

Corn Bioethanol Production Facility Design

A Technical Report submitted to the Department of Chemical Engineering

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On my honor as a University Student, I have neither given nor received unauthorized aid on this
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1. Executive Summary

Bioethanol is a promising alternative to conventional fossil-fuel usage due to the lower environmental strain associated with its production and consumption. Under current conditions, bioethanol is not an economically feasible substitute for conventional fossil-fuels; designing economically viable plant schematics is critical to addressing the global climate change crisis. This report aims to examine the process and design of an optimized production facility that converts corn into fuel-grade ethanol in a profitable manner.

The aim of the plant is to transition a raw corn starting material into two primary products: fuel grade anhydrous ethanol and Distillers Dried Grains and Solubles (DDGS). Plant design is centered around a goal of producing 150 million gallons of ethanol annually, with DDGS being sold as a coproduct to minimize waste and help recuperate costs.

This report includes in-depth analysis of the functionality and design of all equipment necessary in order to reach the plant's production goals. The design of every piece of equipment is first discussed and outlined before the final design considerations are established. The final design proposal clearly delineates the movement of the material streams throughout the process and the operating conditions of all equipment.

In order to adequately analyze the corn in its journey from stalk to barrel, the economics and energy requirements of the plant will serve as key indicators of efficiency. Additional factors, including the importance of worker health and safety, local and global environmental considerations, and overall social impact influenced the decision on the viability of the plant design.

Based on the findings of this report, it has ultimately been recommended that the current plant design not be implemented. As designed, the plant requires a fixed capital input of \$338.5 million. Assuming constant prices of corn and ethanol, after-tax profits are estimated to range from \$100-\$115 million for Years 1-20. The plant requires an input of 21,700 TJ of energy per year. The ethanol produced in the plant annually only contains approximately 11,200 TJ. Although the process outlined allows for the plant to function and turn an economic profit, it operates at an energy deficit. Considering one of the primary purposes of the plant is to provide net “green” energy to the world, using such massive quantities of non-renewable energy to create the ethanol largely defeats the purpose. Though the original design aim was to generate profit, the environmental implications complicate the decision making process. The report establishes that there is validity in converting the stored energy in corn for use in the transportation industry, but a No-Go decision has still ultimately been reached. Further design work should be conducted to investigate the potential for lower weight corn fermentation as a way to minimize external energy input.

2. Background

2.1 Background & Motivation

As concerns about the impact of fossil fuel usage on the environment mount, the United States has devoted significant resources to design alternative renewable energy solutions to meet an ever-growing demand for power. Biofuels, specifically bioethanol produced using corn, are a popular alternative energy solution. Ethanol as engine fuel is far from a recent phenomenon and was first used by Samuel Morey to power an early internal combustion engine in 1826 (Britannica, 2021). Ethanol proceeded to grow in popularity during the 1920s and 1930s as a common octane-boosting gasoline additive. Fuel shortages associated with WWII further increased the demand and usage of ethanol-gasoline blends (EIA, 2021). The modern-day ethanol industry started in the 1970s as scientists and politicians became concerned about the high cost and negative environmental effects of petroleum-based fuels. A useful oxygenate, ethanol reduces exhaust emissions including carbon monoxide and enhances the overall fuel combustion process. Corn became an obvious feedstock for ethanol production due to its abundance and accessibility.

Ethanol's popularity continued to grow as environmental concerns around petroleum fuel usage persisted through the remainder of the century. The first Renewable Fuels Standard (RFS) in 2005 required the production of 4 billion gallons of ethanol in 2006 (EPA, 2021). The Energy Independence and Security Act (EISA) of 2007 further increased the renewable fuel usage requirement to 36 billion gallons annually by 2022 (EPA, 2021). Both the RFS and EISA were intended to not only mitigate environmental harm but also to protect national energy security. Current RFS standards require that 15 billion gallons of ethanol are produced from corn, while the remaining 7 billion are produced from cellulosic feedstocks. The combination of these

domestic policies suggest that corn bioethanol will continue to be a hallmark component of the renewable energy movement. This continued demand suggests that improvements in ethanol production efficiency and economics are highly relevant areas of research.

The global reliance on fossil fuels has been a longstanding issue and requires significant research and development to create cleaner alternatives. Ethanol is the primary biofuel integrated within the United States' energy supply. Given the narrow margin for profitability and energy return on investment, ethanol production facilities must operate under a highly efficient manner to produce a viable biofuel. As the net energy consumption continues to increase, ethanol consumption will mirror the trend. Developing large scale bioethanol facilities will be essential to continue to meet the demand of energy consumption. This can be achieved by integrating economic considerations within process research and development. Developing a process focused on plant economics and energy return on investment will ensure that the plant can operate efficiently while remaining competitive within the market.

This capstone project seeks to design a dry-mill corn ethanol production facility with the capability of producing the cheapest barrel of oil equivalent (BOE) of fuel-grade ethanol. The plant will be designed to produce 150 million gallons of ethanol per year. Plant sizing determination was established given the fact that for the top 15 ethanol production plants in the United States, the most common production output volume is 150 million gallons (Henneberry, 2019). Sizing the plant to produce 150 million gallons annually aligns well with projected increases in domestic demand.

2.2 Product and Coproduct

Fuel grade ethanol is the primary product of the facility design. Ethanol is flammable, volatile, and colorless, and has chemical formula C_2H_5OH (Thermo Fisher Scientific, 2009). The ethanol final product is generated via a fermentation process in which yeast convert corn sugars ($C_6H_{12}O_6$) into ethanol and carbon dioxide (CO_2) as seen in the chemical equation below (Ray & Behera, 2011):



While several alternative industrial applications exist, the scope of this project only focuses on ethanol's use in transportation fuel blends. E10 is the main blend in the United States (EIA, 2016). Since fuel performance is heavily influenced by the quality of blended ethanol, purification is essential. Common impurities characteristic to the production process include water, methanol, and higher molecular weight alcohols (Weaver, 2009). Petroleum companies and gasoline blenders purchasing the fuel ethanol generally follow the industry standard: ASTM D 4806 for fuel grade ethanol. These requirements include comprehensive product purity specifications and dictate the requirements for downstream operations. The ethanol product will meet all specifications set by ASTM D 4806, as detailed in Table 2.2.1. Denaturant in the form of conventional motor gasoline will be added to the final ethanol product stream leaving the molecular sieves.

Table 2.2.1 ASTM D 4806 Performance Requirements

Property	Percent Volume (%)
Ethanol Minimum	92.10
Methanol Maximum	0.50
Water Content Maximum	1.00
Denaturant Content Minimum	1.96
Denaturant Content Maximum	4.76

(Reynolds, 2002)

Distiller's dried grains with solubles (DDGS) is the primary coproduct of the production process used in the facility design. A popular livestock feed, DDGS serves as a highly nutritious and concentrated nutrient source. The overall quality of DDGS is largely dependent on its nutrient profile and can heavily influence the economic value of the coproduct. The nutrient profile and other relevant factors are outlined in Table 2.2.2 below. The DDGS produced in this project will largely adhere to the specifications specified by Table 2.2.2. Sand silica will be added by downstream DDGS processors who add the material in order to keep the pellet-forming dies free from excessive buildup (Behnke, 2007). Sand silica addition will not occur at the bioethanol facility.

Table 2.2.2 Quality Specifications of Corn-Based DDGS

Name	Specification
Protein	28-32%
Moisture	10-12%
Fiber	5-6%
Sand Silica	2-3%
Fat	1-2%
Energy	3,810 kcal/kg
Color	Yellow
Appearance	Powder/Granular

((DDGS (Dried Distillers Grain Soluble))

2.3 Raw Materials

The raw materials involved in corn-based ethanol production include corn, water, enzymes, yeasts, and denaturants. Raw corn will be sourced from nearby farms in the Midwest. Over 60% of the corn kernel is starch, with the remainder being composed of fiber, protein, oil, and water (Ibrahim et al., 2019). Converting the corn starch into sugar is accomplished through enzymatic processing, and vastly improves the overall fermentation productivity. The assumed composition of the sourced corn kernels is detailed in Table 2.3.1.

Table 2.3.1 Corn Kernel Composition

Component	Composition (%)
Starch	62
Protein and Fiber	19
Water	14
Oil	4
Trace Components	~1

(Clifford, 2020)

Process water will be sourced from the surrounding community. This water will be assumed to be at 22°C based on the dew point temperature of Iowa in the peak months of summer. Since process water sourced from the city will meet purity specifications, further on-site treatment will not be employed. Glucoamylase and α -amylase will be the 2 enzymes used to convert the corn starch into readily fermentable sugars as mentioned above. These will be sourced from industrial enzyme suppliers. *Saccharomyces cerevisiae* will serve as the yeast for the fermentation step of the conversion process. *Saccharomyces cerevisiae* will also be purchased

in bulk along with the pre-processing enzymes. Denaturant in the form of conventional motor gasoline will be transported to the facility via rail cars to minimize cost.

2.4 Process Overview

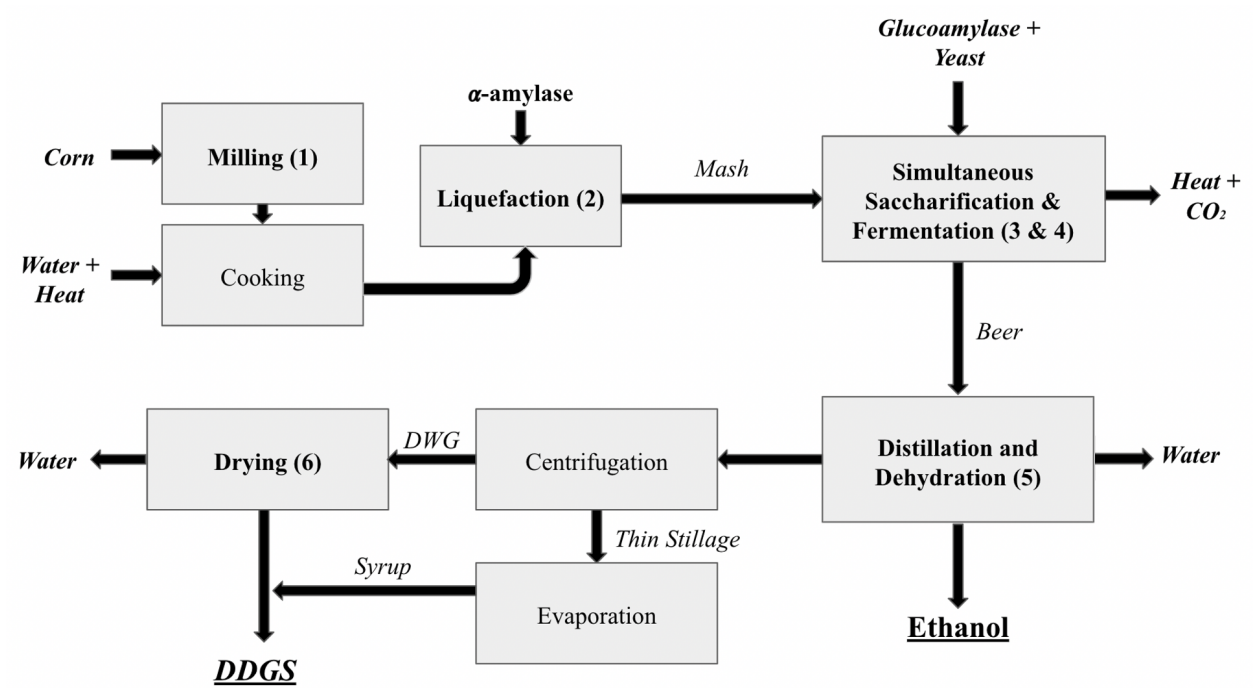
Ethanol production can be accomplished through two different process pathways: dry milling or wet milling. Although wet milling extracts the maximum potential value from the raw corn material, dry milling is more commonly employed at industrial scale (RFA, 2005). Dry milling carries a lower capital cost, and is therefore a more attractive option for large scale production facilities (Devantier et al., 2005). The dry milling ethanol production process consists of six distinct operations. These include milling, liquefaction, saccharification, fermentation, distillation/dehydration, and drying (Figure 2.4.1). The saccharification and fermentation steps can be accomplished simultaneously and are discussed further in the process discussion section. A complete process flow diagram with accompanying stream table is available in Figure 4.1.

Figure 2.4.1 below serves as a broad visual representation of the entire corn bioethanol conversion process. In Figure 2.4.1, the 6 key operations mentioned above are notated and highlighted in bold. Process inputs and outputs are italicized and bolded, while intermediate material flows are simply italicized. Process outputs, namely ethanol and DDGS, are bolded and underlined. As seen in Figure 2.4.1, the conversion process first begins with corn arriving at the plant via rail. This corn is fed into hammermill grinders in “Milling”, notated as **Step (1)**. Milled corn is combined with water and heated before passing through into “Liquefaction”, notated as **Step (2)**. α -amylase enzyme is added and the newly formed “mash” is combined with glucoamylase and yeast in Simultaneous Saccharification & Fermentation (SSF), notated as **Steps (3 & 4)**. After fermentation converts corn sugars into ethanol, “beer” is drained from the fermenter and fed into equipment responsible for Distillation/Dehydration, notated as **Step (5)**. Ethanol exits the overall process following distillation and dehydration. Conventional motor gasoline is used as a denaturant and is combined with the ethanol stream to produce fuel-grade

ethanol. This fuel-grade ethanol meets final product specifications and is ready to be sold. The remaining fermentation products are fed to centrifuges where they are separated into thin stillage and distiller's wet grain (DWG). DWG is fed directly into Drying, notated as **Step (6)**, while thin stillage is further dehydrated to create corn syrup. Corn syrup is applied to the grain exiting **Step (6)** to produce saleable DDGS.

The bottleneck of the production process is fermentation since it occurs on a batch basis and is the only noncontinuous operation. The entire production facility must therefore operate around the material flow rates specified by fermentation capacity. Based on *Devantier et al.*, a 49 hour fermentation window is used to ensure the yeast operate at high levels of productivity. Following literature review, to provide time required for filling, emptying, and routine cleaning, the full batch spans 72 hours total. Throughout the project 'per batch' notation is used to describe the production process during the 72 hour fermentation batch. Given this fermentation batch capacity, 109 batches will be required annually to meet the 150 million gallon production quota. To provide enough time for routine maintenance and thorough cleaning, the plant will operate for 329 days of the calendar year. Every 2 months, a 6 day downtime period will be used to address the maintenance and cleaning requirements.

Figure 2.4.1 Process Overview Schematic



3. Process Design Discussion

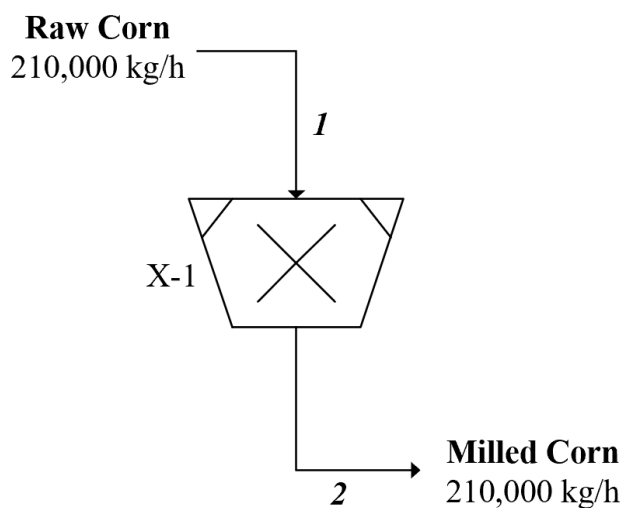
3.1 Milling

Corn milling is an essential step in the corn ethanol production process since it aids in fracturing the tough, outer coating of the corn kernels. Milling the raw corn not only reduces the size of the corn granules, but also increases the amount of grain surface area accessible for enzymatic breakdown. Since the corn ethanol process relies on the conversion of corn starches to corn sugars, and then the subsequent conversion of corn sugar to ethanol by yeast, mechanical digestion is an important first step. This dry mechanical breakdown is characteristic of dry-mill production corn ethanol plants (Indiana Ethanol Producers Association, 2012).

