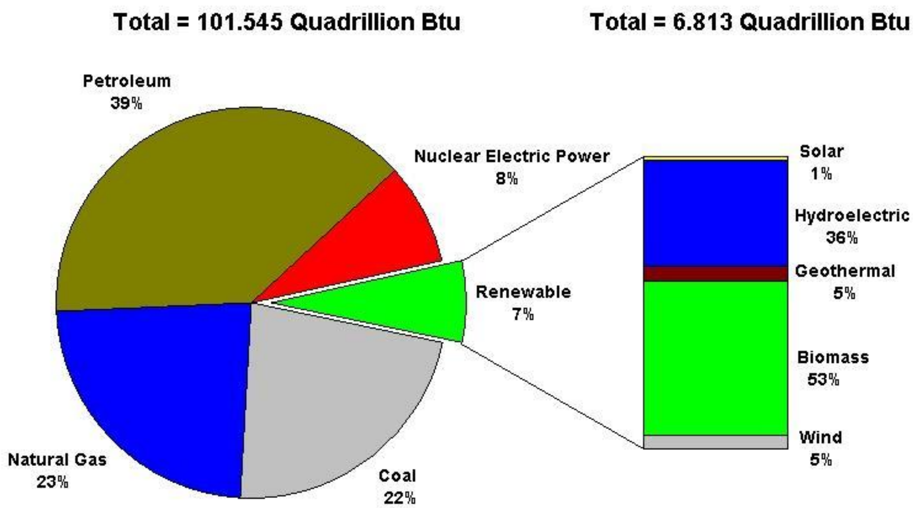

Figure 2.1: Consumption of Energy in United States in 2007 Broken Down by Source

2.2.1 - Coal

Coal forms in layers, and is harvested through mining. There are several types of mining used in the United States. In the first and oldest method, tunnels are dug under the earth to access the coal. Heavy machinery is used to remove the coal to minimize the amount of workers necessary in the mines. This method of mining is very hazardous to workers, due to cave-ins, accidents, and health risks from exposure to coal dust. The other type of mining is surface mining, in which the vegetation, soil, and rock is removed to give increased access to the coal bed. Surface mining involves fewer risks for workers, but is much more detrimental to the environment. A new form of surface mining, mountain top removal is especially harmful. In mountain top removal, the top layer of vegetation, soil, and rock is removed from the top of the mountain and dumped into a nearby valley, effectively destroying the ecosystem of both mountaintop and valley. With a flat surface created, the coal can then be removed.

Once the coal is removed, it can be processed in one of two ways. The first method is the pulverized coal method, in which the coal is ground and burned to create steam, producing large amounts of carbon dioxide as well as sulfur dioxide, nitrogen oxide and ash. The steam is then run through a turbine to generate electricity. Although scrubbers and filters can be used to cut down on the amount of pollutants being released into the atmosphere, they do not reduce the carbon dioxide emissions. For every tonne (1000 kg) of coal burned, 2.11 tonnes of CO2 are produced (Calculations and References, 2009).

The second, newer method is the integrated gasification combined cycle (IGCC), in which the coal is vaporized into syngas. The syngas is then run through a turbine to generate electricity and burned to harvest any remaining heat energy. Using IGCC reduces the level of pollutants released to the environment, and allows easier capture of the carbon dioxide gas.

Currently, coal still provides approximately 22 % of the total energy consumed in the United States (The Role of Renewable Energy Consumption, 2007), though numbers are declining slightly with the increased push toward renewable fuel sources. Although sources vary on the exact number of years, studies have shown that at current consumption rates, coal mines will be depleted in approximately 100 – 150 years.

2.2.2 - Petroleum and Natural Gas

Petroleum and natural gas collect in permeable rock, such as sandstone, often beneath the ocean floor. In order to harvest the petroleum, a well is drilled, after which the petroleum and natural gas are pumped out of the earth. Until the mid 1900's, the natural gas was burned off before the petroleum was recovered, as the pipelines that existed were inadequate to effectively transport natural gas.

The United States imports half of the petroleum it consumes, from Saudi Arabia, Mexico, Canada, Venezuela, and other countries. One of the main environmental hazards of petroleum is due to transportation. This includes both the transportation of the petroleum from the well to refineries and its use as in fuel, such as gasoline in automobiles. The combustion of oil releases nitrous oxides, carbon dioxide, carbon monoxide, sulfur oxides, and other pollutants into the atmosphere. As a point of reference, combustion of oil produces 3.16 tonnes of carbon dioxide per tonne of oil burned (Calculations and References, 2009). Petroleum is burned in the form of gasoline in automobiles, and must be shipped long distances to refineries in other countries. Ocean tanker accidents cause oil spills in the oceans, which coat the surface of the ocean with a film of oil and kill marine wildlife. A damaged ocean tanker can release enough oil to coat thousands of square miles of ocean. According to the National Research Council, almost one billion gallons of oil per year wind up in global waterways, only 13% of which is from tanker oil spills (How Oil Works, 2009).

Currently, petroleum provides 39% of the energy consumed by the United States (The Role of Renewable Energy Consumption, 2007). However, estimates show that globally there is only a 60 to 70 year supply of petroleum. To a country that has built its economy upon petroleum products and consumes over 25% of the world's oil each year, this could be extremely detrimental (How Oil Works, 2009).

Natural gas burns much cleaner than petroleum or coal. Natural gas produces very little sulfur dioxide and no ash upon combustion. It produces 43% fewer carbon pollutants than coal and 30% fewer than petroleum. However, it is a very strong greenhouse gas and still produces nitrous oxide when combusted. Methane gas traps heat 58 times more efficiently than carbon dioxide.

Currently, natural gas provides approximately 23% of the total energy consumed in the United States (The Role of Renewable Energy Consumption, 2007). This number is likely to increase as further developments in hydrogen fuel cells continue to be made, which utilizes natural gas to fuel the production of electricity at a more efficient rate than normal combustion. Estimates based on current technology show that natural gas reserves in the United States alone should last at least sixty years, though this could triple given future improvements in recovery.

2.3 – Different Types of Alternative Energy Sources

There are many energy sources that are cleaner than fossil fuels, including nuclear electric, solar, hydroelectric, geothermal, biofuels, and wind. Nuclear energy is produced by the splitting of uranium ions through fusion to release energy which can be harnessed in the production of steam. Out of all the other alternative energy sources discussed in this section, nuclear energy is the only source that is not considered a renewable source. This is because nuclear energy depends on the element uranium, which is not a renewable resource.

In order to convert uranium to fuel, first it is mined and concentrated into uranium oxide pellets to be transported to the nuclear power plant. Once in the nuclear power plant, the uranium oxide pellets are pelted with neutrons to split them and release heat and neutrons, which collide with other uranium oxide pellets to continue the reaction.

Nuclear energy is often considered “clean” because the reaction doesn’t produce many of the common air pollutants, such as carbon dioxide, sulfur dioxide, and nitrous oxide. However, the process of nuclear fission leaves behind a waste product composed of used uranium fuel, which remains radioactive and very toxic for thousands of years after its generation. At the moment, there is no permanent solution to the problem of what to do with the waste, and it is being stored at the nuclear facilities in which they are generated, though the U.S. Department of Energy wants to create a permanent nuclear waste disposal center in the Yucca Mountains of Nevada.

Solar energy is a clean, renewable resource which can be harvested in a variety of ways, whether through photovoltaic solar panels or solar thermal concentrators for the production of electricity or desiccant coolers and absorption chillers. Federal tax incentives are being offered to citizens who wish to implement solar energy to power their homes and businesses and photovoltaic cells are decreasing in price, making them more affordable for incorporation into homes. However, solar technology generally has a very low efficiency, such as 15% for a photovoltaic cell.

Wind energy is a clean, renewable resource which is harvested by wind turbines, which generate electricity in amounts proportional to the speed of the wind passing through the turbine. One downside to wind energy is that the wind turbines must be built to match the wind speed produced at the wind farm, which can vary seasonally or even daily. If the wind blows too slowly, the turbine won't work, but if the wind blows too fast, the turbine will shut down to avoid being damaged. Hydroelectric energy is similar to wind energy in that it takes advantage of the motion of a fluid to produce electricity.

Geothermal energy takes advantage of the heat produced by the earth. This can be done either by heating homes using water from hot springs, or by pumping air into the ground to be warmed (or cooled) and then pumping it back into the house to be used as a coolant or to be further warmed in a heating system. The second system takes advantage of the fact that the earth remains a constant 10 degrees centigrade a short depth below the ground. Geothermal energy is most efficient in areas that experience temperature extremes, such as the northern United States. However, geothermal technology can be difficult to install in existing buildings. Solar, wind, and geothermal energies do not produce carbon dioxide or other greenhouse gases.

Biological fuels consist of the oil from the fruit of plants such as soybeans, sunflowers, and palms, as well as ethanol produced from the breakdown of cellulosic materials such as corn stocks, sugarcane, or sugar beets. Although the burning of ethanol produces carbon dioxide, a greenhouse gas, biofuels are considered a carbon neutral source. This is because the carbon dioxide produced by the combustion of the ethanol is consumed by the plants which produce the cellulosic materials from which the ethanol is fermented (Harrison, 2003). While the oils are similar to diesel fuel and can be burned in compatible engines, fuel ethanol is often used as an additive to gasoline to cut down on harmful emissions and will be the focus of the next section.

2.4 – Ethanol as a Biofuel for Renewable Energy

Bioenergy is defined as any energy source that is derived from organic matter or biomass, which can then be used to produce heat and electricity, or used for transportation (United Nations, 2007). In particular, liquid biofuels such as ethanol, generally known as bioethanol, and biodiesel are the main producers of bioenergy, especially in transportation (United Nations, 2007). Ethanol used today in biofuels is typically made from fermenting and then distilling starch crops, such as corn or wheat (EPA, 2010). Bioethanol can be produced from any crop that produces fermentable sugars, which also includes sugar cane, sugar beets, and unused portions of other crops such as fruit waste. However, the use of these crops for ethanol threatens the use of land for food (United Nations, 2007). This could potentially be resolved in the future, as ethanol can also be produced by cellulosic biomass such as trees and grass (EPA, 2010), but lignin in their structures restricts access to usable materials for producing ethanol (United Nations, 2007).

A common blend of ethanol sold that can be used to fuel most vehicles is E10, also known as gasohol, which is a 10% mixture of ethanol in gasoline (EPA, 2010). A blend with a higher concentration of ethanol which is used frequently is E85, an 85% mixture of ethanol in gasoline. This mixture can only be used by flex-fuel vehicles (EPA, 2010). Flex-fuel vehicles are capable of operating using any mixture of ethanol and gasoline (EPA, 2010), not just E85, as anhydrous ethanol concentrations are capable of reaching closer to 100% as a fuel where it is not mixed with gasoline.

One benefit from using a higher concentration of ethanol in gasoline is that it is cheaper. For the year 2009, it was estimated that E85 cost on average $2.13/gallon, while regular gasoline cost $2.67/gallon (EPA, 2010). A disadvantage to this comparison is that ethanol has less energy than gasoline. E85 vehicles were estimated to get anywhere from 20-30% worse mileage than comparable gasoline powered vehicles (EPA, 2010). This would mean for a 30 MPG gasoline vehicle, a comparable flex-fuel vehicle running on E85 would get anywhere from 21-24 MPG. In terms of cost per mile, the 30 MPG vehicle would cost $0.089/mile and the comparable flex-fuel vehicle would cost anywhere from $0.089/mile to $0.101/mile. While the average price is lower, the profitability of using ethanol as a fuel is an issue that can be traced to the energy used to produce and distribute the ethanol from its original source, which is currently starch crops (EPA, 2010).

The production of ethanol from a crop usually involves five basic steps: pretreatment of the crops, sugar recovery, fermentation of the sugar to produce ethanol, distillation of the ethanol to higher concentrations, and drying of the ethanol. Once a crop is grown and harvested, it will need to be initially physically treated, such as being cleaned then sliced into thinner pieces or ground into finer material. Sugars are then recovered by various means from different crops, such as through the use of enzymes or simpler extraction methods. These sugars are then fermented by yeasts to produce ethanol. The ethanol is distilled through a series of columns to produce a higher concentration. However, the separation of ethanol from water becomes more difficult as the ethanol concentration rises due to azeotropic condition of the vapor-liquid equilibrium, limiting the distillation potential. This ethanol is then dried further to increase the overall concentration without being hindered by the vapor-liquid equilibrium.

2.5 – The Use of Food for Fuel

Aside from the requirements to produce biofuels from crops, other disadvantages occur during the decision making process on whether to use crops for bioethanol production. Initially, from the farmer's perspective, they have two options to choose from, either continue growing crops for food or re-orient their production to cater to use in ethanol production. Producing ethanol for a farmer could require funds for reconstruction and the building of a new facility. The potential for returns may only be marginal. However, there are subsidies that are available to help fund reorientation of farms toward producing crops for bioethanol (United Nations, 2007).

The leading producers of ethanol as a biofuel are the United States and Brazil (United Nations, 2007). Brazil has seen adequate success using sugar cane as a source for bioethanol (United Nations, 2007), as their climate is conducive for its growth (United Nations, 2007). The United States has more commonly used corn to produce bioethanol, as it requires a more temperate climate that is available in North America (United Nations, 2007). The issue that derives from this is that a competition between using corn as food or fuel has risen. Diverting land use from the production of food to the production of ethanol has the potential to limit the availability of food supplies (United Nations, 2007), making it a moral issue similar to the carbon emissions. The U.S. Food and Agricultural Organization estimated approximately 854 million individuals worldwide are already suffering from undernourishment (United Nations, 2007). An indicator of the limitation of the access of food is the market price for a crop (United Nations, 2007). Since corn has been used for ethanol production, the price for corn in the United States had approximately tripled from its low point in 2005 to its highest point in 2008 (Corn: Monthly Price Chart). The debate over food versus fuel pertaining to corn has resulted in a call for a more economically and socially sustainable crop to be used in the ethanol production process.

2.6 – Alternative Sources for Ethanol Production

Besides corn, there are a number of other options for plant material that can be used as a source for biofuels. All plants have pros and cons; however some are much more viable sources. An ideal plant would grow quickly, thrive in a broad range of environments, be moderately pest resistant, produce a high amount of fuel per unit area of plant growth, and the conversion of the plant material to a biofuel would be easy and low cost in both money and energy. Although all of these factors have to be considered, the last two are the first that need to be considered when determining the viability of a crop as a source of a biofuels.

Plants that are grown as food sources are generally easier to convert into a useful liquid biofuel. These plants contain more easily usable sugars or oils that can be fermented into alcohols or used as biodiesel, respectively. Other plants that are not used as foods, such as grasses, present the problem of being difficult to break down into a usable form. All plants have an irregular polymer called lignin that helps provide structural strength and flexibility in the cell walls. Lignin is very difficult to break down because of the irregularity of the molecule. In plants such as sugarcane, the sugar (sucrose) can be washed away from the fibrous lignin and then used to create useful biofuels. Similarly oil can be removed from algae or soy beans. Research is currently being done by many institutions into ways to break down lignin, but currently it is a huge hurdle for many potential energy crops. This leaves crops that contain more easily usable sugars and oils. Corn or corn oil can be used, however is does not provide very much energy compared to the energy needed to convert it into ethanol or biodiesel.

Essentially any vegetable oil can be used for the production of biodiesel, including soybean oil, rapeseed (canola) oil, olive oil, and sunflower oil among many others. These food sources again raise the issue of using food crops for fuel. However when processing soy beans, soy oil is produced as a byproduct. Although some soy oil is used for food, production far outpaces demand which makes soy oil a good feedstock for production of biodiesel. In 2008 approximately 60% of biodiesel produced in the US was from soy oil (Weber, 2009). Additionally vegetable oil used by restaurants can be cleaned and converted into biodiesel (Yokayo biofuels, 2005). Other non-food oils, notably oil from algae, are also possible sources. Algae needs a significant amount of sunlight to grow, which is somewhat limiting, but otherwise it is a very viable source. All biodiesel sources have issues of limited source materials, although crop production could potentially be increased. Also more restaurants could be made aware that their waste oil may be converted into biodiesel.

Crops used as a source of sugar for food can also be used for fermentation and conversion to alcohols for fuels. The most widely used alcohol for this purpose is ethanol, although butanol is also a potential alternative. Brazil uses significant amounts of ethanol as a liquid fuel for cars. In 2005, all gasoline sold in the country was required to contain at least 26% ethanol. Pure ethanol is also available widely for fuel in Brazil, and can be used by consumers with 'flex fuel' cars. Flex fuel cars are specially designed to run on either traditional gasoline or ethanol (Morgan, 2005). Sugar cane has a much higher energy return compared to the energy put into the conversion than corn. Corn is close to 1:1, although depending on the calculation it is both claimed to be somewhat higher and somewhat lower. Sugar cane is about 8:1, which makes it a considerably better fuel source as well as much more environmentally friendly (Rohter, 2006). While it takes a considerable amount of fossil fuel to create ethanol from corn, the fibrous waste product from sugar cane, called bagasse, can be burned to provide a large portion of the energy needed for conversion. Unfortunately sugar cane can only be grown in very limited areas of the southern United States, and even in those areas that it can be grown the growing conditions are not ideal. However, another crop that could be grown in the United States and potentially converted into biofuels is sugar beets.

2.7 – Sugar Beets as a Source of Ethanol

Sugar beets can grow in much of US, and are produced primarily in the Dakotas and Minnesota. Other states that produce large amounts of sugar beets are Idaho, California, Michigan, Nebraska, Wyoming, Montana, Colorado and Texas. Sugar beets can grow in a wide variety of soil types, from sandy to rich topsoil to clay rich soil (Cattanach et all, 1991). Sugar beets could be grown over a large portion of the United States because they grow well in a wide variety of climates. Sugar beets are used to produce sugar for food, beet sugar accounted for 58.8% of the sugar produced in the United States in 2005 (Jacobs, 2006).

Less expensive imported sugar from both sugar beets and sugar cane has decreased the selling price of sugar in the United States, and has decreased overall sugar production., The average American's yearly sugar intake has fallen from 102 lbs/yr in 1970 to 45 lbs/yr in 2002 (SIC 2063, 2005). The reason for this drastic drop is that corn syrup has taken over a significant amount of the market sugar once had:

> … corn sweeteners—corn syrup, dextrose, and high fructose corn syrup (HFCS)—accounted for more than 55 percent of U.S. sweetener consumption in the early 2000s, according to the Corn Refiners Association. In 2003, an estimated 535 million bushels of corn were used to make HFCS, which since 1980 has been the sweetener of choice for the major U.S. soft drink manufacturers. (SIC 2046, 2005)

An increase in demand for sugar to be fermented to create ethanol might help increase market prices by increasing demand for sugar. Because sugar beets can be grown over so much of the US, the demand for more beet sugar could be met by expanding the amount of land they are grown over.

Sugar beets take more energy to produce sugar from than sugar cane, because unlike sugar cane sugar beets do not have a byproduct like bagasse that can be burned for energy. However, sugar beets do have other byproducts that are used as animal feed. Both the tops of sugar beets and the pulp left after sugar is extracted from the beets are used as feed for cows and sheep. Molasses from sugar beets is also used as an additive to feed. Molasses from sugar beets does not have the same taste as molasses from sugar cane and is not generally consumed by humans. An average of 13 to 25 tons of sugar beets can be grown per acre of un-irrigated farm land. Irrigation increases yield by 15 to 30 percent (Cattanach et all, 1991).

It is relatively easy to extract sugar from sugar beets. The beets are chopped into thin chips called cossettes and washed in a counter current flow with water. The washed cossettes are pressed to remove remaining water and sugar. The sucrose rich water is then cleaned of impurities with lime and filtration and separated by drying and crystallization (Process description, 2000). If the sugar from beets was used to create ethanol instead of sugar, the cleaned sucrose rich water could be put through a fermentation step instead of drying and crystallization. The current sugar extraction process could be duplicated exactly up until this point. The sucrose water mixture might have to be diluted or concentrated to be a more ideal food for the yeast that would be used to ferment it. Next a separation step to recover ethanol, most likely by distillation, would finish the process of creating ethanol from sugar beets.

2.8 – Distillation Methods for Ethanol Production

2.8.1 – Traditional Distillation

In order to purify the ethanol to be used as fuel, distillation is often used. Distillation is the separation of two or more compounds based on their relative volatilities (Wankat, 2006). In traditional distillation, the mixture is fed into a distillation column and heated, causing part of the feed to vaporize. The column is packed with stages which allow vapor to pass up the column and liquid to travel downward. Ideally, at each stage the vapor passing up the column and liquid flowing down are achieving equilibrium. As the mixture progresses up the column, the more volatile component becomes concentrated in the vapor phase and is collected out the top of the column as distillate. The distillate is often condensed, and part of the distillate is recycled back into the tower as reflux. As the mixture progresses down the column, the less volatile component becomes concentrated in the liquid phase and is collected out the bottom of the column. Part of the bottoms are run through a reboiler and recycled as boil-up. The reflux and boil-up allow a greater separation to occur for a given number of stages. The more reflux is recycled to the column; the less stages are needed to achieve the separation. This process is shown in Figure 2.2, where component A is the less volatile component and component B is the more volatile component.

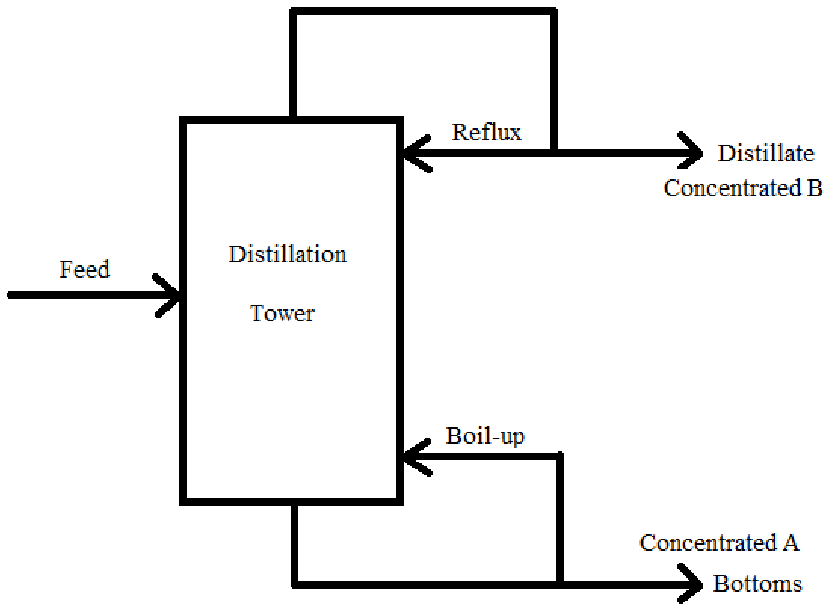

Figure 2.2: Schematic of Traditional Distillation

In the case of the ethanol-water separation, traditional distillation does not give a good enough separation due to the mixture's formation of an azeotrope at 95.6 mass percent ethanol (Clark, 2005). Azeotropic conditions form when, due to the properties of mixing, the boiling point of the mixture drops below the boiling point of both components. In order to be either burned in flex-fuel cars or used as an additive to gasoline, ethanol must be 99.6 percent pure (Mathewson, 1980). Therefore, it is necessary to consider methods of distillation that allow the breaking or bypassing of the azeotrope.

2.8.2 – Reactive Distillation

In reactive distillation, a compound is added to the mixture which reacts reversibly with one of the components of the mixture, giving a product with a differing relative volatility which does not form an azeotrope and can be separated through traditional distillation (Seader & Henley, 1998). Once the product has been separated, the reaction is driven in the opposite direction to allow the added compound to be separated out and recycled. Due to the necessity of separating the reactants, reactive distillation generally involves three distillation columns operated in series with corresponding recycle streams. Due to the necessity of finding an inexpensive compound that reacts selectively and reversibly with water and the presence of other, more feasible, methods of separation, reactive distillation is not used to separate ethanol and water.

2.8.3 – Pressure-Swing Distillation

The composition at which an azeotrope forms generally changes depending upon the pressure under which the system is operated. Pressure-swing distillation takes advantage of this change (Seader & Henley, 1998). Often utilizing a two-column system, the feed goes to a distillation tower at a higher or lower pressure than atmospheric, allowing the mixture to "jump" the azeotrope. The less volatile component leaves the bottoms as a nearly pure liquid. The distillate from the tower with a composition close to azeotropic for the operating pressure is then fed to a second column at a different pressure, generally atmospheric, which gives an azeotropic distillate stream of the less volatile component and a pure bottoms stream of the more volatile component. The distillate from the second tower is recycled back to the first column. This process is shown in Figure 2.3, where component A is the less volatile component and component B is the more volatile component.

