

**Design of Bioethanol Production Facility**

**CHE 433 Group 22**

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## **Abbreviations**

DDGS: Distillers Dried Grains with Solubles

CO<sub>2</sub>: Carbon Dioxide

STEL: Short-term Exposure Limit

TWA: Time-weighted Average

SSF: Simultaneous Saccharification and Fermentation

CSTR: Continuous Stirred Tank Reactor

LMTD: Log Mean Temperature Difference

U: Overall Heat Transfer Coefficient, W/m<sup>2</sup>K

A: Area of Heat Transfer, m<sup>2</sup>

q: Rate of Heat Transfer in Heat Exchanger, W

## Introduction

Ethyl alcohol, or ethanol, is a 2 carbon alcohol that exhibits a wide range of applications in the everyday lives of most people. Of its many uses, its most important applications are for personal drinking, an additive for transportation fuel, sterilization/cleaning, etc [1]. The United States exports on average 3 million gallons of ethanol per day, or approximately 1.2 billion gallons of ethanol per year [1]. However, due to its large abundance and ease of production, its profitability is naturally very low, only selling for about \$1.64 per gallon, making it difficult to design and implement an ethanol production plant [2]. However, the fermentation process used to produce ethanol also has several by-products including a much more useful, 5 carbon alcohol, isoamyl alcohol, or fusel oil. This alcohol is used in the production of industrial solvents, flavoring agents, plasticizers, as well as in biofuel [2]. Due to its low selectivity in most yeast strains and its difficulty in extracting, it is a much more profitable product selling for about \$10.68 per gallon [4]. This project attempts to analyze the profitability of designing an ethanol production plant that is able to produce 75 million gallons of ethanol per year. However, because fusel oil is much more profitable, a secondary goal is to determine the economic impact of fusel oil recovery on an existing ethanol production plant. This requires additional separation equipment that adds to total capital investment, however also adds to total profit. Through this process, it is possible to determine the feasibility of diverting ethanol capacity to fusel oil production using such a fusel oil recovery system.

## Summary

Ethanol has many applications in everyday life as additives to fuel, a disinfectant, as well as for drinking. The production of ethanol involves a series of processing steps including fermentation, centrifugation, as well as several distillation steps. However, due to its large abundance and ease of production, its profitability can vary with market stimuli. Fortunately, fermentation also can produce several other byproducts, including fusel oil. Fusel oil has many uses including the production of industrial solvents, flavoring agents, plasticizers, as well as in biofuel. However, isolating this larger carbon alcohol requires additional unit operations to obtain a purified stream of the more profitable substance. The goal of this project was to first determine the profitability of designing a ethanol production plant to produce 75 million gallons of ethanol per year. However, because economics fluctuate, and fusel oil is much more profitable, a secondary goal was to determine the economic feasibility of diverting ethanol capacity to generating and recovering fusel oil at an assumed ratio of 5 gallons of fusel oil produced for ever 1000 gallons of ethanol.

An ethanol production plant was designed and simulated to produce the desired 75 MM gallons of ethanol per year. This was achieved through a staggered batch fermentation schedule and a continuous ethanol recovery process. Additionally, distiller's dried grains with solubles (DDGS) were recovered from fermentation solids. This plant served as the base case for a comparative analysis in investigating the economic impact of fusel recovery on the plant.

The fermentation section of the ethanol plant was sized to produce 75 million gallons of ethanol per year. The process began with a mass balance with several assumptions including a 90 percent conversion, an alcohol content of 16 vol%, etc. The product stream

resulted in a 200 cubic meter outlet stream with an alcohol concentration of 153 g/L. Using this parameter combined with a productivity of 3 g/(L hr), the fermentation time was determined to be 50 hours with a processing time of 5 hours and cleaning time of 5 hours. To meet the overall capacity in the specifications for the plant, 60 200 m<sup>3</sup> fermenters were required. Each of these fermenters operates in a bank of 5 fermenters, resulting in 12 banks in total, staggered 5 hours apart to maintain a continuous process. The overall purchase cost of the fermenters and two harvest tanks which hold the intermediate product before ethanol recovery came out to \$3.2 MM for the fermenters (the largest cost of any other equipment component in the plant). Additionally, to help maintain the continuous operation of the rest of the plant, two 1000 m<sup>3</sup> harvest tanks were purchased for a combined price of \$230,000. These tanks are necessary to hold volume from the batch fermenters while the continuous recovery process continues to operate between fermentation end times.

Ethanol recovery was performed through the utilization of a stripping column followed by a rectifying column. The stripping column—designed specifically for solids handling with the inclusion of Nutter BDP trays—allows for an initial separation of the liquid fermentation components from the solid components. This column was sized at a 3.1 m diameter and 7 m height to operate at 80% of flooding. This column was able to achieve satisfactory separation of liquids from solids with a bottoms composition at 30 weight-% solids. This column was priced at \$580,000 to purchase. The rectifying column was needed to purify and separate the ethanol product from the water and other liquids present from fermentation. To achieve this goal, the rectifying column was designed as a 26-stage sieve tray distillation column operating at 10 psig (3.5 to 3.9 m diameter and 19 meter height). A marketable concentration of 90% abv was produced in the tops with the 75 MM gallon per

year goal reached. The bottoms of this column (primarily water) are recycled back to fermentation to be diluted and reused as process water. This column was priced at \$520,000. Both columns were decided to be constructed out of stainless steel.

An additional recovered product was the DDGS from the fermentation solids. This was achieved through a drying process using two single-effect evaporators in series with 750 m<sup>2</sup> of heat transfer area each. This combination was able to reduce the moisture content of the DDGS from the 70% water in the feed down to a specified 15% water in the final product. The purchase of these evaporators was costed at \$1.4 MM for the combination of the two. These evaporators were chosen to be constructed out of stainless steel for the increased temperature and weather resistance.

In addition to the base-case ethanol plant, a fusel recovery system was also designed to capture the fusel oils naturally formed in the fermentation process. This fusel recovery was achieved through utilization of a gravity decanter followed by packed distillation column. The gravity decanter, an 8.0 m long vessel with a 0.50 m diameter provided the necessary 43 min residence time to induce spontaneous phase separation and output an enriched fusel oil layer from its tops stream. This enriched stream was then run through a 10 equivalent stage packed distillation column (0.5 m diameter and 6.1 m height) to purify the fusel oils into a marketable product. This column, while operating at 10 psig with a packing of 0.5" Raschig Super-Rings, was able to achieve 99% recovery of fusel oil. The stainless steel decanter was priced at \$16,000 with the stainless steel packed column costing \$64,000.

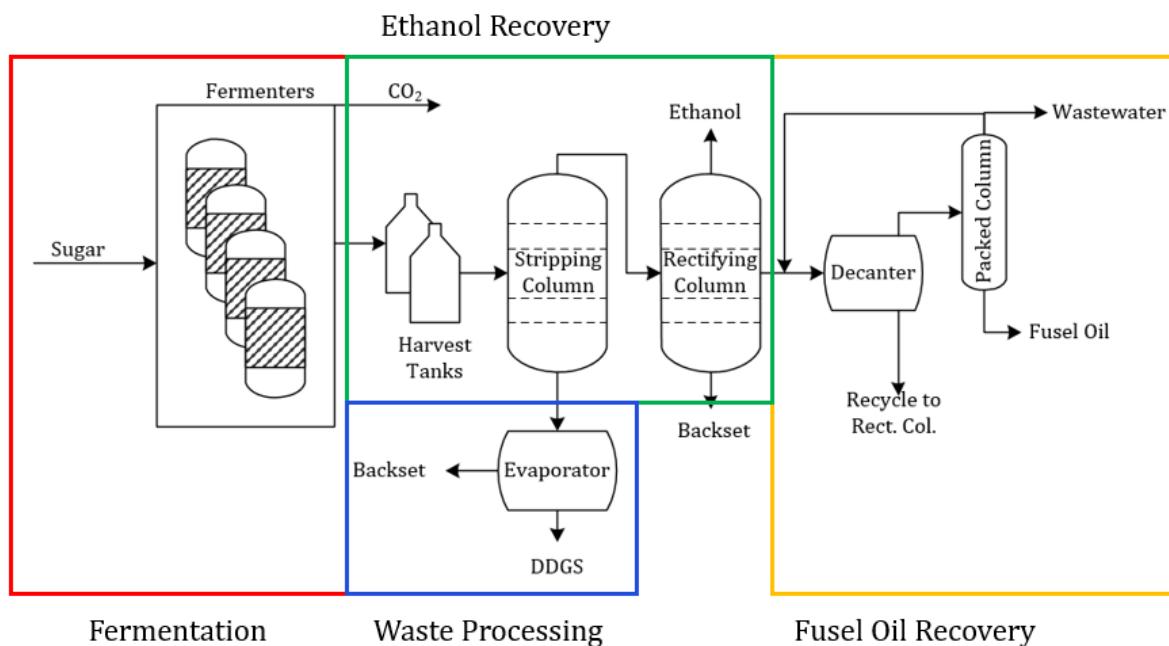
With the equipment costs for both the case with fusel recovery and sole ethanol production, the factorial method of cost estimation was used to arrive at values for capital

investment of \$28 MM and \$27 MM respectively. With this, the cash flow for the plant was analyzed over a hypothetical ten year operating period. Neither of the two plant prices were found to be profitable with the current market price of ethanol and purchase price of corn (\$1.64 per gallon and 3.56 per bushel. The net profit with respect to ethanol production was found to be slightly improved by fusel oil recovery at \$-0.31 per gallon compared to \$-0.33 per gallon without fusel oil recovery. Sensitivity analysis on the different aspects of the plants manufacturing expenses revealed both the price of corn and fusel oil to be the most impactful to the plants profitability and integrally linked. The break-even point for a range of corn purchase prices was found with the corresponding selling prices necessary for profitable operation.

With the very narrow profit margin present at current ethanol market price and corn purchase price, it is not recommended to construct an ethanol plant. If this analysis is related to an already functioning ethanol production facility, the investment into fusel oil recovery equipment is profitable and the additional revenue will help subsidize losses taken in the manufacturing of ethanol. With this consideration stated, any capacity which can be diverted from ethanol production into fusel oil production should be done. Ideally, this could be performed through using different strains of yeast with a higher selectivity for fusel oil. A final consideration to improve the profitability of the plant would be utilize a different starting material as a sugar source. It is possible there are organic waste streams from other local processes which could be used as a supplement to corn.

## Process Description

The bioethanol plant was designed to produce  $75 \times 10^6$  gallons of 90 vol% ethanol per year. The process comprised four main areas: a fermentation section with 60 fermenters and 2 harvest tanks, a ethanol recovery section that included a stripping and a rectifying column, a solids-waste processing section that recovered Distillers Dried Grains with Solubles, (DDGS), and a fusel oil recovery portion with a decanter and a packed column. A simplified flowchart that displays these four areas is shown in **Figure 1** below.