Corn milling in the plant is accomplished using a series of hammer mills in which corn is fed into a chamber and battered by flailing hammers attached to a rotor (Seim, 2015). Hammer blows, particle collision, and collisions with the inner enclosure physically separate the corn particles. Output corn particle size is controlled via a mesh screen which permits particles of no more than 800 μm to exit the mill (Selle et al., 2016). Hammer mills were selected for this application instead of alternatives like roller mills since they are more commonly employed for industrial grinding in recent practice (Bradshaw, 2010).

As seen in Figure 3.1.1, 210,000 kg/h of milled corn is required in order to meet the annual process output quota of 150 million gallons of ethanol. The calculations for required corn input were driven by the corn required for fermentation and are available in Section 3.3.

Figure 3.1.1 Corn Milling Material Balance



Hippo #69 Mill with Air Conveyors (X-1) were selected for use in the plant. Hippo #69 Mills offer relatively low cost and high processing throughput (Pleasant Hill Grain). Each hammermill is able to process approximately 850 kg/h of dry corn (Pleasant Hill Grain). In order to meet the corn input processing requirements, 234 Hippo #69 Mill with Air Conveyors will need to be in continuous operation. An additional 24 mills will be purchased in order to accommodate for equipment breakdowns and any other unforeseen production complications. According to the power consumption specifications provided by the manufacturer, each Hippo #69 Mill with Air Conveyor requires 45 kW to operate. Given this power consumption per mill, 10,500 kW total will be required to meet the 210,000 kg/h corn specification. This energy requirement will be supplied as electricity. A summary of corn milling production details are available below in Table 3.1.1.

Table 3.1.1 Corn Milling Production Summary

Production Specification	Quantity
Corn Milling Requirements	210,000 kg/h
Hippo #69 Milling Capacity	850 kg/h
Hammermills in Operation	234
Spare Hammermills	24
Hippo #69 Power Requirements	45 kW
Total Power Consumption	10,500 kW

3.2 Liquefaction

Liquefaction is an important second operational step in which milled corn is converted into a mash for fermentation. α -amylase enzymes are used to convert the raw corn starch into sugars that are more readily fermented by yeast. Liquefaction can be divided into 3 main stages: primary liquefaction, jet-cooking, and secondary liquefaction. The mash preparation techniques outlined in *Characterization of very high gravity ethanol fermentation of corn. Effect of glucoamylase dosage, pre-saccharification and yeast strain* by Devantier et al. were the primary source for procedures, and are explained in greater detail below.

Primary liquefaction is the first stage of the process and involves the creation of a high gravity slurry. The process is illustrated completely in Figure 3.2.1 below. Forming a high-gravity slurry increases the dry matter (DM) content in the slurry to above 30% (w/w), and improves the plant economics by generating more ethanol output per unit of corn input (Devantier et al., 2005). Though a high gravity mash of 35% (w/w) is ultimately prepared by the end of the overall liquefaction stage, primary liquefaction only contributes approximately 95% of the input water since the remaining 5% are added in the form of steam during jet-cooking. The high gravity mash formed in primary liquefaction is prepared by combining approximately 210,000 kg of milled corn with 372,000 kg of process water per hour. 26.2 kg/h of α -amylase are also added in order to begin the enzymatic-driven digestion of corn starch (Devantier et al., 2005). The 3 feeds are combined in a 250,000 gallon tank (T-1) and agitated. Specific engineering design for agitation was deemed outside the scope of this design brief and is a necessary component of future design work. Water is added via a 7 kW 316 SS centrifugal pump (P-101) while the milled corn is added via conveyor. An operator is responsible for adding the 26.2 kg of α -amylase. The activity of the α -amylase increases with heat so the solution is heated

from 22 °C to 85°C via an external heat exchanger (Devantier et al., 2005). The increase to 85 °C also prepares the slurry for jet-cooking, the second liquefaction step. The external heat exchanger (HX-1) is a floating head shell and tube design manufactured from 316 SS. The exchanger consists of a heat transfer area of approximately 1,100 m² and is composed of 2,700 schedule 40 tubes. The speed of the material slurry flow is 3 m/s. Heating the solution requires 32,400 kW. Steam at 10 bar is used as the heat transfer medium. Pumping is accomplished using a 250 kW 316 SS positive displacement pump (P-201). An additional 250,000 gallon tank was included in the design to accommodate for any necessary cleaning or maintenance, meaning that the total number of tanks dedicated to primary liquefaction is 2. After reaching 85 °C, the slurry is ready to be jet-cooked. A 22 kW 316 SS positive displacement pump (P-202) is used to drain the tank and direct the material stream into the jet-cooker (J-1). A summary of primary liquefaction details is available below in Table 3.2.1.

Table 3.2.1 Primary Liquefaction Process Details

Production Specification	Quantity
Milled Corn Input	210,000 kg/h
Process Water Input	327,000 kg/h
α -amylase Input	26.2 kg/h
Heating Requirements	32,400 kW

Jet-cooking is the second stage of the process and further solubilizes the starch (Devantier et al., 2005). Jet-cooking consists of direct steam injection into the corn slurry mix (Hydro-Thermal Corp). Direct steam injection simultaneously heats and mechanically shears the corn particles, and results in an increased amount of starch availability over traditional corn milling techniques (Pro-Sonix). The corn slurry from primary liquefaction is heated from 85 °C

to 115 °C as it passes through jet-cooker. Hydro-Thermal HydroHeater Jet-Cookers were selected for use in the plant due to their precision and reliability (Hydro-Thermal Corp). In order to adequately heat the 582,000 kg/h of slurry passing through the jet-cookers each hour, 17,600 kg/h of steam at 10 bar is needed. This steam also makes up the remaining water needed in the slurry and results in a final slurry composition of 35% (w/w) corn. Since each unit has a capacity of approximately 50 m³/hr, 32 jet-cookers operate in parallel in order to process the primary liquefaction slurry stream. An additional 4 jet-cookers are included in the design to accommodate for any necessary cleaning or maintenance. A combination of the P-202 pump and steam input force the slurry material through the pipes and into the secondary liquefaction tank (T-2). A summary of jet-cooking details is available below in Table 3.2.2.

Table 3.2.2 Jet-Cooking Process Details

Production Specification	Quantity
Slurry Input	582,000 kg/h
Steam Requirement	17,600 kg/h
Jet Cookers in Operation	32

Secondary liquefaction is the third stage within the liquefaction process and occurs within the 250,000 gallon secondary liquefaction tank (T-2). Secondary liquefaction serves to cool the slurry before yeast and enzymes are added for fermentation. A second dose of α -amylase enzymes is also added to continue the starch breakdown process. Operators are responsible for adding 43.6 kg of α -amylase per hour. At this point in the process, the process water and steam contributed earlier have resulted in a 35% (w/w) corn slurry with an apparent viscosity of 6 cP (Khakbaz, 1979). This elevated viscosity is the result of dissolved starches and other corn

components. Though it is not included in the report design, agitation is used to adequately mix the additional α -amylase. An external floating head shell and tube heat exchanger is used to reduce the slurry temperature from 115 °C to the 32 °C prescribed for fermentation (Devantier et al., 2005). The heat exchanger is manufactured from 316 SS and has a heat transfer area of 1,400 m². The slurry flows at 3 m/s through the 3500 schedule 40 pipes. Heating the solutions requires 47,000 kJ. Cooling water fed at 22 °C is used as the heat transfer medium. A 650 kW 316 SS positive displacement pump (P-301) is used to drive the slurry flow. Once the slurry reaches 32 °C, the tank is drained using a 14 kW 316 SS positive displacement pump (P-302). The slurry is directed into 1 of 4 active fermentation tanks (R-1). In total, the liquefaction process generates 600,000 kg/h of 35% (w/w) slurry mix that flows into fermentation tanks at 32 °C. An additional 250,000 gallon tank was included in the design to accommodate for any necessary cleaning or maintenance, meaning that the total number of tanks (T-2) dedicated to secondary liquefaction is 2. A summary of secondary liquefaction details is available below in Table 3.2.3.

Table 3.2.3 Secondary Liquefaction Process Details

Production Specification	Quantity
Slurry Input	600,000 kg/h
α -amylase Input	43.6 kg/h
Slurry Viscosity	6 cP
Heating Requirements	47,000 kW

3.3 Simultaneous Saccharification and Fermentation

Although saccharification and fermentation have historically been accomplished in separate process steps, the corn ethanol industry is slowly transitioning to performing both processes simultaneously (Olofsson et al., 2008). Independent saccharification primarily involves the addition of glucoamylase to further starch digestion. Independent fermentation mainly includes the introduction of yeast to convert the fermentable sugars into ethanol. Simultaneous saccharification and fermentation (SSF) is an advantageous combination of the two processes since it prevents high concentrations of lower molecular weight saccharides from exposing yeast to significant osmotic stress (Devantier et al., 2005). The delayed saccharification also helps insulate yeast from otherwise toxic levels of ethanol (Devantier et al., 2005). This yeast protection enables the use of higher gravity corn slurries that can reach higher concentrations of ethanol before yeast productivity rapidly declines. Enabling higher concentrations of ethanol during fermentation ultimately allows for more ethanol output per unit of corn input (Devantier et al., 2005). Adding glucoamylase ultimately serves to accelerate and increase overall ethanol production during the yeast's maximum productive period. A comprehensive literature review indicated that *saccharomyces cerevisiae* yeast is widely used for industrial-scale fermentation since it optimizes ethanol production while shortening fermentation time alongside reduced process costs (Tesfaw et al., 2014). Given this background, *saccharomyces cerevisiae* yeast was selected for use in the process.

A visual representation of the fermentation process is available in Figure 3.3.1. Process conditions were modeled after the ideal SSF procedures identified in *Characterization of very high gravity ethanol fermentation of corn mash. Effect of glucoamylase dosage, pre-saccharification, and yeast strain* (Devantier et al., 2005). The literature indicated that

ethanol production output was maximized when fermentation took place over a 49-hour period (Devantier et al., 2005). This window was then extended such that a “batch” consists of a 72 hour period to allow for tank filling, emptying, and cleaning. The continuous material input from liquefaction ultimately requires 5 fermentation tanks (R-1) that operate simultaneously. 4 million gallon 316 SS fermenters (HX-3) are used to accommodate this schedule. Each fermentation tank takes 18 hours to fill since slurry is fed at the output rate of liquefaction of 600,000 kg/h. 2,800 kg of glucoamylase are added to the fermentation vessel at this time via conveyor, approximately 2 hours before *saccharomyces cerevisiae* yeast. This delay in yeast addition aligns with results from the literature (Devantier et al., 2005). The addition of 1,650 kg of *saccharomyces cerevisiae* to the tank serves as the start of the 49-hour fermentation period. Yeast is also added via conveyor. Since each “batch” occurs across 4 fermentation tanks simultaneously, total batch input requirements of glucoamylase and yeast are 11,300 kg and 6,600 kg respectively. The 5th fermentation tank is required to accommodate for the necessary 18 hours of filling time and allow fermentations to begin on schedule.

The optimal process conditions detailed in *Characterization of very high gravity ethanol fermentation of corn mash. Effect of glucoamylase dosage, pre-saccharification, and yeast strain* yielded a beer with an ethanol concentration of 11% (w/w) (Devantier et al., 2005). The mash is kept at 32 °C for the entire 49 hour fermentation cycle. Temperature conditions are maintained using a combination of 2 floating head shell and tube heat exchangers, with a total combined area of 5,700 m². Each is composed of 316 SS and has 3,500 schedule 40 inner tubes. The mash is continuously cycled through the heat exchanger using a 320 kW 316 SS positive displacement pump (P-401). No additional reaction vessels (R-1) or heat exchangers (HX-3) are included in the design due to their capital cost. Cleaning and maintenance for these items is conducted

during the routinely scheduled plant shutdown. The 516,000 kg/h of beer exiting in the fermentation process is pumped into the beer storage tank (T-3) using a 36 kW pump manufactured with 316 SS (P-402). This material flow from fermentation to distillation is available in Figure 4.1. CO₂ is also generated as part of the fermentation reaction. As fermentation proceeds, 600 kg/h of water is lost as vapor, in addition to 83,400 kg/h of carbon dioxide. Stoichiometric equivalents were used to determine the off-gassing carbon dioxide. An open pool model was used to estimate the water vapor lost overhead. Although not included in this design report, CO₂ collection and purification could serve as an additional source of revenue for the plant if applicable equipment was purchased. A summary of fermentation details is available below in Table 3.3.1.

Figure 3.3.1 Fermentation Overall Material Balance & PFD

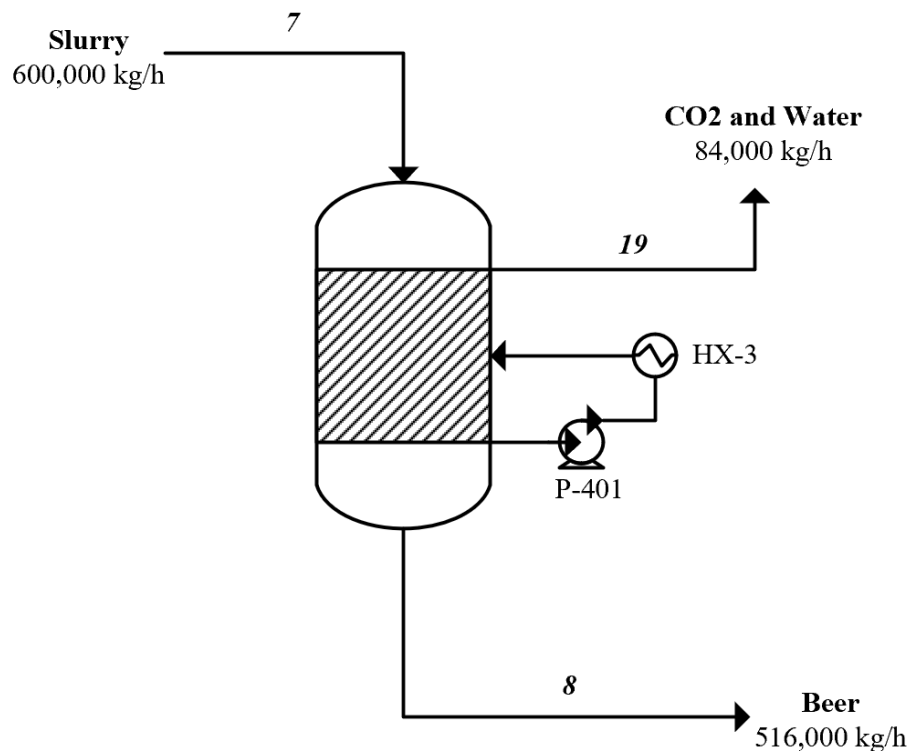


Table 3.3.1 Fermentation Process Details

Production Specification	Quantity
Slurry Requirements	10.7 million kg
Glucoamylase Input	2,800 kg
Yeast Input	1,650 kg
Ethanol Concentration in Beer	11% (w/w)

Tabulated values indicate production requirements are on a per fermenter-batch basis