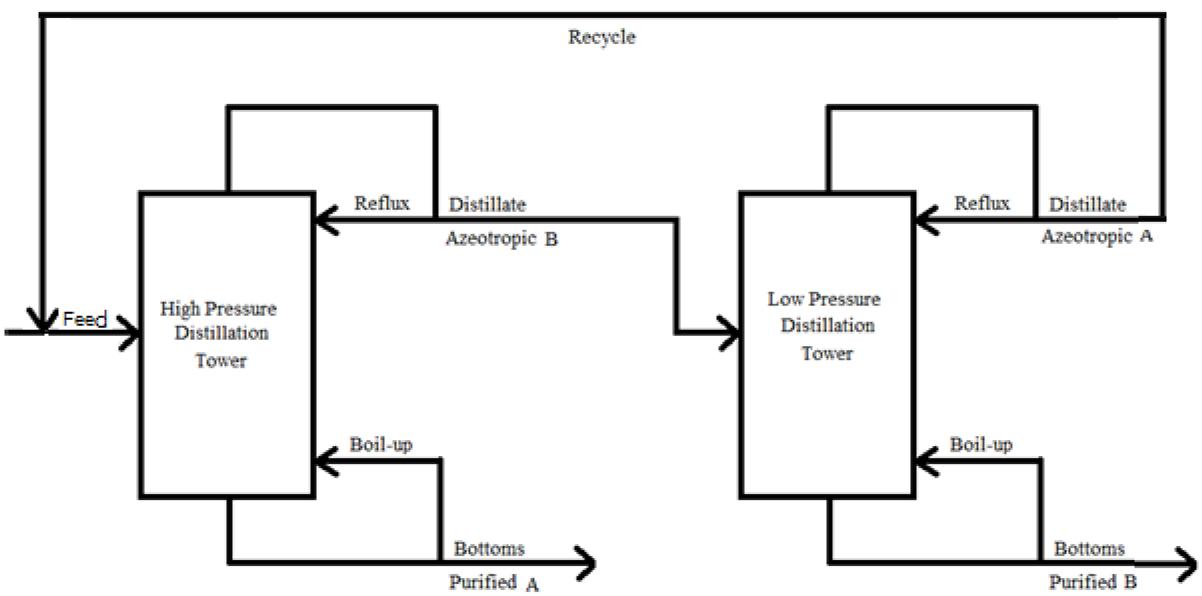

Figure 2.3: Schematic of Pressure-Swing Distillation

Though the azeotrope formed by a mixture of ethanol and water does change over different pressures, it is only changing by a few percentage points. Therefore, although the separation can be completed, it is very expensive. It generally requires very large distillation towers using materials and designs which can withstand conditions of high pressure or vacuum, and a large number of equilibrium stages to complete separations near azeotropic conditions. Due to these reasons, pressure-swing distillation is not the most common method of dehydrating ethanol.

2.8.4 – Extractive Distillation

Extractive Distillation is the most common method of dehydrating ethanol on an industrial scale. In extractive distillation, a solvent is added to the mixture to change the relative volatilities and prevent the formation of an azeotrope (Seader & Henley, 1998). The added solvent generally has a higher boiling point than either of the compounds in the mixture and leaves with the less volatile compound, allowing the more volatile component to be purified. Due to properties of mixing, the addition of the solvent can sometimes switch the volatilities of the compounds, so that the more volatile component leaves in the bottoms of the tower (Seader & Henley, 1998). Extractive distillation usually uses a three-column system. Similar to pressure-swing distillation, the first column separates the distillate to the azeotropic concentration. The distillate is then sent to the second tower, before which the solvent is added. The solvent then leaves with one of the compounds and is separated in the third tower and recycled. Ideally, almost all of the solvent will be recovered and recycled so that the cost of the solvent will be a one-time, start-up cost for the chemical plant. This process is shown in Figure 2.4, where component A is the less volatile component and component B is the more volatile component.

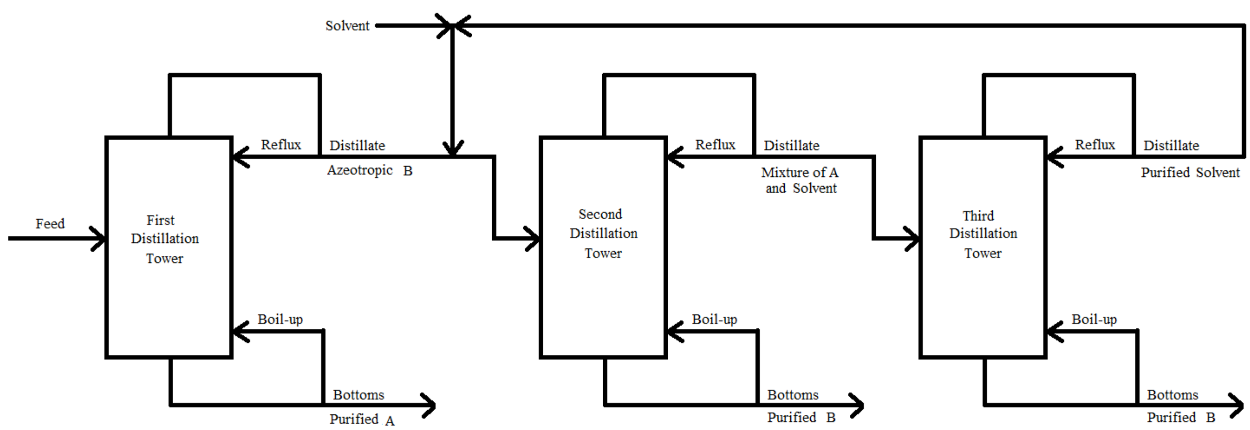

Figure 2.4: Schematic of Extractive Distillation

Solvents for extractive distillation must be non-reactive, easily separated from the components of the mixture, easily available, and inexpensive. Preferably, it should be noncorrosive, nontoxic, and have a low latent heat to aid vaporization in the reboiler of the distillation towers (Doherty & Caldarola, 1985). In the separation of ethanol and water, two solvents are available – benzene and trichloroethylene (Mathewson, 1980). Neither of these compounds are non-toxic, and care needs to be taken to avoid contamination with other parts of the system and the environment. However, in distilling ethanol for use as fuel, the alcohol needs to be denatured to prevent human consumption, so leaving trace amounts of the solvent in the anhydrous ethanol is not an issue.

2.8.5 – Molecular Sieve Adsorption Distillation

A different method utilized in the model designed by the USDA in "Modeling the Process and Costs of Fuel Ethanol Production by the Corn Dry-Grind Process" is molecular sieve adsorption distillation. This method utilizes a column packed with microporous beads. As the ethanol-water mixture flow through the beads, the smaller water molecules are trapped inside, while the larger ethanol molecules are able to flow by unimpeded. This allows the collection of nearly pure (99.6 mass percent) ethanol. The beads can then be heated to drive off the water and reused (Kwiatkowski, 2007). This process is described further in section 2.9.3.

2.8.6 – Salt Separations

Another method of separating ethanol from water is drying with salts. In this method, first distillation is used to bring the ethanol to the azeotropic concentration. To separate the mixture past the azeotrope, the ethanol-water mixture is filtered through dry salt. Since salt will absorb water, but not ethanol, nearly pure ethanol can be collected. The wet salt can then be heated to evaporate off the water and dry the salt, which can then be recycled (Mathewson, 1980).

2.9 – Computer Modeling and Assessment Programs for Chemical Processes

2.9.1 – Aspen Plus

Aspen Plus is a modeling software package used for a variety of process types in chemical engineering. Aspen Plus includes many basic units, such as distillation towers, drums, pumps, and heat exchangers. It does not include many specialty unit operations that are used for particular industries, such as conveyor belts or equipment necessary for the pretreatment of solids. Aspen Plus includes a wide variety of thermodynamic packages, which gives the user many options as to which will most accurately model any process. Aspen can accurately model ideal and non-ideal mixtures as it has a very strong set of available thermodynamic models included. Aspen Plus also includes large databases of information about many chemical components.

Aspen Plus is accurate modeling software to use for processing plants that contain straightforward unit operations without many industry specific pieces of equipment. Aspen Plus can be used to simulate sections of plants modeled with other software in areas where thermodynamically complicated processes take place. An example of this is an azeotropic system, where predicting an ideal system would give wildly inaccurate results.

Aspen Plus can be a difficult program to learn and work with because it contains so much information. This makes the program complicated, but also an incredibly valuable resource. When given inappropriate information about a system Aspen Plus will sometimes give nonsensical results, but more often it will simply give an error message. This often catches user mistakes, which is helpful, if sometimes frustrating. Considerable knowledge of thermodynamics, chemical equipment and processes is necessary to successfully work with Aspen Plus, or the user will end up with many errors and little information.

2.9.2 – SuperPro Designer

SuperPro Designer is also a modeling software package. It is designed specifically for processes including biological components. SuperPro Designer includes unit operations specific to biological operations, such as fermentors and strainers that Aspen lacks. However, SuperPro Designer has significantly less rigorous thermodynamic packages and far less information about components in databases. This often forces the user to research outside of the program in order to gather enough information for a successful simulation.

SuperPro Designer is a much more straightforward program to learn and use than Aspen Plus. However, much more care is required of the user to successfully run a simulation in SuperPro Designer because it will give physically impossible results. SuperPro Designer also includes a costing feature that certain versions of Aspen Plus lack. It is very helpful to get an approximate cost of equipment and plant operating cost as the simulation is completed.

2.9.3 – SuperPro Designer Model Provided by the USDA

In the paper, "Modeling the Process and Costs of Fuel Ethanol Production by the Corn Dry-Grind Process", the USDA modeled the generation of ethanol through the fermentation of corn (Kwiatkowski, 2006). The process was modeled using the commercially available chemical engineering design software SuperPro Designer. The project was a continuation of one previously completed using Aspen Plus for design and Microsoft Excel for compilation and economic analysis. Though the model was designed for corn, it was done so in a way that the model could be adapted to other grains. The benefit that was found in using SuperPro Designer was that they could design and analyze the process using the same software. For specific material and energy balances and cost analysis see Appendices F – H.

2.9.4 – CAPCOST

While SuperPro Designer has the ability to provide a rough estimate for the costs in a process, they are not necessarily accurate, as the program is estimating both the cost and the process outputs simultaneously, which has been stated to be rough estimates as well. Generally, processes that require only a few units are easy to estimate by hand, but a computer aided program would be beneficial in estimating larger processes. For estimating the cost of a larger process, such as the sugar beet to ethanol process, the program CAPCOST, in addition to the modeling program Aspen Plus to find the equipment data necessary, could be used in order to estimate the cost of the process.

CAPCOST takes data given to it, either taken from a proposed process or one modeled in a program such as Aspen Plus, and will provide a detailed capital cost estimate for the plant. CAPCOST comes bundled with Analysis, Synthesis, and Design of Chemical Processes (Turton et al., 2009) in the form of a program created with Microsoft Excel. The program comes with templates to model several common units in a process, such as centrifuges, conveyors, heat exchangers, towers, vessels, and more. Units that it does not have a template for, it can be modeled using another template under certain conditions, such as modeling a reactor as a packed tower.

The information needed by CAPCOST is dependent on the input of the units. For example, CAPCOST will need to know various materials of construction for design, operating pressure, and types of varying units, as well as specific information dependent on unit, such as heat exchange area for a heat exchanger, number of stages, diameter, and height for a distillation column, and size for a vessel. When the cost index is provided, such as the Chemical Engineering Plant Cost Index (CEPCI), CAPCOST will then provide a series of costs for each unit (Turton et. al., 2009).

3. Methodology

The goal of the project was to design a process to produce ethanol using sugar beets as a feedstock and compare the process and corresponding economic analysis with that of using corn. This section describes the gathering of information and the construction of the SuperPro Designer model used to simulate the process.

In the following sections, when using SuperPro Designer to compute large recycle streams, the streams were not connected but iteratively developed so that an inlet stream entered the process in equal composition to the exiting recycle stream. SuperPro Designer is unable to deal with large recycle streams.

3.1 – Pretreatment

3.1.1-Beet Slicing

Sugar beets are bulb-shaped white root vegetables weighing approximately two kilograms each. In order to remove sucrose from the beets, they first must be sliced into much smaller pieces. These pieces are called cossettes, and are shaped approximately like shoe-string French fries. A slicer was used to model this step in SuperPro Designer (Asadi, 2007). Since the flows into and out of the slicer were for the purposes of the model exactly the same as the flow into the extractor, the slicer was not added to the full model, but designed and a cost analysis performed in a separate flow sheet.

3.1.2– Sucrose Extraction

Once the beets are sliced the next step in transforming them into ethanol is extracting the sucrose. In this step, cossettes are washed with a water stream in a counter-current arrangement to extract the sugar. Unlike processes that transform corn into ethanol, sugar beets do not require an enzymatic treatment to produce simpler sugars from starches. This makes the process considerably simpler. The extraction process was modeled as shown in Figure 3.1 (Bogliolo et al., 1996).

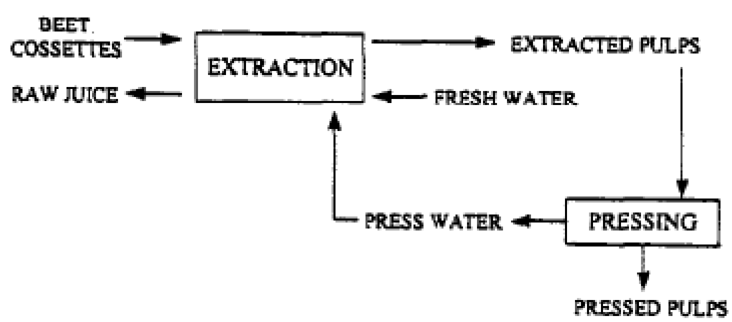

Figure 3.1: Outline of Sugar Extraction and Press Water Production in a Sugar Beet Factory (Bogliolo et al., 1996)

In this section of the process, cossettes, fresh water, and press water are the inlets to the extractor and extracted pulps and raw juice are the outlets. To make the model simpler the beets were represented as only usable sucrose, water, and inert solids. The solid pulp was removed and fed to a press, producing press water that was recycled back into the extractor with fresh water and pressed pulp. In this model the pressed pulp was regarded as waste stream, as it is not used anymore in the process. Often the pulp is sold as a food for animals such as cows, horses, and sheep. In SuperPro Designer, the water entering the extractor was first mixed with the cossettes stream and this combined stream was set to enter the unit. The main product leaving the extractor was the raw juice, which contains sucrose as well as a small amount of other non-usable sugars, ions, and inert compounds that have the potential to be detrimental to the process. The composition of the raw juice is shown in Table 3.1 (Ogbonna et al., 2001)

Table 3.1: Composition of Raw Sugar Beet Juice (Ogbonna et al., 2001)

Components	Weight Percent
Water	65.62%
Solids	17.30%
Sucrose	16.50%
Other Sugars	0.24%
Impurities	0.34%

The extractor was modeled in SuperPro Designer as a mixer/settler liquid extractor. The press was modeled using a two-way component splitter. For the extractor and press, the data from above was used to set the raw juice composition. Additionally information about pulp content was needed to complete this section of the simulation. It was found that the incoming beets are approximately 25% inert solids, and pressed cossettes are about 20% solids (Koppar and Pullammanappallil, 2008). Additionally about 1% of the sugar entering the plant is lost to the cossettes (Asadi, 2007). The cost for the extractor from SuperPro Designer was checked against an estimated cost from the extractor vendor Braunschweigische Maschinenbauanstalt AG (BMA). The value given by BMA was used in the finished model.

3.1.3 – Filtration

Impurities in the sucrose are removed by treating the raw juice with lime. In the filtration process calcium hydroxide is added to the beats. Then carbon dioxide is bubbled through the mixture and, as shown in Equation 3.1, calcium carbonate is formed. The calcium carbonate is allowed to precipitate out in a clarifier and takes with it a vast majority of the impurities found in the raw juice. This carbonation and clarification process may be repeated multiple times, though most commonly it is repeated twice. To simulate the formation of calcium carbonate, a continuous stirred tank reactor was used in order to model the conversion of calcium hydroxide to calcium carbonate (smbsc.com, 2010 and Asadi, 2007). A brief outline of this procedure is shown in Figure 3.2.

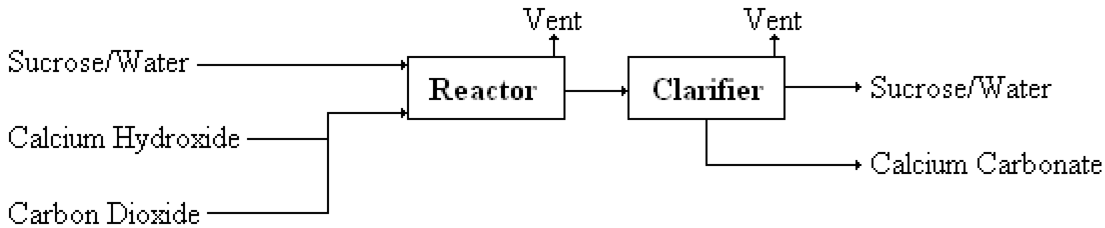

Figure 3.2: The Basic Steps for One Stage of the Filtration Process

The following reaction takes place in the modeled reactor:

$$Ca\ (OH)_2 + CO_2 \rightarrow CaCO_3 + H_2O \qquad \text{Equation 3.1}$$

This process was modeled with two carbonation and clarification steps for completed removal of calcium hydroxide. The modeled reactors in each step were set to consume 95% of the calcium hydroxide. For the first clarification step, 90% of the calcium carbonate was removed using specifications for the particle diameter to be removed. Along with calcium carbonate, some water and some of the remaining calcium hydroxide was removed. The mixture of sucrose, water, carbon dioxide, calcium hydroxide, and calcium carbonate was fed into the second carbonation step, where more carbon dioxide was bubbled through the mixture to produce more calcium carbonate. The second clarifier was set up to remove calcium carbonate, and any remaining calcium hydroxide. Some water and about 1% of the sucrose is also removed in this step (Asadi, 2007). All components other than sucrose and water were assumed to be removed completely in the filtration step. The final product stream contained a mixture exclusively of water and sucrose to be fed to the fermentor.

3.2 – Fermentor and Centrifuge

3.2.1 Yeast and Conversion

An important part of this process is the fermentation step. A biological process is used to convert sucrose to ethanol in a fermentor. The organism used widely in industrial production of ethanol is Saccharomyces cerevisiae, which is a species of yeast. S. cerevisiae is not the only organism that can perform a sugar to ethanol conversion, but to date it is favored for large-scale ethanol production for several reasons. In anaerobic conditions, S. cerevisiae is found to convert approximately 95% of sucrose into ethanol with a retention time of approximately 20 hours with the appropriate fermentor volume (Ogbonna et all, 2000). In industrial applications it is also found to convert 90-93% percent of glucose to ethanol. S. cerevisiae produces byproducts in low levels; the primary byproduct produced is glycerol at about 1% (w/v). Byproduct production can be limited by controlling fermentation conditions (Bai et all, 2008).

Another organism that has been investigated for its ability to produce ethanol from sugars is Zymomonas mobilis. Although Z. mobilis will convert more sugar to ethanol than S. cerevisiae (up to 97% conversion of glucose), it produces very limiting byproducts when fed with sucrose or fructose. Additionally Z. mobilis will only consume D-glucose, D-fructose and sucrose. As feedstocks for industrial ethanol production are either sucrose or sacrificed starches, which produce a variety of sugars, this is extremely undesirable. Additionally many ethanol production processes sell solid byproducts as animal feed, and Z. mobilis, although technically safe, is not acceptable in such feed while S. cerevisiae is acceptable (Bai et all, 2008).

S. cerevisiae produces less ethanol once the ethanol concentration reaches 10%, by weight, and the cells will die once the concentration reaches 18% by weight (Liu and Qureshi, 2009). The conversion of sucrose to ethanol is approximately 95%, although it can be as high as 98% (Ogbonna, 2001). The simulation for this project used 95% conversion, as a conservative number. This means that for each gram of sucrose that enters the fermentor, approximately 0.51 grams of ethanol will be produced (Bai et all, 2008). The reaction that produces ethanol from sucrose and water is shown in Equation 3.2.

$$C_{12}H_{22}O_{11} + H_2O = 4C_2H_5OH + 4CO_2 \qquad \text{Equation 3.2}$$

A high concentration of ethanol will kill essentially any yeast, so the conversion and stoichiometry of the sucrose to ethanol reaction were considered when determining the concentration of the sucrose in the fermentor feed. The fermentor feed was modeled as essentially only water and sucrose. The concentration of sucrose in the feed should be no more than 17.9% by weight in order to keep the exiting ethanol concentration below 10% by weight. In this simulation the concentration of sucrose in the fermentor feed is 14.8% sucrose by weight.

It is ideal for the incoming concentration of sucrose to be close to the maximum while still maintaining a reasonable margin. The difference between the maximum and the amount actually used is important so that the yeast will consume the maximum amount of sucrose without being overwhelmed by ethanol when small fluctuations in concentration occur. However, this difference should not be too large so that the following separation step will not have an unnecessarily high energy cost.

3.2.2 Fermentor Design

Important factors to design of the process in and around the fermentation step that were supplied to SuperPro Designer were: conversion of sucrose to ethanol, feed rate of yeast, and time for conversion. SuperPro Designer uses a heuristic for the maximum size of a single fermentor along with the calculated necessary volume for fermentation to determine the size and number of fermentors required. The fermentors were designed to be large enough to handle the flow of water and sucrose for the necessary retention time of approximately 20 hours. Yeast was fed to the reactor at a rate of 0.25 kg/L of liquid (Mathewson, 1980).

3.2.3 Centrifuge

Yeast must be separated from the fermentor outlet stream so that it does not block or damage equipment later in the process. The easiest way to remove yeast it to take advantage of the very significant difference in size between the tiny molecules of water, ethanol, and sucrose, and the much larger yeast cells. It was determined that the best way to perform this type of separation was with a disk stack centrifuge (Green and Perry, 2008). A disk stack centrifuge was modeled in SuperPro Designer. The only data needed for this equipment to be modeled was the size of a S. cerevisiae cell, which was found to be a minimum of 5 micrometers (Bai et all, 2008).

3.3 – Extractive Distillation

In order to design a separation using extractive distillation, first the solvent must be chosen. For an ethanol-water separation, the solvents benzene and trichloroethylene (TCE) were suggested (Mathewson, 1980). In order to determine the best solvent, two separate simulations were run, first using benzene as the solvent, next using TCE. These simulations were completed in Aspen Plus as shown in the process flow diagram below.

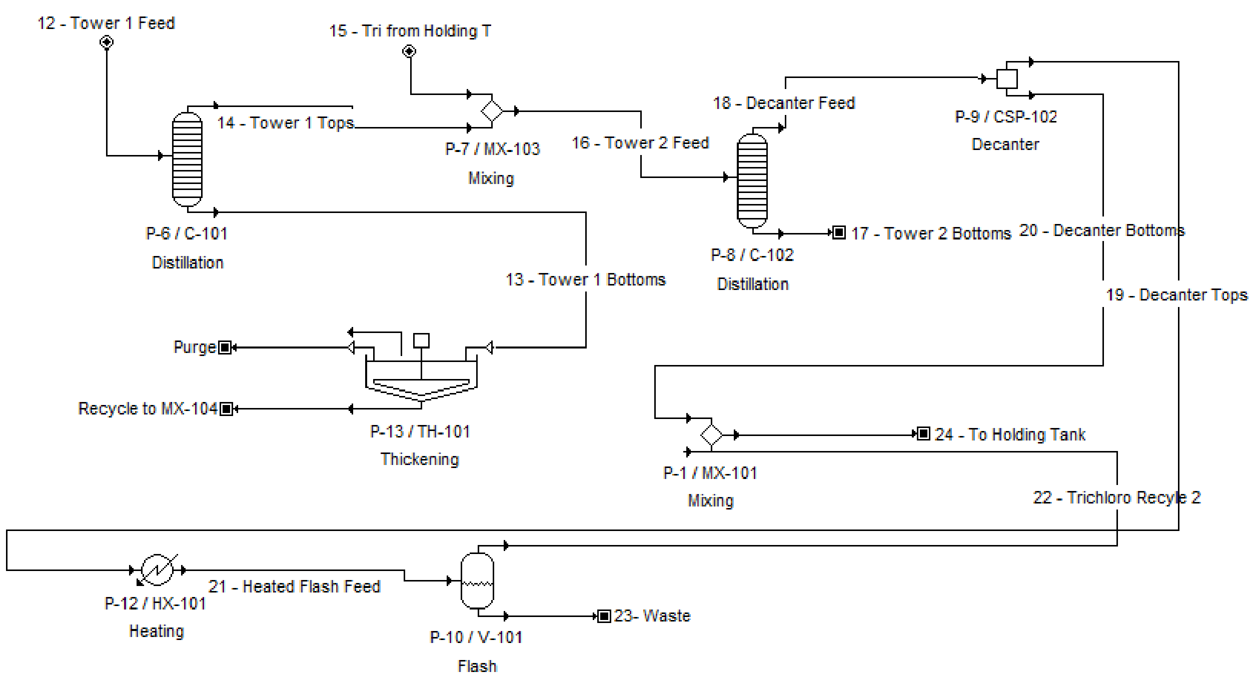

Figure 3.3: Extractive Distillation Process

This process utilizes a pair of distillation towers in series with the ethanol product leaving the bottom of the second distillation column. The first column provides a crude separation of water from ethanol. Ethanol with some water is the distillate from this column, and water with some residual sucrose is the bottoms. A thickening step was added to the bottom of the first tower to allow the recovery of unreacted sucrose that is removed with the water. Since this stream contains too much water to be recycled back to the pre-treatment steps, this thickening step was necessary to drive off some of the water to avoid flooding the extractor. Some of the ethanol is purged with the water in this step.

TCE was added to the more concentrated ethanol stream before the second column. This column produces very high purity ethanol out the bottom, and a mixture of water and TCE as the distillate. Once a majority of the ethanol was purified and removed from the system, the TCE was separated from the water so that the water could be released as waste. Liquid TCE and water form a two-phase liquid-liquid mixture, allowing the majority of the TCE to be removed through the use of a gravity decanter, which separates the two phases based upon the difference in densities. As TCE is denser than water, it is removed from the bottoms of the decanter. There is still enough TCE remaining in the tops of the decanter that the stream does not pass the emissions standards of the United States Environmental Protection Agency. The legal limits set by the EPA for the discharge of TCE and benzene with wastewater is 0.005 mg/L, or 5 parts per billion (EPA, 2010). The waste stream is dilute enough that it doesn't warrant the addition of a third distillation tower. Therefore, the stream is sent through a flash tank. Almost all of the TCE is removed, and a very small amount is lost in the wastewater stream leaving the bottoms of the flash unit.

Once the simulations were running without errors, the reflux ratios and number of stages of the distillation towers were checked to ensure the numbers received were reasonable. The input to the towers was adjusted when necessary to balance the operating cost of the reflux ratio against the equipment cost of the stages. The total and component mass flow rates of each stream were checked to ensure the separations had proceeded as needed.

Once the results of the simulation were determined feasible, the system was transferred into SuperPro Designer and the simulation was refined until the numbers matched those from Aspen Plus to take advantage of Aspen's superior thermodynamic packages. The dimensional specifications of the equipment were also examined.