**Figure 1:** Simplified Process Flow Diagram

The fermentation portion of the process assumed that enzymatic hydrolysis of the corn mash and fermentation took place at the same time in the fermenters. This is known as Simultaneous Sachharification and Fermentation, or SSF. A conversion of 90% was assumed and a production rate of 5 gallons of fusel oil to 1000 gallons of ethanol [7]. The fermentation

times were staggered and stored in harvest tanks, which allowed for a continuous feed into the ethanol recovery portion of the plant. The fermentation schedule is shown in **Table 5** and the entering stream to the ethanol recovery portion of the plant, S-01, is shown in **Table A1.**

The ethanol recovery portion of the plant stripped the solids from the harvest tank using a stripping column. The bottoms product of the column comprising the solids and water fed into a set of evaporators. The tops product of the stripping column fed into a rectifying column that distilled ethanol to 90 vol% of ethanol and water. The bottoms product of the distillation column was made up only water, and an intermediate take-off stream fed into a heat exchanger which was a part of the fusel oil recovery portion of the plant.

The heat exchanger cooled the feed stream into the decanter to a temperature which allowed for the best separation between the oil and aqueous streams. The aqueous stream was fed back into the rectifying column for further separation, and the oil stream was fed into a packed column. The bottoms product of the packed column was made up of production quality fusel oil that can be sold on the market. The tops product of the packed column was made up of wastewater, some of which was recycled back to the feed to the decanter.

The waste processing portion of the plant recovered DDGS by subjecting the bottoms product of the stripping column to be fed into two evaporators. The evaporators removed the water from the solids, and the DDGS was recovered so that it can be used for other purposes that are not mentioned in this design.

**Figure A1** displays a flow chart that shows the ethanol recovery, waste processing, and fusel oil recovery portion of the plant. The fermentation section is not shown in the flow chart because it was designed as a semi-continuous portion of the process, and the modeling software used, *Aspen*, had limited semi-continuous modeling capabilities. The entering stream, S-01, is assumed to have a continuous flow from the harvest tanks with the flow rates shown in **Table A1**, which displays all of the streams in the rest of the process.

### Equipment Design

#### **Fermentation**

Fermentation is the process by which ethanol is produced as a by-product of cell growth from microorganisms. Several species of microorganisms are able to produce ethanol through fermentation such as different strains of yeast or bacteria. Although there are advantages and disadvantages for using various microorganisms, yeast is the most common organism used due to its productivity and lack of process complications [5]. However, microorganisms have several other by-products from fermentation, including isoamyl alcohol, or fusel oil, which is much more profitable than ethanol and has variable selectivity from different microorganisms. For this design project, the yeast strain *Saccharomyces cerevisiae* was used due to its popularity in ethanol production [5]. Designing the fermentation process for such a large scale ethanol production plant began with a mass balance around the fermenters with the goal of producing 75 million gallons of ethanol per year. However, because the plant is to be designed to recover fusel oil produced, it was assumed that one mole of ethanol is sacrificed for every mole of fusel oil generated. Along with this assumption, several other assumptions were made as shown below in **Table 1**.

**Table 1:** Assumptions Made for Mass Balance

**Assumptions [5], [6]**

- 
- Conversion of 90 mole % glucose to alcohol
  - 5 gallons fusel oil/ 1000 gallons ethanol
  - Maximum survivable alcohol content - 16 vol%
  - Negligible mass contribution from yeast and enzyme
  - Corn is 60 mass% starch
  - Corn is 10 mass% water
- 

These assumptions are made due to typical values used in industry for the production of ethanol at large scales. Because the yeast and enzyme mass contribution is small enough relative to the other feed materials, this mass can be assumed to be negligible when performing the mass balance. Using the ratio of 5 gallons of fusel oil produced for every 1000 gallons of ethanol produced, it was found that the fermenters needed to produce 7.1 kg/s ethanol and 0.035 kg/s fusel oil totaling 13,000 kmol alcohol per day. The amount of glucose needed was determined by using the alcohol content in the product stream in **Equation 1** below, assuming a 90 % conversion and a stoichiometric ratio of 2 moles of ethanol produced for every mole of glucose.

$$n_{glucose} = \frac{\frac{1}{2} * n_{alcohol}}{0.9} = \frac{\frac{1}{2} * 13,000}{0.9} = 7400 \frac{kmol}{day} \quad (1)$$

After converting the glucose needed to 1,300,000 kg/day, the amount of corn feed need was determined using **Equation 2** under the assumption that corn is 60 mass% glucose.

$$Corn\ Feed = \frac{Glucose\ Feed}{0.6} = \frac{1,300,000}{0.6} = 2,200,000 \frac{kg}{day} \quad (2)$$

Using this value and the assumption that corn is 10 mass% corn, it was concluded that the 26 kg/s corn feed consisted of 16 kg/s glucose, 8.0 kg/s solids, and 2.0 kg/s water. The exiting glucose flow rate was then calculated by multiplying the inlet glucose flow by

10% due to the 90% conversion, resulting in 1.5 kg/s glucose leaving the fermenter. However, this feed would result in an alcohol concentration much higher than 16 vol%, the threshold for the maximum survivable concentration of the yeast cells. Due to this, the feed was diluted to avoid the abrupt halt of cellular growth before reaching the desired concentration. To determine the amount of water needed, the volume of alcohol produced was first determined by using the densities of each material as shown below in **Equation 3**.

$$V_{alcohol} = \frac{m_{EtOH}}{\rho_{EtOH}} + \frac{m_{fusel}}{\rho_{fusel}} = \frac{7.1 \frac{kg}{s}}{789 \frac{kg}{m^3}} + \frac{0.035 \frac{kg}{s}}{810 \frac{kg}{m^3}} = 0.0090 \frac{m^3}{s} \quad (3)$$

The large amount of corn and excess glucose used in the process also imparts a large volume contribution to the final solution, thus requiring the calculation of its volumetric flow rate as well in determining additional water feed. The volumetric flow rate for this part of the stream was determined using a similar method as **Equation 3** with corn solids and glucose resulting in a volumetric flow rate of 0.016 m<sup>3</sup>/s. More detailed calculations are shown in Appendix B60-63. To obtain the proper alcoholic volume percentage, the remaining volume required must be compensated by additional water. The total amount of additional water required was then calculated to be 44 kg/s as shown in **Equations 4-5**.

$$V_{water} = \frac{V_{alcohol}}{0.16} - V_{alcohol} - V_{solids+glucose} = \frac{0.0090}{0.16} - 0.0090 - 0.016 = 0.046 m^3/s \quad (4)$$

$$m_{water} = V_{water} * \rho_{water} - m_{corn\ water} = 0.046 * 1000 - 2.0 = 44 kg/s \quad (5)$$

Another by-product of fermentation is also CO<sub>2</sub> which is produced in a stoichiometric ratio of 1 mole of CO<sub>2</sub> for every mole of alcohol produced. Although the CO<sub>2</sub> produced is in the gaseous phase and is immediately swept off from the fermenters in its own product stream, it was still important to determine the flow rate to understand the environmental

impact resulting from the fermentation process. The amount of CO<sub>2</sub> produced was determined by converting the number of moles of alcohol produced (13,000 kmol/day) to a mass flow rate shown in Appendix B59-63, which resulted in 6.8 kg/s. More detailed calculations for the entirety of the mass balance is shown in B59-63, however a summary of the mass balance showing inlet flow rates and outlet flow rates is shown in **Table 2**.

**Table 2:** Summary of Mass Balance with Flow Rates into and out of Fermenter

<i>Species</i>	<i>Inlet (kg/s)</i>	<i>Outlet (kg/s)</i>
Corn	26	-
• Glucose	16	1.5
• Solids	8.0	8.0
• Water	2.0	-
Water	44	46
Ethanol	-	7.1
Fusel Oil	-	0.035
Carbon Dioxide	-	6.8

The sizing of the fermenter design first required the calculation of other important processing parameters including the total alcohol concentration, as well as the total volumetric flow rate. The total alcohol concentration was determined by dividing the total alcohol flow rate by the total water volume as shown in **Equation 6**, while the volumetric flow rate was determined by dividing the alcohol concentration by its respective volume percentage as shown in **Equation 7** below.

$$C_{alcohol} = \frac{m_{methanol} + m_{fusel}}{V_{water}} = \frac{7.1 + 0.035}{0.046} * [Conversion] = 153 \frac{g}{L \cdot hr} \quad (6)$$

$$V = \frac{V_{alcohol}}{0.16} = \frac{0.0090}{0.16} * [Conversion] = 200 \frac{m^3}{hr} \quad (7)$$

The fermentation time required for this process was then determined by assuming a productivity of 3 g L<sup>-1</sup> hr<sup>-1</sup>, which is typical for batch fermentation processes with this specific

yeast strain [1]. **Equation 8** below illustrates the procedure for calculating fermentation time required.

$$\text{Fermentation Time} = \frac{C_{alcohol}}{\text{Productivity}} = \frac{153}{3} \sim 50 \text{ hrs} \quad (8)$$

In order to keep the production a continuous process, several banks of fermenters would be required that feed into one storage vessel with a controlled volumetric flow rate of 200 m<sup>3</sup>/hr. Brief economic analysis was performed to determine the storage size that balanced the higher cost characterized by larger storage tanks with the higher maintenance cost required for a greater number of banks of fermenters. This resulted in the conclusion of two storage vessels each 1000 m<sup>3</sup> in size, with one on stand-by to divert flow in the event the first must be cleaned. The processing time for each of these tanks was then determined using **Equation 9** below.

$$\text{Process Time} = \frac{V_{tank}}{V_{flow}} = \frac{1000 \text{ m}^3}{200 \frac{\text{m}^3}{\text{hr}}} = 5 \text{ hrs} \quad (9)$$

The total time for each fermentation process was calculated by adding the fermentation time to the processing time with an additional 5 hour cleaning time assumed between each batch of fermentation. This resulted in a total time of 60 hours required for each bank of fermenters. The number of banks required was determined by using **Equation 10** below.

$$\# \text{ banks} = \frac{\text{Total Time}}{\text{Processing Time}} = \frac{60 \text{ hrs}}{5 \text{ hrs}} = 12 \text{ banks of fermenters} \quad (10)$$

Due to the size of the storage tanks, each bank also required to have a total of 1000 m<sup>3</sup> in volume, however this did not necessarily need to be one vessel. Each bank was split up into five 200 m<sup>3</sup> vessels totaling 1000 m<sup>3</sup> due to the assumption that any bioreactor larger than 200 m<sup>3</sup> would require a cooling system, such as a cooling coil that would increase