3.4 Downstream Separation

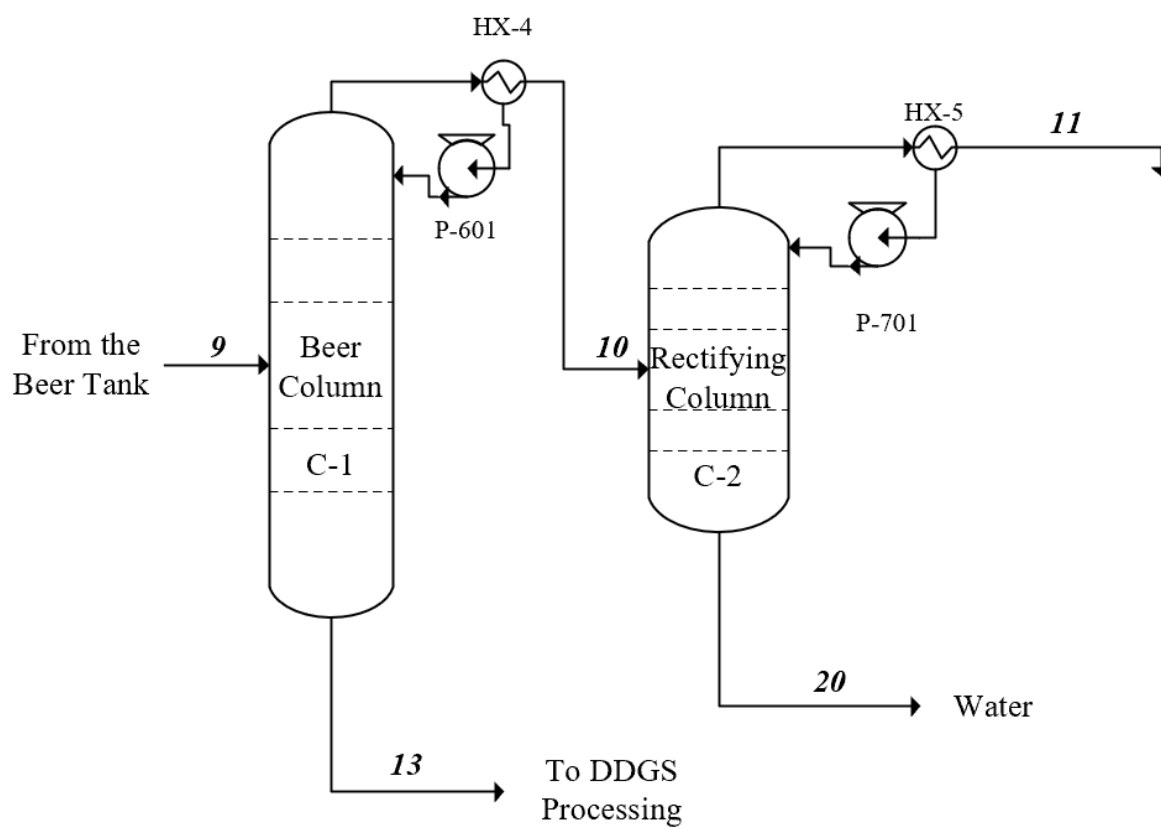
After fermentation, the beer mixture product is 11% (w/w) ethanol and must be separated and purified. This separation is a multistep process that relies on distillation columns and molecular sieves to reach the final fuel ethanol minimum specification (Reynolds, 2002). In order to improve the economic viability of the overall production process, DDGS must also be separated and processed as a valuable coproduct. Preliminary stages of beer separation are accomplished through two separate distillation columns as seen below in Figure 3.4.1.

The first unit of the downstream process design, the beer distillation column (C-1), is responsible for removing the remaining corn solids and a portion of the process water from the beer. The bottoms product is known as whole stillage and is further processed into the DDGS coproduct. The distillate from the beer column is an ethanol-water mixture that is directed to a neighboring rectifying column (C-2). The rectifying column (C-2) further distills the mixture and purifies it near the ethanol-water azeotrope of 95% (w/w) ethanol. The azeotropic composition still does not meet fuel grade ethanol standards and therefore must be dried further. This drying is accomplished through molecular sieves that rely on zeolites to separate water molecules from the overall solution. After further dehydration via the molecular sieves, the ethanol product is ready to be sold. Distillation is addressed in greater detail in Section 3.4.1, while molecular sieves are addressed in Section 3.4.2.

The whole stillage transformation into DDGS relies primarily on water removal and is accomplished in several additional processing units. The whole stillage from the beer column bottoms product is first directed into a centrifuge where it is separated into distillers wet grain (DWG) and thin stillage. Despite the fact that DWG is a marketable coproduct, DWG has a very short shelf life and is less economically lucrative than DDGS. DWG is therefore processed

further to improve the operating economics of the facility. The thin stillage extracted via centrifugation goes through an evaporation process that produces syrup and wastewater. The syrup is then mixed with the DWG with the combined mixture passing through a rotary fired dryer to remove residual water. The DDGS exiting the dryer is market-grade and is ready to be sold. Centrifugation is addressed in greater detail in Section 3.4.3, while drying operations are addressed in Section 3.5.

Figure 3.4.1 Distillation-Based Separation Equipment



3.4.1 Distillation

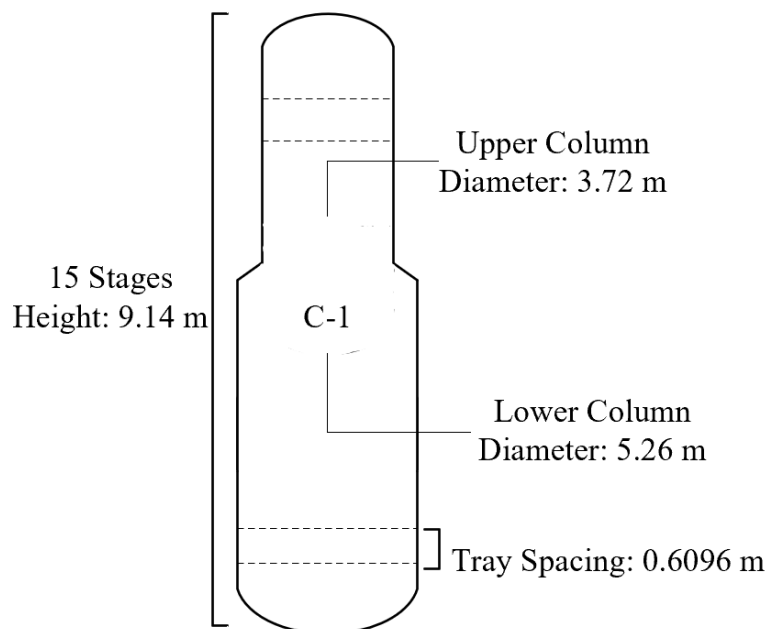
After fermentation, beer is transported to the beer storage tank (T-3) and is fed continuously to the distillation column system via a 90.5 kW positive displacement pump constructed from 316 SS (P-501). The beer column (C-1) serves to crudely separate the fermented solids, water, and ethanol. The column (C-1) concentrates the ethanol into an ethanol-water mixture while removing the solids and a portion of the process water. Crude separation is essential at this stage to prevent unnecessary downstream energy expenditures. Distillation calculations were performed using simulations through AspenPlus software.

The beer column (C-1) was modeled in AspenPlus to identify the appropriate process conditions and product material compositions. This was achieved using the hourly flow rate of materials exiting fermentation, since the amount and composition of feed input governs the overall distillation separation. The column block 'RadFrac' was used to simulate this separation process. The beer was fed to the column (C-1) at a flow rate of approximately 516,000 kg/hr, as supplied by the exit stream from the fermentation material balance. While the components primarily involved in beer column separation are water and ethanol, solid corn remnants from the fermentation process are present and require separation. These solids were modeled as nonconventional solids and all exited in the bottoms product stream. The feed temperature was specified by the fermentation temperature of 32 °C. Within the RadFrac column, Aspen simulations determined that 10 theoretical stages were sufficient to achieve reasonable separation (Figure 3.4.2). A total condenser and kettle reboiler were specified for the Aspen simulation. The column simulation was also operated at 5 bar with a mass-specified reflux ratio of 1.

Since fouling of the trays and reboiler is a significant concern given the high solids content, the beer feed input was introduced above Stage 9. Additionally, anti-fouling fixed valve

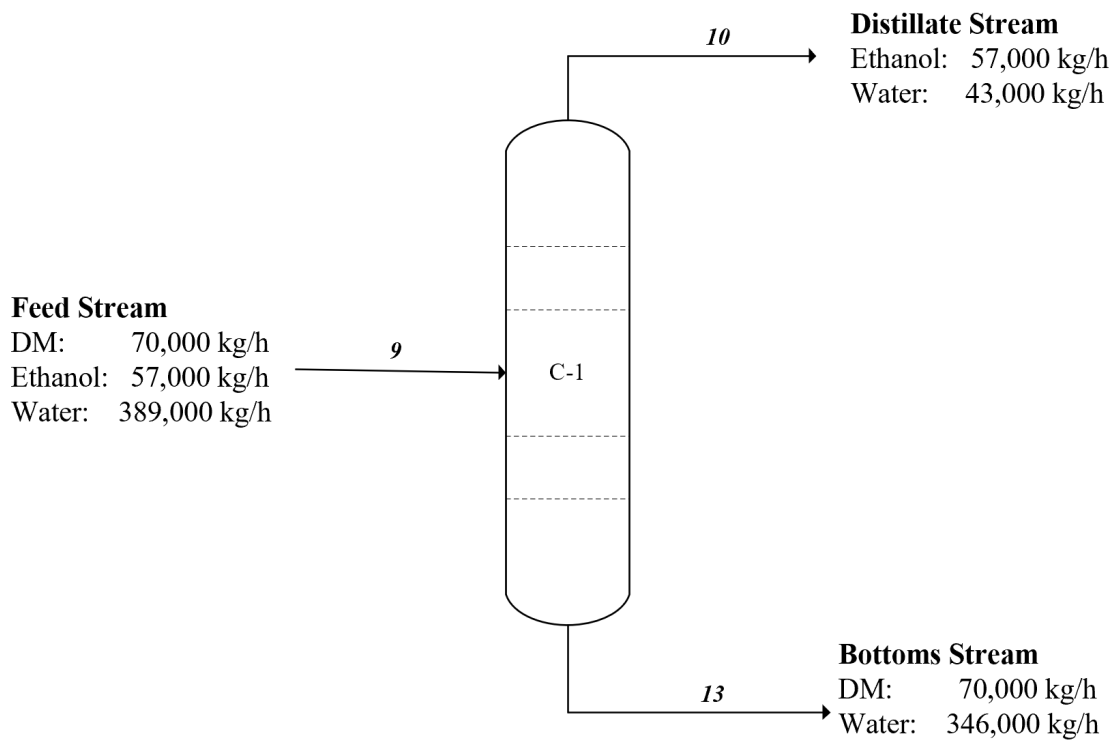
trays were integrated into the design to mitigate fouling and compromised performance. Duplicate reboilers were also included in the pricing of the beer column (C-1) to ensure that fouling would not completely halt production. The Aspen simulation also indicated that a tapered distillation column is required to adequately separate the input materials in an efficient manner. The column diameter for the upper region is 3.72 m, as seen in Figure 3.4.2. The lower portion of the column has a diameter of 5.26 m, as seen in Figure 3.4.2. Tray spacing in the simulation adheres to the 0.6096 m industry standard. Since the Aspen simulation assumes perfect efficiency, an efficiency factor must be integrated into the calculation to determine the real number of trays required for this separation. Tray efficiency was assumed to be 70% for the fixed valve trays, and therefore the actual process will require 15 real trays. The total column height is ultimately 9.14 m tall as seen in Figure 3.4.2.

Figure 3.4.2 Beer Column Design Specifications



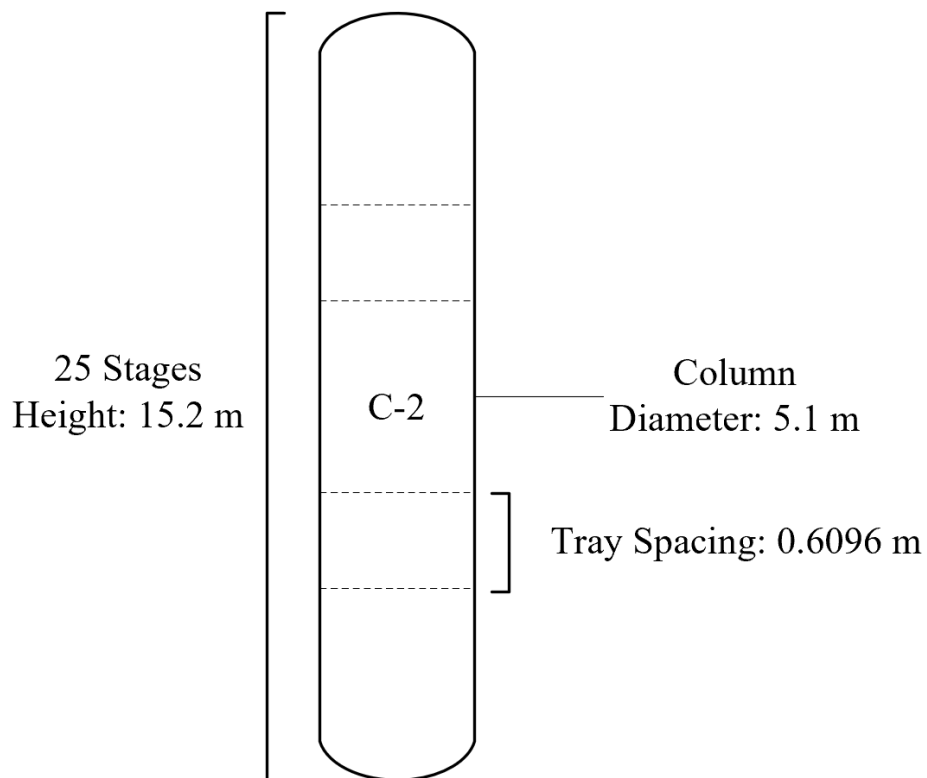
As seen in Figure 3.4.3 below, the input stream is composed of 516,000 kg/h of total material. Of this 516,000 kg, 11% (w/w) is ethanol. The material balance calculated during the AspenPlus simulation indicates that all of the input biomass is separated into the bottoms product. As seen in Figure 3.4.3, the bottoms product composition is 17% (w/w) biomass and 83% (w/w) water, exiting at a flow rate of 416,000 kg/h. This bottoms product stream exits the column at 143°C and will be further processed in centrifugation to produce DDGS. The distillate flow rate is 100,000 kg/h, with a composition of 57% (w/w) ethanol and 43% (w/w) water at 128°C. The distillate is directed to the rectifying column (C-2) to further purify the ethanol. The column and trays are constructed with 316 SS. The AspenPlus simulation reports a condenser duty of 77,200 kW and a reboiler duty of 140,000 kW. These duties require a condenser with an area of 1,000 m² and a reboiler with an area of 1,800 m², both constructed from 316 SS.