Next the recycle streams were incorporated into the system. Water, ethanol, and TCE were recycled whenever possible. No recycle stream from after the addition of TCE was routed back to the pre-treatment, fermentation, or filtration steps to avoid killing yeast due to contamination.

4. Results and Discussion

The following sections describe the results gained through the modeling and analysis of the sugar beets to ethanol process. The detailed SuperPro Designer and Aspen reports can be found in the Appendices, as well as a complete process flow diagram. All of the equipment costs listed in this section were taken from SuperPro Designer cost estimations, with the exception of the extraction unit, where a vender cost was used. All costs from SuperPro Designer were checked with Capcost and were found to be in good agreement.

4.1 – Pretreatment

4.1.1 – Sucrose Extraction

To produce the desired amount of ethanol for the sugar beets process, the amount of cossettes entering the system was approximately 38,600 kg/h. The vendor Braunschweigische Maschinenbauanstalt AG estimated a cost of $5,424,000 for an extractor capable of handling that many beets. This was found to be greater than the cost given by SuperPro Designer, but it was decided that the cost from the vendor would be more accurate for a real plant. Though this is a large cost, the extractor handles a large volume of material and contains many moving parts. The beets are carried upward on a rotating spiral to allow maximum contact with the water passing in the opposite direction while preventing the beets from being swept away with the raw juice. The extractor must be sturdy enough to withstand the flows of beets and water without malfunction.

Figure 4.1 shows this section of the process as modeled in SuperPro Designer. Table 4.1 shows a corresponding stream summary.

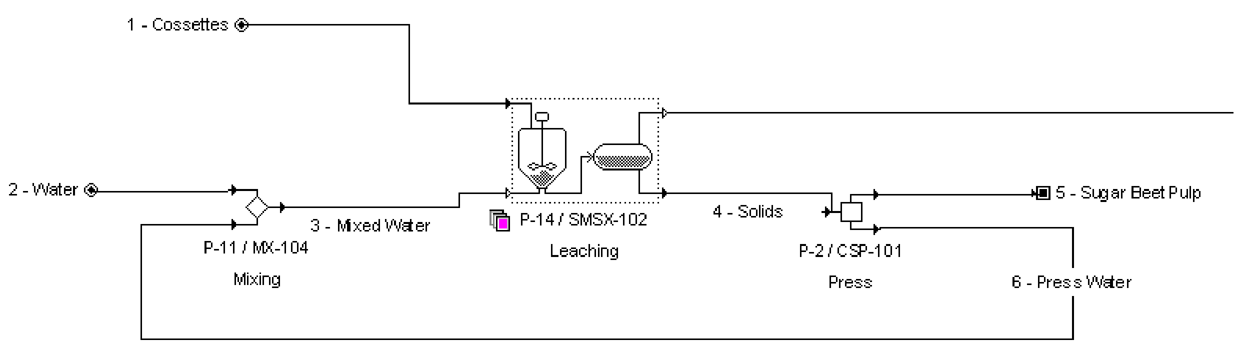

Figure 4.1: The Sucrose Extraction Section Modeled in SuperPro Designer

Table 4.1: Summary of Streams Associated with the Sucrose Extraction Process

	Inlet		Recycle	Outlet	
	Cossette Feed	Water from Recycle Later in Process	Press Water	Sugar Beet Pulp	Sucrose Product
	[kg/hr]	[kg/hr]	[kg/hr]	[kg/hr]	[kg/hr]
Sucrose	30,200	-	1,510	600	29,600
Water	129,000	71,400	57,000	24,800	176,000
Inert Solids	8,400	-	-	8,400	-

Initially, 7.0% of the sucrose entering the system through the extractor is lost with the inert solids. However, by adding the pulp press and press water recovery, more sucrose and utility water was conserved. A total of 5.0% of the sucrose was conserved via this recycle process, so that only 2.0% of the available sucrose is lost via extraction. Recovering sucrose is essential to maximize ethanol production potential. However, the 2.0% loss was required to remove all of the pulp, which cannot be present downstream.

4.1.2 – Filtration

From the extractor, the sucrose/water mixture was fed into the filtration step. The process was broken down into two nearly identical stages consisting of similar steps. This process was modeled in SuperPro Designer as is shown in Figure 4.2. In SuperPro Designer, the reactors were modeled using the stoichiometric continuous stirred tank reactor units and the clarifiers were modeled using the clarification units. A detailed stream summary for Stage 1 is shown in Table 4.2 and a detailed stream summary for Stage 2 is shown in Table 4.3.

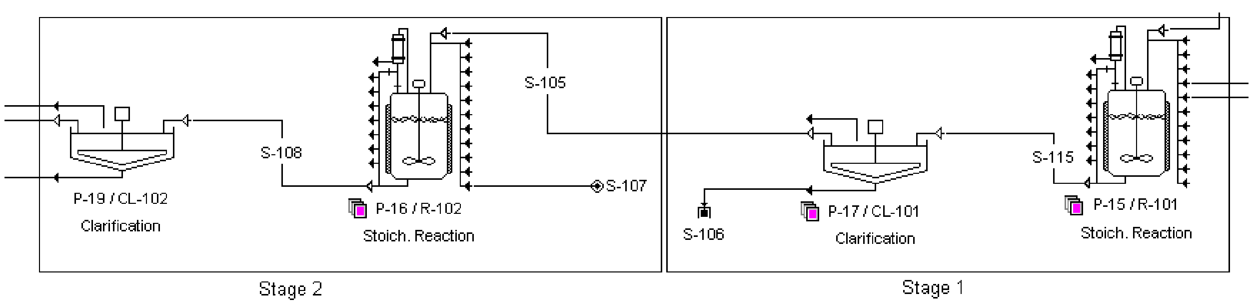

Figure 4.2: The Filtration Section Modeled in SuperPro Designer

Table 4.2: Summary of Streams Associated with Stage 1 in the Filtration Process

	Feed into Reactor R-101		Feed into Clarifier CL-101	Clarifier CL-101 Products	
	Sucrose Inlet Stream	**Reactant Component Feeds**	**Reactor R-101 Product**	**Separated Waste**	**Reactor R-102 Feed**
	[kg/hr]	**[kg/hr]**	**[kg/hr]**	**[kg/hr]**	**[kg/hr]**
Sucrose	29,600	-	29,600	148	29,500
Water	176,000	3,360	186,000	46,141	138,000
Calcium Hydroxide	-	7,840	168	16.8	151
Carbon Dioxide	-	2,500	604	-	604
Calcium Carbonate	-	-	4,310	3,880	431

Table 4.3: Summary of Streams Associated with Stage 2 in the Filtration Process

	Feed into Reactor R-102		Feed into Clarifier CL-102	Clarifier CL-102 Products	
	CL-101 Product	**Reactant Component Feeds**	**Reactor R-102 Product**	**Separated Waste**	**Sucrose Product**
	[kg/hr]	**[kg/hr]**	**[kg/hr]**	**[kg/hr]**	**[kg/hr]**
Sucrose	29,500	-	29,500	147	29,300
Water	138,000	-	138,000	34,600	104,000
Calcium Hydroxide	151	-	7.6	7.6	-
Carbon Dioxide	604	210	729	729	-
Calcium Carbonate	431	-	625	625	-

More calcium hydroxide could be consumed by adding more stages. However, only 0.3% of the total amount entering the first reactor was left unreacted after the two stages. Therefore it would be unprofitable to try to reduce purchase costs for calcium hydroxide by purchasing another set of equipment.

More sucrose and water were lost during filtration. From the previous stage of the process, approximately 75,500 kg/hr of water was lost, while 3,360 kg/hr was recycled back into the filtration steps as well as the first reactor. It might be a viable option to look further into possibly recovering more water for recycling into other parts of the overall process.

Sucrose loss was not as substantial in filtration as it was in extraction, as filtration loses only 48.9% as much sucrose as earlier. The total amount of sucrose lost from the amount originally available from the sugar beets was about 1.0%, 0.5% for each stage of filtration, bringing the total amount of sucrose lost over the entire process to 3.0% so far. Therefore, 97.0% of the sucrose from sugar beets entering the system was conserved and able to be fermented in the next step.

Purchase costs for each filtration unit in this section was taken directly from the SuperPro Designer economic evaluation model rather than private vendors. A summary of costs is shown in Table 4.4.

Table 4.4: A Summary of Costs for the Filtration Section

	Stage 1		**Stage 2**	
	Reactor R-101	**Clarifier CL-101**	**Reactor R-102**	**Clarifier CL-102**
Cost	$1,059,000	$1,874,000	$656,000	$344,000

The cost for the first clarifier is 5.44 times the cost of the second one; however it has to remove 6.2 times the amount of calcium carbonate from the system. One could look into removing the second stage to alleviate this fee, but the cost of the first clarifier would grow exponentially and would not be worth the investment. The second reactor step was essential in order to react the necessary lime within the system and fully clean the sucrose before fermentation.

4.2 – Fermentation

4.2.1 Yeast, Conversion, and Fermentor Design

It was found that the species of yeast best suited for this application is S. cerevisiae. It is also the species of yeast that is most commonly used in biomass to ethanol processes. Equation 4.1 shows the reaction of sucrose to ethanol that occurs in this process.

$$C_{12}H_{22}O_{11} + H_2O = 4C_2H_5OH + 4CO_2 \quad \text{Equation 4.1}$$

One large fermentor was required to handle the total flow of water and sucrose into the fermentation step as modeled by SuperPro Designer. The fermentor was found to be 22.9 meters tall with a diameter of 15.3 meters. The carbon dioxide produced will be vented out of the fermentor, and some of it will be recycled to the carbonation step. The fluid leaving the fermentor will be 8.1% ethanol by weight and 0.8% sucrose by weight remaining. Additionally it contains yeast that will be removed by centrifuging. A summary of streams flows is shown in Table 4.5. The total cost of the fermentor will be 1.83 million USD. At this point in the process the ethanol has been created and now must be separated from the water.

Table 4.5: Stream Summary for the Fermentation Stage

Component and Total Flows (kg/hr)	**Clarified Sucrose and Water**	**Dilution Water**	**Yeast**	**Total Flow to Fermentor**	**Vent**	**Fermentation Products**
Ethanol	-	-	-	-	150	14,800
Water	104,800	64,700	-	168,500	-	167,100
Sucrose	29,300	-	-	-	-	1,470
Yeast	-	-	0.84	0.84	-	0.84
Carbon Dioxide	-	-	-	-	14,300	-
Total	133,100	64,700	0.84	168,500	14,450	183,400

4.2.2 Centrifuge

It was found that a single centrifuge is needed to separate the yeast from the fermentor product stream. Along with all the yeast, about 5 kilograms per hour of water is also removed from the fermentation products stream, along with traces of ethanol and sucrose. The centrifuge is a disk stack centrifuge with required sigma value of 160,000 square meters. A sigma value represents the size of a gravitational settler that would perform an equal separation to the centrifuge. Because many variables affect the separation provided by a centrifuge it is common to give a sigma value instead of a specific size (Green and Perry, 2008). The cost of this centrifuge is $43,800.

4.3 – Distillation

4.3.1 Extractive Distillation

The solvent chosen for use in the extractive distillation steps was trichloroethylene. Although both benzene and trichloroethylene perform similarly throughout the actual distillation steps, trichloroethylene is more easily separated from water in the post-treatment steps. Both benzene and water form a two-phase liquid-liquid mixture with water. This allows the majority of the solvent to be decanted off. A flash unit can be used to remove enough of the remaining trichloroethylene to meet EPA standards. However, the flash unit is not able to remove enough of the benzene to be within legal limits. Either a distillation column or multiple flash units must be purchased, driving up equipment costs, or the water stream needs to be further diluted, resulting in a higher operating cost.

The fermentation steps yielded 14,850 kg/hr of ethanol in a mixture with 1,466 kg/hr of unreacted sucrose and 167,000 kg/hr of water at 30 degrees Centigrade and one atmosphere of pressure. Through the separation process, 14,700 kg/hr of anhydrous ethanol was obtained at a total equipment cost of $769,600. The tables 4.7 through 4.9 summarize the main flows into and out of the each piece of equipment.

Table 4.6: Stream Summary for Tower 1 and Thickener

Component and Total Flows (kg/hr)	**Tower 1 Feed**	**Tower 1 Tops**	**Tower 1 Bottoms**	**Purge**	**Recycle to pre-treatment**
Ethanol	14,850	17,850	1.5	1.1	0.4
Water	167,000	4,176	162,900	116,500	46,100
Sucrose	1,466	-	1,466	-	1,466
Total	183,300	22,030	164,400	116,500	47,570

Table 4.7: Stream Summary for Tower 2

Component and Total Flows (kg/hr)	**TCE from Holding Tank**	**Tower 2 Feed**	**Decanter Feed**	**Tower 2 Bottoms Ethanol Product**
Ethanol	620	15,500	774	14,700
Water	628	4,800	4,800	0.48
Trichloroethylene	84,000	84,000	84,000	-
Total	85,250	104,300	89,570	14,700

Table 4.8: Stream Summary for Decanter and Flash Unit

Component and Total Flows (kg/hr)	**Decanter Tops**	**Decanter Bottoms**	**Waste**	**Trichloroethylene Recycle 2**	**To Holding Tank**
Ethanol	416	357	159	260	614
Water	4,768	36	4,169	598	634
Trichloroethylene	8.4	83,992	.00008	8.3992	83,999
Total	5,192	84,390	4,328	866.4	85,350

The first distillation tower was actually modeled as two towers run in parallel. They had a reflux ratio of 7.894 and 27 stages. The height of the towers was 16.20 meters and the diameter was 2.61 meters. The thickener had a surface area of 564.6 m^2 and held a volume of 1,694,000 liters. The thickener had a diameter of 26.8 meters and a depth of 3 meters. The second distillation tower had a reflux ratio of 0.305 and 73 stages. The height of the tower was 29 meters and the diameter was 1.84 meters. The flash drum was constructed according to ASME standards and held a volume of 453.7 liters. The diameter of the drum was 0.52 meters and the height was 2.10 meters.

From this process, 160.1 kg/hr of ethanol, 120,700 kg/hr of water, and 0.00008 kg/hr of trichloroethylene are lost either in purge or waste streams. Only 1.09 percent of the ethanol and 0.0000001 percent of the trichloroethylene are lost. 72.3 percent of the water is lost, which appears very high. However, as much water as possible was recycled to various parts of the process. All of the water lost in the purge stream after the thickening step had to be driven off because the pre-treatment steps didn't need any more water. The water lost in the waste stream after the flash could not be recycled because it was contaminated with trichloroethylene.

The following table summarizes the cost of each piece of equipment. These costs were taken from SuperPro Designer, with the exception of the cost of the Decanter. Neither SuperPro Designer nor Capcost had a specific model in place for determining the cost of a decanter. However, the decanter is basically a process vessel, and can be costed in Capcost based upon the residence time the liquid spends inside the decanter and the volume of fluid being processed.

A heuristic for liquid-liquid separations sets the residence time to 30 – 60 minutes to allow adequate settling (Turton et. al., 2009). Given a volumetric flow of 65,796 Liters/hour into the decanter and allowing 75% of the vessel volume to be in use at any given time, the total volume of the decanter will be equal to 61,863 L or 61.7 m^3 when using a residence time of 45 minutes. Approximating the width to be equal to the height and the length to be 3 times the width, this gives dimensions of approximately (8.28 x 2.76 w x 2.76 h). This information was then plugged into Capcost to estimate the cost of the decanter.

Table 4.9: Separation Equipment Cost

Equipment	**Tower 1**	**Thickener**	**Tower 2**	**Decanter**	**Heat Exchanger**	**Flash Drum**	**Total**
Cost	$326,000	$407,000	$250,000	$36,000	$40,000	$4,000	$769,600

As can be seen in the table, the majority of the cost for this part of the sugar beets to bioethanol production process is due to the costs of the two towers and the thickener. The first tower was more expensive than the second. However, it was processing almost twice as much material as the second column. The expense of the thickener was due to the volume necessary to accommodate the large amount of fluid flowing through.

4.3.2 Optimization of distillation

Optimization was conducted on the two distillation units in the system. The price of each unit was reiterated for varying values of R/Rmin, a ratio of the actual reflux ratio to the minimum reflux ratio, and plotted to determine which conditions created a global minimum cost. The results for this procedure are shown in the following figures.

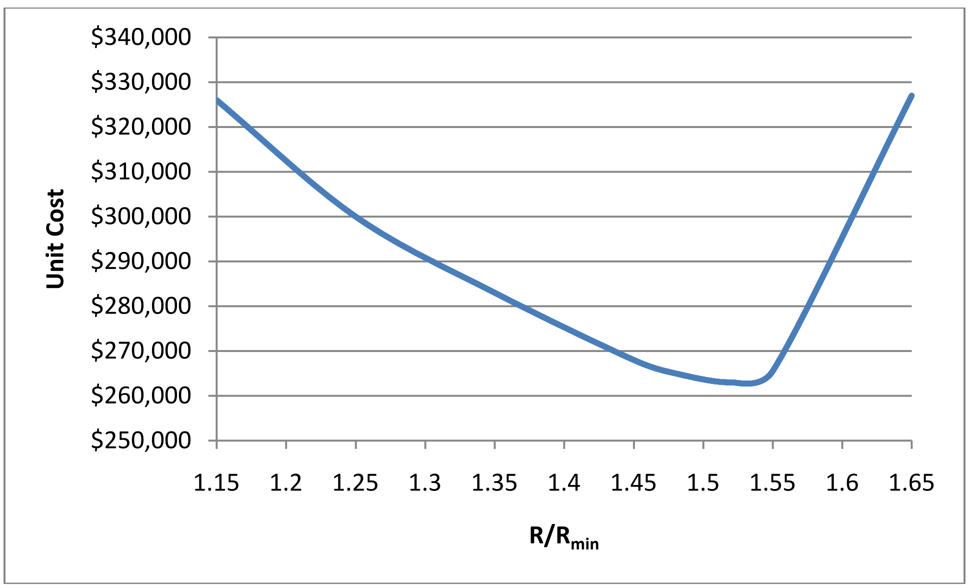

Figure 4.3: Determination of the Optimal Reflux Point for Tower T-101

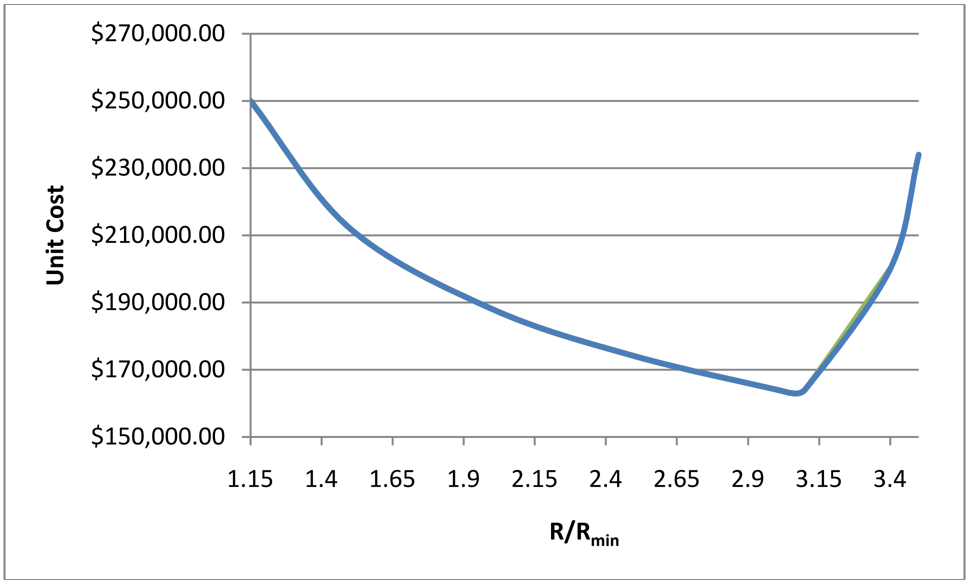

Figure 4.4: Determination of the Optimal Reflux Point for Tower T-102

Our initial approximation of R/R_{min} for both columns was set to 1.15. As seen in Figures 4.3 and 4.4, it is clear that the cost of each unit can be dropped drastically if a different value were chosen. Through SuperPro Designer's cost approximations, it was found that the optimal value for the first tower, T-101, was 1.52 and the optimal value for the second tower, T-102, was 3.05.

The table below compares the cost of the towers before and after optimization. As can be seen below, optimization decreases equipment costs by approximately $50,000 – $100,000 from our earlier approximations of the reflux ratio. This is done without affecting the extent of the separations.

Table 4.20: Optimization of Reflux Ratio

	Tower 1		Tower 2	
	R/Rmin	**Optimal R/Rmin**	**R/Rmin**	**Optimal R/Rmin**
Ratio	1.15	1.52	1.15	3.05
Equipment Cost	$326,000	$274,000	$250,000	$163,000

4.4 – Comparisons of Feedstocks

4.4.1 Sugar Beets

Each gram of sucrose that is recovered from sugar beets can be converted into approximately half a gram of ethanol. Sugar beets are generally about 17% sucrose by mass, although genetic modifications, selective breeding, and irrigation increase this number substantially, up to 21% (Cattanach et all., 1991), (Laboski et al., 2002). An average of 97% of the sucrose in sugar beets can be recovered for fermentation (Asadi, 2007). In 2009 25.8 tons or about 23,400 kg of sugar beets were harvested per acre of land cultivated (Haley and Dohlman, 2009). This means that per acre of land planted with sugar beets approximately 1930 kg of ethanol could be produced. Sugar beets preferably need to be processed soon after harvesting, as the sugar begins to degrade after a short period of time. However, they can be stored up to 180 days, as has been done by existing sugar beet plants when necessary, though this is not optimal for sucrose recovery (Southern Minnesota Sugar Beets Cooperative, 2010).

Sugar beets can be grown in many areas of the United States, from Minnesota to Texas. In recent years the numbers of acres of land that are planted with sugar beets has decreased, however more sugar beets have actually been grown (SIC 2063, 2005). This is due to increased crop density that is made possible by genetic alterations of sugar beets. There is certainly room to grow more sugar beets in the United States, simply by using the areas that were once planted with sugar beets that are no longer in use. In addition sugar beets can be grown in a wide range of soil types and in most areas in the United States (Cattanach et all., 1991).

4.4.2 Sugar Cane

Taking into consideration only the merits of the feedstock, sugar cane is a better choice as a feedstock for ethanol production than sugar beets. When sugar cane is processed the byproduct it produces that is analogous to spent beet pulp is called bagasse. Bagasse is a much more valuable byproduct than beet pulp because it can be burned to provide a large amount of the energy needed to power the ethanol plant (Quinteroa et al., 2008). Sugar beets do not have a byproduct like bagasse that can be burned to provide energy required for plant operation. This means that sugar cane requires less energy for conversion into ethanol than sugar beets will. The corresponding byproduct from sugar beets is useful as an animal feed, but it will not help provide energy (Cattanach et all., 1991).

The reason sugar cane is not a viable feedstock in the United States is that there is a very limited area where sugar cane can be grown. This area includes primarily Florida and Louisiana as well as some areas of Texas (Haley and Dohlman, 2009). Additionally, even in this area conditions are not ideal for growing sugar cane. Sugar cane cannot be transported over long distances, as the sugar will begin to break down if it is not processed soon after the cane is harvested. This means that there is very limited possibility for producing ethanol from sugar cane in the United States. In countries closer to the equator, such as Brazil, sugar cane is an excellent choice as a bioethanol feedstock (Almeida, 2007 and Morgan, 2005).

4.4.3 Corn

Converting corn into ethanol is a more complicated process that converting sugar cane or sugar beets into ethanol. Sugar cane and sugar beets both contain sucrose, which can be removed fairly simply from the plants. The plant material is first crushed or shredded then washed with water to separate sucrose from fiber. In order to get a fermentable sugar from corn a more complicated process is required. Corn contains starch, which must be broken down into sugar before it can be fermented. This requires an enzymatic treatment and enzyme recovery steps.

Approximately 3500 kg of corn can be produced from 1 acre of land, and 1 L of ethanol can be produced from 2.69 kg of corn. This means that only 1300 L of ethanol, or just over 1000 kg of ethanol, can be produced per acre of corn grown (Pimentel and Patzek, 2005). This is just more than half of the amount of ethanol that can be produced from sugar beets grown over the same area.

Additionally, the net energy value (NEV) of corn has been highly debated. Some studies have found that the amount of energy needed to produce ethanol from corn is less than the amount of energy supplied by the ethanol (Shapouri et all, 2002), this finding would correspond to a positive NEV. However other studies have determined that more energy is used to produce ethanol than is supplied by the energy, or a negative NEV (Pimentel and Patzek, 2005). There are many factors involved in calculating the NEV for ethanol from corn. Researchers often estimate different values for specific energy costs, for example the amount of energy needed to produce fertilizer, and some values are neglected entirely in some studies.

Corn stover includes the stalk, leaf, husk, and cobs left over after the corn kernels have been harvested. It does not include the roots of the corn plant, which are essentially always left for soil supplementation. Stover is the analogous fibrous material to bagasse from sugarcane or pressed pulp from sugar beets. Unlike bagasse or pressed pulp, which are used for energy production and animal feed, respectively. Corn stover is rarely used for such purposes. One reason stover is not used for other purposes is that leaving it intact helps improve soil quality. According to USDA guidelines at least 30% of corn stover should be left on the field to supplement the soil, prevent erosion, and contain moisture and therefore improve crops the next year. If stover is removed the soil must be enriched with additional fertilizers, which is estimated to cost approximately $6.40 per acre. (Hettenhaus, 2002)

Cows can be allowed to graze on the stalks, leaves etc that are left on the field, but they have to be monitored carefully and fed supplements as corn stover is not a complete food for cows. Less than 5% of corn stalks were fed to animals or used as bedding as of 2002 (Hettenhaus, 2002). The nutritional value of corn stover decreases significantly after harvest, so if it is to be fed to cows, this should be done soon after harvest. If corn stover is baled it has to be treated with a chemical, usually ammonia, to prevent deterioration and add protein (Kyle, 2009).