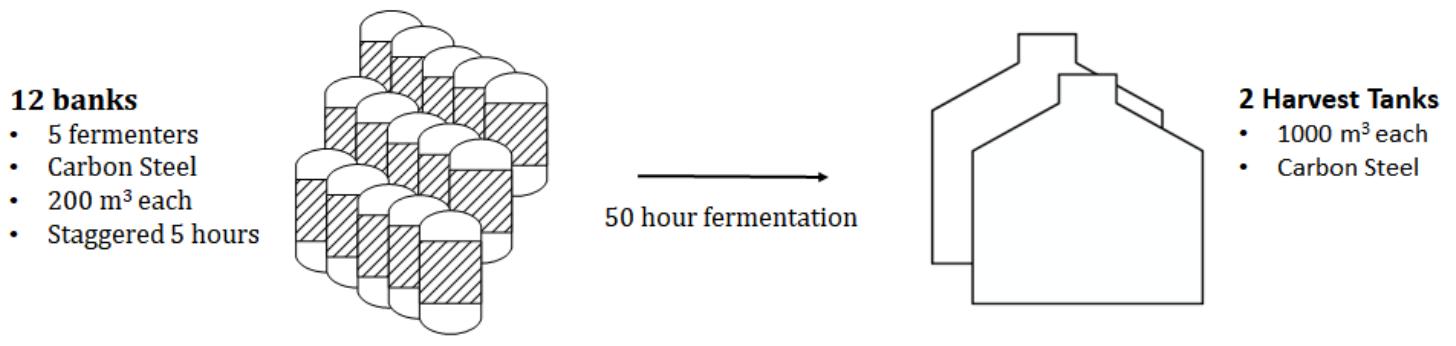
equipment cost significantly [1]. This then resulted in 12 banks of 5 fermenters each that were staggered 5 hours apart in order to keep the process continuous. In total, the process required 60 small, 200 m<sup>3</sup> fermenters that fed into a larger 1000 m<sup>3</sup> storage vessel (2 required, one to keep processing during cleaning). The material chosen for the storage vessels and the fermenters was carbon steel due to the limited risk of corrosion at such low concentrations of ethanol. The fermenters and storage vessels were costed as small field erected storage tanks since there is no mixing required in this batch process. A summary of the sizing/cost analysis of the process is shown below in **Tables 3 & 4**, as well as a process schematic of the design shown in **Figure 2**.

**Table 3:** Sizing and Costing of the Fermenters

<b>Fermenter Volume:</b>	200 m <sup>3</sup>
<b>No. of Vessels:</b>	60
<b>Material of Construction:</b>	Carbon Steel
<b>Purchase Cost:</b>	\$ 3,200,000

**Table 4:** Sizing and Costing of the Storage Vessels

<b>Harvest Tank Volume:</b>	1000 m <sup>3</sup>
<b>No. of Vessels</b>	2
<b>Material of Construction:</b>	Carbon Steel
<b>Purchase Cost:</b>	\$ 230,000



**Figure 2:** Proposed Process Schematic of the Formation Design

As stated earlier, these banks of fermenters were designed to be staggered every 5 hours to keep the process continuous. Due to this aspect, a fermenter schedule was developed to outline the proposed activity of each bank at each time point as shown below in **Table 5**.

**Table 5:** Proposed Fermenter Processing Schedule

Time	Bank 1	Bank 2	Bank 3	Bank 4	Bank 5	Bank 6	Bank 7	Bank 8	Bank 9	Bank 10	Bank 11	Bank 12
0 hrs	Flowing	Ferment	Cleaning									
5 hrs	Cleaning	Flowing	Ferment									
10 hrs	Ferment	Cleaning	Flowing	Ferment								
15 hrs	Ferment	Ferment	Cleaning	Flowing	Ferment							
20 hrs	Ferment	Ferment	Ferment	Cleaning	Flowing	Ferment						
25 hrs	Ferment	Ferment	Ferment	Ferment	Cleaning	Flowing	Ferment	Ferment	Ferment	Ferment	Ferment	Ferment
30 hrs	Ferment	Ferment	Ferment	Ferment	Ferment	Cleaning	Flowing	Ferment	Ferment	Ferment	Ferment	Ferment
35 hrs	Ferment	Ferment	Ferment	Ferment	Ferment	Ferment	Cleaning	Flowing	Ferment	Ferment	Ferment	Ferment
40 hrs	Ferment	Cleaning	Flowing	Ferment	Ferment	Ferment						
45 hrs	Ferment	Cleaning	Flowing	Ferment	Ferment							
50 hrs	Ferment	Cleaning	Flowing	Ferment								
55 hrs	Ferment	Cleaning	Flowing	Flowing								
60 hrs	Flowing	Ferment	Cleaning	Cleaning								

Under these processing conditions, this design could maintain a continuous flow of product from the fermenters that would then lead to a stripping column to remove excess solids from the stream as discussed next. More detailed mass balance and sizing calculations are shown in Appendix B59-63. Other economic parameters such as electricity cost, maintenance, etc. are discussed later in economics.

### ***DDGS Recovery***

The non-fermentable solids leaving the stripping column can be converted into DDGS. According to prior research, DDGS from corn has a moisture content of approximately 14% [7]. In order to achieve the production of DDGS, the water and ethanol must be evaporated from the stream leaving the stripping column, also called wet mash. The desired evaporation can be accomplished by two single-effect evaporators in series. The water and ethanol evaporated from the evaporators can be condensed into the liquid phase to be used elsewhere or recycled back into the process. The DDGS created can be transported and sold to specific buyers to be used for animal feed.

Design of a DDGS production system required an analysis of the feed and desired stream. The incoming feed stream from the stripping column had a moisture content of 70%. The desired moisture content for the final DDGS stream was 15%. It was evident that the water and ethanol needed to be evaporated from the non-fermentable solids that make up the DDGS. A mass balance around the evaporator was performed where it was assumed that all the solids are present in the more concentrated stream and no solids are present in the vapor stream. The mass balance was performed with an assumed exiting moisture content of 45%. The moisture content of 45% was used for the initial mass balance in the hand

calculations, (see App. B64-65), and may change during process optimization. It seemed unrealistic for one evaporator to evaporate such a large amount of the wet mash stream to 15% moisture content. This is where the addition of a second evaporator was incorporated into the DDGS production design. The second evaporator was added after the first evaporator to bring the moisture content of the wet mash stream from 45% to 15%. Similarly, it was assumed that no solids entered the vapor stream of the second evaporator and all the solids exited as the DDGS stream.

An energy balance on the evaporator was performed. The stream being used to evaporate the liquid from the wet mash steam was saturated steam at 150 °C. The temperature of the saturated steam was chosen as a saturated steam with a higher temperature could result in a greater stream enthalpy and greater heat transfer between the wet mash stream and the saturated steam. For the energy balance, it was assumed that all the steam was condensed into a liquid, and therefore the heat transferred between the two streams was equivalent to the heat of vaporization of the saturated steam. In addition, an assumption was made that no heat was lost to the surroundings during the heat transfer. The area of heat transfer could be calculated using **Equation 11** shown below.

$$q = U * A * LMTD \quad (11)$$

The overall heat transfer coefficient was obtained for stainless steel 304 using an online source [8]. The overall heat transfer used in the design of the evaporators was 680 W/m<sup>2</sup> K [8]. Using the amount of heat transfer, overall heat transfer coefficient (U) and the log mean temperature difference (LMTD), the area of heat transfer of the evaporators was

calculated. A summary table of the important design parameters for the DDGS production system is outlined in **Table 6**.

**Table 6:** DDGS Production Design Parameters

<b>Heat Transfer:</b>	23000 kW
<b>Overall heat transfer coefficient:</b>	680 W <sup>2</sup> /m <sup>2</sup> K
<b>Heat Transfer Area:</b>	750 m <sup>2</sup>
<b>Log Mean Temperature Difference(LMTD):</b>	44 K
<b>Moisture Content of Wet Mash:</b>	70 mass-%
<b>Moisture Content of Wet Solids:</b>	50 mass-%
<b>Moisture Content of DDGS:</b>	15 mass-%
<b>Material of Construction:</b>	304 SS
<b>Purchase Cost for 1 Evaporator:</b>	\$1,200,000

The designed DDGS production section of the designed bioethanol plant was modeled using the *Aspen Plus* simulator. A flash drum was used to model both evaporators. The heat duty parameter for the flash drum was optimized by minimizing the heat duty of both evaporators to obtain the desired 15% moisture content in the final DDGS stream. The pressures of both evaporators was set to 1 atm. Instead of including an actual saturated steam stream to enter the evaporator, the heat duty was specified. This is due to the fact that the evaporator was modeled as a flash drum in *Aspen*.

Costing the DDGS production system was accomplished by using a costing chart found in Timmerhaus [9]. The cost for a single evaporator with a heat transfer area of approximately 750 m<sup>3</sup> is \$590,000. Due to the similarity of design parameters of both the evaporators, the cost for each evaporator is assumed to be the same. The cost of the

evaporators needed for DDGS production is expensive, however, the revenue from selling DDGS is a big part of the total revenue that the designed bioethanol plant was designed to make. The DDGS is sold at \$135 per ton [10].

There are improvements that could be made with the DDGS production system. A double-effect evaporator can replace the current two single-effect evaporators in series. This would reduce the price for utility costs, and with the lower utilities, a larger and more effective evaporator could be purchased or designed. A double-effect evaporator design was difficult to properly model in *Aspen*, and further research and project time would allow for the proper design and integration of a double-effect evaporator in the DDGS production system.

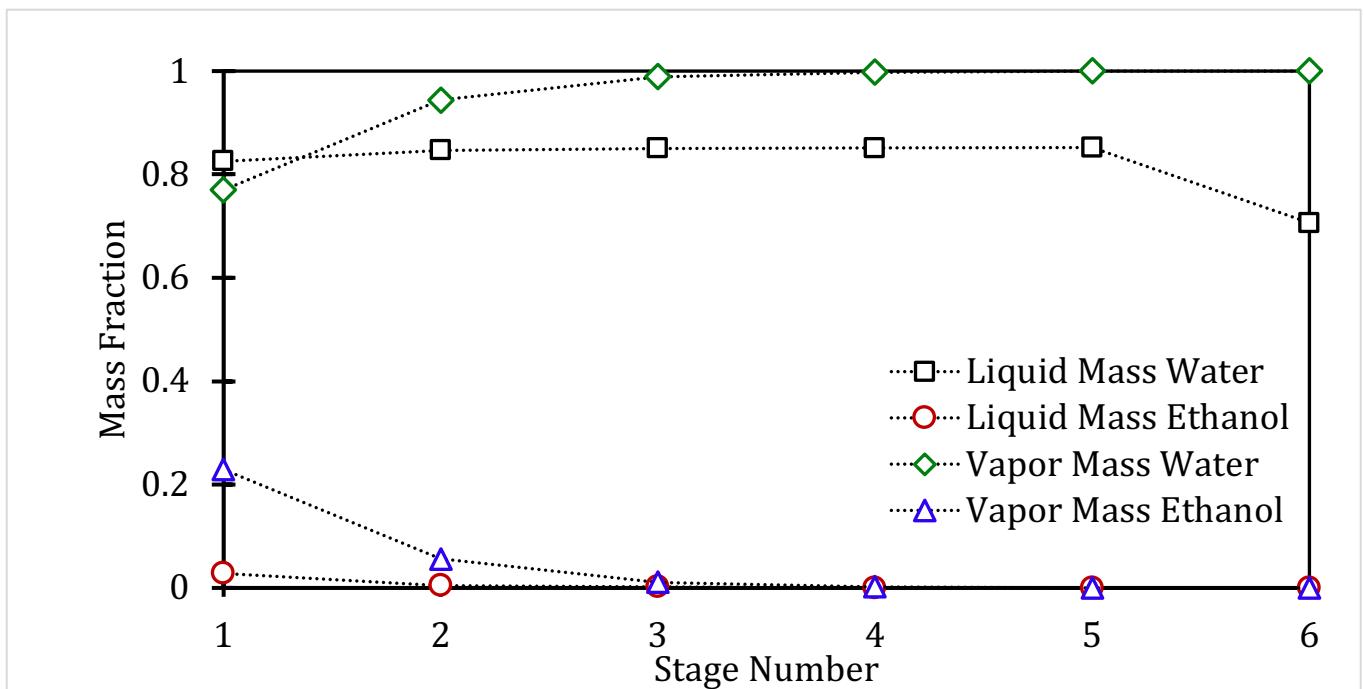
### **Stripping Column**

The initial feed into the ethanol recovery is comprised of both solids and liquids. Due to the solids being dissolved in the solution, filtration could not be used. Therefore, a stripping column was introduced into the process. The purpose of the stripping column was to separate the fusel oil and ethanol from the rest of the materials.