Figure 3.4.3 Beer Column Material Balance



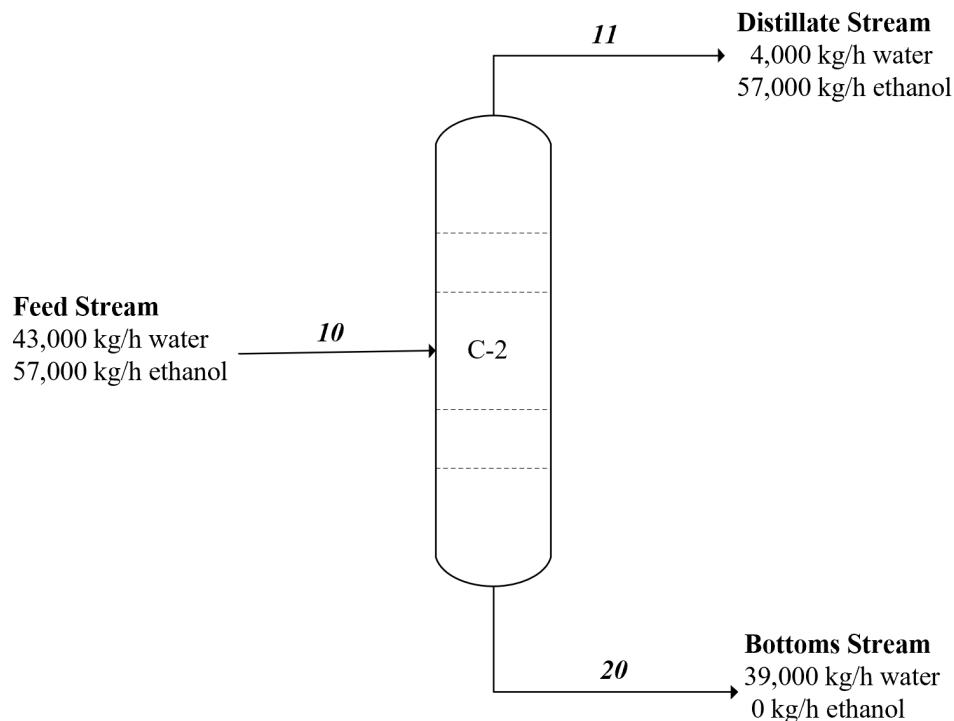
The water-ethanol mixture exiting as the beer column distillate serves as the feed of the rectifying column (C-2), where the ethanol is further purified. A total condenser and kettle reboiler were specified for the Aspen simulation. The column simulation was operated at 3 bar with a mass-specified reflux ratio of 1. The AspenPlus simulation reports a continuous column diameter of 5.1 m. The AspenPlus simulation also reports that the column operates using 20 theoretical stages. Since fouling is not a concern and the column uses sieve trays, the tray efficiency was assumed at 80%. Factoring in the tray efficiency results in a 25 tray design. The industry standard tray spacing of 0.6096 m ultimately results in a column height of 15.2 m. The complete rectifying column (C-2) design specifications are available in Figure 3.4.4 below.

Figure 3.4.4 Rectifying Column Design Specifications



In the rectifying column (C-2), the mixture is separated to near the azeotropic composition limit, achieving a distillate composition of 93.5% (w/w) ethanol and 6.5% (w/w) water. The rectifying column (C-2) is fed the distillate output from the beer column (C-1) and therefore receives 100,000 kg/h of the water-ethanol mixture. This stream is added to the column at Stage 19. The rectifying column (C-2) ultimately separates the input mixture into a 61,000 kg/h distillate stream of 93.5 % (w/w) ethanol and a 39,000 kg/h bottoms stream of nearly-pure water (Figure 3.4.5). The distillate stream exits the column at 108 °C while the bottoms stream exits at 134 °C. The column and trays are constructed with 316 SS. The AspenPlus simulation reports a condenser duty of 74,500 kW and a reboiler duty of 78,800 kW. These duties require a condenser with an area of approximately 1,000 m² and a reboiler with an area of approximately 1,000 m², both constructed from 316 SS. The mass-specified reflux ratio in the simulation is 4.

Figure 3.4.5 Rectifying Column Material Balance

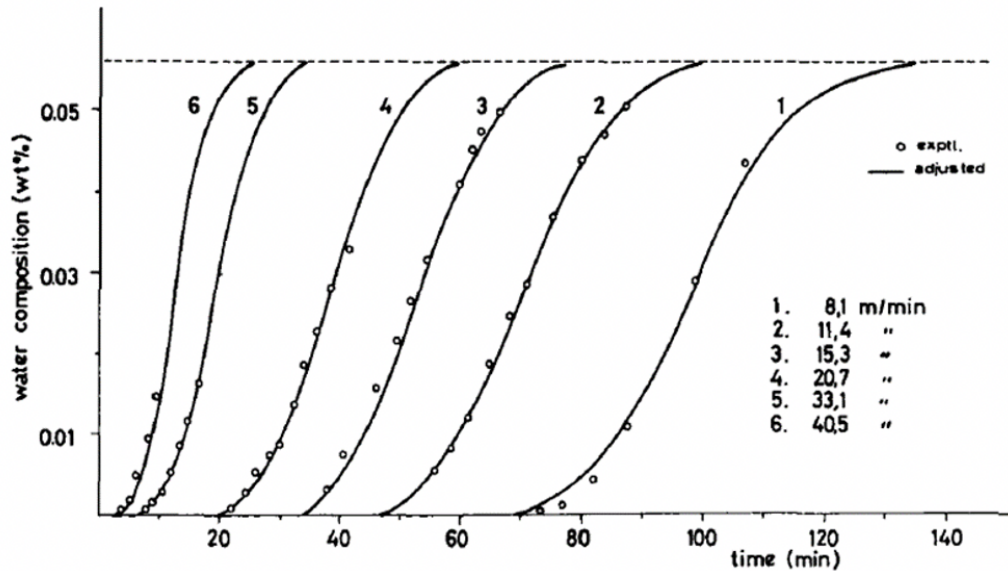


3.4.2 Molecular Sieves

To reach the desired level of product purity, the azeotropic distillate product of the rectifying column (C-2) is fed in vapor form into a molecular sieve system (X-2). The sieve operates as a packed bed adsorption column filled with a porous alumino-silicate zeolite. The specific zeolite being used is class 3A with the chemical formula $K_{12}[(AlO_2)_{12}(SiO_2)_{12}]-NH_2O$. Due to the inherent porosity of the material, water is adsorbed onto the bed, effectively drying the vapor stream to 99.5% (w/w) ethanol.

The vapor outlet stream of the rectifying column (C-2) is 61,000 kg/h and is the input feed for the molecular sieve beds (X-2). Of this 61,000 kg/h of vapor, 57,000 kg are ethanol with the remainder as water. 3,600 kg/h of water must be removed from this stream in order to meet ethanol product specifications. The configuration of an appropriate column for this task is based upon the maximum observed level of sieve adsorption capacity in literature of 21.9 g water/100 g zeolite (Cartón et al., 1987). The dimensions of the column are 29.7 m in height and 0.928 m in diameter, yielding a total volume of 20.1 m³. The volume is packed with 16,500 kg of zeolitic material. The column will be operated at a pressure of 3 bar during adsorption. Vapor flows through the column at a linear velocity of 8.34 m/s, and therefore the residence time of the vapor in the column will be approximately 3.5 seconds. Based on the governing kinetics of the adsorption by the zeolite, breakthrough of water into the product stream will be minimized as seen in Figure 3.4.6.

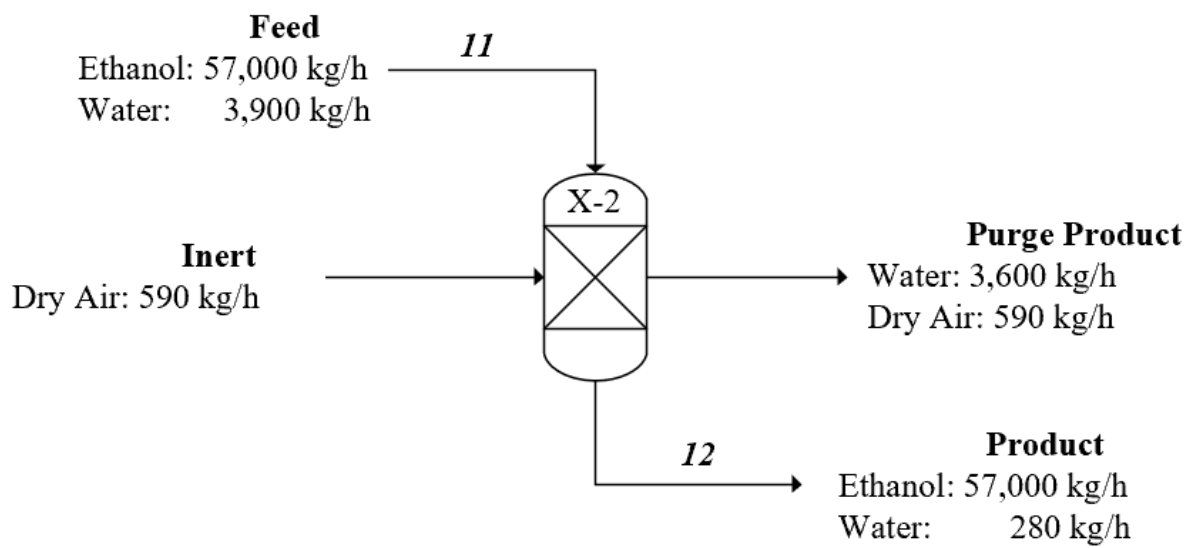
Figure 3.4.6 Water Breakthrough Curve in Type 3A Zeolite



(Carton et al., 1987)

For successful implementation of this column within the overall ethanol process, it should be built in duplicate. This redundancy allows for the regeneration of the zeolite in one column, while the other continues to dehydrate the continuous vapor stream from the rectifying column (C-2). To adequately purge the packed bed, a hot air stream will run through the column in order to purge the water clinging to the bed particles. The air stream will be pushed through the column at a temperature of 230°C at a mass flow rate of 590 kg/h. By running the adsorption and desorption cycles concurrently in two-hour increments, the effective removal of water from the vapor stream can be attained (Figure 3.4.7). The associated heat duties of running each column can be anticipated at roughly 4,200 kW heat duty for the adsorption process and 14,566 kW for the desorption process.

Figure 3.4.7 Material Balance of Molecular Sieve Column

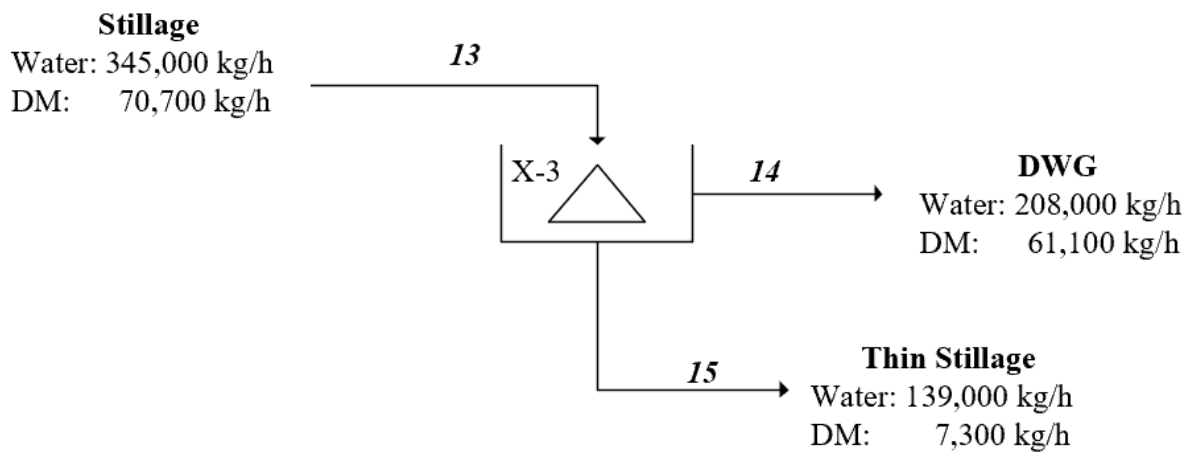


3.4.3 Centrifugation

Centrifugation offers a cost-effective way to begin reducing the moisture content of the whole stillage beer column bottoms product. Reducing the moisture content allows for the transformation of whole stillage waste into the valuable DDGS coproduct. The total material input to centrifugation is 416,00 kg/h of whole stillage. This whole stillage is pumped from the beer column (C-1) to the decanter centrifuges (X-3) via a 0.7 kW positive displacement pump constructed from 316 SS (P-801). Of the 416,000 kg/h of whole stillage, 17% (w/w) is dry matter while the remainder is water. The decanter centrifuge ultimately separates the input into 2 separate streams: thin stillage and DWG. Literature review was used to inform the following centrifugation design process (Ganesan et al., 2008). 65% of the whole stillage feed exits the centrifuge as DWG with a water content of 77 % (w/w). The remaining 35% of whole stillage feed exits the centrifuge as thin stillage with a water content of 95% (w/w). The DWG is directed to the drying process while thin stillage is fed to the falling film evaporators to produce syrup. A complete summary of the centrifugation process stage is available in Figure 3.4.8 below.

The centrifuge selected for the design has a diameter of 750 mm and rotational speed of 2,500 rpm (ZK Separation). With a capacity of 60 m³/h per centrifuge, and the whole stillage feed occupying 470 m³/h, 8 centrifuges are needed. An additional 1 centrifuge is included in the plant plans to address potential malfunctions and cleaning. Each centrifuge is driven by 2 motors that require 162 kW. This results in a total energy requirement of 1270 kW to power the 8 centrifuges operating at any point in time.

Figure 3.4.8 Centrifugation Material Balance



3.5 Drying

Drying serves to reduce the water content present in the remaining corn solids to prepare them for sale as DDGS. Although higher moisture content animal feeds like DWG exist, transportation difficulties and the limited shelf life complicate effective distribution (Birkelo et al., 2004). Thin stillage processing is addressed specifically in Section 3.5.1 while DWG processing is addressed in Section 3.5.2.

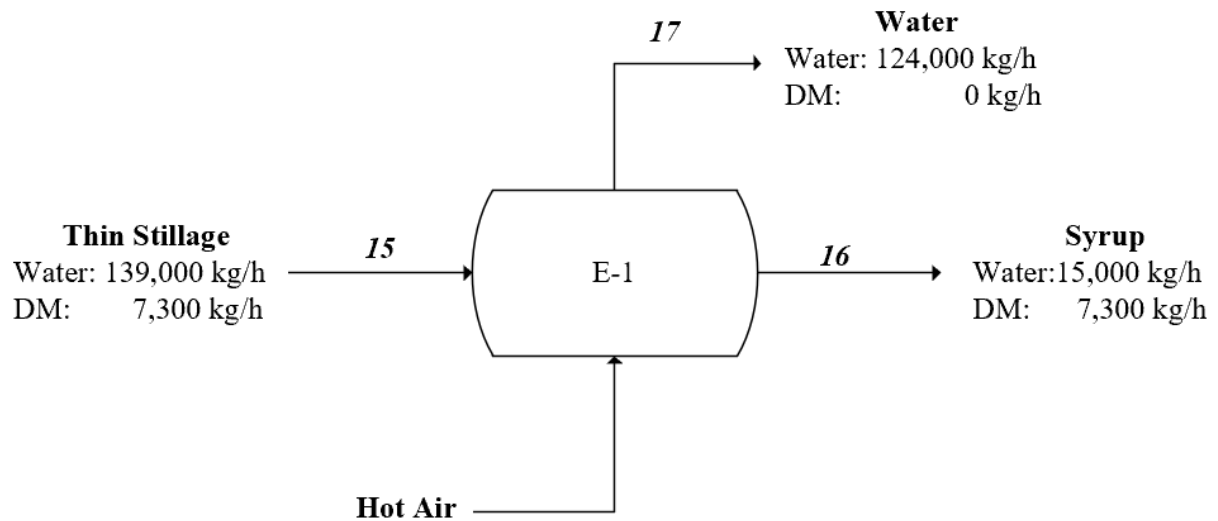
3.5.1 Thin Stillage

As seen above in Figure 3.4.8, centrifugation produces 146,000 kg/h of thin stillage. The thin stillage consists of 95% (w/w) water and 5% (w/w) DM. This water-rich solution is passed through a series of evaporators (E-1) that ultimately concentrate the syrup until it is approximately 33% (w/w) DM. This sugar rich syrup is then mixed with the processed DWG to produce the commercial-grade DDGS.