The other possible use for corn stover is to use it as a feedstock for conversion into ethanol. As there is a very large quantity of corn stover produced there is a large amount of this cellulosic byproduct. However, like all cellulosic materials, corn stover contains lignin, which significantly limits its possible use as an ethanol feedstock (Chapel et al., 2007). More information about lignin is included in section 2.6: Alternative Sources for Ethanol Production. As a result of these limitations to the uses of corn stover, it is very common for farmers to simply leave stover on the field, or to plow it underground to enrich the soil (Kyle, 2009).

4.4.4 Energy required for conversion

The amount of energy necessary to produce ethanol from a feedstock depends both on the chemical process required for the conversion and on factors such as transportation costs, irrigation requirements, fertilizer needs, and pesticide applications. Transportation costs are impossible to assess comparatively since they would depend on the location of a sugar beet to ethanol plant, which does not exist at this time. Pesticide use varies widely from farm to farm and from year to year because it is often applied in response to a problem. As a result an accurate comparison was not possible.

Sugar beets require approximately 560 mm of water to grow to maturity (Efetha, 2008). They also require certain levels of potassium, phosphate, and especially nitrogen. The level of each of these nutrients in the soil already has a very high effect on how much fertilizer must be added. Maximum amounts of each nutrient necessary are: 80 lbs potassium/acre, 100 lbs phosphate/acre, and 200 lbs nitrogen/acre (Cattanach et all., 1991). In general however, lower levels will be required because the soil will not be completed depleted of these nutrients, which the above numbers assume. Amounts as low as one half of the amount listed above are required for soil with just moderate amounts of each nutrient.

Corn requires approximately 26 inches of water to grow to maturity (Kranz, 2008), which is almost exactly equal to 560mm. This means that there will be essentially no energy difference per acre due to irrigation needs. Per year corn requires approximately the following amounts of nutrients as fertilizer: 160 lbs potassium/acre, 70 lbs phosphate/acre, and 140 lbs nitrogen/acre (ProCrop, 2009). It is difficult to comment on which crop requires more energy as fertilizer per acre, since fertilizer needs depend so heavily on the levels of nutrients already in the soil. Corn definitely requires more potassium, but the other nutrients may be higher or lower for either crop. However, both amount of irrigation and fertilizer required are on a per acre basis and sugar beets will produce almost twice as much ethanol per acre. This means that per liter of ethanol produced the energy cost for both irrigation and fertilizer will likely be much lower for sugar beets.

The amount of energy used to convert corn to ethanol was compared to the amount of energy used to convert sugar beets into ethanol. This comparison was made between the USDA SuperPro Designer corn to ethanol simulation (see section 2.9.3 for more information), and the sugar beet to ethanol SuperPro Designer simulation created in this project (Kwiatkowski, 2006). The energy used in the sugar beet process falls into one of four categories: electricity, cooling water, chilled water, and steam. The sugar beet simulation calculates an annual use of 28,940,000 kW/h of electricity, 93,600,000,000 kg of cooling water, 437,000,000 kg cooled water and 1,265,000,000 kg of steam. The corn to ethanol process has a longer list of utilities used, including natural gas, cooling tower water, and more varieties of steam. The USDA corn simulation calculates an annual use of 29,400,000 kW/h of electricity, 37,300,000,000 kg of cooling water, 0 kg cooled water 11,800,000 kg natural gas, and 21,850,000,000 kg of cooling tower water. Table 4.11 shows the different types of steam used in this process. Total steam usage for the corn process 240,000,000 kg, which is much lower that the sugar beet steam usage. However, the amount of energy required to produce high pressure steam is much greater that medium pressure steam, which is used exclusively in the sugar beet process.

Table 4:31: Steam Usage in Corn to Ethanol Process

Type of steam	Amount used (kg)
Steam 50 PSI	31,700,000
Steam 6258 BTU	32,500,000
Steam 2205 BTU	176,000,000
Steam (High P)	1,110,000

Since the utilities used in each process are different it is difficult to determine which will have an overall higher energy usage without studying the total amount of fossil fuels used to produce each utility. Further research should be done to determine which feedstock has a better overall net energy value.

4.4.5 Cost per gallon of ethanol

A summary of the required costs to produce a gallon of ethanol was produced based on the information produced from each process in SuperPro Designer. The cost for feedstocks were used by the most recent data produced from the EPA, in 2009, for corn, where the average was $4.80/bushel of corn, which is greater than in 2007 when the original corn to ethanol plant design was made, but less than the average in 2008. A value of $45/ton of sugar beets was used to determine the cost of sugar beets per gallon produced, as it included a slight safety factor from the average costs of the 2004 U.C. Cooperative Extension case study on sugar beet costs, where the average was approximately $41/ton. Process stream costs were found by recent ICIS costs. Utility costs were taken directly from those estimated in SuperPro Designer. The resulting costs on a per gallon basis are shown in Table 4.12.

Table 4.42: Cost to Produce Ethanol on a Per Gallon Basis

	Corn	Sugar Beets
Feedstock	$1.76	$1.66
Process Streams	$0.10	$0.07
Utilities	$0.38	$0.57
Total	$2.24	$2.30

All costs in Table 4.12 are based on annual costs and do not factor in fixed capital costs such as the cost to build the plant and equipment costs. What the table shows is that while costs of feedstocks and process streams are relatively comparable and that sugar beets are slightly more cost effective in both, netting $0.13 less to produce a gallon of ethanol. However, the process utilities cost, which has been discussed in Section 4.4.4, provides a much larger discrepancy, one that could be improved with refinement to the sugar beets to ethanol plant design. As it stands now, sugar beets costs $0.06 more to produce a gallon of ethanol, but this difference could be alleviated with improvements to the system.

5. Conclusions

The previously described process was modeled to produce the same amount of ethanol as was produced by the USDA model. For sugar beets, 168,000 kg of raw beet produces 14,700 kg of ethanol. A process modeled by the USDA for producing ethanol from corn required 46,350 kg of corn to produce the same amount of ethanol. This is a very large difference in the amounts of raw materials necessary. However, when the amounts of raw materials are reduced to the approximate amount of acres, it can be seen that it requires 7.2 acres of sugar beets versus 13.2 acres of corn to produce the same amount of ethanol. Therefore, although it requires approximately 10 times as much raw sugar beet to produce the same amount of ethanol, as far as land use sugar beets can produce approximately 1.8 times as much ethanol. Knowing that an acre of sugar beets and an acre of corn require approximately as much water, and assuming comparable fertilizer and pesticide requirements, it can be determined that sugar beets are almost twice as efficient when it comes to land usage requirements.

The cost of purchasing the equipment for a sugar beets processing facility comes out to approximately $15,876,000, based on estimates from SuperPro Designer. The cost of purchasing the equipment for the process modeled by the USDA was $18,556,000. The majority of the cost of the sugar beets plant came from the extractor used to wash the sucrose from the cossettes. Further research into the use and refinement of the extractor could yield a lower cost. Although costs would be cut considerably if a sugar beet to table sugar plant were available for conversion into a sugar beet to ethanol plant, as all the equipment except those required for fermentation and separation would already be in place, this is not possible at this time. Currently all sugar beet plants in the United States are running at capacity. Besides not leaving any room for new sugar beet farmers unless they are able to build a new facility or take the spot of another farmer in an existing collective, this also does not leave any sugar beet plants free for partial or full conversion. However, if in the future demands on domestic sugar decrease enough that the plants begin to operate under capacity, this would be an economic option for the creation of a sugar beets to ethanol plant.

The proposed sugar beets plant required total annual utilities of 9,620,000 kilowatt hours of electricity, 1,272,000,000 kg of steam, and 93,595,000,000 kg of cooling water. The corn model provided by the USDA required total annual utilities of 29,430,000 kilowatt hours of electricity, 37,312,000,000 kg of cooling water, 11,770,000 kg of natural gas, and 241,300,000 kg of steam. The corn model consumes 67% more electricity than the sugar beets process, and requires a large amount of natural gas. The sugar beets process requires nearly 70% more cooling water and 81% more steam. The majority of this extra cost is likely to come from the utilities required by the distillation towers. The corn model used molecular sieves and packed towers with salt to separate ethanol and water. If a similar separation method was put in place for the sugar beet process, the utility cost could be cut significantly.

From the 30,200 kg of sucrose available from 168,000 kg of raw beets, 900 kg is lost during the process. Therefore, only about 3% of the total available sugar is being lost throughout the process. Of the 14,700 kg of ethanol produced from the same amount of sugar beets, 190 kg are lost during separation. This is a loss of a little more than one percent. Out of the 84,000 kg of trichloroethylene added to the second distillation tower to separate the 14,700 kg of ethanol from the water, only 0.00008 kg is lost during the separation from the water.

Taking these factors into consideration, as well as the growing area, crop density, and food versus fuel problem mentioned earlier, it is clear that sugar beets are the superior feedstock for producing ethanol.

6. Recommendations

Though through our research and simulations, we determined that sugar beets are the superior feedstock, there is still a lot of work to be done before a sugar beets to ethanol plant would be ready to be built. There was uncertainty built into both our data and the data gained from the corn model provided by the USDA, due to the use of SuperPro Designer, especially with data relying on thermodynamic principles. As was stated in the background, SuperPro Designer lacks rigorous thermodynamic packages. Though our simulations were run through Aspen Plus as often as possible to check the accuracy of the data, we were unable to model the pretreatment, filtration, and fermentation steps in Aspen Plus. We recommend that the process be modeled using software with rigorous thermodynamic packages similar to Aspen Plus but with the ability to model these steps.

We also recommend that further studies on sugar beets to ethanol plants find and incorporate additional data about processing and growing sugar beets that can be used to modify the process simulation and increase accuracy. A partnership with the USDA would yield statistical data on the annual fertilizer and pesticide requirements of sugar beets, as well as current growth patterns and ranges. With this information, an accurate NEV can be calculated for sugar beets, and the optimal location for a sugar beets to ethanol plant can be found. This location would be one where the growth of the sugar beets does not limit the growth of other key food crops, while allowing the minimization of necessary fertilizers and pesticides.

We also recommend that further research be conducted into utility optimization for the sugar beet to ethanol process. Unlike with sugar cane, the pressed beet pulp cannot be burned to recovered energy and help fuel the process. To prevent the plant from running unsustainably by relying on energy from non-renewable sources to drive the production of ethanol, we recommend that research on alternative fuel sources be researched and considered, as well as the energy demands of the sugar beets plant minimized. Part of this research could involve utilizing the molecular sieve adsorption separation mechanism used in the USDA model to cut back on utility costs. This would minimize not only the steam and cooling water utilities necessary for the extractive distillation system, but also eliminate the electricity usage necessary to pump the large volumes of trichloroethylene through the system.

We recommend that further studies be conduction on minimizing the cost per gallon, including investigating if there is an optimum size at which equipment and utility costs become minimized. If it can be shown that sugar beets can be more economical than corn, even with current government-sponsored subsidies on corn, sugar beets will be more likely to become mainstream as a feedstock for bioethanol production.

We recommend that further research be conducted into utilizing sugar beets as a feedstock for producing ethanol, and that a pilot scale plant be built. Although studies have been done researching the theoretical benefits of using sugar beets over corn, there are currently no plants in place for converting the sucrose washed from the sugar beets into ethanol. Were a sugar beet to table sugar plant become unused in the future, we recommend it's conversion to a bioethanol plant, though we recognize that this is not possible at this time.

Any future researchers can contact the contributing engineers for SuperPro Designer files and Simulation data used for this project at the following e-mail addresses:

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Additionally, Professor William M. Clark has access to all of the files and data used for this project.

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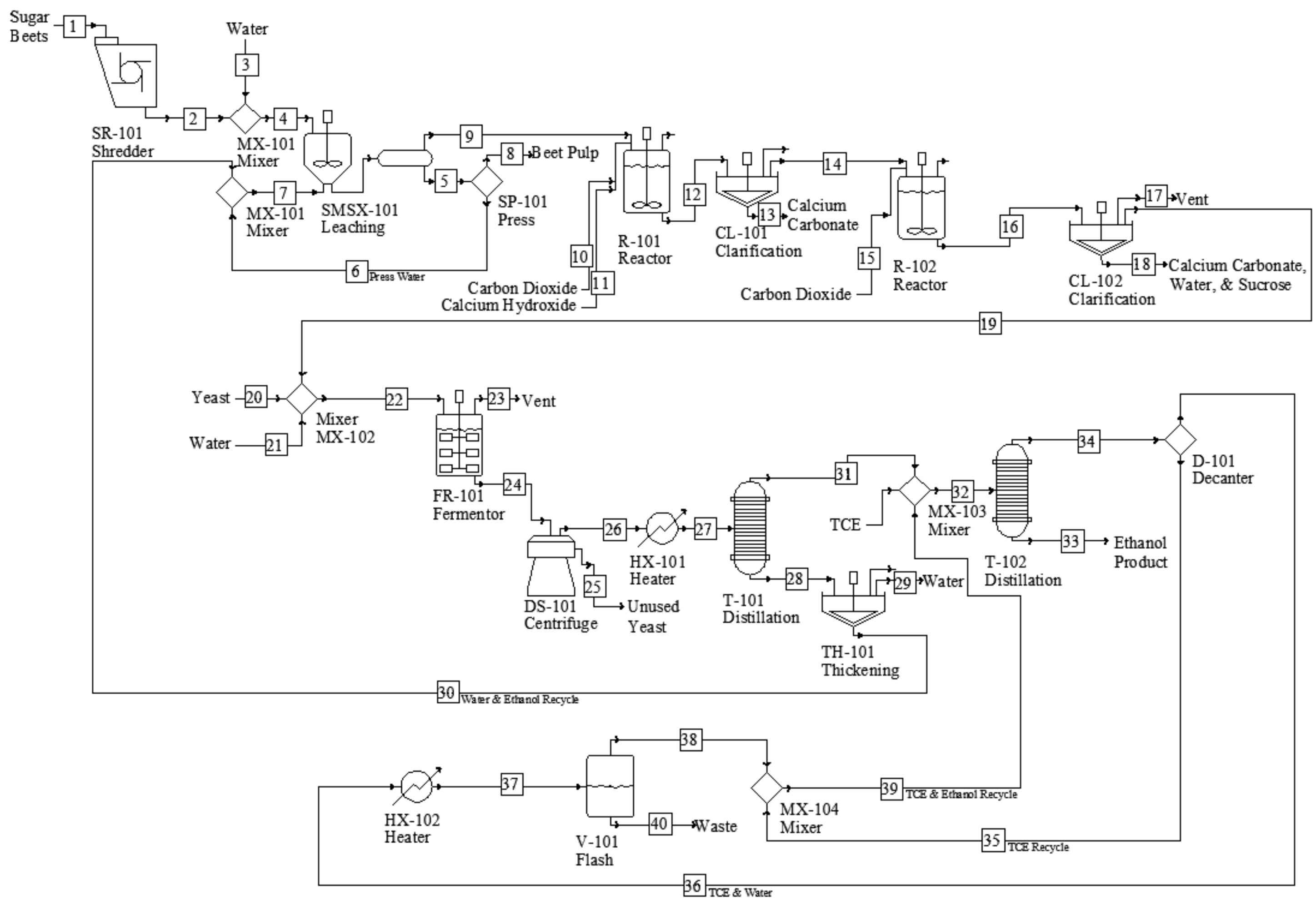
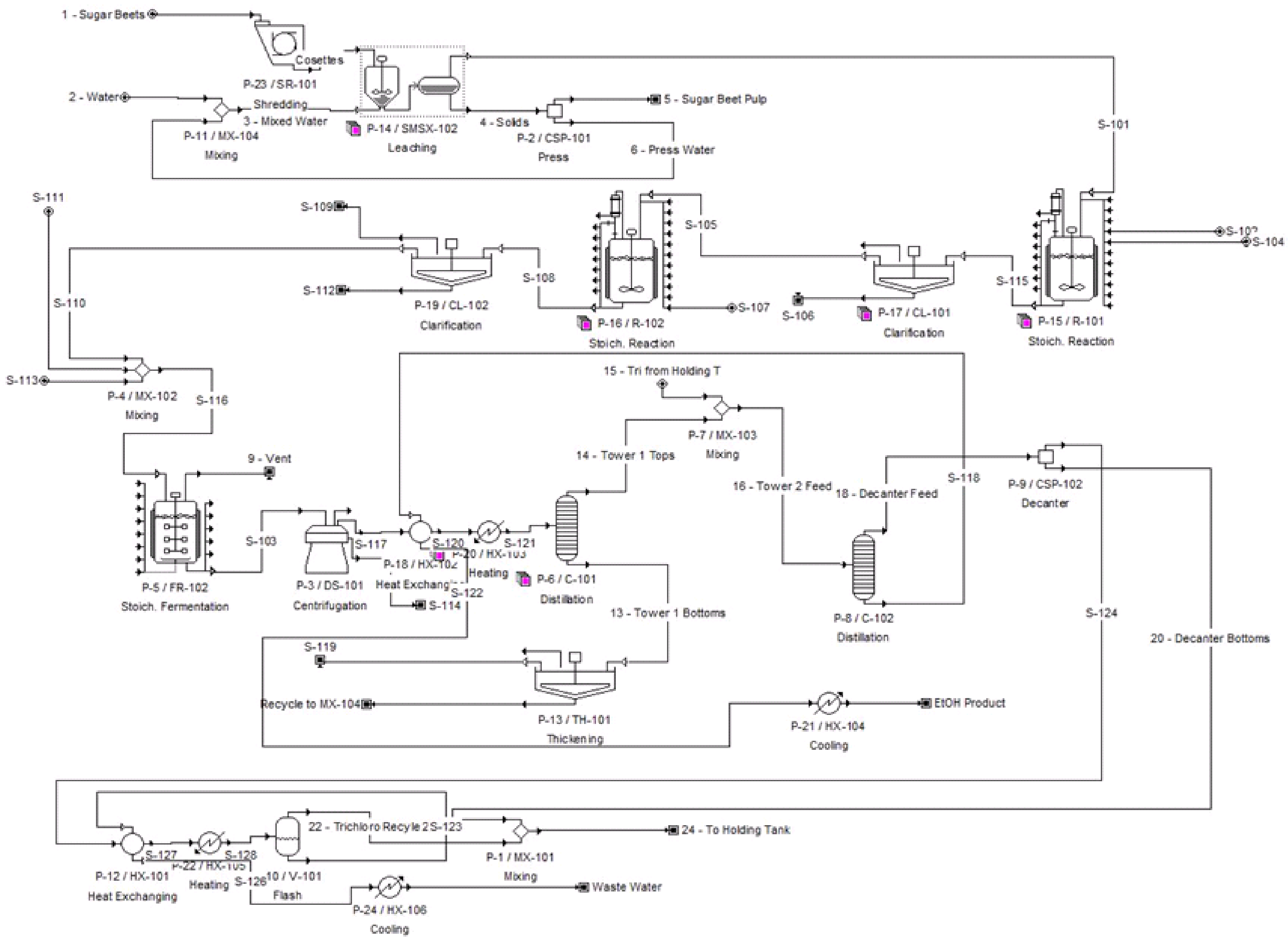
8. Appendix A – Process Flow Diagram of Sugar Beets Process from AutoCad



9. Appendix B – Process Flow Diagram of Sugar Beets Process from SuperPro Designer



10. Appendix C – Sugar Beets Material and Energy Balance from SuperPro Designer

Materials & Streams Report

for Sugar Beets backup from 4-26-10

April 26, 2010

1. OVERALL PROCESS DATA

Annual Operating Time	8,000.00h
Annual Throughput	0.00kg MP
Operating Days per Year	333.33

MP = Main Product = Undefined

2.1 STARTING MATERIAL REQUIREMENTS (per Section)

Section	Starting Material	Active Product	Amount Needed (kg Sin/kg MP)	Molar Yield (%)	Mass Yield (%)	Gross Mass Yield (%)
Sugar Beets Simulation	(none)	(none)	Unknown	Unknown	Unknown	Unknown

Sin = Section Starting Material, Aout = Section Active Product

2.2 BULK MATERIALS (Entire Process)

Material	kg/yr	kg/h	kg/kg MP
1,1,2-TriChEth	671,999,934	83,999.992	
Ethyl Alcohol	4,966,080	620.760	
Water	2,191,264,106	273,908.013	
Inert Solids	67,199,993	8,399.999	
Sucrose	241,919,976	30,239.997	
Ca Hydroxide	26,879,997	3,360.000	
Carb. Dioxide	21,680,000	2,710.000	
Yeast	6,720	0.840	
TOTAL	3,225,916,806	403,239.601	

2.3 BULK MATERIALS (per Section)

SECTIONS IN: Main Branch

Sugar Beets Simulation

Material	kg/yr	kg/h	kg/kg MP
1,1,2-TriChEth	671,999,934	83,999.992	
Ethyl Alcohol	4,966,080	620.760	
Water	2,191,264,106	273,908.013	
Inert Solids	67,199,993	8,399.999	
Sucrose	241,919,976	30,239.997	
Ca Hydroxide	26,879,997	3,360.000	
Carb. Dioxide	21,680,000	2,710.000	
Yeast	6,720	0.840	
TOTAL	3,225,916,806	403,239.601	

2.4 BULK MATERIALS (per Material)

1,1,2-TriChEth

1,1,2-TriChEth	% Total	kg/yr	kg/h	kg/kg MP
Sugar Beets Simulation (Main Branch)				
P-7	100.00	671,999,934	83,999.992	
TOTAL	100.00	671,999,934	83,999.992	

Ethyl Alcohol

Ethyl Alcohol	% Total	kg/yr	kg/h	kg/kg MP
Sugar Beets Simulation (Main Branch)				
P-7	100.00	4,966,080	620.760	
TOTAL	100.00	4,966,080	620.760	

Water

Water	% Total	kg/yr	kg/h	kg/kg MP
Sugar Beets Simulation (Main Branch)				
P-7	0.23	5,026,560	628.320	
P-11	26.07	571,199,944	71,399.993	
P-14	47.23	1,034,879,899	129,359.987	
P-15	2.86	62,717,754	7,839.719	
P-4	23.61	517,439,949	64,679.994	
TOTAL	100.00	2,191,264,106	273,908.013	

Inert Solids

Inert Solids	% Total	kg/yr	kg/h	kg/kg MP
Sugar Beets Simulation (Main Branch)				
P-14	100.00	67,199,993	8,399.999	
TOTAL	100.00	67,199,993	8,399.999	

Sucrose

Sucrose	% Total	kg/yr	kg/h	kg/kg MP
Sugar Beets Simulation (Main Branch)				
P-14	100.00	241,919,976	30,239.997	
TOTAL	100.00	241,919,976	30,239.997	

Ca Hydroxide

Ca Hydroxide	% Total	kg/yr	kg/h	kg/kg MP
Sugar Beets Simulation (Main Branch)				
P-15	100.00	26,879,997	3,360.000	
TOTAL	100.00	26,879,997	3,360.000	

Carb. Dioxide

Carb. Dioxide	% Total	kg/yr	kg/h	kg/kg MP
Sugar Beets Simulation (Main Branch)				
P-15	92.25	20,000,000	2,500.000	
P-16	7.75	1,680,000	210.000	
TOTAL	100.00	21,680,000	2,710.000	

Yeast

Yeast	% Total	kg/yr	kg/h	kg/kg MP
Sugar Beets Simulation (Main Branch)				
P-4	100.00	6,720	0.840	
TOTAL	100.00	6,720	0.840	

2.5 BULK MATERIALS: SECTION TOTALS (kg/h)

Raw Material	Sugar Beets Simulation
1,1,2-TriChEth	83,999.992
Ethyl Alcohol	620.760
Water	273,908.013
Inert Solids	8,399.999
Sucrose	30,239.997
Ca Hydroxide	3,360.000
Carb. Dioxide	2,710.000
Yeast	0.840
TOTAL	403,239.601

2.6 BULK MATERIALS: SECTION TOTALS (kg/yr)

Raw Material	Sugar Beets Simulation
1,1,2-TriChEth	671,999,934
Ethyl Alcohol	4,966,080
Water	2,191,264,106
Inert Solids	67,199,993
Sucrose	241,919,976
Ca Hydroxide	26,879,997
Carb. Dioxide	21,680,000
Yeast	6,720
TOTAL	3,225,916,806

3. STREAM DETAILS

Stream Name	1 - Cossettes	3 - Mixed Water	S-101	4 - Solids
Source	**INPUT**	**P-11**	**P-14**	**P-14**
Destination	**P-14**	**P-14**	**P-15**	**P-2**
Stream Properties				
Activity (U/ml)	0.000	0.000	0.000	0.000
Temperature (°C)	25.000	25.514	26.152	26.152
Pressure (bar)	1.013	1.013	1.013	1.013
Density (g/L)	1,059.789	998.478	1,045.720	1,002.118
Component Flowrates (kg/h averaged)				
Inert Solids	8,399.99918	0.00000	0.00000	8,399.99918
Sucrose	30,239.99704	1,511.39505	29,634.59230	2,116.79979
Water	129,359.98734	128,472.57782	175,949.37318	81,883.19198
TOTAL (kg/h)	167,999.98355	129,983.97288	205,583.96548	92,399.99096
TOTAL (L/h)	158,522.057	130,182.086	196,595.701	92,204.713

Stream Name	5 - Sugar Beet Pulp	6 - Press Water	2 - Water	S-102
Source	**P-2**	**P-2**	**INPUT**	**INPUT**
Destination	**OUTPUT**	**P-11**	**P-11**	**P-15**
Stream Properties				
Activity (U/ml)	0.000	0.000	0.000	0.000
Temperature (°C)	26.152	26.152	25.000	25.000
Pressure (bar)	1.013	1.013	1.013	1.013
Density (g/L)	1,000.396	1,003.115	994.704	1.799
Component Flowrates (kg/h averaged)				
Carb. Dioxide	0.00000	0.00000	0.00000	2,500.00000
Inert Solids	8,399.99918	0.00000	0.00000	0.00000
Sucrose	605.40474	1,511.39505	0.00000	0.00000
Water	24,810.60717	57,072.58481	71,399.99301	0.00000
TOTAL (kg/h)	33,816.01109	58,583.97987	71,399.99301	2,500.00000
TOTAL (L/h)	33,802.641	58,402.071	71,780.117	1,389,753.433