The first step in the column design, was to assume that all of the solids went to the bottoms off the column. This is a fair assumption as the vaporization temperature of the solids is greater than that of all of the liquids. To keep a negligibly changing viscosity within the column it was ideal to have a bottoms flowrate of 70 wt-% liquids to 30 wt-% solids. A simple mass balance around the column was then performed to obtain the mass flowrate of the distillate. Using these flowrates, a Fenske calculation was done to obtain the required stage number for the set separation. This calculation is shown in Appendix B66. The

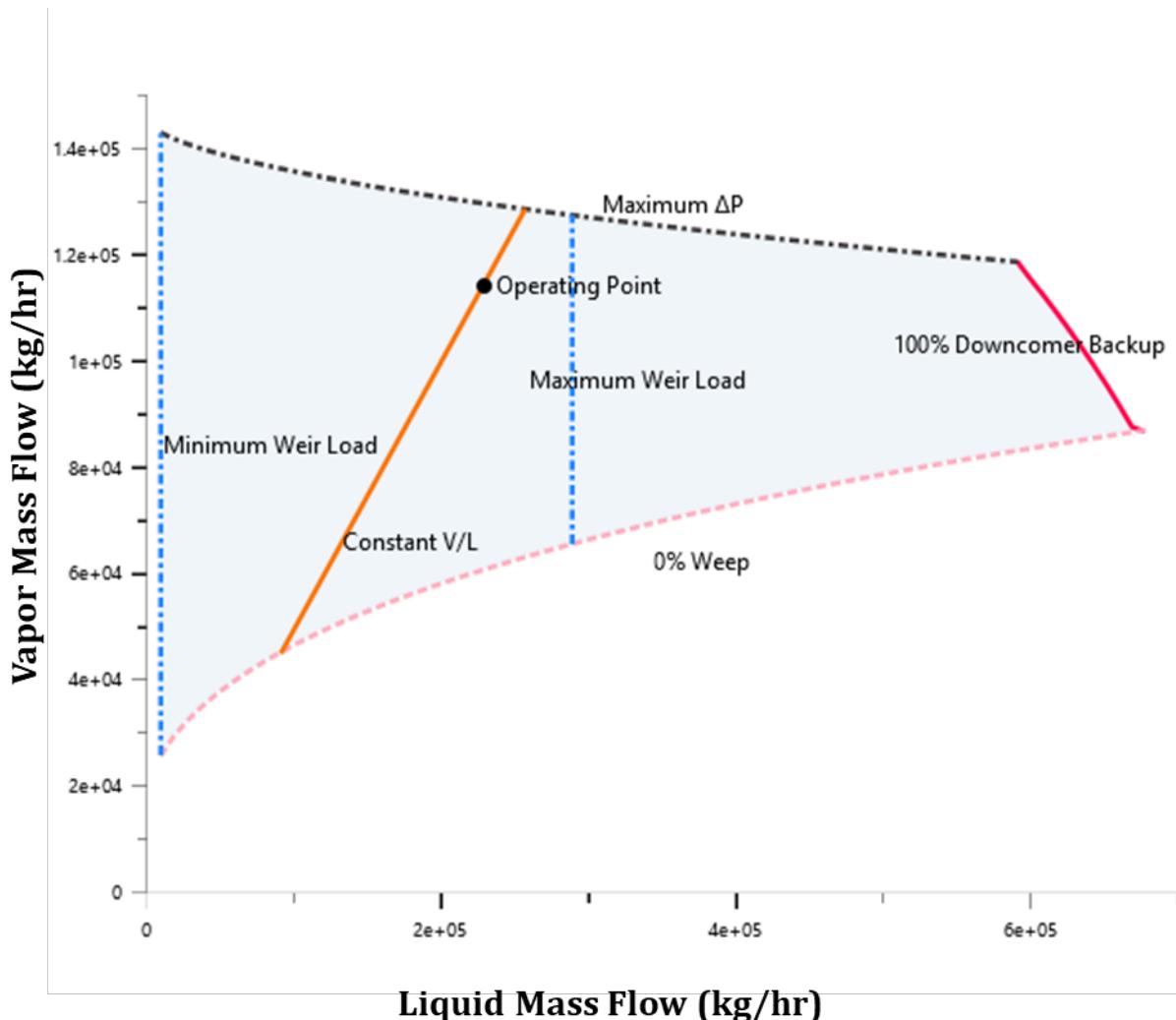
minimum number of stages necessary was found with the Fenske equation to be 5. Since, the small necessity for a precise separation in this column the actual number of stages could be the theoretical number required.

A design pressure of 10 psig was decided to operate at. This pressure is high enough to ensure driven flow through equipment, while not increasing the pressure to a point where equipment costs would increase. At these operating parameters an *Aspen* computational analysis was performed. A stage composition diagram was generated to ensure 5 stages was ideal. As you see in **Figure 3** below, the ethanol mass fraction is minimal at Stage 4 and therefore this is a better bottom of the column than Stage 5. Therefore, this number was based with Fenske calculation and optimized in *Aspen*.



**Figure 3:** Stripping Column Stage Compositions for Water and Ethanol

With the size and feeds of the column specified, a tray spacing and tray type were then decided. This was done through an iterative process to ensure the column would be within a proper operating zone on a hydraulic plot as well keep a minimal diameter, which directly correlated to cost of the column. A tray spacing of 3 feet with a Nutter-BDP tray type was optimal. The Nutter-BDP tray is a rectangular shaped valve set. The valves are oriented parallel to the liquid flow, allowing lateral vapor release, while allowing maximum solid flow to occur down the column. Hydraulic charts of the column with the same parameters but with a bubble cap valve vs a Nutter-BDP are shown in **Figures B2 & B3**, respectively. **Figure 4** is shown below and represents the final hydraulic plot for the designed column.



**Figure 4:** Stripping Column Optimized Hydraulic Plot

Using Figure 16-28 in Timmerhaus, a cost of \$580,000 for the stripping column could be obtained [7]. The overall parameters and specifications of the stripping column are listed in **Table 7**.

**Table 7:** Stripping Column Design

Operating Pressure	10 psig
Number of Stages	3 plus a reboiler
Number of Sections	1
Height(s), m	7
Tray Spacing, m	0.9
Diameter(s), m	3.2
% Flooding	80
Tray Perforation	NUTTER-BDP
Tray Thickness	10 gauge; 3.4 mm
Boilup Ratio (Mass Basis)	0.98
Tray and Column Material	Stainless Steel, 316, SA-240
Reboiler Duty, kW	69,000
Cost of Installed Column & Auxiliaries	\$580,000

### Rectifying Column

A distillation column was designed to separate the tops product stream coming out of the stripping column to a distillate product that contains 90 vol% ethanol and a distillate rate of  $75 \times 10^6$  gal/yr. The column was originally designed without the intent of recovering the fusel oil, and had fusel oil exiting with water in the bottoms product. The same column was used with the addition of the fusel oil recovery system to the plant, but included an intermediate take-off and organic feed recycle in the stripping section of the column. The simulation software *Aspen* was used to emulate both scenarios, and the RadFrac option was chosen so that a stage-by-stage analysis could be conducted. Detailed specifications for the column in the case with and without the recovery system are shown in Appendix B68-71. Below will describe the methodology of designing the column with the recovery system.

The streams, operating parameters, and design specifications were initially chosen by comparison of relevant studies in literature [8] [9] [10]. The intermediate take-off rate was varied until a maximum concentration of isoamyl alcohol was reached in the stripping

section of the column. The main feed stage was chosen to be above tray 9 because it provided the most optimal distillate and bottoms composition at a mass reflux ratio of 3. It was found that 26 stages allowed for the distillate rate and composition to be achieved. Upon inspection of the liquid composition plot, (see **Figure B4**), of the column that did not have intermediate take-off, isoamyl alcohol was found to be most concentrated above the 24<sup>th</sup> tray at a mass fraction of 0.082. This was due to the heterogeneous azeotrope that isoamyl alcohol and water form at 95.1 °C at atmospheric pressure [8]. An organic recycle stream was returned from the decanter to the column below the intermediate take-off stage to further separate alcohol-rich contents. It was returned to the stage below the take-off stage because the relative volatility of isoamyl alcohol and water is very high at the stage temperature (see **Figure B6**). This will cause the isoamyl alcohol to distill up the column and provide a higher concentration of isoamyl alcohol in the rectifying section. Pressure was determined to be constant throughout the column at 10 psig.

The distillation column was designed by using known correlations for the design of trayed columns and was also designed by using *Aspen* to determine the optimal parameters [11]. **Table 8** outlines key similarities and differences between each design.