Thin film evaporators constructed from 316 SS are used to process the thin stillage. The evaporator reduces moisture content by dropping the thin stillage through a rotor enveloped in a heating jacket (Aptia Engineering, 2021). The heated jacket warms the falling water drops and produces a water vapor stream. The combination of the heated jacket and rotor maximize particle exposure to thermal energy while minimizing residence time and fouling potential (Aptia Engineering, 2021). The evaporator stack operates in a series where the stream from one tower is fed into the next. The water vapor stream exits co-current to the solution. This exit stream consists of 124,000 kg/h of pure water, as seen in Figure 3.4.9. B&P falling film evaporators with a maximum evaporation capacity of 56,000 kg/h of water will be operated simultaneously (B&P Engineering). A fifth evaporator is included in the design to accommodate for any maintenance or equipment issues. The falling film evaporators are operated below their production capacities to prevent fouling issues. The ~33 % (w/w) of syrup is pumped with a 316 SS positive displacement pump (P-999) with a power requirement of 0.62 kW. A complete evaporation material balance is available below in Figure 3.4.9. Heated air is used in the falling film evaporators. Natural gas is used to meet the heating requirement. ~119,000 GJ of energy is needed annually in order to remove the water exit stream. Multi-effect evaporation in the falling film evaporators was not considered as part of the energy calculations, and would reduce the

heating burden if calculated. This reduction in natural gas consumption is a potential area of plant refinement in future design work.

Figure 3.4.9 Evaporation Material Balance



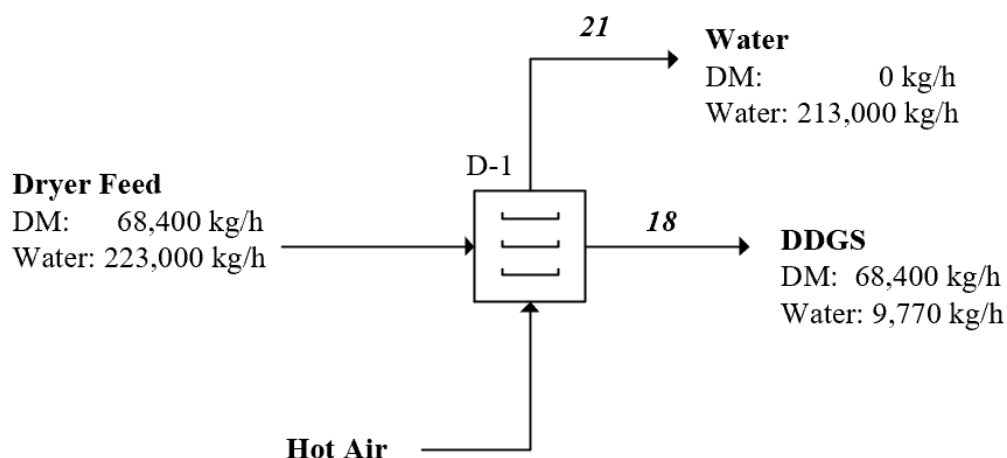
3.5.2 DWG

The DWG centrifuge output (X-3) is combined with the evaporated syrup from E-1 before being dried in D-1. The total moisture content of this stream is approximately 73% (w/w). This combined stream has a moisture content that is far too high for sale as DDGS. The moisture must be reduced to ~12% (w/w). Several rotary drum dryers are used for this purpose. A material balance around the dryer is available below in Figure 3.4.10.

The dryers reduce moisture via hot air that is blasted counter-currently into the rotating drum (FEECO International, 2022). The drum continues to rotate as the DWG passes through it, with lifting flights tumbling the grain through the hot air stream. Lifting and dropping the DWG increases the surface area of the grain that is exposed to the hot air, and directly increases the drying efficiency (FEECO International, 2022). The drum is also at a slight incline which promotes the movement of the DWG through the drum.

Model ONL-165 Rotary Drum Dryers were sourced in the design. Each dryer has dimensions of 18.3 m in length and 4 m in diameter (Onix Corporation, 2021). Given the capacity limitations of the dryer, 24 are included in the design to accommodate the entire DWG input. An additional 3 are included in the design specifications to accommodate for maintenance and cleaning issues.

Figure 3.4.10 Drying Material Balance



3.6 Heat Transfer Equipment

For this production facility, heat exchangers serve as the primary method of heat transfer. Each heat exchanger followed a similar design process. While several heat exchangers were designed to meet different production requirements, each was designed in a similar fashion. Schedule 40 pipes were selected for the ‘tubes’ of the heat exchangers given their cost, thermal transfer, and the low pressure operating conditions. 316 SS was used as the construction material to prevent corrosion and impurities. The thermal conductivity was specified by locating 316 SS property data in Perry’s Handbook. Convective heat transfer coefficients were estimated using Equation 4.6.1. The estimations relied on a fluid velocity of 3 m/s. Based upon the Reynolds number calculations, each heat exchanger operates under turbulent conditions. The total rate of heat transfer was calculated using the specific heat capacity of the slurry. The logarithmic mean temperature was determined and used to calculate the area required to achieve the necessary heat transfer rate. Given different requirements for each heat transfer scenario, either steam or water was used for heat transfer.

Equation 4.6.1 Heat Transfer Equations

$$\frac{f}{2} = 0.125 \left[0.79 \ln(\text{Re}) - 1.64 \right]^{-2}$$

$$\text{Nu} = \frac{(f/2)(\text{Re} - 1,000)\text{Pr}}{1.0 + 12.7\sqrt{f/2}(\text{Pr}^{2/3} - 1)} \left(\frac{\mu}{\mu_s} \right)^{0.11}$$

$$h_i = \frac{\text{Nu } k}{D_i}$$

$$h_o = \frac{\text{Nu } k}{D_H}$$

$$U_o = \left[\frac{1}{h_o} + \frac{r_o \ln(r_o/r_i)}{k} + \frac{1}{h_i} \frac{r_o}{r_i} \right]^{-1}$$

$$A = \frac{U_o \Delta T_{lm}}{Q}$$

Where f is the friction factor, Re is the Reynolds number, Nu is the Nusselt number, Pr is the Prandtl number, μ is the viscosity, h_i and h_o are convective heat transfer coefficients, D_i and D_H are diameters, r_o and r_i are tube radii, k is the thermal conductivity, U_o is the overall heat transfer coefficient, ΔT_{lm} is the logarithmic mean temperature, Q is the rate of heat transfer, and A is the heat transfer area

To accurately evaluate the heat transfer requirements and necessary area, the overall heat transfer coefficient needed to be approximated. To achieve this, estimations based on the Nusselt number were used. The friction factor, Prandtl number, and Reynolds number were determined for each scenario, then used to determine h_o and h_i . All heat exchangers were operated in a countercurrent configuration to improve the effectiveness of the heat transfer fluid and area. The logarithmic mean temperature was highly influenced by the process fluid and heat transfer fluid employed. Cooling requirements were met using cooling water that was assumed to enter the

exchanger at 22°C. This temperature was determined using the dew point temperature in Iowa during summer months. While using cooling water as opposed to a refrigerant requires more cooling fluid and a larger heat exchanger surface area, refrigerants are significantly more expensive as process fluids. To maintain the economics of the facility, refrigerants were not implemented. Heating requirements were met using steam as the heat transfer fluid. Saturated steam is supplied at 10 bar. An on-site boiler is needed to generate this steam but was not designed or included as part of this design brief.

3.7 Material Handling - Pumps

Transporting materials around the plant to various process units is essential to the overall function of the facility. Pumps serve as the primary means of transferring the large process volumes needed to meet the product specification aims described above. Two types of pumps, centrifugal and positive displacement pumps, will serve to meet these needs. 13 pumps are needed to sufficiently transport the process fluids throughout the plant. Exact pump locations are detailed in Figure 4.1. The process design includes a spare pump for each pumping application to minimize any potential downtime stemming from equipment problems. This brings the total number of pumps purchased for the facility to 26. The pumps were sized to determine the hydraulic power and pump/utility costs. This was accomplished by determining the differential pressure and volumetric flow rate characteristic to each required pump, then taking the product (Equation 4.7.1.).

Equation 4.7.1 Pump Sizing

$$P_h = \Delta P \times V$$

This calculation was repeated for each pump, and the pump sizing for the process water pump (P-101) will be discussed below. The remaining pumps in the design followed a similar calculation procedure and are included in Table 4.7.1. While the hydraulic powers of each pump align with expectations, there are some minor deviations. The hydraulic power for the second condenser pump is greater than that of the first. This is largely due to the fact that the reflux ratio for C-2 is 4 times the reflux ratio of C-1, thus increasing the hydraulic power required.

The differential pressure, the actual pressure difference, frictional losses, and gravity head were considered to determine pump power requirements. For the primary liquefaction process water pump, the pressure differential between the source and destination is 0 atm. 0.5

atm was allotted for frictional losses in pipes. Since the process water pump is a centrifugal pump, $\frac{1}{3}$ of the total frictional losses, or 0.17 atm, were allocated for contributions from the control valve. Collectively, these pressure differentials resulted in a total differential pressure of 0.67 atm. The volumetric flow rate of the process water is 0.1 m³/s, resulting in a total hydraulic power requirement of approximately 7 kW. The hydraulic power is used in addition to material of construction to determine pricing in Section 5, Project Economics.

Table 4.7.1 Pump Sizing

Pump Number	Hydraulic Power (kW)
P-101	7.0
P-201	246.0
P-202	21.5
P-301	653.0
P-302	14.3
P-401	321.0
P-402	35.6
P-501	90.5
P-601	8.1
P-701	25.2
P-801	0.7
P-901	4.2
P-999	0.6

3.8 Water Treatment and Cooling

Water treatment is essential in the overall ethanol production process. Water is particularly important in the slurry synthesis and as a heat transfer medium. Contaminated process water added to the slurry mix can compromise the efficiency of the fermentation process and significantly reduce operational output. Water used for heat transfer can also become a significant burden to plant operations should excessive pipe scaling or microbial growth impede flow. Facilities, equipment, and scheduling for water treatment processes were deemed outside the scope of this analysis. Further investigation into onsite process water treatment and cooling tower design should be carried out prior to moving forward.

4. Final Process Design

Figure 4.1 Process Flow Diagram

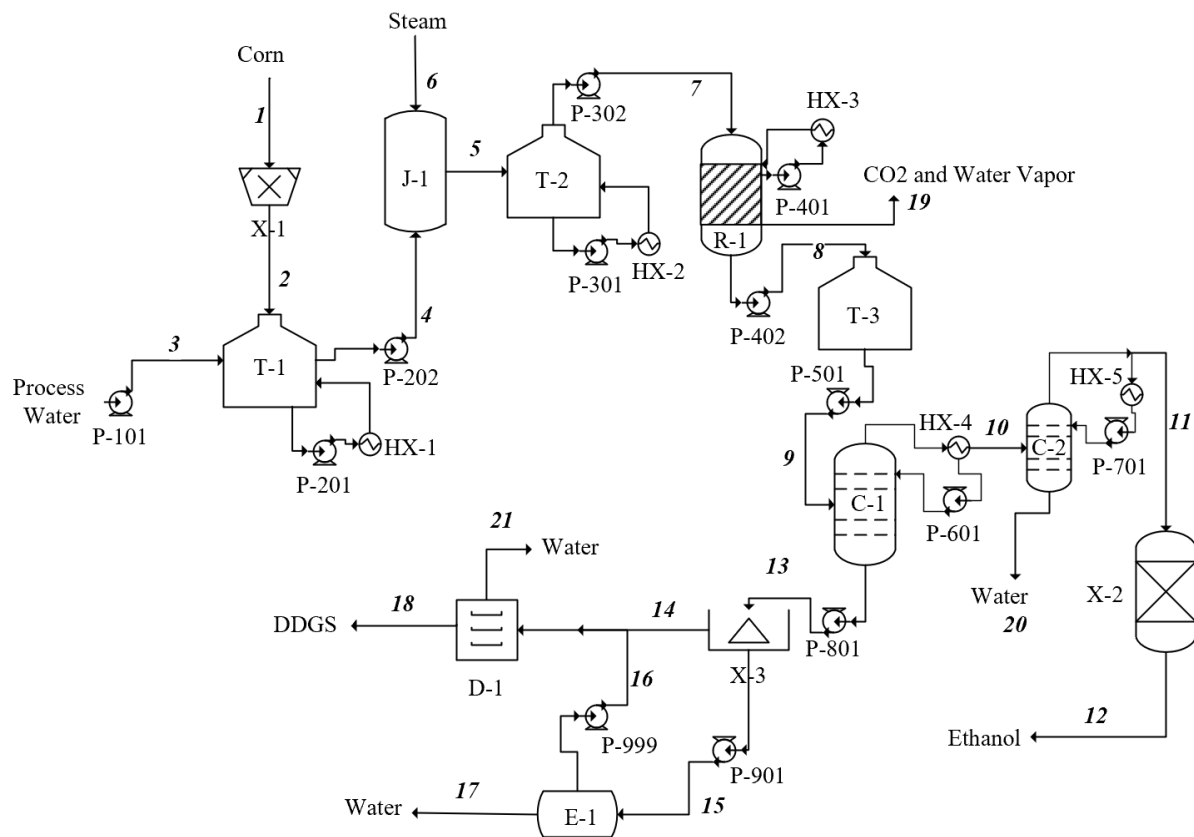


Figure 4.2 Stream Table

	Stream						
	1	2	3	4	5	6	7 *
Component							
Dry Matter (Corn)	210,000	210,000	-	210,000	210,000	-	210,000
Water	-	-	372,000	372,000	389,600	17,600	389,600
Ethanol	-	-	-	-	-	-	-
Alpha-amylase	-	-	-	26.2	26.2	-	69.8
Glucoamylase	-	-	-	-	-	-	-
Yeast	-	-	-	-	-	-	-
CO2	-	-	-	-	-	-	-
Total Mass Flow (kg/h)	210,000	210,000	372,000	582,000	600,000	17,600	600,000
	Stream						
	8 *	9	10	11	12	13	14
Component							
Dry Matter (Corn)	70,000	70,000	-	-	-	70,000	61,100
Water	389,000	389,000	43,000	4,000	280	346,000	208,000
Ethanol	57,000	57,000	57,000	57,000	57,000	-	-
Alpha-amylase	69.8	69.8	-	-	-	69.8	69.8
Glucoamylase	157	157	-	-	-	157	157
Yeast	92	92	-	-	-	92	92
CO2	-	-	-	-	-	-	-
Total Mass Flow (kg/h)	516,000	516,000	100,000	61,000	57,300	416,000	270,000
	Stream						
	15	16	17	18	19	20	21
Component							
Dry Matter (Corn)	7,300	7,300	-	68,400	-	-	-
Water	139,000	15,000	124,000	9,770	600	39,000	213,000
Ethanol	-	-	-	-	-	-	-
Alpha-amylase	-	-	-	69.8	-	-	-
Glucoamylase	-	-	-	157	-	-	-
Yeast	-	-	-	92	-	-	-
CO2	-	-	-	-	83,400	-	-
Total Mass Flow (kg/h)	146,000	22,300	124,000	78,500	84,000	39,000	213,000

* delineates streams associated with fermentation batch process

The final process converts the raw corn material into fuel grade ethanol via the six step process discussed in Section 3 (Figure 4.1 & Figure 4.2). 210,000 kg of raw corn material are processed hourly to supply the milled grain necessary for the process (Stream 1). 234 hammer mills (X-1) are actively required to mill the necessary corn. To prevent downtime in the event of malfunctions or machinery issues, an additional 24 hammermills will be included, thus requiring 258 hammermills total for this phase of the process. Given that milling requires the mechanical breakdown of the corn, this portion of the process is energy intensive. Milling requires approximately 1,000 kilowatts of energy, supplied as electricity. The milled corn is then transferred to a 250,000 gallon tank (T-1) to undergo liquefaction.