Stream Name	S-104	S-115	S-105	S-106
Source	**INPUT**	**P-15**	**P-17**	**P-17**
Destination	**P-15**	**P-17**	**P-16**	**OUTPUT**
Stream Properties				
Activity (U/ml)	0.000	0.000	0.000	0.000
Temperature (°C)	25.000	28.377	28.377	28.377
Pressure (bar)	1.013	1.013	1.013	1.013
Density (g/L)	1,201.196	401.150	339.074	1,046.959
Component Flowrates (kg/h averaged)				
Ca Hydroxide	3,359.99967	167.99998	151.19999	16.80000
CaCO3	0.00000	4,311.73052	431.17305	3,880.55747
Carb. Dioxide	0.00000	604.05687	604.05687	0.00000
Sucrose	0.00000	29,634.59230	29,486.41934	148.17296
Water	7,839.71923	184,565.17547	138,423.88160	46,141.29387
TOTAL (kg/h)	11,199.71890	219,283.55514	169,096.73085	50,186.82429
TOTAL (L/h)	9,323.803	546,637.942	498,702.131	47,935.811

Stream Name	S-107	S-108	S-109	S-110
Source	**INPUT**	**P-16**	**P-19**	**P-19**
Destination	**P-16**	**P-19**	**OUTPUT**	**P-4**
Stream Properties				
Activity (U/ml)	0.000	0.000	0.000	0.000
Temperature (°C)	25.000	28.376	28.376	28.376
Pressure (bar)	1.013	1.013	1.013	1.013
Density (g/L)	1.799	297.635	1.779	1,074.239
Component Flowrates (kg/h averaged)				
Ca Hydroxide	0.00000	7.56000	0.00000	0.00000
CaCO3	0.00000	625.20093	0.00000	0.00000
Carb. Dioxide	209.99998	728.73941	728.73941	0.00000
Sucrose	0.00000	29,486.41934	0.00000	29,338.98724
Water	0.00000	138,458.80534	0.00000	103,844.10401
TOTAL (kg/h)	209.99998	169,306.72501	728.73941	133,183.09125
TOTAL (L/h)	116,739.277	568,839.508	409,694.125	123,979.000

Stream Name	S-112	S-111	S-113	S-116
Source	P-19	INPUT	INPUT	P-4
Destination	OUTPUT	P-4	P-4	P-5
Stream Properties				
Activity (U/ml)	0.000	0.000	0.000	0.000
Temperature (°C)	28.376	25.000	25.000	27.144
Pressure (bar)	1.013	1.013	1.013	1.013
Density (g/L)	1,006.498	1,562.000	994.704	1,046.898
Component Flowrates (kg/h averaged)				
Ca Hydroxide	7.56000	0.00000	0.00000	0.00000
CaCO3	625.20093	0.00000	0.00000	0.00000
Sucrose	147.43210	0.00000	0.00000	29,338.98724
Water	34,614.70134	0.00000	64,679.99367	168,524.09767
Yeast	0.00000	0.84000	0.00000	0.84000
TOTAL (kg/h)	35,394.89436	0.84000	64,679.99367	197,863.92491
TOTAL (L/h)	35,166.383	0.538	65,024.341	189,000.208

Stream Name	9 - Vent	S-103	S-117	S-114
Source	P-5	P-5	P-3	P-3
Destination	OUTPUT	P-3	P-18	OUTPUT
Stream Properties				
Activity (U/ml)	0.000	0.000	0.000	0.000
Temperature (°C)	33.000	33.000	33.065	33.065
Pressure (bar)	1.013	1.013	1.013	1.013
Density (g/L)	1.753	972.888	972.859	1,029.433
Component Flowrates (kg/h averaged)				
Carb. Dioxide	14,334.23279	0.00000	0.00000	0.00000
Ethyl Alcohol	150.04857	14,854.80874	14,854.40978	0.39895
Sucrose	0.00000	1,466.94936	1,466.90996	0.03940
Water	0.00000	167,057.20831	167,052.72169	4.48662
Yeast	0.00000	0.84000	0.00000	0.84000
TOTAL (kg/h)	14,484.28136	183,379.80641	183,374.04143	5.76497
TOTAL (L/h)	8,264,051.670	188,490.055	188,489.886	5.600

Stream Name	S-118	S-120	S-122	S-121
Source	**P-8**	**P-18**	**P-18**	**P-20**
Destination	**P-18**	**P-20**	**P-21**	**P-6**
Stream Properties				
Activity (U/ml)	0.000	0.000	0.000	0.000
Temperature (°C)	79.000	34.825	43.065	100.000
Pressure (bar)	1.013	1.013	1.013	1.013
Density (g/L)	737.738	972.095	769.785	18.252
Component Flowrates (kg/h averaged)				
Ethyl Alcohol	14,700.00007	14,854.40978	14,700.00007	14,854.40978
Sucrose	0.00000	1,466.90996	0.00000	1,466.90996
Water	0.48046	167,052.72169	0.48046	167,052.72169
TOTAL (kg/h)	14,700.48053	183,374.04143	14,700.48053	183,374.04143
TOTAL (L/h)	19,926.411	188,638.040	19,096.856	10,046,574.815

Stream Name	14 - Tower 1 Tops	13 - Tower 1 Bottoms	15 - Tri from Holding T	16 - Tower 2 Feed
Source	**P-6**	**P-6**	**INPUT**	**P-7**
Destination	**P-7**	**P-13**	**P-7**	**P-8**
Stream Properties				
Activity (U/ml)	0.000	0.000	0.000	0.000
Temperature (°C)	82.000	100.000	78.250	78.250
Pressure (bar)	1.013	1.013	1.013	1.013
Density (g/L)	2.024	970.340	1,356.513	10.992
Component Flowrates (kg/h averaged)				
1,1,2-TriChEth	0.00000	0.00000	83,999.99178	83,999.99178
Ethyl Alcohol	14,852.92434	1.48544	620.75994	15,473.68428
Sucrose	0.00000	1,466.90996	0.00000	0.00000
Water	4,176.31804	162,876.40364	628.31994	4,804.63798
TOTAL (kg/h)	19,029.24238	164,344.79905	85,249.07166	104,278.31404
TOTAL (L/h)	9,399,991.643	169,368.208	62,844.290	9,487,064.465

Stream Name	18 - Decanter Feed	S-119	Recycle to MX-104	EtOH Product
Source	**P-8**	**P-13**	**P-13**	**P-21**
Destination	**P-9**	**OUTPUT**	**OUTPUT**	**OUTPUT**
Stream Properties				
Activity (U/ml)	0.000	0.000	0.000	0.000
Temperature (°C)	58.000	100.000	100.000	30.000
Pressure (bar)	1.013	1.013	1.013	1.013
Density (g/L)	1,361.447	961.736	972.101	781.436
Component Flowrates (kg/h averaged)				
1,1,2-TriChEth	83,999.99178	0.00000	0.00000	0.00000
Ethyl Alcohol	773.68421	1.06284	0.42260	14,700.00007
Sucrose	0.00000	0.00000	1,466.90996	0.00000
Water	4,804.15752	116,538.42103	46,337.98261	0.48046
TOTAL (kg/h)	89,577.83351	116,539.48387	47,805.31518	14,700.48053
TOTAL (L/h)	65,796.033	121,176.138	49,177.293	18,812.130

Stream Name	S-124	20 - Decanter Bottoms	S-123	S-127
Source	**P-9**	**P-9**	**P-10**	**P-12**
Destination	**P-12**	**P-1**	**P-12**	**P-22**
Stream Properties				
Activity (U/ml)	0.000	0.000	0.000	0.000
Temperature (°C)	58.000	58.000	93.074	78.251
Pressure (bar)	1.013	1.013	1.013	1.013
Density (g/L)	960.099	1,397.398	958.025	239.011
Component Flowrates (kg/h averaged)				
1,1,2-TriChEth	8.40000	83,991.59178	0.00008	8.40000
Ethyl Alcohol	416.78369	356.90053	158.87794	416.78369
Water	4,768.12634	36.03118	4,169.72648	4,768.12634
TOTAL (kg/h)	5,193.31002	84,384.52349	4,328.60451	5,193.31002
TOTAL (L/h)	5,409.141	60,386.893	4,518.256	21,728.376

Stream Name	S-125	S-128	22 - Trichloro Recyle 2	Waste Water
Source	**P-12**	**P-22**	**P-10**	**P-23**
Destination	**P-23**	**P-10**	**P-1**	**OUTPUT**
Stream Properties				
Activity (U/ml)	0.000	0.000	0.000	0.000
Temperature (°C)	68.000	95.000	93.074	30.000
Pressure (bar)	1.013	1.013	1.013	1.013
Density (g/L)	968.028	18.537	0.740	983.118
Component Flowrates (kg/h averaged)				
1,1,2-TriChEth	0.00008	8.40000	8.39992	0.00008
Ethyl Alcohol	158.87794	416.78369	257.90574	158.87794
Water	4,169.72648	4,768.12634	598.39986	4,169.72648
TOTAL (kg/h)	4,328.60451	5,193.31002	864.70552	4,328.60451
TOTAL (L/h)	4,471.569	280,151.930	1,168,357.378	4,402.937

Stream Name	24 - To Holding Tank
Source	**P-1**
Destination	**OUTPUT**
Stream Properties	
Activity (U/ml)	0.000
Temperature (°C)	78.251
Pressure (bar)	1.013
Density (g/L)	837.088
Component Flowrates (kg/h averaged)	
1,1,2-TriChEth	83,999.99169
Ethyl Alcohol	614.80627
Water	634.43104
TOTAL (kg/h)	85,249.22900
TOTAL (L/h)	101,840.220

4. OVERALL COMPONENT BALANCE (kg/yr)

COMPONENT	IN	OUT	OUT-IN
1,1,2-TriChEth	671,999,934	671,999,934	- 0
Ca Hydroxide	26,879,997	194,880	- 26,685,117
CaCO3	0	36,046,067	36,046,067
Carb. Dioxide	21,680,000	120,503,778	98,823,778
Ethyl Alcohol	4,966,080	125,004,938	120,038,858
Inert Solids	67,199,993	67,199,993	- 0
Sucrose	241,919,976	18,943,673	- 222,976,303
Water	2,191,264,106	2,186,017,045	- 5,247,061
Yeast	6,720	6,720	0
TOTAL	3,225,916,806	3,225,917,029	222

5. EQUIPMENT CONTENTS

This section will be skipped (overall process is continuous)

11. Appendix D – Sugar Beets Economic Evaluation Report from SuperPro Designer

Economic Evaluation Report

April 18, 2010

for Sugar Beets backup from 4-12-10

1. EXECUTIVE SUMMARY (2010 prices)

Total Capital Investment	179,061,000 $
Capital Investment Charged to This Project	179,061,000 $
Operating Cost	908,050,000 $/yr

THE MAIN REVENUE STREAM HAS NOT BEEN IDENTIFIED. PRICING AND PRODUCTION/PROCESSING UNIT COST DATA HAVE NOT BEEN PRINTED

Main Revenue	0 $/yr
Gross Margin	- 1.00 %
Return On Investment	- 502.12 %
Payback Time	- 1.00 years
IRR (After Taxes)	Out of search interval (0-1000%)
NPV (at 7.0% Interest)	0 $

MT = Metric Ton (1000 kg)

2. MAJOR EQUIPMENT SPECIFICATION AND FOB COST (2010 prices)

Quantity/ Standby/ Staggered	Name	Description	Unit Cost ($)	Cost ($)
1 / 0 / 0	FR-102	Fermentor	1,830,000	1,830,000
		Vessel Volume = 4200004.62 L		
1 / 0 / 0	C-102	Distillation Column	163,000	163,000
		Column Volume = 55017.94 L		
1 / 0 / 0	V-101	Flash Drum	4,000	4,000
		Vessel Volume = 453.71 L		
2 / 0 / 0	C-101	Distillation Column	137,000	274,000
		Column Volume = 81829.26 L		
1 / 0 / 0	TH-101	Thickener	407,000	407,000
		Surface Area = 564.56 m2		
2 / 0 / 0	SMSX-102	Solids Mixer-Settler Extractor	2,712,000	5,424,000
		Rated Throughput = 144352.07 L/h		
2 / 0 / 0	CL-101	Clarifier	937,000	1,874,000
		Surface Area = 2262.94 m2		
1 / 0 / 0	CL-102	Clarifier	344,000	344,000
		Surface Area = 426.63 m2		
1 / 0 / 0	DS-101	Disk-Stack Centrifuge	438,000	438,000
		Sigma Factor = 160117.27 m2		
3 / 0 / 0	R-101	Stirred Reactor	353,000	1,059,000
		Vessel Volume = 38133.44 L		
2 / 0 / 0	R-102	Stirred Reactor	328,000	656,000
		Vessel Volume = 29463.49 L		
1 / 0 / 0	HX-102	Heat Exchanger	5,000	5,000
		Heat Exchange Area = 10.43 m2		
2 / 0 / 0	HX-103	Heat Exchanger	20,000	40,000
		Heat Exchange Area = 69.98 m2		
1 / 0 / 0	HX-104	Heat Exchanger	2,000	2,000
		Heat Exchange Area = 3.02 m2		
1 / 0 / 0	HX-101	Heat Exchanger	4,000	4,000
		Heat Exchange Area = 6.75 m2		
1 / 0 / 0	HX-105	Heat Exchanger	2,000	2,000
		Heat Exchange Area = 1.93 m2		
1 / 0 / 0	HX-106	Heat Exchanger	2,000	2,000
		Heat Exchange Area = 3.20 m2		
		Unlisted Equipment		3,132,000
			TOTAL	15,659,000

3. FIXED CAPITAL ESTIMATE SUMMARY (2010 prices in $)

3A. Total Plant Direct Cost (TPDC) (physical cost)	
1. Equipment Purchase Cost	15,659,000
2. Installation	6,115,000
3. Process Piping	5,481,000
4. Instrumentation	6,264,000
5. Insulation	470,000
6. Electrical	1,566,000
7. Buildings	7,047,000
8. Yard Improvement	2,349,000
9. Auxiliary Facilities	6,264,000
TPDC	51,214,000

3B. Total Plant Indirect Cost (TPIC)	
10. Engineering	12,803,000
11. Construction	17,925,000
TPIC	30,728,000

3C. Total Plant Cost (TPC = TPDC+TPIC)	
TPC	81,942,000

3D. Contractor's Fee & Contingency (CFC)	
12. Contractor's Fee	4,097,000
13. Contingency	8,194,000
CFC = 12+13	12,291,000

3E. Direct Fixed Capital Cost (DFC = TPC+CFC)	
DFC	94,233,000

4. LABOR COST - PROCESS SUMMARY

Labor Type	Unit Cost ($/h)	Annual Amount (h)	Annual Cost ($)	%
Operator	69.00	96,000	6,624,000	100.00
TOTAL		96,000	6,624,000	100.00

5. MATERIALS COST - PROCESS SUMMARY

Bulk Material	Unit Cost ($/kg)	Annual Amount (kg)	Annual Cost ($)	%
1,1,2-TriChEth	0.990	671,999,934	665,279,935	77.13
Ethyl Alcohol	0.750	4,966,080	3,724,560	0.43
Water	0.000	2,191,264,106	0	0.00
Inert Solids	0.000	67,199,993	0	0.00
Sucrose	0.800	241,919,976	193,535,981	22.44
Ca Hydroxide	0.000	26,879,997	0	0.00
Carb. Dioxide	0.000	21,680,000	0	0.00
Yeast	2.300	6,720	15,456	0.00
TOTAL		3,225,916,806	862,555,932	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 (2010 prices) - PROCESS SUMMARY

THE CONSUMABLES COST IS ZERO.

7. WASTE TREATMENT/DISPOSAL COST (2010 prices) - PROCESS SUMMARY

THE TOTAL WASTE TREATMENT/DISPOSAL COST IS ZERO.

8. UTILITIES COST (2010 prices) - PROCESS SUMMARY

Utility	Annual Amount	Reference Units	Annual Cost ($)	%
Electricity	9,619,990	kWh	961,999	4.58
Steam	1,264,859,153	kg	15,178,310	72.30
Steam (High P)	0	kg	0	0.00
Cooling Water	93,595,250,084	kg	4,679,763	22.29
Chilled Water	436,946,888	kg	174,779	0.83
TOTAL			20,994,850	100.00

9. ANNUAL OPERATING COST (2010 prices) - PROCESS SUMMARY

Cost Item	$	%
Raw Materials	862,556,000	94.99
Labor-Dependent	6,624,000	0.73
Facility-Dependent	17,875,000	1.97
Consumables	0	0.00
Waste Treatment/Disposal	0	0.00
Utilities	20,995,000	2.31
Transportation	0	0.00
Miscellaneous	0	0.00
Advertising/Selling	0	0.00
Running Royalties	0	0.00
Failed Product Disposal	0	0.00
TOTAL	908,050,000	100.00

Economic Evaluation Report

April 18, 2010

for Sugar Beets backup from 4-18-10

1. EXECUTIVE SUMMARY (2010 prices)

Total Capital Investment	1,952,000 $
Capital Investment Charged to This Project	1,952,000 $
Operating Cost	2,328,000 $/yr

THE MAIN REVENUE STREAM HAS NOT BEEN IDENTIFIED. PRICING AND PRODUCTION/PROCESSING UNIT COST DATA HAVE NOT BEEN PRINTED

Main Revenue	0 $/yr
Gross Margin	- 1.00 %
Return On Investment	- 111.02 %
Payback Time	- 1.00 years
IRR (After Taxes)	Out of search interval (0-1000%)
NPV (at 7.0% Interest)	0 $

MT = Metric Ton (1000 kg)

2. MAJOR EQUIPMENT SPECIFICATION AND FOB COST (2010 prices)

Quantity/ Standby/ Staggered	Name	Description	Unit Cost ($)	Cost ($)
1 / 0 / 0	SR-101	Shredder	217,000	217,000
		Size/Capacity = 38640.00 kg/h		
		Unlisted Equipment		54,000
			TOTAL	271,000

3. FIXED CAPITAL ESTIMATE SUMMARY (2010 prices in $)

3A. Total Plant Direct Cost (TPDC) (physical cost)	
1. Equipment Purchase Cost	271,000
2. Installation	136,000
3. Process Piping	95,000
4. Instrumentation	109,000
5. Insulation	8,000
6. Electrical	27,000
7. Buildings	122,000
8. Yard Improvement	41,000
9. Auxiliary Facilities	109,000
TPDC	917,000

3B. Total Plant Indirect Cost (TPIC)	
10. Engineering	229,000
11. Construction	321,000
TPIC	550,000

3C. Total Plant Cost (TPC = TPDC+TPIC)	
TPC	1,467,000

3D. Contractor's Fee & Contingency (CFC)	
12. Contractor's Fee	73,000
13. Contingency	147,000
CFC = 12+13	220,000

3E. Direct Fixed Capital Cost (DFC = TPC+CFC)	
DFC	1,687,000

4. LABOR COST - PROCESS SUMMARY

Labor Type	Unit Cost ($/h)	Annual Amount (h)	Annual Cost ($)	%
Operator	69.00	1,143	78,857	100.00
TOTAL		1,143	78,857	100.00

5. MATERIALS COST - PROCESS SUMMARY

THE COST OF ALL MATERIALS IS ZERO. PLEASE CHECK THE MATERIAL BALANCES AND THE PURCHASING COST OF RAW MATERIALS.

6. VARIOUS CONSUMABLES COST (2010 prices) - PROCESS SUMMARY

THE CONSUMABLES COST IS ZERO.

7. WASTE TREATMENT/DISPOSAL COST (2010 prices) - PROCESS SUMMARY

THE TOTAL WASTE TREATMENT/DISPOSAL COST IS ZERO.

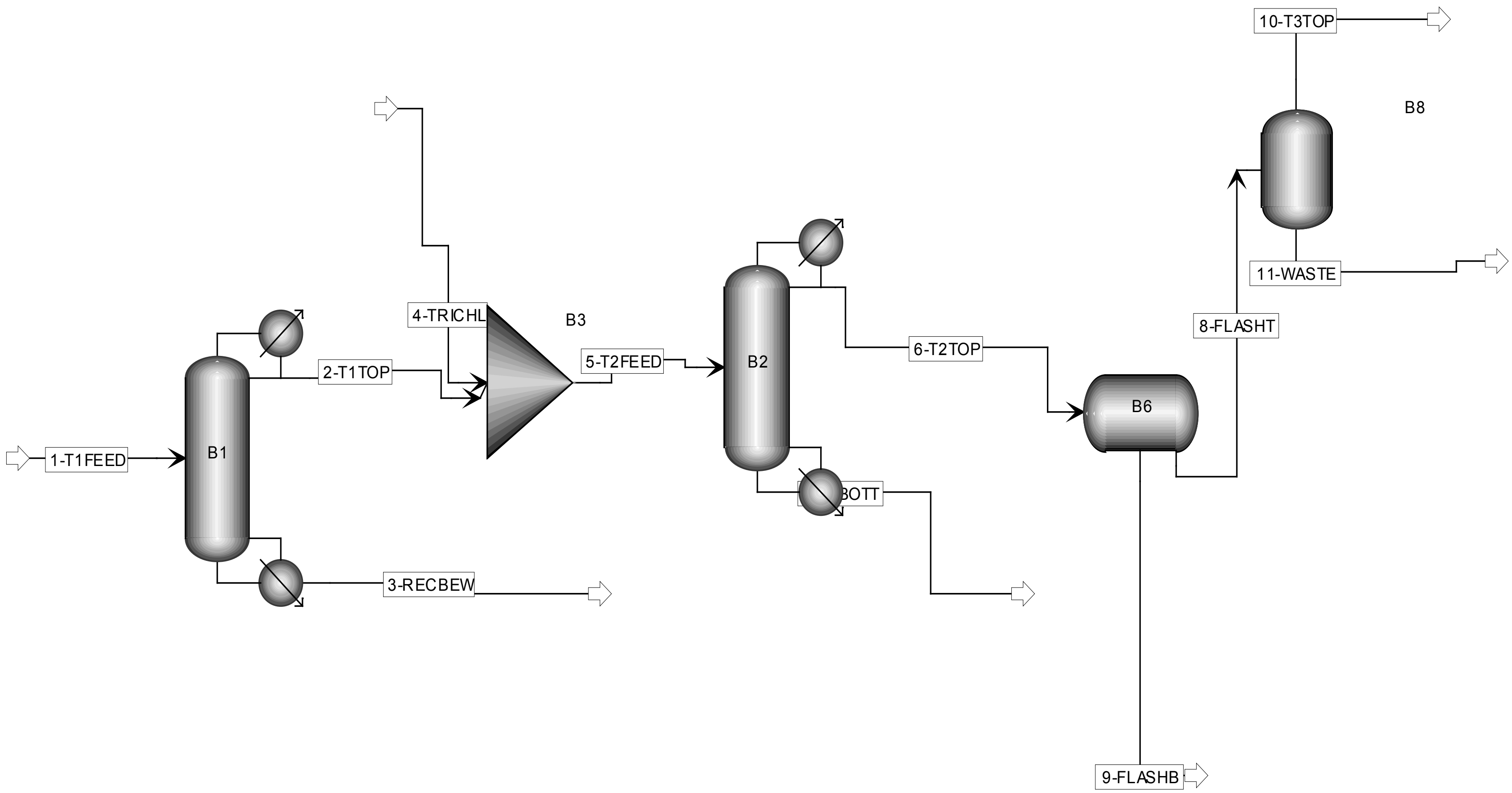
8. UTILITIES COST (2010 prices) - PROCESS SUMMARY

Utility	Annual Amount	Reference Units	Annual Cost ($)	%
Electricity	19,319,998	kWh	1,932,000	100.00
Steam	0	kg	0	0.00
Steam (High P)	0	kg	0	0.00
Cooling Water	0	kg	0	0.00
Chilled Water	0	kg	0	0.00
TOTAL			1,932,000	100.00

9. ANNUAL OPERATING COST (2010 prices) - PROCESS SUMMARY

Cost Item	$	%
Raw Materials	0	0.00
Labor-Dependent	79,000	3.39
Facility-Dependent	317,000	13.61
Consumables	0	0.00
Waste Treatment/Disposal	0	0.00
Utilities	1,932,000	83.00
Transportation	0	0.00
Miscellaneous	0	0.00
Advertising/Selling	0	0.00
Running Royalties	0	0.00
Failed Product Disposal	0	0.00
TOTAL	2,328,000	100.00

13. Appendix E – Sugar Beets Aspen Simulation Data


10-T3TOP
B8
4-TRICHL
B3
2-T1TOP
5-T2FEED
B2
6-T2TOP
8-FLASHT
11-WASTE
B6
1-T1FEED
B1
BOTT
3-RECBEW
9-FLASHB

Stream Summary Data

	1-T1FEED	2-T1TOP	3-RECBEW	4-TRICHL	5-T2FEED
Mole Flow mol/hr					
ETHANOL	387.701	387.701	0.000	16.041	403.742
WATER	10940.657	273.516	10667.141	41.520	315.037
SUCROSE	5.150452	7.50E-05	5.15037665	0	7.50E-05
TRICH-01	0	0	0	761.104283	761.104283
Mass Flow kg/hr					
ETHANOL	17861	17861	0.0001786	739	18599.9998
WATER	197099	4927.475	192171.525	748	5675.475
SUCROSE	1763	0.025661	1762.97434	0	0.02566054
TRICH-01	0	0	0	100000	100000
Total Flow kmol/hr	11333.51	661.2172	10672.291	818.665734	1479.88295
Total Flow kg/hr	216723	22788.5	193934.5	101487	124275.5
Total Flow l/min	3789.173	492.6703	3511.43924	1193.45048	1650.53959
Temperature K	322.15	352.3908	373.181043	320.15	330.63997
Pressure atm	1	1	1	1	1
Vapor Frac	0	0	0	0	0
Liquid Frac	1	1	1	1	1
Solid Frac	0	0	0	0	0
Enthalpy cal/mol	-68140.42	-65560.2	-67284.237	-13735.414	-36890.935
Enthalpy cal/gm	-3563.397	-1902.26	-3702.6777	-110.79954	-439.30031
Enthalpy cal/sec	-2.15E+08	-1.2E+07	-199465822	-3123531.4	-15165074
Entropy cal/mol-K	-40.20001	-59.2048	-36.269272	-41.501911	-49.373546
Entropy cal/gm-K	-2.102256	-1.71785	-1.9959122	-0.3347837	-0.5879443
Density mol/cc	0.04985	0.022368	0.05065487	0.01143275	0.01494342
Density gm/cc	0.953256	0.770918	0.92048912	1.41727706	1.25489769
Average MW	19.12232	34.46447	18.1717777	123.966346	83.9765742
Liq Vol 60F l/min	3687.786	458.1764	3229.61002	1171.43884	1629.61528