**Table 8:** Rectifying Column Design

Specification	Correlation Value	Aspen Value
Number of Trays	25	25
Number of Sections	1	3
Height(s), m	12	4.3, 9.1, 1.2
Tray Spacing, m [12]	0.61	0.61
Diameter(s), m	3.2	3.9, 3.9, 3.5
Tray Perforation	Bubble-Cap	Sieve
Tray Thickness	12 gauge; 2.77 mm	10 gauge; 3.40 mm
Reflux Ratio (Mass Basis)	3.0	3.0
Boilup Ratio (Mass Basis)	0.58	0.58
Tray and Column Material	Stainless Steel, 316, SA-240	Stainless Steel, 316, SA-240
Cost of Installed Column & Auxiliaries	\$1,100,000	\$520,000
Condenser Duty, kW	-29,000	-29,000
Reboiler Duty, kW	31,000	31,000

The material of construction chosen for the trays and column was stainless steel because of its resistance against corrosion. It was also chosen because it has more thermal resistance compared to carbon steel, this would minimize heat loss over the column [13]. The column designed using correlations was \$547,690 more, primarily because of the use of bubble-cap trays. It was determined that sieve trays could be used for the operating conditions of the column, because there was minimal risk of column weeping. If the column were to be used for other processes or at other operating conditions, it is recommended that bubble-cap trays are used because they maximize tray efficiency by maximizing vapor-liquid contact [14]. A more detailed table that outline all design specifications of both columns is shown in **Table B3**. The ratio of the area of the holes to the active tray area was determined to be 0.1 to allow for a good enough turndown ratio and a low enough pressure drop inside of the column [14]. The fraction of flooding in each column was determined to be 0.80 [14].

It was chosen that the design specifications from *Aspen* are used for this process with the caution that this column can only be used for this process. The operating points of the first 6 stages are close to the entrainment region; the hydrodynamics of the rectifying section of the column need to undergo a more rigorous review before purchasing. The sections of the column may need to be separated and run in parallel. A plot that depicts the vapor and liquid volumetric flow rates throughout each stage is shown in **Figure B5**. **Figures B7 & B8** shows the liquid and vapor composition profiles, respectively. A hydraulic analysis report is available upon request that displays hydraulic plots of each stage, as well as other hydraulic parameters concerning the column<sup>1</sup>. Hand calculations that show column design are in Appendix B74-75 and energy balances are shown in Appendix B72-73.

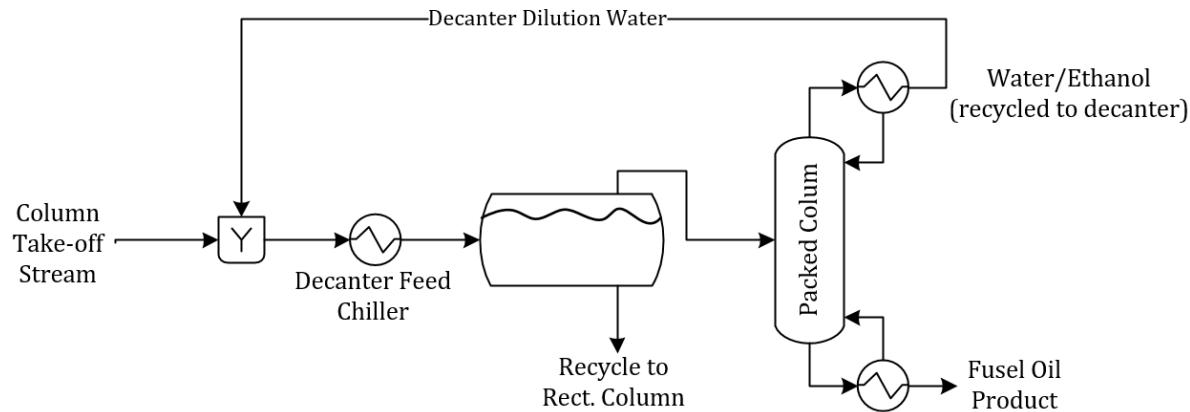
### Fusel Oil Recovery

To investigate the impact of recovering the fusel oils naturally present in ethanol production, a recovery system, serving as an add-on to the existing plant, was designed. As mentioned prior, fusel oil was shown through simulation to accumulate above Tray 24 of the rectifying column (COL-002). An intermediate take-off placed on this stage took a continuous stream of this fusel rich mixture through a series of unit operations for further separation and purification of the fusel oil product. The recovery strategy for this component consists of first passing the stream through a gravity decanter—allowing an initial spontaneous phase separation to enrich the concentration of fusel oil—and then continuing through a packed distillation column where the fusel oil could be concentrated to

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a marketable purity. The simplified layout for this add on to the existing ethanol production plant is presented below in **Figure 5**.



**Figure 5:** Simplified Diagram of the Fusel Oil Recovery Add-On to the Existing Ethanol Production Plant

*Not pictured: there is a take-off and make-up water stream present in the decanter dilution water recycle loop.*

#### Fusel Recovery Decanter

The first step in the recovery process of fusel oil from the intermediate take-off stream in the rectifying column, was to cool it down and pass it through a gravity decanter. A gravity decanter is essentially a vessel providing the residence time necessary for a spontaneous phase separation to occur between different liquid components in a mixture. As reported by Guymon et. al, the degree of separation between the fusel oil and the other components present in the mixture (ethanol and water) can be increased by operating the decanter at low temperature and by introducing a water stream to dilute the feed into the

decanter [6]. Both considerations serve to increase the levels of separation by taking advantage of ethanol's higher miscibility in water compared to fusel oil.

The design specifications for the decanter can be referenced below in **Table 9**. These operating parameters were simulated in *Aspen* Plus to produce an increase in fusel oil concentration from 2.0 mol-% in the decanter feed stream to 38 mol-% in the tops outlet of the decanter. This level of separation was found to be in the range reported by Guymon et. al. for a similar fusel recovery scheme which also utilized gravity decantation [6]. Sizing calculations for the gravity decanter were performed based on a calculated residence time necessary for the fusel oil to disperse from the continuous aqueous phase. This value was then taken and related to the length and diameter necessary to perform a successful phase separation as reported for petrochemical decanter design [14]. The length of the decanter was increased by 1-meter from the calculated value as a way to provide extra residence time in the case of process disturbances. Sample calculations for this sizing method are shown in Appendix B77-79.

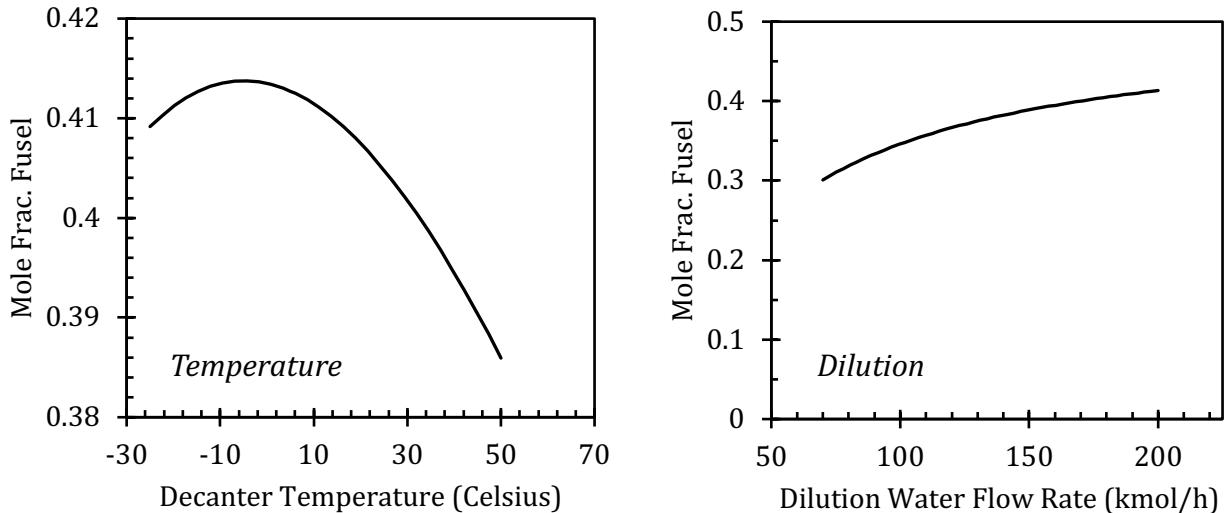
**Table 9:** Specifications and Cost for the Fusel Recovery System's Gravity Decanter  
*Costing is performed as a horizontal vessel with charts published by Peters and Timmerhaus [7].*

<b>Decanter length:</b>	8.0 m
<b>Decanter Diameter:</b>	0.5 m
<b>Decanter Volume:</b>	1.6 m <sup>3</sup>
<b>Residence Time:</b>	43 minutes
<b>Material of construction:</b>	316 SS
<b>Purchase Cost:</b>	\$ 16,000

Two of the key operating parameters for improving the separation observed in simulations of the gravity decanter were the temperature of the incoming feed mixture and

the extent to which the feed was diluted with water prior to entering the decanter. Simulations were performed to analyze the sensitivity these two parameters had on the mole fraction of fusel oil present in the tops outlet in the gravity decanter. The resulting relationships from this analysis are presented below in **Figure 6**. There is a clear maximum for the relationship with temperature (-8.0 °C), although it is present at a point which would put the system at risk for solids formation. To mitigate this risk and still operate near this ideal value, a heat exchanger was designed to chill the decanter's feed stream down to 5.0 °C. Although the presence of alcohol in the process fluid will depress the freezing point to some extent, safer operation is to be had at the chosen temperature. Looking into the impact of dilution on the outlet fusel oil concentration showed a positive relationship with increased fusel oil separation with increased dilution water flow rates. This is likely due to the increased tendency of larger volumes of water to draw out the more of the miscible ethanol into the aqueous layer in the separation. This relationship was seen as a trade-off with increased equipment and operating costs necessary for processing larger volumes of liquids through the decanter. It was chosen to use a recycle stream as a diluent for this purpose using the maximum flow rate possible, as increasing the size of this piece of equipment did not hugely impact the final purchase cost.

Start-up for this piece of equipment would require an initial input of a fresh water make-up stream to provide the necessary levels of dilution required for normal operation until the recycle stream can be brought up to full capacity. In this scenario, this added water will need to be slowly decreased in flow rate until the normal operating capacity in the recycle stream is reached.



**Figure 6:** Key Operating Design Choices for the Fusel Oil Recovery Gravity Decanter

*Left: relationship between temperature of the decanters incoming feed stream the mole fraction of fusel oil present in the tops of the gravity decanter. Right: relationship between the input flow rate of dilution water and the mole fraction of fusel oil present in the tops stream of the decanter. These figures were developed using sensitivity analyses in Aspen Plus.*

As a note on the simulation of this piece of equipment, the DECANTER block used in Aspen Plus does not allow the specification of residence time. Aspen's simulation instead assumes a decanter with a residence time long enough to result in a phase separation to the point of liquid-liquid equilibrium. With this consideration, the simulation would represent an ideal case, while the decanter which was sized by hand was found with the minimum separation residence time and then a safety factor added.

#### *Fusel Recovery Decanter Feed Stream Chiller*

The operating temperature for the gravity decanter was chosen at 5.0 °C for both satisfactory separation of fusel oil and safe operation. To achieve this feed stream

temperature, a heat exchanger was designed to provide the cooling necessary to bring the feed stream down to this value. The *Aspen Plus EDR* console was used to integrate heat exchanger design and optimization into the process simulation. This tool allowed for sizing of an ideal heat exchanger for each process change at the cost of computational simulation time. The results for the heat exchanger present in the final design are presented below in **Table 10**. This exchanger utilizes a coolant flow rate of just under 11 gpm to cool the heat exchanger feed down from 86 °C to the desired temperature of 5.0 °C.