In primary liquefaction, the 210,000 kg/h of milled corn (Stream 2) is combined with 372,000 kg/h of process water (Stream 3) to begin forming the viscous slurry. Two primary liquefaction tanks are included to provide adequate cleaning and maintenance time. Process water is pumped in using a centrifugal pump operating at a hydraulic power of 7 kW (P-101). As the slurry is formed, 26.2 kg/h of α -amylase enzyme is added to break down the starches and convert them into fermentable sugars. Slurry pH and temperature are closely monitored to promote further starch degradation (Clifford, 2020). During primary liquefaction the slurry temperature is increased from 22°C to 85°C using a single pass shell and tube heat exchanger (HX-1). The slurry is pumped through the heat exchanger using a positive displacement pump operating at a hydraulic power of 246 kW (P-201). The heat source for the heat exchanger is steam at 180°C and 10 bar. The heat exchanger is composed of 316 SS and has an area of 1,100 m². The tubes are Schedule 40 pipes. Energy in this step of the process is supplied as steam and an equivalence of 32,400 kW is required.

The slurry is then fed into the steam-powered jet cooker (J-1) at a rate of 210,000 kg/h (Stream 4). In the jet cooker the corn is hydrolyzed to form mash, and is then subsequently cooled (ProSonix, 2021). 17,600 kg/h of steam (Stream 6) at 180°C and 10 bar is required to jet cook the slurry to a temperature of 115°C. To meet production requirements, 36 jet cookers will be included. 32 will be operated simultaneously, with 4 extras to account for potential malfunctions and issues.

After jet cooking, the slurry is transferred to the 250,000 gallon tank (T-2) for secondary liquefaction. Two tanks are included to account for cleaning and maintenance. A positive displacement pump operating at a hydraulic power of 21.5 kW is used to transport the slurry (P-202). The slurry is cooled to 32°C to prepare the mixture for fermentation. This is achieved using a single pass shell and tube heat exchanger (HX-2) with an area of 2,200 m². The heat exchanger is composed of 316 SS schedule 40 piping. The slurry is pumped into the heat exchanger using a positive displacement pump operating at a hydraulic power of 653 kW (P-301). The heat transfer fluid is cooling water at 22°C. Energy in this step of the process is supplied as cooling water and 47,000 kW is required. During secondary liquefaction, 43.6 kg/h of α -amylase is added to facilitate the continued breakdown of starches into digestible sugars.

600,000 kg/h of the prepared slurry (Stream 7) is then transferred to one of the 4 fermenters to undergo SSF. A positive displacement pump (P-302) operating at a hydraulic power of 14.3 kW is used to pump the slurry from the tank to the fermenter (R-1). SSF couples saccharification and fermentation into a single batch operation. SSF, pumping, and cleaning are completed in a 72 hour window. The fermentation window itself is 49 hours. 2 hours are included for pre-saccharification, and the remainder for pumping and cleaning. Given that this is the only noncontinuous portion of the overall process, this step dictates the operations for all

other processes involved in the plant. All material additions occurring during this phase of the process will be discussed in terms of batches. To meet the production quota, 5 total 4 million gallon fermentation vessels are included, each equipped with its own pump and heat exchanger. During fermentation, the yeast produces heat as a by-product that must be removed to maintain the desired 32°C fermentation temperature. A single pass shell and tube heat exchanger (HX-3) with an area of 2,500 m² is used to achieve this. Each fermenter is equipped with 2 identical heat exchangers. The heat exchangers are composed of 316 SS with schedule 40 piping serving as the tubes. Cooling water at 22°C will serve as the heat transfer fluid. 61,600 kW is required to meet cooling requirements. A positive displacement pump is employed to circulate the slurry through the heat exchanger, and requires hydraulic power of 320 kW (P-401).

2 hours prior to starting SSF, the pre-saccharification process introduces 2,800 kg of glucoamylase per fermenter to further promote raw material breakdown. In SSF and fermentation, all of the contents are added at once since this is a batch process. 1,650 kg of yeast are also added per tank. 0.7 million kg of the slurry is fermented in each fermenter every 72 hours. Following the pre-saccharification step, fermentation proceeds for 49 hours. 600 kg/h of water vapor leaves the fermenter, alongside 83,400 kg/h of carbon dioxide. The ethanol-rich beer is removed from the fermenter into the beer tank (T-3) at a rate of 516,000 kg/h (Stream 8). A positive displacement pump (P-402) with a hydraulic power of 35.6 kW is used to transfer beer from the fermenter to the beer tank. The beer exiting the fermenter has a composition of 11% (w/w) ethanol. The beer tank is 500,000 gallons and allows for continuous distillation downstream.

To convert the beer into fuel grade ethanol, the ethanol must be separated from the other components. To achieve this, beer is fed from the beer storage tank to the beer column (C-1) at a

mass flow rate of 516,000 kg/h (Stream 9) at a temperature of 32°C, as specified by fermentation. The beer is transferred using a positive displacement pump (P-501) operating at a hydraulic power of 90.5 kW. The beer column (C-1) serves to roughly separate the dry matter from the ethanol-water mixture to allow for less energy intensive, finer separation later in the process. To remove the dry matter and a portion of the process water a distillation column equipped with a total condenser and kettle reboiler is used (HX-4). The beer column (C-1) is operated at 5 bar and a reflux ratio of 1 is employed. 15 fixed valve trays with an efficiency of 70% are required to achieve the preliminary separation. The beer is fed to the column above stage 9. The trays are spaced by 0.6096 m, as suggested by current industry standards. The column is 9.14 m tall and has a tapered structure. The lower column diameter is 5.26 m and the upper column diameter is 3.72 m.

The beer column bottoms product is removed at a flow rate of 416,000 kg/h (Stream 13). The composition of the bottoms product is 17 wt% dry matter and 83 wt% water. All dry matter is removed from the column as bottoms product. The bottoms product is removed from the column at a temperature of 143°C and is later processed into DDGS. The mass specified reflux ratio is 1. The distillate is removed at a rate of 100,000 kg/h (Stream 10) with a composition of 43 % (w/w) water and 57 % (w/w) ethanol at 128°C. The condenser heat duty is -77,200 kW and the reboiler heat duty is 140,000 kW. The beer column (C-1) is equipped with duplicate reboilers to facilitate routine cleaning throughout the process, given the potential for fouling due to the large amount of solid content. The reflux pump for this column is a centrifugal pump (P-601) operating at a hydraulic power of 8.13 kW.

From the beer column, the ethanol-water distillate is fed to the rectifying column (C-2) at the specifications set by the distillate flow rate and distillate composition (Stream 10). The

rectifying column (C-2) seeks to separate the mixture to near its azeotropic composition limit. Given the requirements of separation in this column, the rectifying column operates with 25 sieve trays, each having a tray efficiency of 80%. The ethanol-water mixture is fed above stage 19. The same 0.6096 m tray spacing is also used in this column. The column is 15.2 m tall and has a column diameter of 5.10 m. A column pressure of 3 bar is used for this operation. The mass specified reflux ratio is 4. The rectifying column bottoms product is removed at a mass flow rate of 39,000 kg/h and is almost entirely composed of water. Only 27.2 kg/h of ethanol is lost in the bottoms product. This stream is further treated and removed from the plant as wastewater. The distillate is removed at a flow rate of 61,000 kg/h (Stream 11) with a composition of 93.6 % (w/w) ethanol and 6.4 % (w/w) water. A partial condenser (HX-5) is used to maintain the vapor phase of the distillate, which is needed for further dehydration using molecular sieves. The distillate is removed at 108°C and the bottoms product is removed at 134°C. The reboiler is a kettle reboiler. The condenser heat duty is -74,500 kW and the reboiler heat duty is 78,800 kW. The reflux pump on this column is a centrifugal pump which operates at a hydraulic power of 25.2 kW (P-701).

The ethanol-water vapor is fed at 61,000 kg/h (Stream 11) into a series of molecular sieves (X-2) to achieve a product purity of 99.5 wt% ethanol. The molecular sieve is a packed bed composed of $K_{12}[(AlO_2)_{12}(SiO_2)_{12}]-NH_2O$, or class 3A zeolite. The column is 29.7 m tall with a column diameter of 0.928 m. Thus, the total column volume is 20.1 m³ which is packed with 16,500 kg of class 3A zeolite. During adsorption, the column is operated at 3 bar. The ethanol-water vapor moves through the column at a linear velocity of 8.34 m/s. Two columns are included in the design to allow for simultaneous dehydration and regeneration. The regeneration process requires a hot air stream flowing at 598 kg/h to expel collected water from the bed. The

air stream is at 230°C, which requires 14,600 kW to heat using natural gas. 4,200 kW of energy is required for adsorption to heat the adsorbent. Adsorption and desorption run concurrently in 2 hour intervals to maintain the integrity of the beds. 57,000 kg/h of ethanol at 99.5 wt% (Stream 12) exits the molecular sieve. Conventional motor gasoline serves as a denaturant and is mixed with the ethanol stream at a rate of 1,080 kg/h. Following denaturant addition, the ethanol product is regarded as fuel-grade and is ready to be sold.

As the ethanol-water mixing is dehydrated using the molecular sieves, the bottoms product of the beer column (C-1) is further processed to produce DDGS as a coproduct. 416,000 kg/h of the stillage (Stream 13) is transferred from the bottom of the beer column (C-1) to the centrifuge (X-3) using a positive displacement pump operating (P-801) at a hydraulic power of 0.7 kW. 8 centrifuges are required to process the stillage, with 1 additional centrifuge procured to account for possible breakdowns and complications. The centrifuges process the stillage and separate into DWG and thin stillage. 270,000 kg/h exits the centrifuge as DWG (Stream 14), which has a water content of 77 wt% and 23 wt% dry matter. 146,000 kg/h exits the centrifuge as thin stillage (Stream 15) and has a water content of 95 wt%. To achieve the desired separations, the 750 mm diameter centrifuges are operated at a rotational speed of 2,500 rpm. 1270 kW of energy is required to power the collection of centrifuges, which is supplied as electricity.

The 146,000 kg/h of thin stillage is further processed to separate it into syrup and water. The thin stillage is transferred from the centrifuge to the falling film evaporator (E-1) using a positive displacement pump (P-901) that requires a hydraulic power of 4.19 kW. The falling film evaporator is implemented to concentrate a portion of the mixture to 33 wt% solids, which is then mixed with the DWG to prepare a higher quality DDGS. 14,400 kg/h of syrup (Stream 16) is taken off the falling film evaporator and combined with 270,000 kg/h of DWG (Stream 14). To

process this quantity of thin stillage, 4 falling film evaporators must be operated continuously. One additional falling film evaporator is allocated to allow for routine cleaning given the likelihood of fouling. The falling film evaporators are operated below their production capacities to prevent frequent fouling. The syrup is removed using a positive displacement pump operating (P-999) at a hydraulic power of 0.62 kW.

To remove remaining water from the grains, 291,000 kg/h of the mixture is processed using rotary fired dryers (D-1). Given the large quantity of water removed, this is an energy intensive portion of the process. The rotary fired dryers operate in a countercurrent configuration, introducing the grain mixture to a hot air stream at 200°C. 24 dryers operate continuously to handle the grain stream, with 3 additional dryers allotted for potential failures and routine maintenance. DDGS exits the dryer with a water content of 12.5 wt%, significantly reduced from the input stream. To achieve this level of water removal, 197,000 kW of energy is required to heat the air. Natural gas is used to prepare the air stream in a cost effective manner. Additionally, electrical power is required to meet the 1,230 kW energy requirement for the mechanical operation of the dryers. DDGS exits the dryer at a mass flow rate of 78,500 kg/h (Stream 18). Water is removed in the form of water vapor at a rate of 213,000 kg/h. The DDGS composition is 87.5 wt% grain content and the remainder water. This product is processed and ready to be marketed as animal feedstock. Downstream DDGS feed processors can pelletize the material and improve its commercial potential. These feed processors usually add sand silica to the final product in order to prevent buildup in the pellet-forming dies (Behnke, 2007). The remaining nutritional components illustrated in Table 2.2.2 are satisfied by the dried corn product generated during the overall process.

5. Project Economics

5.1 Revenue

The profitability of ethanol production has largely increased over the years due to imposed mandates and technological advancements. The fuel mandates implemented by the US government serve to ensure that the demand for ethanol will remain stable as long as the government stance remains unchanged. The current wholesale price for ethanol is at \$2.22 per gallon (Business Insider). Given that the plant yields 150 million gallons of ethanol annually, the revenue generated by the primary product is \$333 million. Net revenue is increased through the production of DDGS. DDGS retails at \$229.20 per short ton, given that the product is fully dried and meets the specifications outlined in Table 2.2.2 (Kennedy, 2022). Based on the plant production capabilities, \$156 million in revenue is made from DDGS annually. The total revenue for the plant is approximately \$489 million per year. Table 5.1.1 below details the plant revenue sources.

Table 5.1.1 Annual Plant Revenue

Revenue Source	Revenue (\$ MM)
Ethanol	333
DDGS	156
Total	489

5.2 Raw Materials

In ethanol production, the primary raw material is corn. Approximately 1.65 billion kg of corn is required annually to meet production requirements. Given that the average wholesale price of corn is \$150.40 per ton, the annual cost of corn is \$254 million. Over the past ten years, corn prices have steadily risen by approximately 20% (*Corn and soybean prices*, 2022).

Therefore, there is likely significant volatility in corn prices, which will influence the future economics of the plant. Process water is purchased from the city, at a cost of approximately \$6.00 per 1,000 gallons (*Water Rates*, 2022). 3.06 billion kg of water is required annually, and costs the plant \$4.86 million. *Saccharomyces cerevisiae* yeast are procured at \$6.75 per ton (Alibaba). Based on the annual yeast requirement of 718,000 kg, \$4,850 is spent annually on yeast. Enzymes are also purchased in bulk, with α -amylase costing \$2.93 per kg (*Thermostable alpha amylase for alcohol/beer brewing*). Glucoamylase costs \$2.20 per kg (*Beer enzyme glucoamylase/thermostable /alpha-amylase/ beta -amylase / glucoamylase*). Based on bulk costs and annual enzyme requirements, \$4.32 million is allocated to purchase enzymes per year. Conventional motor gasoline sourced for denaturant arrives via rail car and is purchased for \$12.5 million per year. The required materials sum to a total annual cost of \$276 million, representing a large portion of the overall operating costs. Table 5.2.1 below illustrates the plant raw material costs.

Table 5.2.1 Annual Plant Raw Material Costs

Process Input	Cost (\$ MM)
Corn	254.0
Process Water	4.9
Enzymes & Yeast	4.3
Denaturant	12.5
Total	276

5.3 Utilities

Utilities are separated into mechanical power, natural gas, steam, and cooling water. Mechanical power is detailed below in Table 5.3.1, and yields a total mechanical power requirement of 25.6 MW. Mechanical power is used to operate the equipment that enables milling, drying, pumping, and centrifugation. Specific process equipment uses are detailed in Table 5.3.1. Applying the cost of electrical energy of \$0.07 per kWh, mechanical power costs the plant \$14.1 million annually. Other energy-intensive operations are detailed below and were supplied with alternative utilities given the high cost of electricity.