	6-T2TOP	7-T2BOTT	8-FLASHT	9-FLASHB	10-T3TOP
Mole Flow mol/hr					
ETHANOL	20.187	383.555	10.875	9.312	6.746
WATER	315.005	0.032	312.632	2.373	39.224
SUCROSE	0	7.50E-05	0	0	0
TRICH-01	761.1043	9.88E-07	0.02908909	761.075291	0.02907751
Mass Flow kg/hr					
ETHANOL	930	17670	500.99548	429.004412	310.768356
WATER	5674.907	0.567548	5632.15709	42.7486404	706.632615
SUCROSE	0	0.025661	0	0	0
TRICH-01	100000	0.00013	3.82195894	99996.1908	3.82043796
Total Flow kmol/hr	1096.297	383.5863	323.536199	772.760408	45.9988575
Total Flow kg/hr	106604.9	17670.59	6136.97453	100467.944	1021.22141
Total Flow l/min	1264.381	401.5395	107.262439	1181.59009	23222.518
Temperature K	328.6804	351.4601	320.15	320.15	369.15
Pressure atm	1	1	3	3	1
Vapor Frac	0	0	0	0	1
Liquid Frac	1	1	1	1	0
Solid Frac	0	0	0	0	0
Enthalpy cal/mol	-27070.42	-64531.6	-67804.074	-10640.031	-56815.173
Enthalpy cal/gm	-278.385	-1400.83	-3574.5745	-81.838987	-2559.1248
Enthalpy cal/sec	-8243670	-6875952	-6093631.3	-2283943	-725953.64
Entropy cal/mol-K	-40.32843	-77.482	-38.988888	-41.55956	-14.090518
Entropy cal/gm-K	-0.414727	-1.68195	-2.0554618	-0.31966	-0.6346789
Density mol/cc	0.014451	0.015921	0.05027174	0.0109	3.30E-05
Density gm/cc	1.405232	0.733452	0.95357618	1.41712911	0.00073292
Average MW	97.24094	46.06679	18.9684324	130.011764	22.2010168
Liq Vol 60F l/min	1257.732	371.8834	104.637583	1153.09443	18.3838592

Equipment Reports

	Tower 1	Tower 2
Minimum reflux ratio:	3.289	1.606
Actual reflux ratio:	3.782	8.032
Minimum number of stages:	1.12E+01	28.6590725
Number of actual stages:	2.31E+01	32.6990552
Feed stage:	8.715907	-29.774353
Number of actual stages above feed:	7.715907	-30.774353
Reboiler heating required (cal/sec):	11345407	23022755.5
Condenser cooling required (cal/sec):	8333325	22977302.8
Distillate temperature (K):	352.3908	328.68036
Bottom temperature (K):	373.181	351.460089
Distillate to feed fraction:	0.058342	0.74079954

	Decanter	Flash Unit
Temperature (K)	320.150	369.150
Pressure (atm)	3	1
Heat Duty (cal/sec)	-133904.72	2.08E+05

13. Appendix F – Simplified Process Flow Diagram of Corn Process from "Modeling the Process and Costs of Fuel Ethanol Production by the Corn Dry-Grind Process (Kwiatkowski, 2006)

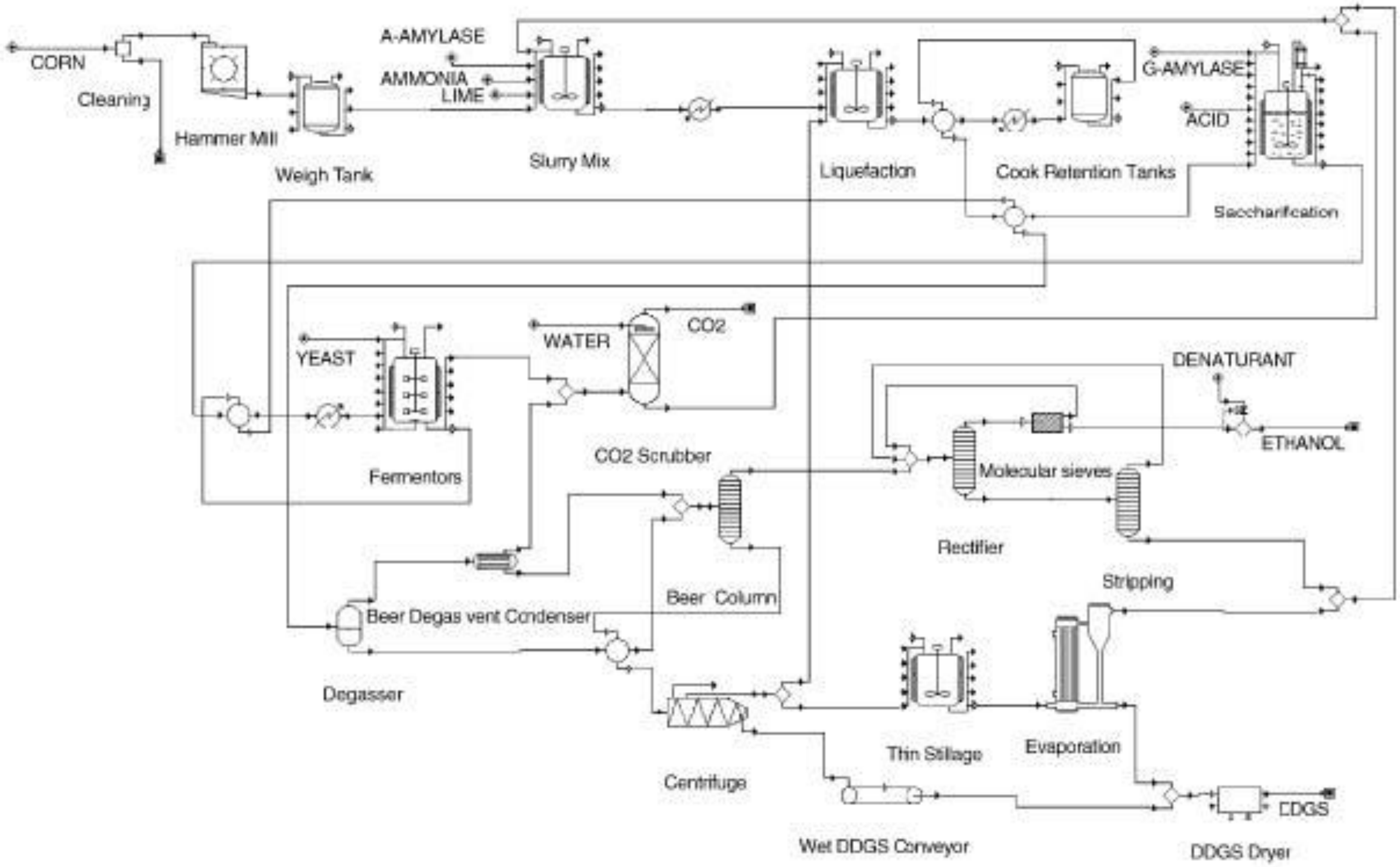


Fig. 1. Simplified flow diagram of the dry-grind ethanol from corn process.

15. Appendix G – Corn Material and Energy Balance from SuperPro Designer (Kwiatkowski, 2006)

Materials & Streams Report
for Corn_to_EtOH_40MGY_v7_5

April 26, 2010

1. OVERALL PROCESS DATA

Annual Operating Time	7,920.00h
Annual Throughput	119,171,455.05kg MP
Operating Days per Year	330.00

MP = Main Product = Total flow of stream ETHANOL

2.1 STARTING MATERIAL REQUIREMENTS (per Section)

Section	Starting Material	Active Product	Amount Needed (kg Sin/kg MP)	Molar Yield (%)	Mass Yield (%)	Gross Mass Yield (%)
Main Section	(none)	(none)	0.00	Unknown	Unknown	Unknown
Grain Handling & Milling	(none)	(none)	0.00	Unknown	Unknown	Unknown
Starch to Sugar Conversion	(none)	(none)	0.00	Unknown	Unknown	Unknown
Fermentation	(none)	(none)	0.00	Unknown	Unknown	Unknown
Ethanol Processing	(none)	(none)	0.00	Unknown	Unknown	Unknown
CoProduct Processing	(none)	(none)	0.00	Unknown	Unknown	Unknown
Common Suport Systems	(none)	(none)	0.00	Unknown	Unknown	Unknown

Sin = Section Starting Material, Aout = Section Active Product

2.2 BULK MATERIALS (Entire Process)

Material	kg/yr	kg/h	kg/kg MP
Corn	367,097,738	46,350.724	3.080
Lime	438,190	55.327	0.004
Liq. Ammonia	733,337	92.593	0.006
Alpha-Amylase	257,139	32.467	0.002
Glucoamylase	371,408	46.895	0.003
Sulfuric Acid	733,337	92.593	0.006
Caustic	18,423,742	2,326.230	0.155
Yeast	96,466	12.180	0.001
Water	159,933,643	20,193.642	1.342
Octane	2,383,429	300.938	0.020
Air	275,269,874	34,756.297	2.310
TOTAL	825,738,300	104,259.886	6.929

2.3 BULK MATERIALS (per Section)

SECTIONS IN: Main Branch

Grain Handling & Milling

Material	kg/yr	kg/h	kg/kg MP
Corn	367,097,738	46,350.724	3.080
TOTAL	367,097,738	46,350.724	3.080

Starch to Sugar Conversion

Material	kg/yr	kg/h	kg/kg MP
Lime	438,190	55.327	0.004
Liq. Ammonia	733,337	92.593	0.006
Alpha-Amylase	257,139	32.467	0.002
Glucoamylase	371,408	46.895	0.003
Sulfuric Acid	733,337	92.593	0.006
Caustic	18,423,742	2,326.230	0.155
TOTAL	20,957,152	2,646.105	0.176

Fermentation

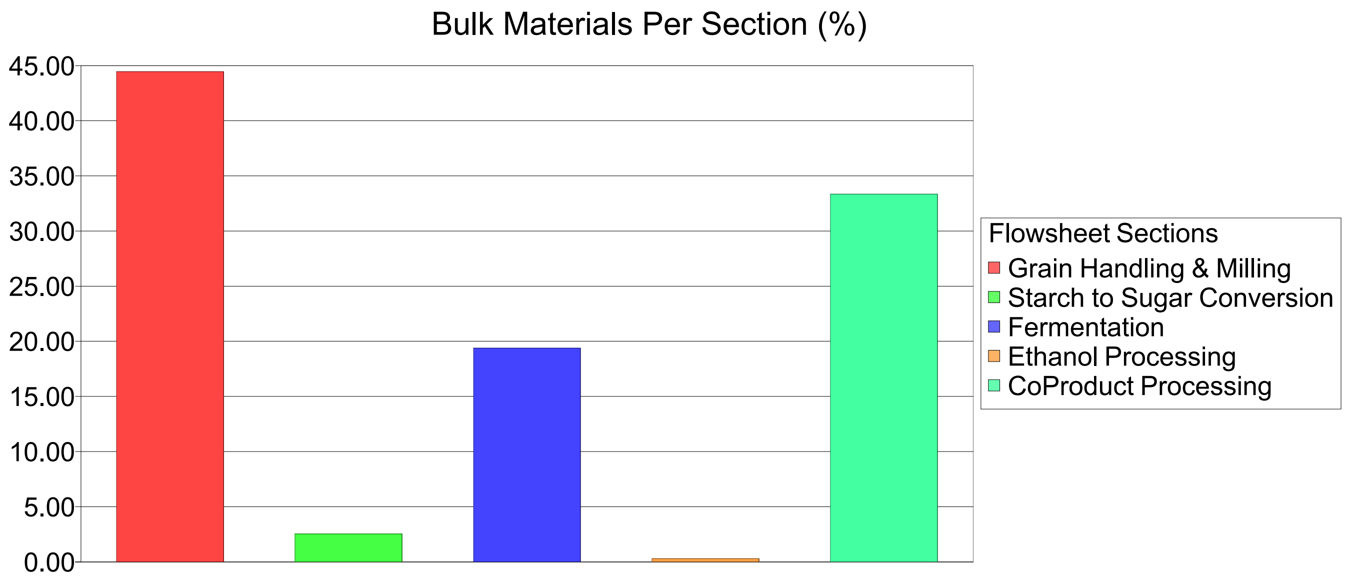
Material	kg/yr	kg/h	kg/kg MP
Yeast	96,466	12.180	0.001
Water	159,933,643	20,193.642	1.342
TOTAL	160,030,108	20,205.822	1.343

Ethanol Processing

Material	kg/yr	kg/h	kg/kg MP
Octane	2,383,429	300.938	0.020
TOTAL	2,383,429	300.938	0.020

CoProduct Processing

Material	kg/yr	kg/h	kg/kg MP
Air	275,269,874	34,756.297	2.310
TOTAL	275,269,874	34,756.297	2.310



2.4 BULK MATERIALS (per Material)

Corn				
Corn	**% Total**	**kg/yr**	**kg/h**	**kg/kg MP**
Grain Handling & Milling (Main Branch)				
101MH	100.00	367,097,738	46,350.724	3.080
TOTAL	100.00	367,097,738	46,350.724	3.080

Lime				
Lime	**% Total**	**kg/yr**	**kg/h**	**kg/kg MP**
Starch to Sugar Conversion (Main Branch)				
305V	100.00	438,190	55.327	0.004
TOTAL	100.00	438,190	55.327	0.004

Liq. Ammonia				
Liq. Ammonia	**% Total**	**kg/yr**	**kg/h**	**kg/kg MP**
Starch to Sugar Conversion (Main Branch)				
303V	100.00	733,337	92.593	0.006
TOTAL	100.00	733,337	92.593	0.006

Alpha-Amylase

Alpha-Amylase	% Total	kg/yr	kg/h	kg/kg MP
Starch to Sugar Conversion (Main Branch)				
301V	100.00	257,139	32.467	0.002
TOTAL	100.00	257,139	32.467	0.002

Glucoamylase

Glucoamylase	% Total	kg/yr	kg/h	kg/kg MP
Starch to Sugar Conversion (Main Branch)				
317V	100.00	371,408	46.895	0.003
TOTAL	100.00	371,408	46.895	0.003

Sulfuric Acid

Sulfuric Acid	% Total	kg/yr	kg/h	kg/kg MP
Starch to Sugar Conversion (Main Branch)				
319V	100.00	733,337	92.593	0.006
TOTAL	100.00	733,337	92.593	0.006

Caustic

Caustic	% Total	kg/yr	kg/h	kg/kg MP
Starch to Sugar Conversion (Main Branch)				
306P	100.00	18,423,742	2,326.230	0.155
TOTAL	100.00	18,423,742	2,326.230	0.155

Yeast

Yeast	% Total	kg/yr	kg/h	kg/kg MP
Fermentation (Main Branch)				
403V	100.00	96,466	12.180	0.001
TOTAL	100.00	96,466	12.180	0.001

Water

Water	% Total	kg/yr	kg/h	kg/kg MP
Fermentation (Main Branch)				
403V	0.27	429,264	54.200	0.004
409V	99.73	159,504,379	20,139.442	1.338
TOTAL	100.00	159,933,643	20,193.642	1.342

Octane

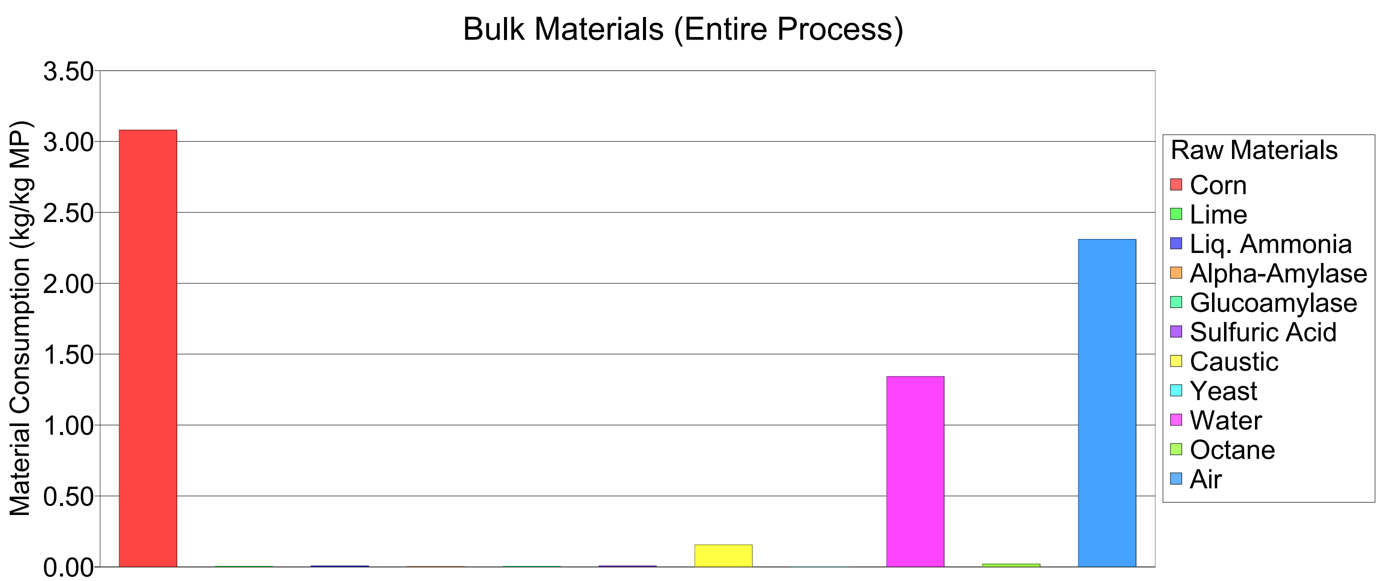
Octane	% Total	kg/yr	kg/h	kg/kg MP
Ethanol Processing (Main Branch)				
509V	100.00	2,383,429	300.938	0.020
TOTAL	100.00	2,383,429	300.938	0.020

Air				
Air	**% Total**	**kg/yr**	**kg/h**	**kg/kg MP**
CoProduct Processing (Main Branch)				
610D	99.80	274,720,484	34,686.930	2.305
611X	0.20	549,389	69.367	0.005
TOTAL	100.00	275,269,874	34,756.297	2.310

2.5 BULK MATERIALS: SECTION TOTALS (kg/kg MP)

Raw Material	Main Section	Grain Handling & Milling	Starch to Sugar Conversion	Fermentation
Corn	0.000	3.080	0.000	0.000
Lime	0.000	0.000	0.004	0.000
Liq. Ammonia	0.000	0.000	0.006	0.000
Alpha-Amylase	0.000	0.000	0.002	0.000
Glucoamylase	0.000	0.000	0.003	0.000
Sulfuric Acid	0.000	0.000	0.006	0.000
Caustic	0.000	0.000	0.155	0.000
Yeast	0.000	0.000	0.000	0.001
Water	0.000	0.000	0.000	1.342
Octane	0.000	0.000	0.000	0.000
Air	0.000	0.000	0.000	0.000
TOTAL	0.000	3.080	0.176	1.343

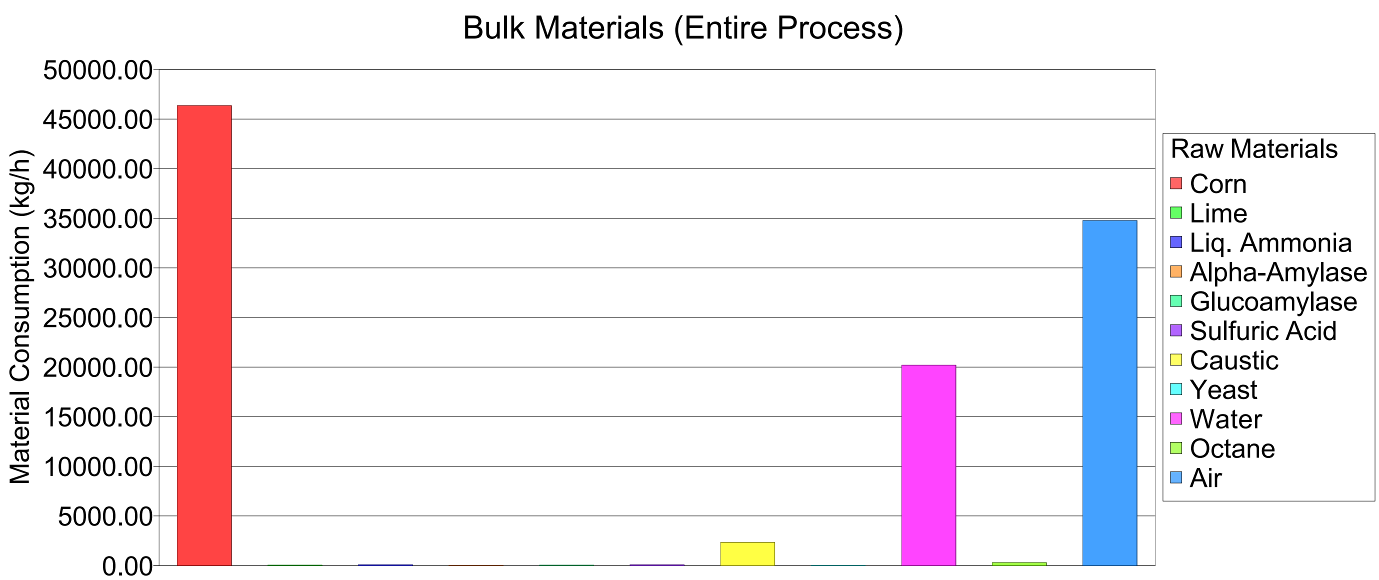
Raw Material	Ethanol Processing	CoProduct Processing	Common Suport Systems
Corn	0.000	0.000	0.000
Lime	0.000	0.000	0.000
Liq. Ammonia	0.000	0.000	0.000
Alpha-Amylase	0.000	0.000	0.000
Glucoamylase	0.000	0.000	0.000
Sulfuric Acid	0.000	0.000	0.000
Caustic	0.000	0.000	0.000
Yeast	0.000	0.000	0.000
Water	0.000	0.000	0.000
Octane	0.020	0.000	0.000
Air	0.000	2.310	0.000
TOTAL	0.020	2.310	0.000



2.6 BULK MATERIALS: SECTION TOTALS (kg/h)

Raw Material	Main Section	Grain Handling & Milling	Starch to Sugar Conversion	Fermentation
Corn	0.000	46,350.724	0.000	0.000
Lime	0.000	0.000	55.327	0.000
Liq. Ammonia	0.000	0.000	92.593	0.000
Alpha-Amylase	0.000	0.000	32.467	0.000
Glucoamylase	0.000	0.000	46.895	0.000
Sulfuric Acid	0.000	0.000	92.593	0.000
Caustic	0.000	0.000	2,326.230	0.000
Yeast	0.000	0.000	0.000	12.180
Water	0.000	0.000	0.000	20,193.642
Octane	0.000	0.000	0.000	0.000
Air	0.000	0.000	0.000	0.000
TOTAL	0.000	46,350.724	2,646.105	20,205.822

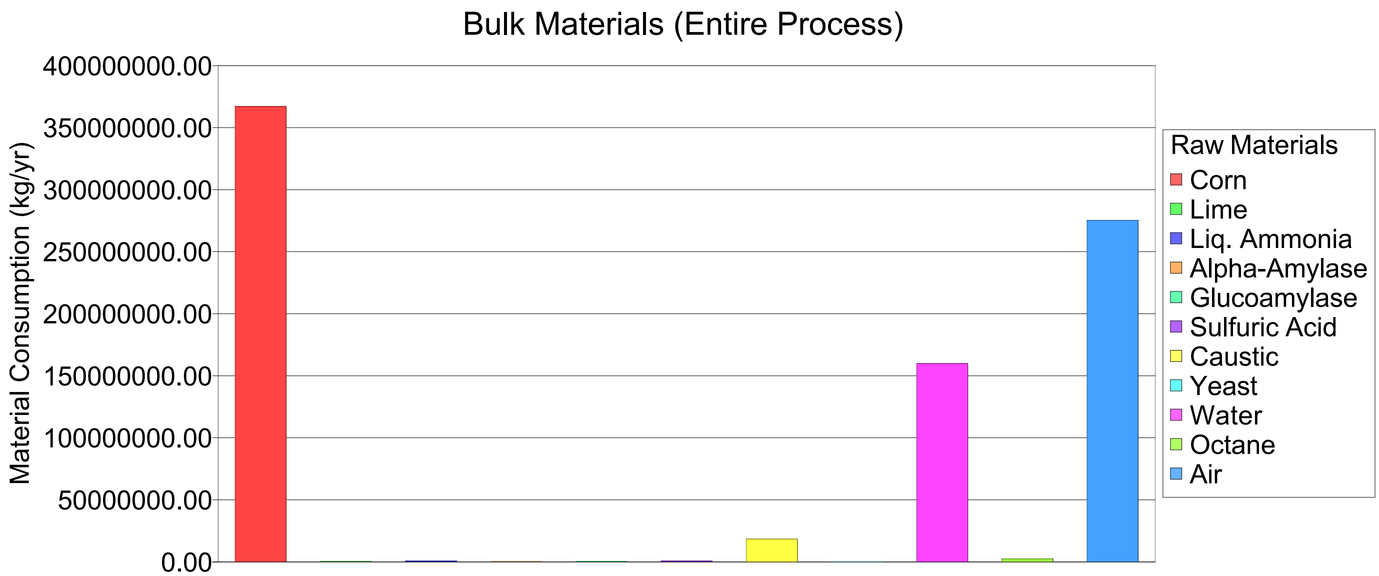
Raw Material	Ethanol Processing	CoProduct Processing	Common Suport Systems
Corn	0.000	0.000	0.000
Lime	0.000	0.000	0.000
Liq. Ammonia	0.000	0.000	0.000
Alpha-Amylase	0.000	0.000	0.000
Glucoamylase	0.000	0.000	0.000
Sulfuric Acid	0.000	0.000	0.000
Caustic	0.000	0.000	0.000
Yeast	0.000	0.000	0.000
Water	0.000	0.000	0.000
Octane	300.938	0.000	0.000
Air	0.000	34,756.297	0.000
TOTAL	300.938	34,756.297	0.000