**Table 10:** Specifications and Cost for the Fusel Recovery System's Feed Stream Chiller

*The exchanger utilizes a 1:1 mixture of ethylene glycol and water at -20°C as a coolant [7].*

<b>Tube-side Fluid:</b>	Water/Alcohol	<b>Shell Side Fluid:</b>	Coolant
<b>Tube-side Pressure Drop (psi):</b>	0.50	<b>Shell-side Pressure Drop (psi):</b>	0.20
<b>Heat Duty (kW):</b>	160	<b>Corrected LMTD (°C):</b>	24.0
<b>Exchanger Area (m<sup>3</sup>):</b>	50.	<b>Exchanger Length (m):</b>	6
<b>UA (kW/m<sup>3</sup>-K):</b>	0.1357	<b>Shell Diameter (m):</b>	0.36
<b>Head Type:</b>	Fixed	<b>Pitch:</b>	Square
<b>Tube OD (in)</b>	0.75	<b>No. of Tube/Shell Passes:</b>	1/1
<b>Material of Construction</b>	316 SS	<b>Max Operating Pressure (psia):</b>	150
<b>Exchanger Cost (USD): \$17,000</b>			

This heat exchanger is small and simple in the design with a minimal pressure drop both tube and shell side. Having the dilution stream mix with the decanters feed prior to entering the heat exchanger helps reduce the total heat duty required for the exchanger. The process fluid was chosen as the tube-side liquid, as there is a higher likelihood for fouling compared to the ethylene glycol coolant used in the shell-side of the exchanger. The water and alcohol mix alone is not a danger for fouling as it is essentially a dilute cleaning agent in itself, but with the nature of the process there is the possibility solids from the corn mash

could make their way into this portion of the system and foul the tubes. Although unlikely, errors in operation of the distillation columns could have this result and it is better to be safe in equipment design and place this fluid in the more easily cleaned portion of this exchanger. Additionally, it is safer to have the hot fluid present inside the tubes of the exchanger compared to the shell. If the shell were to be punctured, it is better for cold fluid to leak compared to alcohols near their vaporization temperatures. A more detailed specification list for this heat exchanger is shown in Appendix B80.

#### *Fusel Recovery Packed Distillation Column*

Following the initial separation performed by the gravity decanter, the enriched fusel oil stream passes through a small packed distillation column where it is purified into a marketable product. Simulations of this packed column showed the capacity for the equipment to take the incoming 38 mol-% fusel oil stream and purify it to 95 mol-% while achieving 99% recovery of the component. Specifications for the design of this distillation column are presented below in **Table 11**.

**Table 11:** Design Specifications for the Fusel Recovery Packed Distillation Column

*This piece of equipment was costed as the price of the vertical column and the price of packing in [7].*

<b>Total Column Height (m):</b>	6.1	<b>Operating Pressure (psig):</b>	10
<b>Packed Height (m):</b>	3.5	<b>Column Pressure Drop (psi):</b>	0.015
<b>Column Diameter (m):</b>	0.50	<b>Top Stage Temperature (°C):</b>	95
<b>Packing Type:</b>	½" Raschig Super-Ring	<b>Bottom Stage Temperature (°C):</b>	138
<b>HETP (ft)</b>	1.5	<b>Boil-up Ratio</b>	9
<b>Material of Construction</b>	316 SS	<b>Reflux Ratio:</b>	3
<b>Column Cost (USD):</b> \$64,000			

The addition of this distillation column as the final element of fusel product recovery deviates from the original design published by Guymon et. al. which has been extensively referenced for the fusel oil recovery section of this plant [6]. The mentioned publication included an extraction column to wash the fusel oil with water to further separate the final fusel product from the water and ethanol present in the process mixture. This extraction column was found difficult to simulate, with the robustness of the separation reliability also coming into question. At the cost of increased operating expenses due to the utility usage of distillation compared to extraction, it was decided distillation is more suitable for this process. The increased reliability of distillation will help account for disturbances in the continuous feed stream to the distillation column. Inconsistency of fermentation will result in fluctuations of the composition in the intermediate take-off stream entering the fusel recovery system and by effect, the separation needed to be performed by the packed column may also vary. To account for these variations, two additional stages were specified to be added to the distillation column. This would add increased separation capacity in the case a stream more dilute were to be delivered to the feed of the distillation column.

*Aspen* simulations performed on the column's operation verified the stability of each of the equivalent packed stages, ensuring the column was not flooding and the pressure drop was not too large. This was a key factor in the decision to use Raschig Super-Rings as the packing in the column. Compared to more traditional packing types (i.e. Berl saddles or Raschig Rings), this new high-efficiency random packing serves to increase the surface area present in the column while also minimizing the pressure drop [14]. This allows for the columns to be more efficient and operate in an operating region more stable and resistant to

flooding. An example hydraulic plot for the packed distillation column operating with this type of packing is shown in **Figure B10**.

### ***Overall System Simulation and Performance***

Simulation of the recovery system as a whole proved its robustness for handling many different capacities. It was found utilities usage was a large area to improve based on the initial plant's design. One of the key improvements made to the recovery system's design was the recycle stream from the packed column's distillate to the feed of the decanter to act as diluent. Without changing the size or specification of the column, dilution rates ranging from 0% up to 99% of this distillate stream resulted in stable operation (this process flow is shown in **Figure 6**). It was decided on to recycle 95% of this stream back into the decanter as diluent. This allowed any ethanol present in the product to have a second chance to be recycled back into the bottoms stream of the decanter (this stream is recycled back into the feed for the rectifying column).

### **Health and Safety Considerations**

There are multiple factors to take into consideration for the health and safety of the individuals involved in the process. Exposure and flammability limits of hazardous component of the process are shown in **Table 12** below.

**Table 12:** Exposure and Flammability Limits of Hazardous Components

<b>Process Component</b>	<b>TWA</b>	<b>STEL, ppm</b>	<b>Upper Explosion Limit</b>	<b>Lower Explosion Limit</b>	<b>Autoignition Temperature, °C</b>
Ethanol [1]	1000 ppm	1000 ppm	19.0%	3.3%	422.7
Isoamyl Alcohol [2]	100 ppm	125 ppm	9%	1.2%	350
DDGS [3] [4]	5 mg/m <sup>3</sup>	8 mg/m <sup>3</sup>	-	-	240
CO <sub>2</sub> [5]	5000 ppm	30000 ppm	-	-	-

Appropriate personal protective equipment should be worn when handling the above components, which can be found in the sources listed. The maximum time-weighted average value, (TWA), represents the maximum concentration of the substance in air for an 8-hour workday, and the maximum short-term exposure limit, (STEL), represents the maximum concentration of a substance in air for any 15-minute period [6].

All of the areas of the plant where alcohols are transported will be well ventilated so that the lower explosion limits will not be reached. A nitrogen blanket will be placed on top of the storage components in the vessels that contain the alcohols to ensure there is no oxidizing substance that will allow for combustion to take place. The storage tanks will be placed far enough apart from each other so if an explosion did occur, the other tanks will not catch fire as well.

The milling of corn allows introduces combustible dust into the process. There should not be any exposed ignition sources in the corn milling and drying areas, and the building design will be built up to electrical code that takes into consideration the combustible dust present. Proper cooling of process equipment and adequate ventilation that will minimize the confinement and dispersion of milling particulates. All of the milling by-products should be kept well under 240 °C to ensure no spontaneous combustion.

The fermentation portion of the process involves the processing of biological material. The fermentation will be ran at relatively low temperatures and moderate pH. The fermentation reaction evolves carbon dioxide, (CO<sub>2</sub>), which introduces the risk of the fermentation tanks to be over-pressurized. The tanks will have pressure gauges and a pressure relief system that is controlled by feedback control loops.

A process control scheme including thermocouples, pressure sensors, and flow meters will be implemented to ensure the process is maintained at operating conditions. The control system will lead into a control room with a backup computer in case of a lightning strike. All settings and start up procedures will be documents, and staff will be notified via email and phone to alert them of the computer restart procedure. The control scheme will minimize environmental contamination by maintaining it at optimum conditions by ensuring all recycle streams are operating efficiently so that the process is sustainable. All streams of the process will be used in other parts of the plant. What-if analyses about each equipment type is shown in Appendix D92-95.

### Aspen Simulation

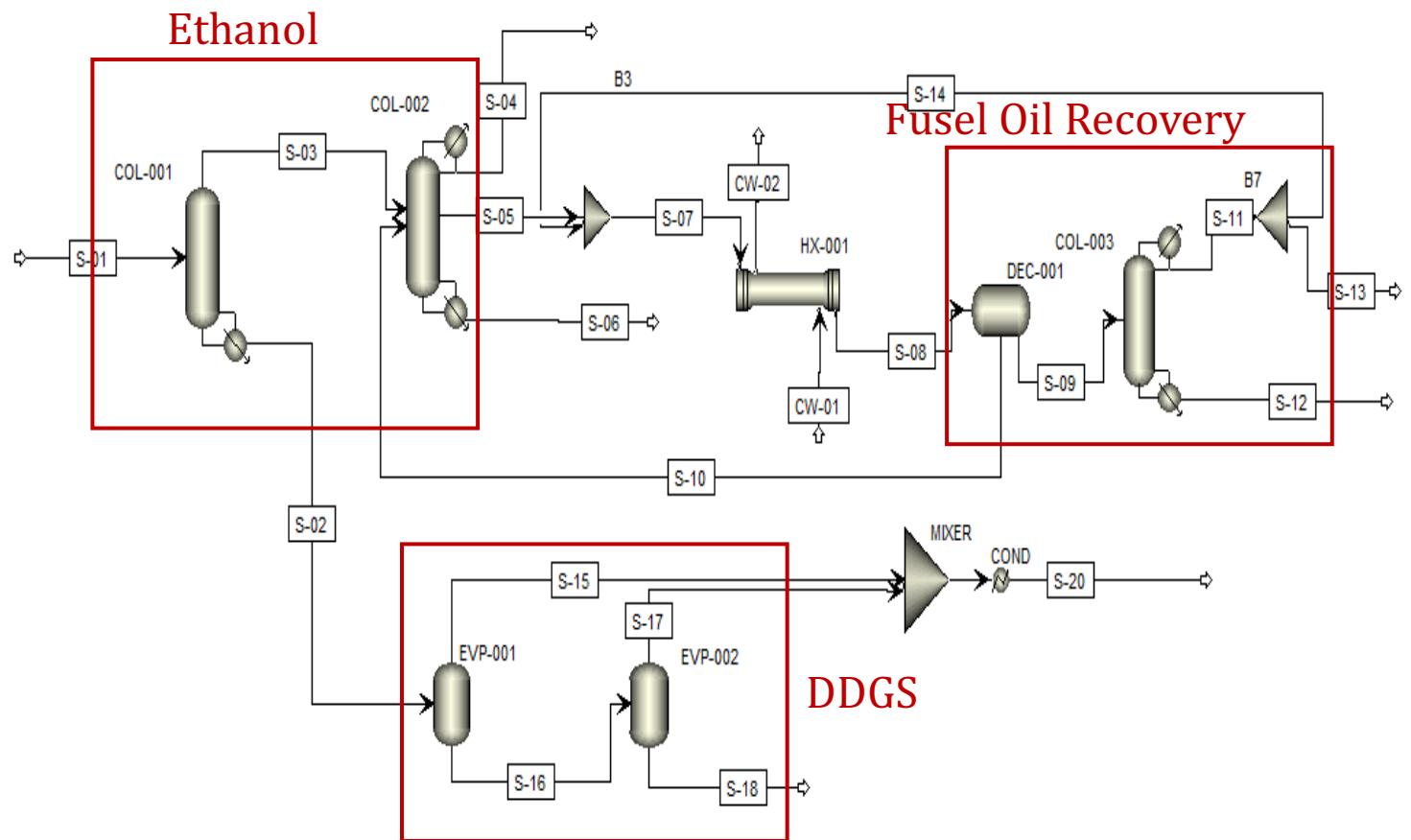
The overall bioethanol continuous process was simulated in *Aspen*. To do this, a thermodynamic model must be utilized. The team decided to model the process using the Universal Quasichemical model with a Redlich-Kwong equation of state correction (UNIQUAC-RK).