Table 5.3.1 Mechanical Work Requirements

Process Equipment	Total Energy Required (MW)
Hammermills	10.5
Rotary Fired Dryer	12.4
Decanter Centrifuges	1.3
Pumps	1.4
Total	25.6

Steam was used to meet the heating requirements of pretreatment, jet cooking, and reboiler operation. Steam based heating requirements were calculated with steam that costs \$5.66 per GJ. The energy requirement for steam-based heating is detailed in Table 5.3.2 in GJ/year. 6.54 million GJ of steam-based energy are required annually. The largest portion of steam usage is accounted for by the reboilers that supply heat to the beer and rectifying columns. The total steam requirement at the plant costs \$37 million per year.

Table 5.3.2 Steam Heating Energy Requirements

Process Equipment	Energy (GJ/year)
Primary Liquefaction Heat Exchanger	921
Jet Cooking	314,000
Beer Column Reboiler	3,990,000
Rectifying Column Reboiler	2,240,000
Total	6,540,000

Natural gas is used to heat the air required for drying, the molecular sieves, and the evaporators. Since these processes require large amounts of energy, natural gas serves as the most economically viable energy source. Natural gas costs \$3.16 per GJ. Table 5.3.3 details the energy requirements of each process stage step. The heat required in drying is the largest portion of the overall natural gas usage at the plant. The total cost of natural gas energy is \$19.4 million per year. Alternative and more efficient approaches to drying should be explored in future work to address the large energy burden.

Table 5.3.3 Natural Gas Energy Requirements

Process Equipment	Energy (GJ/year)
Heated Air for Rotary Fired Dryers	5,600,000
Molecular Sieve Adsorption	414,000
Molecular Sieve Desorption	2,760,000
Heated Air for Evaporator	119,000
Total	8,890,000

Cooling water at 22 °C is used to meet production cooling requirements. Cooling is needed for the fermentation heat exchanger, the pretreatment cooling stage, and the column condensers. Cooling water is obtained at a cost of \$0.378 per GJ. Annual energy requirements

were determined and are detailed in Table 5.3.4. Total cooling water energy requirements sum to 7.4 million GJ annually. The total cost per year is \$2.8 million. Despite the large quantities of cooling water needed in fermentation and pretreatment, cooling water is still more economical than refrigeration. It is estimated that in the summer cooling water will enter at approximately 22°C. This water temperature will likely be lower during the winter season and therefore require less cooling water during these months.

Table 5.3.4 Cooling Water Energy Requirements

Process Equipment	Energy (GJ/year)
Fermentation Heat Exchanger	1,750,000
Secondary Liquefaction Heat Exchanger	1,340,000
Beer Column Condenser	2,200,000
Rectifying Column Condenser	2,120,000
Total	7,400,000

The total annual cost of utilities is \$73.3 million, and accounts for approximately 15% of the total annual revenue. The primary driver of utility cost and consumption is heating. Steam and natural gas expenses combined account for slightly over 75% of the total utility cost. Overall utility requirements by category and their cost are detailed below in Table 5.3.5.

Table 5.3.5 Overall Utility Requirements

Utility Category	Cost of Utility	Total GJ Required	Total Cost (\$/yr)
Mechanical	\$0.07/kWh	726,000	14,100,000
Steam	\$5.66/GJ	6,540,000	37,000,000
Natural Gas	\$3.16/GJ	8,890,000	19,400,000
Cooling Water	\$0.378/GJ	7,400,000	2,800,000
Total	—	23,600,000	72,900,000

5.4 Capital Expenses

The purchased equipment cost is the estimated capital needed to purchase all the equipment included in this design brief. As shown in Table 5.4.1, the equipment costs are estimated at \$71.3 million. These price values were obtained using the CAPCOST spreadsheet available in *Analysis Synthesis and Design of Chemical Processes*. The \$71.3 million CAPCOST estimate only includes the base equipment purchase price. Total installed cost for this equipment is calculated using a Lang Factor of 4.74. The Lang Factor is an approximate ratio of total installed cost to purchased equipment cost and accounts for additional expenses including site preparation, utilities connections, and process piping. The \$71.3 million of purchased equipment is multiplied by the Lang Factor to obtain a product of \$338 million. This \$338 million is used as the total plant cost estimate. Table 5.4.1 below details a complete breakdown of equipment and associated costs by process function.

Table 5.4.1 Capital Cost for Ethanol Plant

Section	Total Units	Capital Cost (\$ MM)	Total (\$ MM)
Milling			4.85
- Hammermill	258	4.85	
Liquefaction			5.6
- Storage Tank	4	0.70	
- Heat Exchanger	4	4.74	
- Jet Cooker	36	0.18	
SSF			27.9
- Fermentor	5	10.8	
- Heat Exchanger	10	16.9	
- Storage Tank	1	0.25	
Distillation & Dehydration			16.7
- Distillation Column	2	6.61	
- Condenser	2	1.37	
- Reboiler	2	3.42	
- Molecular Sieve	2	5.30	

Drying			15.0
- Centrifuge	9	0.29	
- Evaporator	5	0.29	
- Rotary Dryer	27	14.45	
Pumps			1.17
- Plant Pumps	26	1.17	
Plant Equipment			71.3
Lang Factor			4.74
Total Plant Capital Cost Estimate			338.0

The largest portion of equipment expenses are from the equipment used to perform SSF. The high cost is primarily associated with the large-scale heat exchangers used to remove excess heat from the sizable fermenter volumes. Rotary dryers are the second most expensive category of equipment. This high cost is driven by the 27 units needed to dry the large volume of grain involved in the process. Exploring alternative fermentation and drying technologies in future work could potentially reduce the overall equipment cost and annual depreciation expense.

5.5 Operating Expenses

Operating expenses are incurred throughout the year during normal operation. The primary operating expenses for the plant includes raw materials, utilities, and labor. Raw materials account for \$276 million annually as discussed in Section 5.2. Utilities account for \$73.3 million as discussed in Section 5.3. Labor is estimated at a total cost of \$2.4 million per year. A complete breakdown of labor requirements is detailed below in Table 5.5.1. The operating labor requirement was estimated using process equipment quantity. Values were sourced from Turton to determine reasonable salary and benefits costs. Supervisor requirement estimates were also drawn from Turton. It was estimated that 12 operators are needed to run the plant at any given time. At least one plant supervisor must also be on duty at any point in time. The plant labor structure of 32 operators and 4 supervisors offers redundancies at both positions and allows for rotating shift schedules. This enables management to call in additional operators and supervisors to perform additional duties like downtime maintenance. Operators primarily serve to perform routine checks, maintenance, and ensure that the process is running smoothly. Little supervisory support is required since most of the plant functions are automated. Table 5.5.2 details the annual expenses of the production facility. The total annual operating expenses are estimated at \$351 million per year.

Table 5.5.1 Labor Demand

Position	Salary + Benefits (USD/yr)	Quantity	Total cost (\$/yr)
Operator	60,000	32	1,920,000
Supervisor	120,000	4	480,00
Total			2,400,000

Table 5.5.2 Overall Expenses

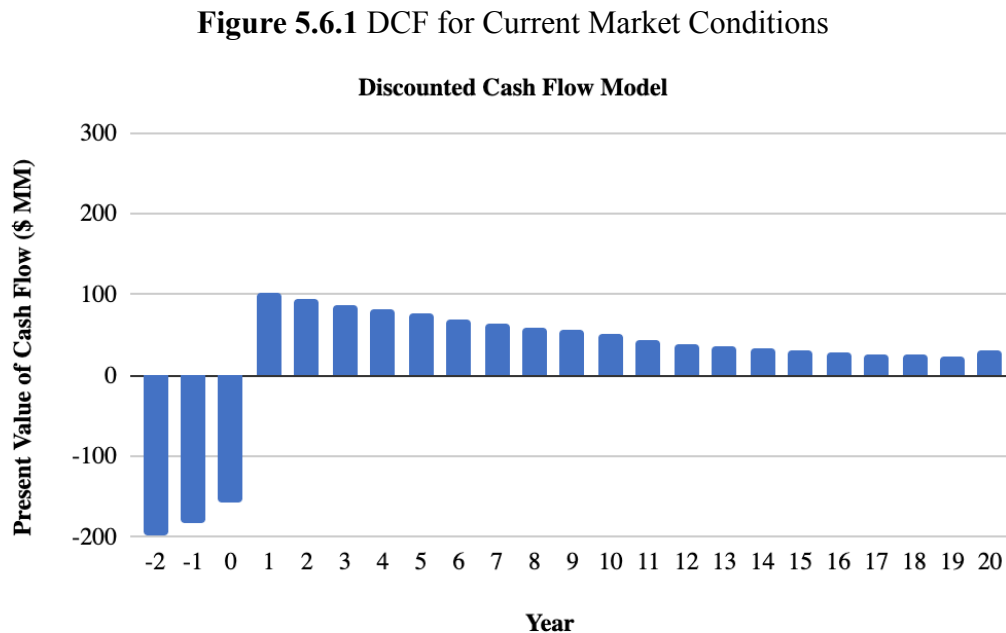
Expense Category	Cost (\$/yr)
Utilities	72,900,000
Raw Materials	276,000,000
Labor	2,400,000
Total	351,000,000

5.6 Economic Analysis

Economic analysis was conducted to evaluate the financial merits of the design. Land cost was estimated at \$500,000 and calculated as 50 acres at \$10,000 per acre. The purchase cost per acre was estimated with Iowa as the plant location. This \$500,000 for land was added to the plant capital cost from Section 5.4 (\$338 million) to obtain a total fixed capital input cost of \$338.5 million. This fixed capital input was divided in half and applied in the 2 years before the plant begins operation. An additional \$51 million of working capital was estimated in Year 0 as the plant slowly begins to operate. The plant is assumed to operate for only 6 months in Year 0. Revenues in Year 0 were assumed to be half of the annual revenue reported in Section 5.1, while the operational expenses reported in Section 5.5 were held constant for Years 0-20. Annual revenue reported in Section 5.1 (\$489 million) was assumed to be constant for Year 1-20. Year 20 also assumed additional cash inputs (\$66 million) for equipment salvage and the sale of the site land. The plant capital cost was depreciated using 10-year straight line depreciation. The depreciation expense reduced the tax expense present in Years 1-10. Taxes were estimated at 27%. The after-tax profit in Years 1-10 was a constant \$111 million while Years 11-19 saw after-tax profits of \$100 million. Year 20 saw additional after-tax profits of \$148 million given the additional revenues from plant liquidation. The plant has an IRR of 17.8% if current corn and ethanol prices are fixed over the 20-year plant life. The plant also has a cumulative cash position that turns positive in Year 5.

A discounted cash flow (DCF) analysis was used to evaluate the present value of estimated plant cash flows over time. Revenues and expenses were held constant for the scenario. Holding each value constant primarily assumes that corn and ethanol prices remain

unchanged over the next 20 years. The results of the DCF analysis are available below in Figure 5.6.1. When using a discount rate of 8%, the NPV of the plant is \$530 million.



The economic analysis indicates that the plant is a strong financial investment. The plant offers an extremely high IRR with a 5-year payback period. The primary drawback of these financial calculations is that corn and ethanol prices are unlikely to remain constant over the 20-year plant life. Since ethanol price is largely dependent on government mandate, changes in federal policy can quickly ruin the plant's economic prospects. Changes in corn price are also highly relevant considerations when evaluating the plant's financial outlook. Large increases in corn prices that are not met with corresponding increases in ethanol prices will quickly unravel the business case.

6. Safety, Health, and Environmental Considerations

Safety is an important consideration in any chemical process environment. Identifying and characterizing the hazards present in a process allows for enhanced recognition and engineering design to accommodate for and mitigate the potential impacts to plant operators and the surrounding community. Careful risk assessment and design also serves to mitigate capital risk and ensure plant longevity and ongoing financial performance.

The primary chemical safety hazards present in the facility are explosions and fires that stem from 2 major areas: fine-particle corn, and high concentration ethanol mixes. Since dry-mill corn ethanol facilities have seen prolonged operation and can be found throughout the United States, process hazards are generally well understood. Commonly referenced industry hazard understanding is used to guide the following process hazard analysis.

The most common hazard identified for corn ethanol production was dust accumulation from corn milling. The danger of dust build up is that the particles can combust, specifically corn dust can catch on fire and even combust if levels are high enough. This hazard can be mitigated by regular cleaning, and the most common measure that ethanol plants utilize is to isolate the milling operation from the rest of the plant, and then to minimize difficult to clean surfaces to prevent dust build up. Since there are a large number of hammermills, the danger of static electricity generated sparks can be mitigated using traditional grounding techniques, though the regular cleaning will help lower risk.

The second largest hazard is ethanol storage and transportation. The finished ethanol will be transported using rail car, meaning that a mass casualty event could arise from transportation failure. While this safety hazard falls on the shipping industry, there is still a hazard of leakage at the ethanol loading site. Spillage of ethanol can contribute to a critical safety incident, and is one

of the most common historical ethanol plant safety incidents. This critical safety event can be mitigated with proper storage and handling practices, however to ensure that these measures are implemented, routine inspections and training is required. The danger of ethanol spills and subsequent fires can be minimized using common safety practices, though the complete elimination of risk cannot be guaranteed. Another system of mitigation is constructing a moat around the ethanol storage to prevent seepage of the ethanol after a spill occurs.

In an effort to help mitigate the risk of particularly dangerous events, a Layers of Protection Analysis (LOPA) can be performed. This analysis provides guidance for the plant on the implementation of additional safeguards to minimize damage. Here, an example LOPA is given detailing the safeguards recommended for the Maximum Credible Event (MCE) of the plant. It was determined that a dust explosion created from the buildup of particulate matter surrounding the hammer mills is the most likely dangerous scenario to occur (Table 6.1).

Table 6.1 LOPA Process for MCE (Dust Explosion)

<i>Description of event: Dust explosion caused by small fire near hammer mills</i>		
Stage	LOPA Step	Process Attribute
1	Initiating Event (IE) Frequency	Small External Fire (10^{-1})
2	Severity Level	Very Serious (Employee Fatality Possible)
3	Likelihood	Unlikely (10^{-1} - 10^{-2} years)
4	Risk Level	Risk Level B
5	Target Mitigated Event Frequency (TMEF)	TMEF = 1×10^{-5}
6	Enabling Conditions	N/A
7	Conditional Modifiers	Probability of Personnel Being in Affected Area = 50% (0.5)

8	Adjusted Initiating Event Frequency	Adjusted = 0.05
9	Existing Layers of Protection	N/A
10	Frequency with Existing Layers of Protection	Current Frequency = 0.05
11	Additional Layers of Protection Required	Additional Layers of Protection = 0.0002

To alleviate the risk from this scenario, it would be wise to install blast walls around the hammer mill operating area. This would be exceptionally helpful in reducing the consequences of any explosions that occur, and help keep the equipment and workers safe. For purposes of the LOPA, the blast walls would be assigned a value of 10^{-3} (Crowl and Louvar, 2019). If these methods of protection were incorporated into the LOPA and used to reassess the risk matrix, the resulting risk would be reduced down to Risk Level C which is far more acceptable. For further guidance, the National Fire Protection Association (NFPA) publication *NFPA 652: Standard on the Fundamentals of Dust Explosions* can provide additional recommendations.