2.7 BULK MATERIALS: SECTION TOTALS (kg/yr)

Raw Material	Main Section	Grain Handling & Milling	Starch to Sugar Conversion	Fermentation
Corn	0	367,097,738	0	0
Lime	0	0	438,190	0
Liq. Ammonia	0	0	733,337	0
Alpha-Amylase	0	0	257,139	0
Glucoamylase	0	0	371,408	0
Sulfuric Acid	0	0	733,337	0
Caustic	0	0	18,423,742	0
Yeast	0	0	0	96,466
Water	0	0	0	159,933,643
Octane	0	0	0	0
Air	0	0	0	0
TOTAL	0	367,097,738	20,957,152	160,030,108

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
Yeast	0	0	0
Water	0	0	0
Octane	2,383,429	0	0
Air	0	275,269,874	0
TOTAL	2,383,429	275,269,874	0



Bulk Materials (Entire Process)
Material Consumption (kg/yr)
400000000.00
350000000.00
300000000.00
250000000.00
200000000.00
150000000.00
100000000.00
50000000.00
0.00
Raw Materials
Corn
Lime
Liq. Ammonia
Alpha-Amylase
Glucoamylase
Sulfuric Acid
Caustic
Yeast
Water
Octane
Air

3. STREAM DETAILS

Stream Name	CIP	S-193	LIME	S-113
Source	INPUT	306P	INPUT	305V
Destination	306P	307V	305V	307V
Stream Properties				
Activity (U/ml)	0.00	0.00	0.00	0.00
Temperature (°C)	82.20	82.22	25.00	25.00
Pressure (bar)	1.01	3.29	1.01	1.01
Density (g/L)	980.70	980.70	1,173.66	1,173.66
Component Flowrates (kg/h averaged)				
Other Solids	116.312	116.312	55.327	55.327
Water	2,209.919	2,209.919	0.000	0.000
TOTAL (kg/h)	2,326.230	2,326.230	55.327	55.327
TOTAL (L/h)	2,371.998	2,372.020	47.141	47.141

Stream Name	AMMONIA	S-117	S-161	A-AMYLASE
Source	INPUT	303V	304P	INPUT
Destination	303V	304P	307V	301V
Stream Properties				
Activity (U/ml)	0.00	0.00	0.00	0.00
Temperature (°C)	25.00	25.00	25.13	25.00
Pressure (bar)	1.01	1.01	7.91	1.01
Density (g/L)	1,173.66	1,173.66	1,173.57	994.70
Component Flowrates (kg/h averaged)				
Other Solids	92.593	92.593	92.593	0.000
Water	0.000	0.000	0.000	32.467
TOTAL (kg/h)	92.593	92.593	92.593	32.467
TOTAL (L/h)	78.893	78.893	78.899	32.640

Stream Name	S-119	S-120	ACID	S-121
Source	**301V**	**302P**	**INPUT**	**319V**
Destination	**302P**	**307V**	**319V**	**320P**
Stream Properties				
Activity (U/ml)	0.00	0.00	0.00	0.00
Temperature (°C)	25.00	25.08	21.00	21.00
Pressure (bar)	1.01	4.46	1.01	1.01
Density (g/L)	994.70	994.67	1,832.36	1,832.36
Component Flowrates (kg/h averaged)				
Sulfuric Acid	0.000	0.000	92.593	92.593
Water	32.467	32.467	0.000	0.000
TOTAL (kg/h)	32.467	32.467	92.593	92.593
TOTAL (L/h)	32.640	32.641	50.532	50.532

Stream Name	S-126	G-AMYLASE	S-168	S-169
Source	**320P**	**INPUT**	**317V**	**318P**
Destination	**321V**	**317V**	**318P**	**321V**
Stream Properties				
Activity (U/ml)	0.00	0.00	0.00	0.00
Temperature (°C)	21.27	21.00	21.00	21.00
Pressure (bar)	7.91	1.01	1.01	1.01
Density (g/L)	1,832.11	996.16	996.16	996.16
Component Flowrates (kg/h averaged)				
Sulfuric Acid	92.593	0.000	0.000	0.000
Water	0.000	46.895	46.895	46.895
TOTAL (kg/h)	92.593	46.895	46.895	46.895
TOTAL (L/h)	50.539	47.076	47.076	47.076

Stream Name	YEAST	Water4	S-134	S-135
Source	**INPUT**	**INPUT**	**403V**	**404P**
Destination	**403V**	**403V**	**404P**	**405V**
Stream Properties				
Activity (U/ml)	0.00	0.00	0.00	0.00
Temperature (°C)	25.00	25.00	42.34	42.80
Pressure (bar)	1.01	1.01	1.01	11.36
Density (g/L)	994.70	994.70	988.39	988.22
Component Flowrates (kg/h averaged)				
Water	9.135	54.200	63.335	63.335
Yeast Dry Matte	3.045	0.000	3.045	3.045
TOTAL (kg/h)	12.180	54.200	66.380	66.380
TOTAL (L/h)	12.245	54.489	67.160	67.172

Stream Name	DENATURANT	S-114	S-188	CORN
Source	**INPUT**	**509V**	**510P**	**INPUT**
Destination	**509V**	**510P**	**MX-105**	**101MH**
Stream Properties				
Activity (U/ml)	0.00	0.00	0.00	0.00
Temperature (°C)	25.00	25.00	25.10	26.70
Pressure (bar)	1.01	1.01	4.46	1.01
Density (g/L)	699.19	699.19	699.12	1,335.34
Component Flowrates (kg/h averaged)				
Non-starch Poly	0.000	0.000	0.000	3,230.645
Octane	300.938	300.938	300.938	0.000
Oil	0.000	0.000	0.000	1,575.925
Other Solids	0.000	0.000	0.000	3,151.849
Protein - insol	0.000	0.000	0.000	2,285.091
Protein - solub	0.000	0.000	0.000	1,575.925
Starch	0.000	0.000	0.000	27,578.681
Water	0.000	0.000	0.000	6,952.609
TOTAL (kg/h)	300.938	300.938	300.938	46,350.724
TOTAL (L/h)	430.409	430.409	430.455	34,710.793

Stream Name	S-105	S-167	S-106	Trash
Source	**101MH**	**102V**	**103MH**	**103MH**
Destination	**102V**	**103MH**	**104M**	**OUTPUT**
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	3,230.645	3,230.645	3,220.954	9.692
Oil	1,575.925	1,575.925	1,571.197	4.728
Other Solids	3,151.849	3,151.849	3,142.394	9.456
Protein - insol	2,285.091	2,285.091	2,278.235	6.855
Protein - solub	1,575.925	1,575.925	1,571.197	4.728
Starch	27,578.681	27,578.681	27,495.945	82.736
Water	6,952.609	6,952.609	6,931.751	20.858
TOTAL (kg/h)	46,350.724	46,350.724	46,211.672	139.052
TOTAL (L/h)	34,710.793	34,710.793	34,606.661	104.132

Stream Name	S-109	S-110	S-111	S-112
Source	104M	105V	106W	107V
Destination	105V	106W	107V	307V
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	3,220.954	3,220.954	3,220.954	3,220.954
Oil	1,571.197	1,571.197	1,571.197	1,571.197
Other Solids	3,142.394	3,142.394	3,142.394	3,142.394
Protein - insol	2,278.235	2,278.235	2,278.235	2,278.235
Protein - solub	1,571.197	1,571.197	1,571.197	1,571.197
Starch	27,495.945	27,495.945	27,495.945	27,495.945
Water	6,931.751	6,931.751	6,931.751	6,931.751
TOTAL (kg/h)	46,211.672	46,211.672	46,211.672	46,211.672
TOTAL (L/h)	34,606.661	34,606.661	34,606.661	34,606.661

Stream Name	S-137	S-153	S-160	S-125
Source	**407P**	**315E**	**316E**	**316E**
Destination	**316E**	**316E**	**321V**	**412V**
Stream Properties				
Activity (U/ml)	0.00	0.00	0.00	0.00
Temperature (°C)	47.52	97.80	59.52	78.25
Pressure (bar)	6.18	2.00	2.00	6.18
Density (g/L)	975.98	516.01	1,036.45	239.13
Component Flowrates (kg/h averaged)				
Carb. Dioxide	0.000	14.038	14.038	0.000
Ethyl Alcohol	14,744.373	431.867	431.867	14,744.373
Non-starch Poly	3,291.497	3,291.497	3,291.497	3,291.497
Oil	1,724.069	1,724.069	1,724.069	1,724.069
Other Solids	5,040.753	4,525.380	4,525.381	5,040.754
Protein - insol	2,328.132	2,328.132	2,328.132	2,328.132
Protein - solub	2,561.333	2,139.664	2,139.664	2,561.333
Starch	275.018	27,501.839	27,501.839	275.018
Sulfuric Acid	119.005	26.412	26.412	119.005
Water	105,161.770	108,319.239	108,319.229	105,161.760
Yeast Dry Matte	624.694	45.753	45.753	624.694
TOTAL (kg/h)	135,870.645	150,347.890	150,347.882	135,870.636
TOTAL (L/h)	139,214.947	291,367.336	145,060.951	568,196.912

Stream Name	S-122	S-182	S-157	S-131
Source	**321V**	**322P**	**406P**	**401E**
Destination	**322P**	**401E**	**401E**	**402E**
Stream Properties				
Activity (U/ml)	0.00	0.00	0.00	0.00
Temperature (°C)	60.00	60.03	32.24	45.40
Pressure (bar)	1.01	3.91	4.46	3.91
Density (g/L)	971.18	1,013.42	983.26	1,020.38
Component Flowrates (kg/h averaged)				
Carb. Dioxide	14.038	14.038	0.000	14.038
Ethyl Alcohol	431.867	431.867	14,744.373	431.867
Glucose	30,252.023	30,252.023	0.000	30,252.023
Non-starch Poly	3,291.497	3,291.497	3,291.497	3,291.497
Oil	1,724.069	1,724.069	1,724.069	1,724.069
Other Solids	4,525.381	4,525.381	5,040.753	4,525.381
Protein - insol	2,328.132	2,328.132	2,328.132	2,328.132
Protein - solub	2,139.664	2,139.664	2,561.333	2,139.664
Starch	275.018	275.018	275.018	275.018
Sulfuric Acid	119.005	119.005	119.005	119.005
Water	105,340.922	105,340.922	105,161.827	105,340.922
Yeast Dry Matte	45.753	45.753	624.694	45.753
TOTAL (kg/h)	150,487.370	150,487.370	135,870.701	150,487.370
TOTAL (L/h)	154,953.563	148,494.786	138,184.296	147,481.073

Stream Name	S-124	S-136	S-199	S-129
Source	401E	402E	P-2	405V
Destination	407P	P-2	405V	MX-103
Stream Properties				
Activity (U/ml)	0.00	0.00	0.00	0.00
Temperature (°C)	47.50	32.20	32.20	32.20
Pressure (bar)	4.46	3.91	3.91	1.01
Density (g/L)	975.99	1,026.67	1,026.67	1.72
Component Flowrates (kg/h averaged)				
Carb. Dioxide	0.000	14.038	14.038	14,055.135
Ethyl Alcohol	14,744.373	431.867	431.867	385.820
Glucose	0.000	30,252.023	30,252.023	0.000
Non-starch Poly	3,291.497	3,291.497	3,291.497	0.000
Oil	1,724.069	1,724.069	1,724.069	0.000
Other Solids	5,040.753	4,525.381	4,525.381	0.000
Protein - insol	2,328.132	2,328.132	2,328.132	0.000
Protein - solub	2,561.333	2,139.664	2,139.664	0.000
Starch	275.018	275.018	275.018	0.000
Sulfuric Acid	119.005	119.005	119.005	0.000
Water	105,161.770	105,340.922	105,340.922	242.430
Yeast Dry Matte	624.694	45.753	45.753	0.000
TOTAL (kg/h)	135,870.645	150,487.370	150,487.370	14,683.384
TOTAL (L/h)	139,213.564	146,578.004	146,578.004	8,548,878.403

Stream Name	S-132	S-128	S-175	S-143
Source	**405V**	**412V**	**412V**	**408E**
Destination	**406P**	**408E**	**413E**	**MX-103**
Stream Properties				
Activity (U/ml)	0.00	0.00	0.00	0.00
Temperature (°C)	32.20	81.00	81.00	37.80
Pressure (bar)	1.01	1.01	1.01	1.00
Density (g/L)	983.28	0.92	946.13	821.56
Component Flowrates (kg/h averaged)				
Ethyl Alcohol	14,744.373	560.286	14,184.087	10.341
Non-starch Poly	3,291.497	0.000	3,291.497	0.000
Oil	1,724.069	0.000	1,724.069	0.000
Other Solids	5,040.753	0.000	5,040.754	0.000
Protein - insol	2,328.132	0.000	2,328.132	0.000
Protein - solub	2,561.333	0.000	2,561.333	0.000
Starch	275.018	0.000	275.018	0.000
Sulfuric Acid	119.005	0.000	119.005	0.000
Water	105,161.827	483.744	104,678.016	3.694
Yeast Dry Matte	624.694	0.000	624.694	0.000
TOTAL (kg/h)	135,870.701	1,044.030	134,826.605	14.035
TOTAL (L/h)	138,181.605	1,133,823.317	142,503.998	17.084

Stream Name	S-140	S-133	WATER	CO2
Source	**408E**	**MX-103**	**INPUT**	**409V**
Destination	**MX-104**	**409V**	**409V**	**OUTPUT**
Stream Properties				
Activity (U/ml)	0.00	0.00	0.00	0.00
Temperature (°C)	37.80	37.80	12.80	16.06
Pressure (bar)	1.00	1.00	1.00	1.00
Density (g/L)	861.94	1.78	999.15	1.84
Component Flowrates (kg/h averaged)				
Carb. Dioxide	0.000	14,055.135	0.000	14,041.079
Ethyl Alcohol	549.945	396.161	0.000	0.792
Water	480.050	246.124	20,139.442	100.911
TOTAL (kg/h)	1,029.995	14,697.420	20,139.442	14,142.783
TOTAL (L/h)	1,194.968	8,257,433.707	20,156.550	7,671,787.315

Stream Name	S-142	S-144	S-183	S-139
Source	409V	410P	502P	413E
Destination	410P	608T	413E	MX-104
Stream Properties				
Activity (U/ml)	0.00	0.00	0.00	0.00
Temperature (°C)	16.06	16.09	115.33	99.38
Pressure (bar)	1.00	3.90	4.46	1.01
Density (g/L)	726.02	907.81	8,055.98	939.43
Component Flowrates (kg/h averaged)				
Carb. Dioxide	14.055	14.055	0.000	0.000
Ethyl Alcohol	395.369	395.369	44.202	14,184.087
Non-starch Poly	0.000	0.000	3,291.497	3,291.497
Oil	0.000	0.000	1,724.069	1,724.069
Other Solids	0.000	0.000	5,040.754	5,040.754
Protein - insol	0.000	0.000	2,328.132	2,328.132
Protein - solub	0.000	0.000	2,561.333	2,561.333
Starch	0.000	0.000	275.018	275.018
Sulfuric Acid	0.000	0.000	119.005	119.005
Water	20,284.655	20,284.655	92,076.402	104,678.016
Yeast Dry Matte	0.000	0.000	624.694	624.694
TOTAL (kg/h)	20,694.079	20,694.079	108,085.107	134,826.605
TOTAL (L/h)	28,503.443	22,795.716	13,416.754	143,519.100

Stream Name	S-162	S-141	S-146	S-148
Source	**413E**	**MX-104**	**411P**	**501T**
Destination	**601V**	**411P**	**501T**	**MX-101**
Stream Properties				
Activity (U/ml)	0.00	0.00	0.00	0.00
Temperature (°C)	93.00	78.25	78.29	104.00
Pressure (bar)	4.46	1.00	4.10	1.01
Density (g/L)	938.76	16.57	56.12	0.86
Component Flowrates (kg/h averaged)				
Ethyl Alcohol	44.202	14,734.032	14,734.032	14,689.830
Non-starch Poly	3,291.497	3,291.497	3,291.497	0.000
Oil	1,724.069	1,724.069	1,724.069	0.000
Other Solids	5,040.754	5,040.754	5,040.754	0.000
Protein - insol	2,328.132	2,328.132	2,328.132	0.000
Protein - solub	2,561.334	2,561.333	2,561.333	0.000
Starch	275.018	275.018	275.018	0.000
Sulfuric Acid	119.005	119.005	119.005	0.000
Water	92,076.349	105,158.065	105,158.065	13,081.663
Yeast Dry Matte	624.694	624.694	624.694	0.000
TOTAL (kg/h)	108,085.054	135,856.600	135,856.600	27,771.493
TOTAL (L/h)	115,135.990	8,197,495.675	2,420,781.131	32,342,491.331

Stream Name	S-147	S-150	S-155	S-149
Source	501T	MX-101	503T	503T
Destination	502P	503T	504X	506P
Stream Properties				
Activity (U/ml)	0.00	0.00	0.00	0.00
Temperature (°C)	115.33	100.00	95.00	114.36
Pressure (bar)	1.01	1.01	1.01	1.01
Density (g/L)	8,178.74	1.01	1.69	902.00
Component Flowrates (kg/h averaged)				
Ethyl Alcohol	44.202	17,627.164	17,528.452	98.712
Non-starch Poly	3,291.497	0.000	0.000	0.000
Oil	1,724.069	0.000	0.000	0.000
Other Solids	5,040.754	0.000	0.000	0.000
Protein - insol	2,328.132	0.000	0.000	0.000
Protein - solub	2,561.333	0.000	0.000	0.000
Starch	275.018	0.000	0.000	0.000
Sulfuric Acid	119.005	0.000	0.000	0.000
Water	92,076.402	16,614.352	1,904.004	14,710.340
Yeast Dry Matte	624.694	0.000	0.000	0.000
TOTAL (kg/h)	108,085.107	34,241.516	19,432.456	14,809.052
TOTAL (L/h)	13,215.371	34,003,660.717	11,495,786.280	16,418.018

Stream Name	S-156	S-171	S-186	S-184
Source	504X	504X	505P	506P
Destination	505P	511V	MX-101	507T
Stream Properties				
Activity (U/ml)	0.00	0.00	0.00	0.00
Temperature (°C)	95.00	95.00	95.04	114.40
Pressure (bar)	1.01	1.01	4.46	4.46
Density (g/L)	2.51	1.53	11.03	2.50
Component Flowrates (kg/h averaged)				
Ethyl Alcohol	2,839.609	14,688.843	2,839.609	98.712
Water	1,846.884	57.120	1,846.884	14,710.340
TOTAL (kg/h)	4,686.493	14,745.963	4,686.493	14,809.052
TOTAL (L/h)	1,863,904.720	9,631,881.561	424,908.313	5,912,791.494

Stream Name	S-151	S-152	S-185	S-163
Source	**507T**	**507T**	**508P**	**601V**
Destination	**MX-101**	**508P**	**608T**	**602P**
Stream Properties				
Activity (U/ml)	0.00	0.00	0.00	0.00
Temperature (°C)	90.00	114.00	114.04	93.04
Pressure (bar)	2.00	1.01	4.46	1.00
Density (g/L)	52.83	902.00	2.50	784.35
Component Flowrates (kg/h averaged)				
Ethyl Alcohol	97.725	0.987	0.987	44.202
Non-starch Poly	0.000	0.000	0.000	3,291.497
Oil	0.000	0.000	0.000	1,724.069
Other Solids	0.000	0.000	0.000	5,040.754
Protein - insol	0.000	0.000	0.000	2,328.132
Protein - solub	0.000	0.000	0.000	2,561.334
Starch	0.000	0.000	0.000	275.018
Sulfuric Acid	0.000	0.000	0.000	119.005
Water	1,685.805	13,024.535	13,024.535	92,076.349
Yeast Dry Matte	0.000	0.000	0.000	624.694
TOTAL (kg/h)	1,783.530	13,025.522	13,025.522	108,085.054
TOTAL (L/h)	33,760.047	14,440.712	5,216,783.519	137,801.363

Stream Name	S-164	S-165	S-177	S-154
Source	602P	split	split	603
Destination	split	603	P-4	P-6
Stream Properties				
Activity (U/ml)	0.00	0.00	0.00	0.00
Temperature (°C)	93.08	93.08	93.08	93.65
Pressure (bar)	4.45	4.45	4.45	4.45
Density (g/L)	938.56	923.80	1,303.06	930.84
Component Flowrates (kg/h averaged)				
Ethyl Alcohol	44.202	44.202	0.000	0.988
Non-starch Poly	3,291.497	263.320	3,028.177	0.000
Oil	1,724.069	1,724.069	0.000	570.632
Other Solids	5,040.754	5,040.754	0.000	112.625
Protein - insol	2,328.132	186.251	2,141.881	0.000
Protein - solub	2,561.334	2,561.334	0.000	57.228
Starch	275.018	22.001	253.017	0.000
Sulfuric Acid	119.005	119.005	0.000	2.659
Water	92,076.349	92,076.349	0.000	2,057.251
Yeast Dry Matte	624.694	206.149	418.545	4.606
TOTAL (kg/h)	108,085.054	102,243.434	5,841.620	2,805.988
TOTAL (L/h)	115,159.941	110,676.946	4,482.995	3,014.466

Stream Name	S-174	S-197	S-198	S-196
Source	**603**	**603**	**P-6**	**FSP-103**
Destination	**P-4**	**P-6**	**FSP-103**	**310V**
Stream Properties				
Activity (U/ml)	0.00	0.00	0.00	0.00
Temperature (°C)	93.65	93.65	93.65	93.65
Pressure (bar)	4.45	4.45	4.45	4.45
Density (g/L)	924.32	923.09	923.35	923.35
Component Flowrates (kg/h averaged)				
Ethyl Alcohol	7.583	35.632	36.619	9.810
Non-starch Poly	0.000	263.320	263.320	70.543
Oil	1,153.437	0.000	570.632	152.872
Other Solids	864.731	4,063.398	4,176.023	1,118.754
Protein - insol	0.000	186.251	186.251	49.896
Protein - solub	439.392	2,064.714	2,121.942	568.467
Starch	0.000	22.001	22.001	5.894
Sulfuric Acid	20.415	95.931	98.590	26.412
Water	15,795.514	74,223.583	76,280.835	20,435.597
Yeast Dry Matte	35.364	166.179	170.785	45.753
TOTAL (kg/h)	18,316.436	81,121.009	83,926.997	22,484.000
TOTAL (L/h)	19,816.199	87,879.765	90,894.230	24,350.518

Stream Name	S-176	S-101	S-102	S-103
Source	**FSP-103**	**605V**	**606P**	**607Ev**
Destination	**605V**	**606P**	**607Ev**	**608T**
Stream Properties				
Activity (U/ml)	0.00	0.00	0.00	0.00
Temperature (°C)	93.65	93.71	93.74	79.00
Pressure (bar)	4.45	1.00	4.45	1.01
Density (g/L)	923.35	765.13	923.30	721.42
Component Flowrates (kg/h averaged)				
Ethyl Alcohol	26.809	26.809	26.809	26.208
Non-starch Poly	192.777	192.777	192.777	0.000
Oil	417.760	417.760	417.760	0.000
Other Solids	3,057.269	3,057.269	3,057.269	0.000
Protein - insol	136.354	136.354	136.354	0.000
Protein - solub	1,553.475	1,553.475	1,553.475	0.000
Starch	16.107	16.107	16.107	0.000
Sulfuric Acid	72.178	72.178	72.178	0.000
Water	55,845.238	55,845.238	55,845.238	45,494.933
Yeast Dry Matte	125.031	125.031	125.031	0.000
TOTAL (kg/h)	61,442.997	61,442.997	61,442.997	45,521.141
TOTAL (L/h)	66,543.712	80,304.322	66,547.083	63,098.973

Stream Name	S-172	S-104	S-187	S-127
Source	**607Ev**	**608T**	**609P**	**FSP-101**
Destination	**MX-102**	**609P**	**FSP-101**	**307V**
Stream Properties				
Activity (U/ml)	0.00	0.00	0.00	0.00
Temperature (°C)	79.00	100.00	100.04	100.04
Pressure (bar)	1.01	1.01	4.46	4.46
Density (g/L)	1,000.67	5.45	2.60	2.60
Component Flowrates (kg/h averaged)				
Carb. Dioxide	0.000	14.055	14.055	14.038
Ethyl Alcohol	0.601	422.564	422.564	422.057
Non-starch Poly	192.777	0.000	0.000	0.000
Oil	417.760	0.000	0.000	0.000
Other Solids	3,057.269	0.000	0.000	0.000
Protein - insol	136.354	0.000	0.000	0.000
Protein - solub	1,553.475	0.000	0.000	0.000
Starch	16.107	0.000	0.000	0.000
Sulfuric Acid	72.178	0.000	0.000	0.000
Water	10,350.305	78,804.123	78,804.123	78,709.558
Yeast Dry Matte	125.031	0.000	0.000	0.000
TOTAL (kg/h)	15,921.856	79,240.742	79,240.742	79,145.653
TOTAL (L/h)	15,911.135	14,551,488.393	30,485,762.061	30,449,179.146