The UNIQUAC model is used for strongly non-ideal liquid solutions and liquid-liquid equilibria. This is seen in part in the stripping column, due to the large amount of solute dissolved in the liquid, a highly non-ideal solution exists. Also, the liquid-liquid equilibria is a strong analysis for the fusel recovery decanter. The decanter contains non-miscible oil and aqueous solutions that do not follow fluid ideality. This analytic approach was confirmed from the similarities between the simulated results and those found in the Guymon publication. This will be discussed in further detail during the discussion of the decanter later on.

The Redlich-Kwong equation of state acts as a non-ideal vapor correction. This correction is necessary for the stripping column. Due to the large differences in component mass flow rates, ideal vaporization cannot occur. Due to fusel oil's higher boiling point than water, if an ideal calculation was performed, all of the water would vaporize before any of the fusel oil. However, due to the dilute nature of the fusel oil in the column, it vaporized almost completely. Therefore, a non-ideal correction must be included.

The process simulation is pictured below in **Figure 7**. Recycle streams were used to minimize product waste and to minimize the utilities used. The distillate stream from the fusel recovery plant, stream "S-14", to dilute the incoming feed to the decanter and acts as

an initial coolant before reaching the heat exchanger to save utilities. The decanted stream, "S-10", feeds back to the packed column in the ethanol recovery plant to also act as an initial coolant and take away need for a condenser on the distillate of the stripping comumn,"COL-001". Lastly, the bottoms of the packed column, "S-06", being almost pure water was fed back to fermentation to reduce fresh water needed. These recycle streams are pictured for "S-14" and "S-10", but not in "S-06" due to its feedback to the non-continuous portion of the facility.



**Figure 7:** Overall Process Simulation

## Process Economics

Rigorous economic calculations were performed for the designed bioethanol plant for base case of ethanol production and the case examining the addition of fusel recovery equipment. In both cases, the plant is assumed to be built in an initial year zero with its economics studied over ten years of operation. A conservative stream factor of 0.9 was utilized to account for shutdowns and equipment maintenance which would require plant downtime. With the large fluctuations in both ethanol and raw material prices, cash flow payback period was not a feasible comparison metric for the cases analyzed. Rather, both the net profit in reference to the ethanol capacity being produced and the net present worth for the plant after ten years of operation were used. Ethanol is a satisfactory vantage point as it is the primary product of the plant and any additional revenue will be used to subsidize losses taken in its production.

### ***Capital Investment***

After pricing each respective piece of equipment (values presented below in **Table 13**), this overall major equipment cost was related to the capital investment for the plant. The factorial method of capital investment estimation was used to relate the total equipment cost to the total capital investment to construct the plant. This was done for both the case of the ethanol production plant (major equipment cost of \$6.3 MM) and after the addition of fusel recovery equipment (major equipment cost of \$6.5 MM).

**Table 13:** Major Equipment Cost for the Ethanol Production Facility and the Fusel Oil Recovery System

<b>Fusel Recovery Plant Add-on</b>	
<b>Major Equipment</b>	<b>Purchase Cost</b>
<i>Fusel Take-off stream heat exchanger</i>	\$ 17,116
<i>Fusel Oil Decanter</i>	\$ 16,491
<i>Packed Distillation Column</i>	\$ 63,705
<i>Fusel Oil Product Storage Tank</i>	\$ 108,306
<b>Total Major Equipment Cost:</b>	<b>\$ 210,000</b>
<b>Ethanol Production Plant</b>	
<b>Major Equipment</b>	<b>Purchase Cost</b>
<i>Fermenters</i>	\$ 3,223,506
<i>Harvest Tanks</i>	\$ 229,506
<i>Stripping Column</i>	\$ 581,071
<i>Double-effect Evaporator</i>	\$ 1,568,902
<i>Rectifying Column</i>	\$ 367,000
<i>Ethanol Product Storage Tanks</i>	\$ 339,373
<b>Total Major Equipment Cost:</b>	<b>\$ 6,300,000</b>

The factorial method of cost estimation multiplies the base equipment cost by a series of factors to account for things like installation, piping, electronics, etc [20]. Performing this for both the cases mentioned, the base-case ethanol production plant has a total capital investment of \$27 MM while the plant with the added fusel recovery system totals \$28 MM. A more detailed breakdown of each factor and their components can be referenced in Appendix C87-89. In both cases, the corresponding Lang factor for the plant is 4.27 which is typical for plants that operate with both solids and liquids processing.

### **Variable Manufacturing Costs**

The variable manufacturing costs considered for the plant were the costs of raw materials and the cost of utilities necessary for the plants operation. For the process utilities, the *Aspen Plus* simulation for the plant was used to track the magnitude of steam, cooling water, and electricity required in the various pieces of equipment. With initial designs for the plant, operating costs were seen to make a very large contribution to the plants manufacturing expenses and used as a point of interest for reducing the cost of operation. Relationships published for the cost of these utilities by Peters and Timmerhaus were used to obtain annual utilities costs (**Table 14**) [7].

**Table 14:** Summary of Plant Utilities Cost for Both Fusel Recovery and Ethanol Production

	<b>Fusel Recovery Plant Add-on</b>	<b>Ethanol Production Plant</b>	<b>Total Plant</b>
<b>Steam @ 100 psi (USD/Year):</b>	\$ 55,000	\$ 41,000,000	\$ 41,055,000
<b>Cooling Water (USD/Year):</b>	\$ 110,000	\$ 22,000,000	\$ 22,110,000
<b>Electricity (USD/Year):</b>	\$ 26,000	\$ 38,000	\$ 64,000
<b>Total Utilities Cost (USD/Year):</b>	<b>\$ 190,000</b>	<b>\$ 63,000,000</b>	<b>\$ 63,000,000</b>

The other component of variable costs being the cost of raw materials were determined based on the mass balance first calculated around the system. Three main components were considered raw materials in this case: corn, process water, and the yeast necessary to start the fermentation. Bulk market pricing for these materials was identified for the most recent date the information could be obtained. The raw material prices for water, corn, and yeast are as follows: \$3.56 per bushel of corn, \$0.76/ $10^3$  kg for water, and \$1.13 per kg of yeast [22] [7] [23]. Taking all these prices into account, the raw material

costs for the plant totaled at \$120 MM per year in both cases. The total variable manufacturing cost for the process was then used as the sum of the utilities cost and raw materials cost for the plant.

### ***Fixed Manufacturing Costs***

Fixed manufacturing costs were taken as labor and any miscellaneous costs not dependent on the variability of production. For labor, the amount of man-hours needed for equipment operation were taken as a function of plant capacity and number of processing steps in the plant as detailed by Peters and Timmerhaus [7]. Using this relationship, it was determined the base-case ethanol production plant would required 300 man-hours per day of operation and the plant with the addition of the fusel recovery steps would need 350 man-hours. With this base value decided, costs for supervisory, facilities cost, and supervisory expenses were determined. **Table 15** displays the fixed costs for the ethanol plant with and without the recovery system. A more detailed layout of the calculations for fixed cost are shown in **Table C5**.

**Table 15:** Summary of Fixed Costs for the Ethanol Production Plant w & w/o Recovery

*This table includes labor costs, depreciation, and miscellaneous costs for the plant.*

	Ethanol Production Plant	Total Plant with Fusel Recovery Equipment
	(USD/year)	(USD/year)
<b>Operators:</b>	\$ 4,400,000	\$ 5,100,000
<b>Supervisory:</b>	\$ 1,300,000	\$ 1,500,000
<b>Benefits:</b>	\$ 230,000	\$ 270,000
<b>Plant overhead:</b>	\$ 470,000	\$ 550,000
<b>Maintenance Materials:</b>	\$ 60,000	\$ 62,000
<b>Operating Materials:</b>	\$ 440,000	\$ 511,000
<b>Local Taxes and Insurance:</b>	\$ 54,000	\$ 56,000
<b>Total Fixed Cost:</b>	\$ 6,900,000	\$ 8,100,000
<b>Depreciation (10 year fixed straight line):</b>	\$ 2,700,000	\$ 2,800,000

### Cash Flow

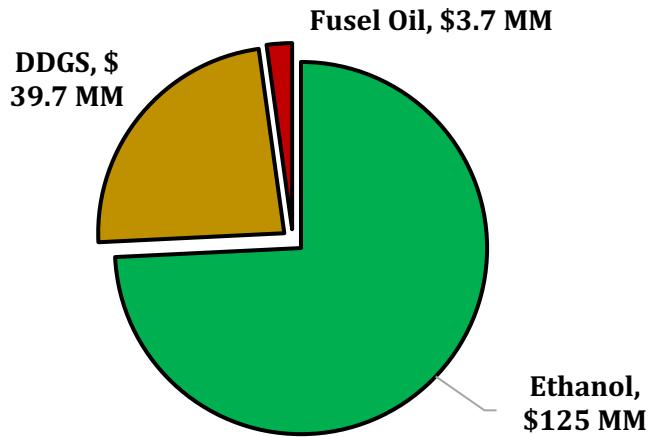
With the annual manufacturing costs figured out, a cash flow over a ten-year operating period was mapped out for both the case of the base ethanol plant and the plant with the added fusel recover equipment. This analysis was performed assuming all equipment costs were depreciable capital with straight-line depreciation for the full ten-year period. Current selling prices for raw materials and fusel oil were used when estimating the revenue for the plant. The market prices found for ethanol and fusel oil were \$1.64 per gallon and \$10.28 per gallon, respectively. The revenue for the plant in both cases of operation can be referenced in **Table 16** along with key results obtained from looking into the plant's cash flow. The addition of the fusel recovery system was found to improve the economics of the ethanol production plant with an increased net profit and increased NPW after ten years of operation. The plant is still not profitable under these circumstances, but the additional

revenue generated by recovery and sales of fusel oil helps subsidize the loss taken during the production of ethanol. The full cash flow analysis for the ten speculative years of operation can be referenced in Appendix C90-91.