Along with taking measures to mitigate the chance of large-scale disasters, several safety measures will be enforced among the staff to avoid other incidents, including minor injuries as well as fatalities. All employees will be required to wear proper personal protective equipment (PPE) when within the confines of the plant, unless in certain designated areas such as offices and restrooms. The correct PPE will include safety glasses, a hard hat, pants, and safety boots. This gear will help to ensure that employees are protected from falling materials, projectiles, and other potential safety hazards that are common in a plant environment. Employees will also be required to wear a reflective vest, or a bright neon shirt, to ensure that they are highly visible to other employees. When cleaning dusty areas such as the hammer mills or silos, employees must

wear a dust mask to protect their throat and lungs. When handling ethanol, workers will be required to wear safety goggles and chemical-resistant gloves.

In order to dissuade employees from violating safety protocols, a three-strike system will be implemented where a strike will be received for certain safety violations, with three strikes resulting in immediate termination. Minor violations, such as not wearing or incorrectly wearing PPE, will be worth one strike. More severe incidences, such as smoking close to flammable materials, will be worth two strikes. So that all workers are aware of the safety protocols of the plant, no one will be permitted to work until viewing the initial safety video. This video will contain information on PPE, information on the three-strike policy, specific safety details about certain areas and process units of the plant, information regarding the chemicals in use in the plant, and procedures to be taken in case of various emergencies. Additionally, free CPR certification classes will be offered to all employees.

In addition to operation-based safety concerns, there are also several environmental implications that must be considered. When operated in its current configuration, the plant generates waste streams that must be treated before exiting the facility. In addition, large amounts of carbon dioxide and water vapor exit the fermenter. Since corn bioethanol is intended to aid in combating climate change and reduce overall emissions, this excessive emission is concerning. The carbon dioxide emissions must be addressed in order to improve the plant's environmental outlook. Future work and research should investigate carbon capture methods that would reduce the impact of ethanol production. The water vapor stream may also contain trace amounts of ethanol and should be processed appropriately. Given the large quantities of water

and heat consumed in the plant, this stream could likely be directed for use in other regions of the process.

The rotary fired dryers and evaporators also expel significant amounts of water vapor. Although likely devoid of ethanol contaminants, the water contains organic particulates from the grain and must be treated and processed appropriately. This water vapor stream may be integrated into the process to supply heat or process water, similarly to the water vapor expelled in fermentation. Integrating these waste streams and redirecting them throughout the plant would serve to improve the energy and material balances of the overall process. Limiting water consumption and net energy consumption would partially alleviate the significant environmental burden currently posed by the plant.

As the plant is currently designed, a significant amount of energy is lost in converting the corn into ethanol. The current plant design requires 21,700 TJ of energy while the annual ethanol product output only contains 11,200 TJ. This imbalance in energy input and output directly contradicts the “green” mission of the facility. Utility integration between process steps or integrating with a secondary facility would significantly aid in minimizing the overall energy loss. Improved heat integration and carbon capture can be areas of future work that will likely lead to significant improvements in the environmental viability of the plant.

Land-use impacts relating to the project are also of significant concern. The plant demands massive volumes of corn which in turn require large-scale industrial farming techniques. Industrial farming is generally seen as environmentally harmful because of its excessive land consumption and heavy use of fertilizer and pesticides. The soil degradation and chemical runoff associated with large-scale agriculture are also thought to be harmful. Directly

supporting the agricultural practices associated with corn farming is an important consideration when analyzing the full life-cycle impacts of corn bioethanol.

7. Social Impact and Considerations

One of the primary motivations of creating a bioethanol production facility is the manufacturing of “green” energy. Creating a viable energy source from a renewable and abundant material like corn can help to decrease dependence on the oil and gas markets that heavily contribute to global pollution and greenhouse gas emissions. Bioethanol blended into conventional gasoline helps to reduce the amount of harmful chemicals expelled into the atmosphere when the fuel is burned.

In its current state, the design for this particular ethanol production facility is off-target in meeting this energy goal. By producing 150,000,000 gallons of ethanol per year, the equivalent energy output of the output fuel corresponds to approximately 11,200 TJ per year. In order to operate the facility at this manufacturing capacity, calculations for utilities indicate that 21,700 TJ per year of energy will be required. Attempting to operate the facility at such a significant energy deficit appears to defeat the point of creating “green” energy, unless the utility energy is also sourced from renewables. Additional consideration needs to be put towards the energy that is required to harvest and grow the corn, which is not considered within the scope of this project. Even without the additional energy requirements from corn growth, the plant does not appear to be viable from the standpoint of an energy balance.

It is worth considering the validity of transitioning energy into a usable form. The energy that is stored in corn is unlocked through the bioethanol production process, and put towards an industry that can greatly benefit from it. The transportation sector accounts for roughly 35% of U.S. energy consumption totaling roughly 24.3 quadrillion Btu in the year 2020 (EIA, 2021). It has been stated that bioethanol serves as “the cleanest and most affordable source of octane on the global market today” (Renewable Fuels Association, 2017). This cleanliness suggests that

ethanol as an additive to the global fuel supply can have massive environmental ramifications on the battle against climate change.

Further environmental considerations revolve around the land usage that is necessary for a viable corn ethanol production process. In order to produce corn on the scale necessary for a feasible operation, natural conditions in the plant's location are altered severely. Corn requires a large amount of nitrogen in order to grow, which is often provided through potentially harmful ammonium nitrate based fertilizers. Groundwater usage and soil erosion occur at significantly higher rates, and overall biodiversity is disrupted due to the reallocation of naturally occurring nutrients (De Oliveira, 2005). The gradual degradation of the land used for such intense corn production is a concern for both the ethanol manufacturer and the farmer alike. Abusing the soil ensures that either costly reparations need to be made in order to revitalize the growing conditions, or lose out on periods of production.

For further development of the bioethanol industry, an increasing amount of land will be required to contribute to the domestic and global fuel market. To meet the needs of the entire fleet of U.S. automobiles, it is estimated that all of the currently available cropland will be allocated towards corn growth (De Oliveira, 2005). The need for such immense amounts of land and corn spell disaster on existing supply chains. In fact, corn-based bioethanol production already impacts the corn supply chain and leads to higher prices for consumers (Gardebroek, 2013). The creation of an additional American bioethanol plant will further increase the scarcity of corn, and result in even higher prices for domestic consumers.

Creating disruptions in the corn supply chain and raising prices will have a significant impact on the food security of families across the United States. Issues with food security will have a disproportionately more pronounced impact in lower income and ethnic minority

communities (Coleman-Jensen, 2014). Families in Iowa tend to have a household median income near the national average, but this might be slightly lower in the rural regions surrounding the plant. Native Iowans will be afforded the opportunity to work in both skilled and unskilled labor roles within the factory, which can help add vitality to the local economy. The creation of jobs may help to offset some of the food scarcity issues proposed in the local area of the factory.

8. Conclusion and Recommendations

Though the current operational capabilities of this plant allow it to fulfill the design basis and turn a small profit, it is not recommended that the project move forward in its current state. The economic benefits of the plant come with real environmental cost, and directly conflict with the mission of “greener” alternative energy sources. This environmental cost largely stems from the net negative energy balance of the plant, and the fact that this lost energy is sourced from fossil fuels. The high energy consumption is largely a result of using corn slurries with high solid content in the fermentation process. Using higher-weight corn slurries allows for more ethanol production per unit of corn input, and therefore reduces raw material costs. Unfortunately, the advantage in reduced raw material costs is negated by a corn slurry that requires more energy to transport, heat, and cool. Although the slurries are composed of less water, the heat needed to adequately heat the additional corn solids content is too large. Other facilities employ lower weight corn slurries that require less energy to heat and cool, though they do require more input corn. Since most of this energy demand for the facility is met using fossil fuels due to their low cost, the life cycle advantage of corn ethanol production is reduced. This failure is of particular significance given that one of the aims of ethanol production is to reduce greenhouse gas emissions and help combat climate change.

To reduce the net energy consumption and environmental impacts of the plant, additional research must be conducted. Some areas of improvement include the exploration of alternative technologies for DDGS drying, as well as sourcing “cleaner” energy from renewables. Onsite yeast propagation and yeast recycling will likely reduce material input costs. Carbon dioxide capture from fermentation should also be researched to determine if an additional revenue stream could be generated. Water streams obtained in the later stages of the processing steps can also be

“backset” and redirected back for use in manufacturing corn slurry mixes in liquefaction.

“Backsetting” downstream water waste streams into earlier process stages can significantly reduce the plant water consumption and aid in the plant’s environmental aim. By successfully balancing the economic viability of ethanol production with environmentalism, the corn ethanol plant may ultimately become a “Go” decision.

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11. Appendix

Figure 11.2 SSF Ethanol Production Overview (Devantier et al., 2005)

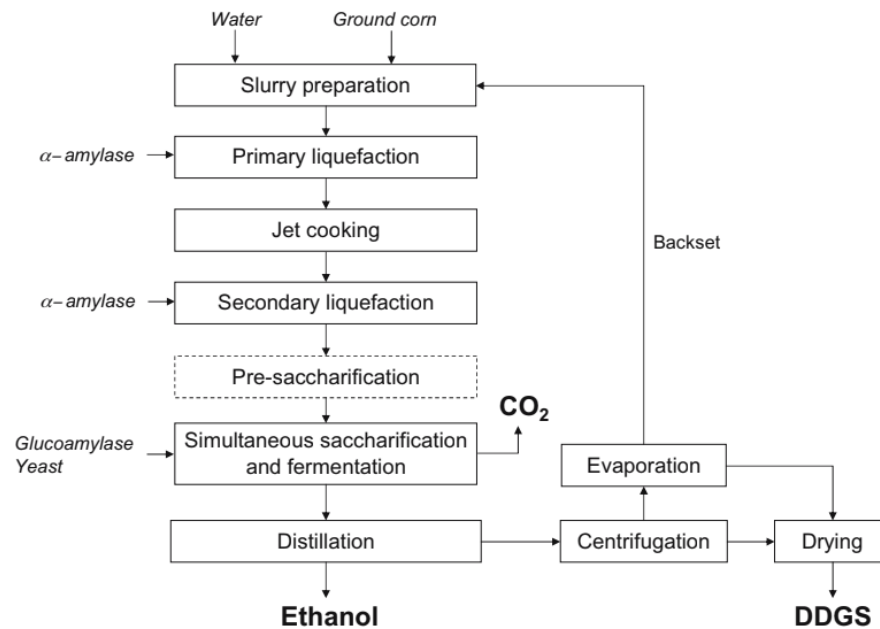


Figure 11.3 Ethanol Water Azeotrope (Volpato Filho, 2008)

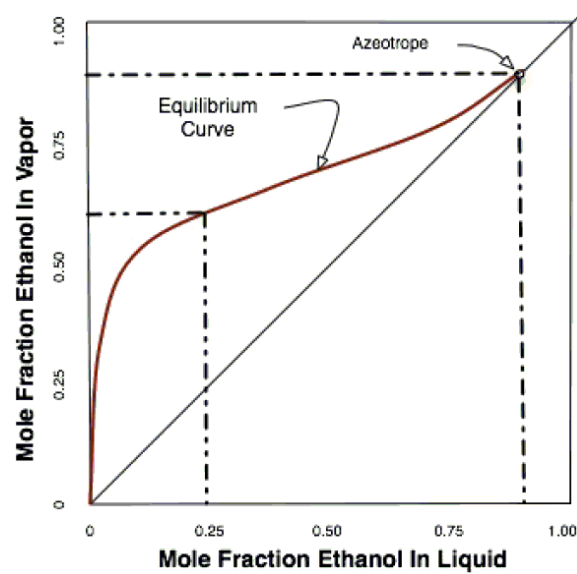


Figure 11.4 Ethanol Production and Cellular Kinetics (Devantier et al., 2005)

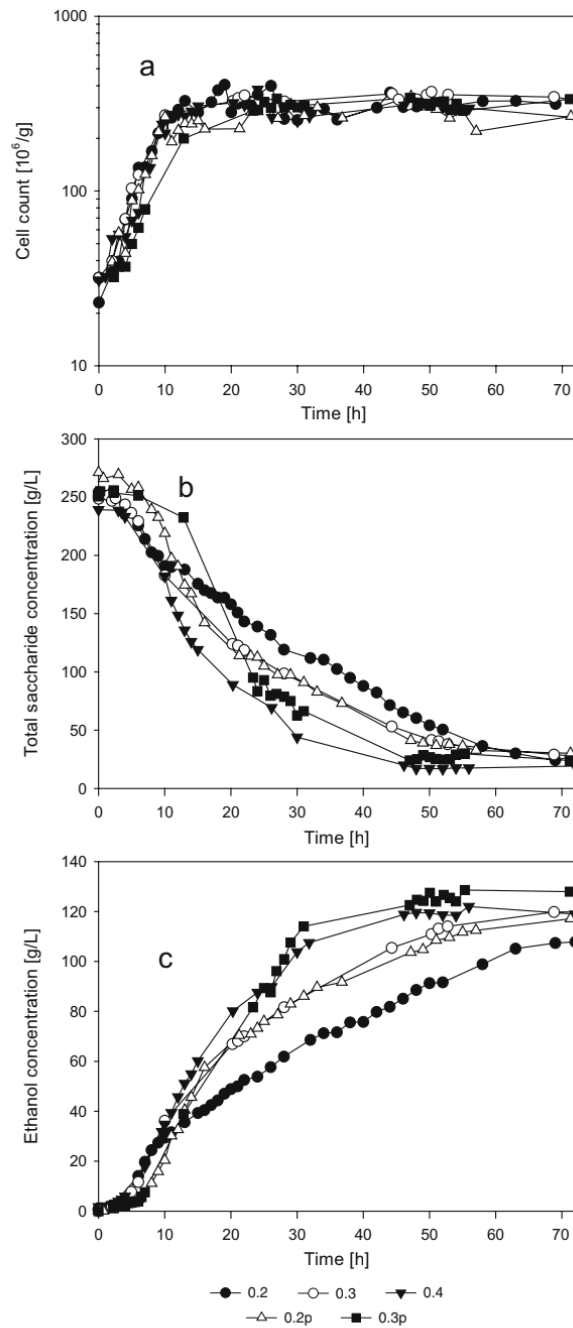


Fig. 2a-c Representative time courses of concentrations during fermentations with *Saccharomyces cerevisiae* Fermiol Super HA using a varying enzyme dosage with or without pre-saccharification. **a** Cell count. **b** Total saccharides. **c** Ethanol. *Numbers in the key* refer to glucoamylase dose [amyloglucosidase units (AGU)/g dry matter] and *suffix p* indicates the fermentations with pre-saccharification

Table 11.1 Glucoamylase Requirements (Devantier et al., 2005)

Glucoamylase dose (AGU/g dry matter)	0.2	0.3	0.4	0.2p ^a	0.3p	P ^b	Average ^c
$\mu_{\max}$ (h ⁻¹)	0.28	0.27	0.21	0.25	0.25	0.02	
Y_{sx} (C mol/C mol)	0.035	0.038	0.032	0.037	0.031	0.09	0.035
Y_{se} (C mol/C mol)	0.60	0.63	0.61	0.63	0.57	0.34	0.61
Y_{sg} (C mol/C mol)	0.039	0.045	0.041	0.044	0.032	0.07	0.040
Max (EtOH g/kg)	106	117	119	110	126	0.01	
Time of max EtOH (h)	73	57	46	57	49	0.01	
Productivity (g/kg/h)	1.4	2.1	2.6	1.9	2.6	0.001	
Cell count (10 ⁶ /g) ^d	279	328	278	282	317	0.33	296

^aSuffix p indicates the fermentations with pre-saccharification

^bFrom single-factor ANOVA

^cAverage shown if there was no significant difference within row. The significance level used was $P \leq 0.05$

^dAverage of stationary phase