Stream Name	PC	S-115	S-108	S-116
Source	FSP-101	307V	308P	P-1
Destination	OUTPUT	308P	P-1	310V
Stream Properties				
Activity (U/ml)	0.00	0.00	0.00	0.00
Temperature (°C)	100.04	100.20	100.31	87.00
Pressure (bar)	4.46	1.01	4.18	4.18
Density (g/L)	2.60	0.85	3.52	686.05
Component Flowrates (kg/h averaged)				
Carb. Dioxide	0.017	14.038	14.038	14.038
Ethyl Alcohol	0.507	422.057	422.057	422.057
Non-starch Poly	0.000	3,220.954	3,220.954	3,220.954
Oil	0.000	1,571.197	1,571.197	1,571.197
Other Solids	0.000	3,406.625	3,406.625	3,406.625
Protein - insol	0.000	2,278.235	2,278.235	2,278.235
Protein - solub	0.000	1,571.197	1,571.197	1,571.197
Starch	0.000	27,495.945	27,495.945	27,495.945
Water	94.565	87,883.694	87,883.694	87,883.694
TOTAL (kg/h)	95.089	127,863.942	127,863.942	127,863.942
TOTAL (L/h)	36,582.914	149,730,693.500	36,285,337.062	186,376.044

Stream Name	S-118	S-138	S-189	S-123
Source	**310V**	**311P**	**314V**	**312E**
Destination	**311P**	**312E**	**312E**	**313E**
Stream Properties				
Activity (U/ml)	0.00	0.00	0.00	0.00
Temperature (°C)	88.77	88.81	110.00	101.01
Pressure (bar)	1.01	4.18	2.00	4.18
Density (g/L)	350.20	711.99	668.60	645.48
Component Flowrates (kg/h averaged)				
Carb. Dioxide	14.038	14.038	14.038	14.038
Ethyl Alcohol	431.867	431.867	431.867	431.867
Non-starch Poly	3,291.497	3,291.497	3,291.497	3,291.497
Oil	1,724.069	1,724.069	1,724.069	1,724.069
Other Solids	4,525.380	4,525.380	4,525.380	4,525.380
Protein - insol	2,328.132	2,328.132	2,328.132	2,328.132
Protein - solub	2,139.664	2,139.664	2,139.664	2,139.664
Starch	27,501.839	27,501.839	27,501.839	27,501.839
Sulfuric Acid	26.412	26.412	26.412	26.412
Water	108,319.291	108,319.291	108,319.291	108,319.291
Yeast Dry Matte	45.753	45.753	45.753	45.753
TOTAL (kg/h)	150,347.942	150,347.942	150,347.942	150,347.942
TOTAL (L/h)	429,320.809	211,164.408	224,870.429	232,925.134

Stream Name	S-191	S-130	S-181	S-107
Source	**312E**	**313E**	**P-4**	**604MH**
Destination	**315E**	**314V**	**604MH**	**MX-102**
Stream Properties				
Activity (U/ml)	0.00	0.00	0.00	0.00
Temperature (°C)	97.80	110.00	93.60	93.60
Pressure (bar)	2.00	4.18	4.45	4.45
Density (g/L)	516.01	1,220.86	994.18	994.18
Component Flowrates (kg/h averaged)				
Carb. Dioxide	14.038	14.038	0.000	0.000
Ethyl Alcohol	431.867	431.867	7.583	7.583
Non-starch Poly	3,291.497	3,291.497	3,028.177	3,028.177
Oil	1,724.069	1,724.069	1,153.437	1,153.437
Other Solids	4,525.380	4,525.380	864.731	864.731
Protein - insol	2,328.132	2,328.132	2,141.881	2,141.881
Protein - solub	2,139.664	2,139.664	439.392	439.392
Starch	27,501.839	27,501.839	253.017	253.017
Sulfuric Acid	26.412	26.412	20.415	20.415
Water	108,319.239	108,319.291	15,795.514	15,795.514
Yeast Dry Matte	45.753	45.753	453.909	453.909
TOTAL (kg/h)	150,347.890	150,347.942	24,158.057	24,158.057
TOTAL (L/h)	291,367.336	123,148.832	24,299.365	24,299.365

Stream Name	S-166	HOT AIR	S-178	S-170
Source	**MX-102**	**INPUT**	**610D**	**610D**
Destination	**610D**	**610D**	**611X**	**612MH**
Stream Properties				
Activity (U/ml)	0.00	0.00	0.00	0.00
Temperature (°C)	87.69	104.00	70.00	70.00
Pressure (bar)	1.01	1.01	1.01	1.01
Density (g/L)	911.46	0.93	1.76	1,166.25
Component Flowrates (kg/h averaged)				
Ethyl Alcohol	8.183	0.000	7.752	0.431
Nitrogen	0.000	26,608.964	26,608.964	0.000
Non-starch Poly	3,220.954	0.000	0.000	3,220.954
Oil	1,571.197	0.000	0.000	1,571.197
Other Solids	3,922.000	0.000	0.000	3,922.000
Oxygen	0.000	8,077.966	8,077.966	0.000
Protein - insol	2,278.235	0.000	0.000	2,278.235
Protein - solub	1,992.866	0.000	0.000	1,992.866
Starch	269.124	0.000	0.000	269.124
Sulfuric Acid	92.593	0.000	0.000	92.593
Water	26,145.819	0.000	24,768.626	1,377.193
Yeast Dry Matte	578.941	0.000	0.000	578.941
TOTAL (kg/h)	40,079.913	34,686.930	59,463.308	15,303.535
TOTAL (L/h)	43,973.352	37,210,741.444	33,873,175.263	13,121.986

Stream Name	DDGS	S-180	EXHAUST	S-158
Source	**612MH**	**INPUT**	**611X**	**511V**
Destination	**OUTPUT**	**611X**	**OUTPUT**	**512P**
Stream Properties				
Activity (U/ml)	0.00	0.00	0.00	0.00
Temperature (°C)	70.00	25.00	76.05	42.00
Pressure (bar)	1.01	1.01	1.01	1.01
Density (g/L)	1,166.25	1.18	1.72	771.39
Component Flowrates (kg/h averaged)				
Carb. Dioxide	0.000	0.000	14.812	0.000
Ethyl Alcohol	0.431	0.000	0.000	14,688.843
Nitrogen	0.000	53.213	26,662.177	0.000
Non-starch Poly	3,220.954	0.000	0.000	0.000
Oil	1,571.197	0.000	0.000	0.000
Other Solids	3,922.000	0.000	0.000	0.000
Oxygen	0.000	16.154	8,077.966	0.000
Protein - insol	2,278.235	0.000	0.000	0.000
Protein - solub	1,992.866	0.000	0.000	0.000
Starch	269.124	0.000	0.000	0.000
Sulfuric Acid	92.593	0.000	0.000	0.000
Water	1,377.193	0.000	24,777.723	57.120
Yeast Dry Matte	578.941	0.000	0.000	0.000
TOTAL (kg/h)	15,303.535	69.367	59,532.677	14,745.963
TOTAL (L/h)	13,121.986	58,827.198	34,542,440.734	19,116.154

Stream Name	S-159	S-192	S-179	ETHANOL
Source	**512P**	**MX-105**	**513V**	**514P**
Destination	**MX-105**	**513V**	**514P**	**OUTPUT**
Stream Properties				
Activity (U/ml)	0.00	0.00	0.00	0.00
Temperature (°C)	42.08	41.77	41.77	41.81
Pressure (bar)	4.46	4.46	1.00	2.72
Density (g/L)	771.32	769.67	769.67	769.64
Component Flowrates (kg/h averaged)				
Ethyl Alcohol	14,688.843	14,688.843	14,688.843	14,688.843
Octane	0.000	300.938	300.938	300.938
Water	57.120	57.120	57.120	57.120
TOTAL (kg/h)	14,745.963	15,046.901	15,046.901	15,046.901
TOTAL (L/h)	19,117.873	19,549.804	19,549.804	19,550.686

4. OVERALL COMPONENT BALANCE (kg/h)

COMPONENT	IN	OUT	OUT-IN
Carb. Dioxide	0.000	14,055.908	14,055.908
Ethyl Alcohol	0.000	14,690.573	14,690.573
Nitrogen	26,662.177	26,662.177	0.000
Non-starch Poly	3,230.645	3,230.645	- 0.000
Octane	300.938	300.938	0.000
Oil	1,575.925	1,575.925	- 0.000
Other Solids	3,416.081	3,931.455	515.375
Oxygen	8,094.120	8,077.966	- 16.154
Protein - insol	2,285.091	2,285.091	- 0.000
Protein - solub	1,575.925	1,997.594	421.670
Starch	27,578.681	351.860	- 27,226.821
Sulfuric Acid	92.593	92.593	0.000
Water	29,444.666	26,428.370	- 3,016.296
Yeast Dry Matte	3.045	578.941	575.896
TOTAL	104,259.886	104,260.037	0.150

5. EQUIPMENT CONTENTS

This section will be skipped (overall process is continuous)

16. Appendix H – Corn Economic Evaluation Report from SuperPro Designer (Kwiatkowski, 2006)

Economic Evaluation Report

April 23, 2010

for Corn_to_EtOH_40MGY_v7_5

1. EXECUTIVE SUMMARY (2007 prices)

Total Capital Investment	58005000.00	$
Capital Investment Charged to This Project	58005000.00	$
Main Product Rate	119171455.05	kg MP/yr
Operating Cost	80003000.00	$/yr
Product Unit Cost	0.67	$/kg MP
Main Revenue	46534000.00	$/yr
Other Revenues	12648990.00	$/yr
Total Revenues	59183000.00	$/yr
Gross Margin	- 35.18	%
Return On Investment	- 25.90	%
Payback Time	- 1.00	years
IRR (After Taxes)	Out of search interval	(0-1000%)
NPV (at 5.0% Interest)	- 170,884,000	$

MP = Total Flow of Stream ETHANOL

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

Quantity/ Standby/ Staggered	Name	Description	Unit Cost ($)	Cost ($)
1 / 0 / 0	101MH	Belt Conveyor Belt Length = 100.00 m	121000.00	121000.00
1 / 0 / 0		Silo/Bin Vessel Volume = 18540.29 m3	979000.00	979000.00
1 / 0 / 0	104M	Grinder Size/Capacity = 46211.67 kg/h	98000.00	98000.00
1 / 0 / 0	105V	Receiver Tank Vessel Volume = 76.90 m3	33000.00	33000.00
1 / 0 / 0	106W	Hopper Vessel Volume = 100.93 m3	51000.00	51000.00
1 / 0 / 0	107V	Receiver Tank Vessel Volume = 76.90 m3	44000.00	44000.00
1 / 0 / 0	307V	Blending Tank Vessel Volume = 15.28 m3	69000.00	69000.00
1 / 0 / 0	305V	Hopper Vessel Volume = 4.02 m3	9000.00	9000.00
1 / 0 / 0	303V	Receiver Tank Vessel Volume = 8.77 m3	28000.00	28000.00
1 / 0 / 0	301V	Receiver Tank Vessel Volume = 12.19 m3	50000.00	50000.00
1 / 0 / 0	302P	Gear Pump Power = 0.20 kW	4000.00	4000.00
1 / 0 / 0	310V	Blending Tank Vessel Volume = 141.43 m3	161000.00	161000.00
1 / 0 / 0	321V	Stirred Reactor Vessel Volume = 52.13 m3	103000.00	103000.00
1 / 0 / 0	317V	Receiver Tank	84000.00	84000.00

		Vessel Volume = 17.57 m3		
1 / 0 / 0	319V	Receiver Tank	19000.00	19000.00
		Vessel Volume = 18.87 m3		
1 / 0 / 0	401E	Heat Exchanger	285000.00	285000.00
		Heat Exchange Area = 197.97 m2		
1 / 0 / 0	402E	Heat Exchanger	85000.00	85000.00
		Heat Exchange Area = 190.14 m2		
1 / 0 / 0	404P	Gear Pump	7000.00	7000.00
		Power = 0.06 kW		
1 / 0 / 0	405V	Fermentor	2812000.00	2812000.00
		Vessel Volume = 10451.09 m3		
1 / 0 / 0	409V	Absorber	91000.00	91000.00
		Absorber Volume = 13.41 m3		
1 / 0 / 0	410P	Centrifugal Pump	7000.00	7000.00
		Power = 3.21 HP-E		
1 / 0 / 0	413E	Heat Exchanger	331000.00	331000.00
		Heat Exchange Area = 232.86 m2		
1 / 0 / 0	608T	Receiver Tank	96000.00	96000.00
		Vessel Volume = 485.54 m3		
1 / 0 / 0	411P	Centrifugal Pump	16000.00	16000.00
		Power = 50.00 HP-E		
1 / 0 / 0	501T	Distillation Column	597000.00	597000.00
		Column Volume = 96.61 m3		
1 / 0 / 0	MX-101	Mixer	0.00	0.00
		Size/Capacity = 34241.52 kg/h		
1 / 0 / 0	503T	Distillation Column	254000.00	254000.00
		Column Volume = 113.57 m3		
1 / 0 / 0	507T	Distillation Column	168000.00	168000.00
		Column Volume = 3.73 m3		
1 / 0 / 0	511V	Flat Bottom Tank	93000.00	93000.00
		Vessel Volume = 481.39 m3		
1 / 0 / 0	509V	Flat Bottom Tank	34000.00	34000.00
		Vessel Volume = 339.24 m3		
1 / 0 / 0	513V	Flat Bottom Tank	308000.00	308000.00
		Vessel Volume = 3392.22 m3		
1 / 0 / 0	601V	Blending Tank	197000.00	197000.00
		Vessel Volume = 755.41 m3		
1 / 0 / 0	605V	Blending Tank	230000.00	230000.00
		Vessel Volume = 481.39 m3		
1 / 0 / 0	604MH	Belt Conveyor	56000.00	56000.00
		Belt Length = 100.00 m		
1 / 0 / 0	610D	Rotary Dryer	2278000.00	2278000.00
		Drying Area = 1244.73 m2		
1 / 0 / 0	612MH	Belt Conveyor	123000.00	123000.00
		Belt Length = 100.00 m		
1 / 0 / 0	FSP-101	Flow Splitter	0.00	0.00
		Size/Capacity = 79240.74 kg/h		
1 / 0 / 0	MX-103	Mixer	0.00	0.00
		Size/Capacity = 14697.42 kg/h		
1 / 0 / 0	MX-104	Mixer	0.00	0.00
		Size/Capacity = 135856.60 kg/h		
1 / 0 / 0	313E	Heat Exchanger	14000.00	14000.00
		Heat Exchange Area = 36.67 m2		
1 / 0 / 0	312E	Heat Exchanger	201000.00	201000.00
		Heat Exchange Area = 332.90 m2		
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 = 0.02 kW		
1 / 0 / 0	403V	Blending Tank	115000.00	115000.00
		Vessel Volume = 2.97 m3		
1 / 0 / 0	611X	Wet Air Oxidizer	925000.00	925000.00
		Vessel Volume = 13.33 m3		
1 / 0 / 0	514P	Gear Pump	40000.00	40000.00
		Power = 1.79 HP-E		
1 / 0 / 0	510P	Gear Pump	5000.00	5000.00
		Power = 5.00 HP-E		
1 / 0 / 0	314V	Receiver Tank	174000.00	174000.00
		Vessel Volume = 14.16 m3		
1 / 0 / 0	412V	Flash Drum	62000.00	62000.00
		Vessel Volume = 14.29 m3		
1 / 0 / 0	408E	Condenser	19000.00	19000.00
		Condensation Area = 58.70 m2		
1 / 0 / 0	316E	Heat Exchanger	614000.00	614000.00
		Heat Exchange Area = 405.68 m2		
1 / 0 / 0	308P	Centrifugal Pump	25000.00	25000.00
		Power = 3.55 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	406P	Centrifugal Pump	76000.00	76000.00
		Power = 25.34 HP-E		
1 / 0 / 0	407P	Centrifugal Pump	15000.00	15000.00
		Power = 12.76 HP-E		
1 / 0 / 0	502P	Centrifugal Pump	13000.00	13000.00
		Power = 50.00 HP-E		
1 / 0 / 0	506P	Centrifugal Pump	7000.00	7000.00
		Power = 20.00 HP-E		
1 / 0 / 0	508P	Centrifugal Pump	5000.00	5000.00
		Power = 10.00 HP-E		
1 / 0 / 0	505P	Centrifugal Pump	4000.00	4000.00
		Power = 0.35 HP-E		
1 / 0 / 0	606P	Centrifugal Pump	11000.00	11000.00
		Power = 20.00 HP-E		
1 / 0 / 0	602P	Centrifugal Pump	13000.00	13000.00
		Power = 50.00 HP-E		
1 / 0 / 0	609P	Centrifugal Pump	12000.00	12000.00
		Power = 20.00 HP-E		
1 / 0 / 0	512P	Centrifugal Pump	5000.00	5000.00
		Power = 10.00 HP-E		
1 / 0 / 0	315E	Heat Exchanger	51000.00	51000.00
		Heat Exchange Area = 179.12 m2		
1 / 0 / 0	306P	Gear Pump	4000.00	4000.00
		Power = 5.00 HP-E		
1 / 0 / 0	MX-105	Mixer	0.00	0.00
		Size/Capacity = 15046.90 kg/h		
1 / 0 / 0	MX-102	Mixer	0.00	0.00
		Size/Capacity = 40079.91 kg/h		
1 / 0 / 0	607Ev	Evaporator	3415000.00	3415000.00
		Evaporation Area = 664.18 m2		
1 / 0 / 0	103MH	Flow Splitter	61000.00	61000.00
		Size/Capacity = 46350.72 kg/h		
1 / 0 / 0	504X	Component Splitter	1718000.00	1718000.00

		Size/Capacity = 19432.46 kg/h		
1 / 0 / 0	603.00	Disk-Stack Centrifuge	852000.00	852000.00
		Throughput = 1735.13 L/min		
1 / 0 / 0	split	Component Splitter	0.00	0.00
		Size/Capacity = 108085.05 kg/h		
1 / 0 / 0	MX-106	Mixer	0.00	0.00
		Size/Capacity = 24158.06 kg/h		
1 / 0 / 0	FSP-103	Flow Splitter	0.00	0.00
		Size/Capacity = 83927.00 kg/h		
1 / 0 / 0	MX-107	Mixer	0.00	0.00
		Size/Capacity = 83927.00 kg/h		
6 / 0 / 0	HX-101	Heat Exchanger	23000.00	138000.00
		Heat Exchange Area = 98.85 m2		
		Unlisted Equipment		0.00
			TOTAL	18556000.00

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

Section Name	DFC ($)
Main Section	552000.00
Grain Handling & Milling	4161000.00
Starch to Sugar Conversion	4956000.00
Fermentation	11760000.00
Ethanol Processing	9752000.00
CoProduct Processing	24624000.00
Common Suport Systems	2200000.00
Plant DFC	58005000.00

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	367097738.00	50567713.00	92.96
Lime	0.09	438190.00	39437.00	0.07
Liq. Ammonia	0.22	733337.00	161334.00	0.30
Alpha-Amylase	2.25	257139.00	578562.00	1.06
Glucoamylase	2.25	371408.00	835669.00	1.54
Sulfuric Acid	0.11	733337.00	80667.00	0.15
Caustic	0.01	18423742.00	223296.00	0.41
Yeast	1.86	96466.00	179426.00	0.33
Water	0.00	159933643.00	7037.00	0.01
Octane	0.72	2383429.00	1722266.00	3.17
Air	0.00	275269874.00	0.00	0.00

TOTAL		825738300.00	54395407.00	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.

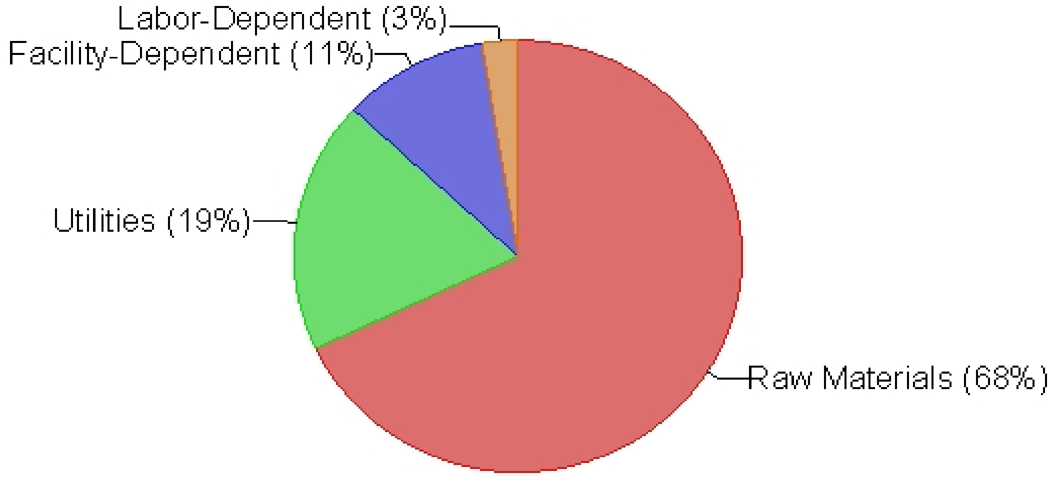
7. UTILITIES COST (2007 prices) - PROCESS SUMMARY

Utility	Annual Amount	Reference Units	Annual Cost ($)	%
Electricity	29429053.00	kWh	1471453.00	9.74
Steam	0.00	kg	0.00	0.00
Cooling Water	37311511255.00	kg	3731151.00	24.69
Chilled Water	0.00	kg	0.00	0.00
Natural Gas	11773735.00	kg	4140705.00	27.40
CT Water	5295747600.00	kg	370702.00	2.45
CT Water NoCost	13035843322.00	kg	0.00	0.00
CT Water 35Cout	361163231.00	kg	25281.00	0.17
CT Water 31Cout	3160638730.00	kg	221245.00	1.46
Well Water	0.00	kg	0.00	0.00
Steam 50 PSI	31661847.00	kg	675980.00	4.47
Steam 6258 BTU	32538865.00	kg	694705.00	4.60
Steam 2205 BTU	175987794.00	kg	3757339.00	24.87
Rectifier OH	85806902.00	kg	0.00	0.00
Steam (High P)	1110832.00	kg	22217.00	0.15
TOTAL			15110778.00	100.00

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

Cost Item	$	%
Raw Materials	54395000.00	67.99
Labor-Dependent	2059000.00	2.57
Facility-Dependent	8437000.00	10.55
Consumables	0.00	0.00
Utilities	15111000.00	18.89
Advertising/Selling	0.00	0.00
Running Royalties	0.00	0.00
Failed Product Disposal	0.00	0.00
TOTAL	80003000.00	100.00

Annual Operating Cost Breakdown (%)



9. PROFITABILITY ANALYSIS (2007 prices)

A.	Direct Fixed Capital	58005000.00	$
B.	Working Capital	0.00	$
C.	Startup Cost	0.00	$
D.	Up-Front R&D	0.00	$
E.	Up-Front Royalties	0.00	$
F.	Total Investment (A+B+C+D+E)	58005000.00	$
G.	Investment Charged to This Project	58005000.00	$
H.	Revenue/Credit Stream Flowrates		
	Total flow in DDGS (Other Revenue)	121203994.00	kg/yr
	Total flow of stream ETHANOL (Main Revenue)	119171455.00	kg/yr
I.	Annual Operating Cost		
	AOC	80003000.00	$/yr
J.	Product Unit Cost		
	ETHANOL	0.67	$/kg
K.	Selling / Processing Price		
	Total flow in DDGS	0.10	$/kg
	Total flow of stream ETHANOL	0.39	$/kg
L.	Revenues		
	DDGS (Other Revenue)	12649000.00	$/yr
	ETHANOL (Main Revenue)	46534000.00	$/yr
	Total Revenues	59183000.00	$/yr
M.	Gross Profit (L-I)	- 20,820,000	$/yr
N.	Taxes (40%)	0.00	$/yr
O.	Net Profit (M-N + Depreciation)	- 15,022,000	$/yr

Gross Margin Negative

Return On Investment Negative

Payback Time Negative

Appendix I – Cost Analysis Calculations

Corn

Feedstock

$$\left(\frac{\$4.80^{*}}{bushel}\right)\left(\frac{bushel}{25.4\ kg}\right)\left(\frac{372{,}451{,}731\ kg}{40{,}000{,}000\ gallons\ ethanol}\right)=\frac{\$1.76}{gallon\ ethanol}$$

*2009 USDA estimate

Utilities

$$\left(\frac{\$15{,}110{,}000}{1\ year}\right)\left(\frac{1\ year}{40{,}000{,}000\ gallons\ ethanol}\right)=\frac{\$0.38}{gallon\ ethanol}$$

Process Streams

$$\left(\frac{\$3{,}827{,}694}{1\ year}\right)\left(\frac{1\ year}{40{,}000{,}000\ gallons\ ethanol}\right)=\frac{\$0.10}{gallon\ ethanol}$$

Total

$$\frac{\$1.76}{gallon\ ethanol}+\frac{\$0.38}{gallon\ ethanol}+\frac{\$0.10}{gallon\ ethanol}=\frac{\$2.24}{gallon\ ethanol}$$

Sugar Beets

Feedstock

$$\left(\frac{\$45^{*}}{ton}\right)\left(\frac{ton}{907.18\ kg}\right)\left(\frac{1{,}344{,}000{,}000\ kg}{40{,}000{,}000\ gallons\ ethanol}\right)=\frac{\$1.66}{gallon\ ethanol}$$

*2004 U.C. Cooperative Case Study

Utilities

$$\left(\frac{\$22{,}927{,}000}{1\ year}\right)\left(\frac{1\ year}{40{,}000{,}000\ gallons\ ethanol}\right)=\frac{\$0.57}{gallon\ ethanol}$$

Process Streams

$$\left(\frac{\$2{,}431{,}699}{1\ year}\right)\left(\frac{1\ year}{40{,}000{,}000\ gallons\ ethanol}\right)=\frac{\$0.07}{gallon\ ethanol}$$

Total

$$\frac{\$1.66}{gallon\ ethanol} + \frac{\$0.57}{gallon\ ethanol} + \frac{\$0.07}{gallon\ ethanol} = \frac{\$2.30}{gallon\ ethanol}$$