**Table 16:** Fusel Recovery Economic Comparison

	<b>Ethanol Production Plant</b>	<b>Plant with Fusel Recovery Equipment</b>
<b>Revenue (USD/year):</b>	\$160 MM	\$170 MM
<b>Manufacturing Cost (USD/year):</b>	\$190 MM	\$190 MM
<b>Net Profit (USD/gal ethanol):</b>	\$-0.34	\$-0.31
<b>NPW (10 years):</b>	\$-290 MM	\$-270 MM

Net present worth (NPW) was chosen as a comparison metric for both cases. With the current operating expenses for the plant, net profit was often negative, making comparison values such as cash flow payback period inadequate for analysis. This negative net profit for the plant is heavily related to both the purchase price of corn (\$3.56 per bushel) and the selling price of ethanol (\$1.64 per gallon). Both cost components are the largest contributor to their respective categories. The relative sizes of each component of the plant's revenue can be compared by viewing **Figure 9**.

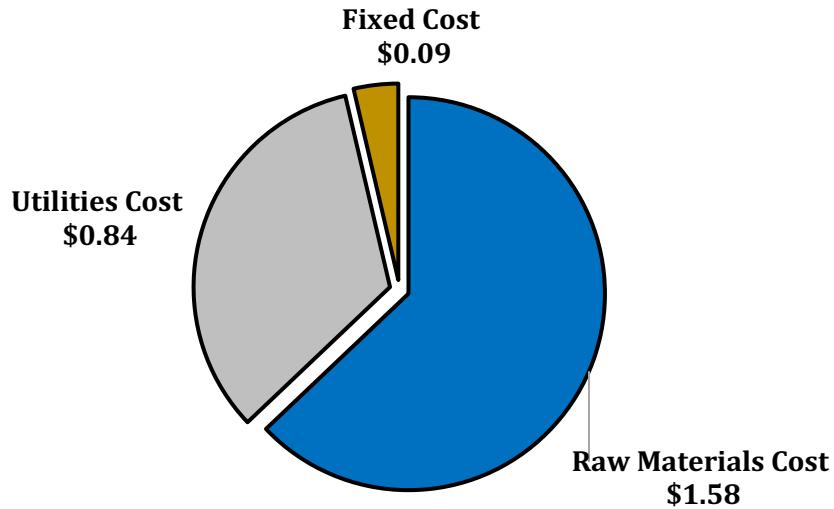


**Figure 9:** Relative Magnitudes for the Three Main Sources of Revenue for the Ethanol Production Plant

*This figure is generated only for the case that the plant is recovering fusel oil.*

Fusel oil is a small but significant contributor to the plant's revenue and well worth the increased operating costs it brings with it. The addition of the distillation column, heat exchanger, and gravity decanter bring with them an increased manufacturing cost which is negligible (within two significant figures). When the capital investment of the fusel oil recovery equipment is analyzed with respect to the amount of value it is generating for the plant, it pays itself off after three months of operation.

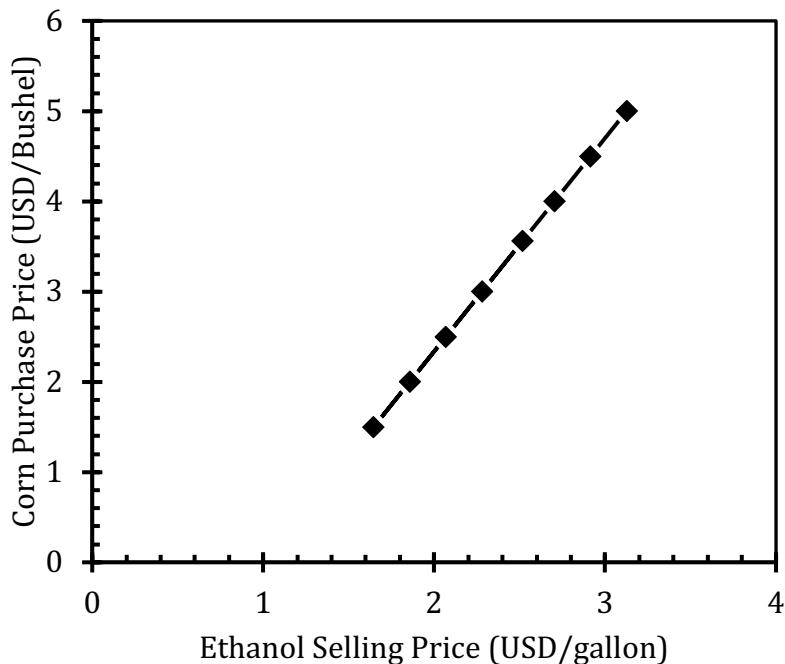
The selling price of ethanol—the primary product of this facility—and the purchase cost of corn were found to be integrally linked in determining the profitability of the facility. The relative magnitudes of each component is shown in **Figure 10**. Raw material costs are the largest component of manufacturing costs and are primarily comprised of corn.



**Figure 10:** Relative Magnitudes of the Different Costs for Manufacturing One Gallon of Ethanol Product

*The USD values are given relative to the gallons of ethanol produced (Ex. Raw material costs are \$1.58/gallon of ethanol produced).*

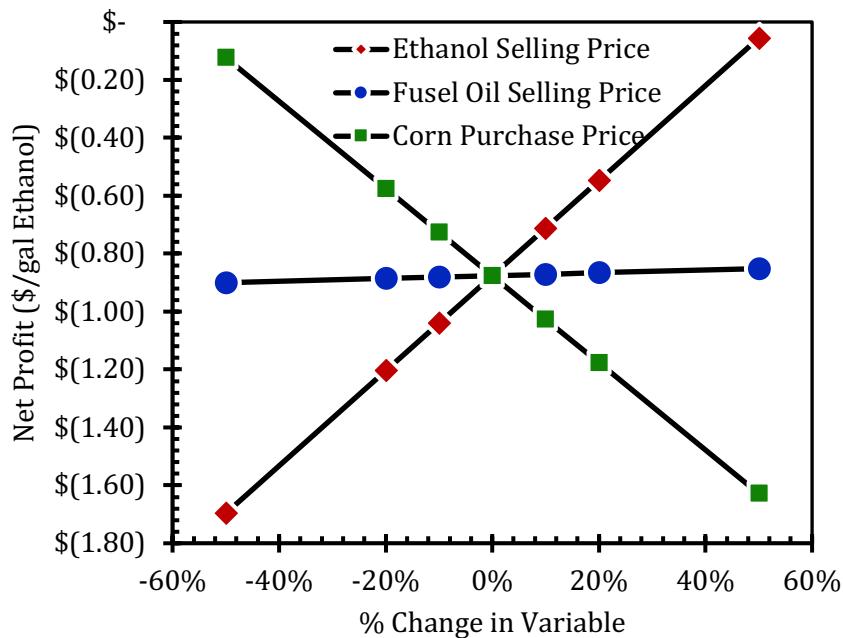
Both corn and ethanol prices are noted for their fluctuations in the past years, making bioethanol production viable some years and a poor investment for others. For the plant which was designed, the cost of purchasing corn was iterated and compared against the selling price for ethanol required to reach the break-even point for production. The resulting relationship between these variables is presented in **Figure 11**. In order to keep the plant's operation profitable, the selling price must be adjusted to accommodate for fluctuations in the selling price of corn. Choosing a selling price anywhere to the right of the line relating the two variables will ensure the plant is able to at least cover its manufacturing costs for the year and all additional revenue will be considered net profit.



**Figure 11:** Break-even Production Chart Relating the Selling Price of Ethanol and the Purchase Cost of Corn

### *Sensitivity Analysis*

With the volatility of the selling prices of its products (market price fluctuations) and the purchase price of corn, a sensitivity analysis on the plant's net profit with respect to the volume of ethanol produced was performed. The variables tested, were the purchase price of corn, market price of fusel oil, and the market price of ethanol. The results can be seen below in **Figure 12**. Each variable was investigated out to plus or minus 50% of its original value.



**Figure 12:** Sensitivity Analysis

With the volatility of corn prices, and the very small profit margin on ethanol production, it is not recommended to construct a brand-new ethanol plant at this time. There are three main ways to increase the profitability of the plant if it is already constructed. First, the addition of the fusel recovery equipment was shown to be a good investment with the fusel oil product useful to subsidize the losses taken during ethanol manufacturing. With this consideration, any capacity that can be diverted from ethanol production to fusel oil production should be heavily considered. This was achieved through utilizing different strains of yeast or possibly putting different environmental stressors to the current fermentation to boost the selectivity of the outcome for fusel oils instead of ethanol. Another consideration is reducing raw material costs. The SSF process brings with it the flexibility of being able to break down starches into sugars for the fermentation. With this, alternative sugar sources should be considered either to replace corn as the primary sugar source or as

an additive. It is possible there are other plants in the area whose waste streams could be useful as a starting material.

A final consideration could be increasing the plants operating efficiency and decreasing the contribution of utilities usage to the operating costs of the plant. As shown in **Figure 10** that the plants utilities usage are a large component of the total manufacturing cost. These utilities usages are primarily in the operation of the stripping column (steam usage) and the operation of the rectifying column (steam and cooling water usage). Finding a way to decrease these significantly through process optimization and high efficiency equipment would positively affect the economics of the plant and increase the very narrow margin for profitable operation.

### Conclusion and Recommendations

A proposed design for a bioethanol plant was created and economically analyzed to determine its feasibility and profitability. The plant consists of an ethanol recovery system, a DDGS production section, and a possible addition of a fusel recovery system which was further analyzed to determine its profitability and value for the plant. The economic analysis showed that the construction and operation of a new plant would not be profitable, however, the addition of a fusel oil recovery system can help offset the losses of the plant. The addition of a fusel oil recovery system in an already existing plant is a good investment. There is a net loss of \$0.34 for every gallon of ethanol produced. The fusel oil recovery system lowers the net loss to \$0.31 per gallon. The total equipment cost for the bioethanol plant with and without fusel oil recovery are \$6,500,000 and \$6,300,000 respectively. The plant revenue increases from \$160 million dollars to \$170 million dollars with the addition of the fusel oil recovery system. The net present worth of the bioethanol plant after 10 years of operation

is \$-290 million dollars. The fusel oil recovery part of the plant increases the net present worth to about \$-270 million dollars. For both cases, the net present worth is still negative and the proposed plant is not profitable. The factorial method outlined in Perry's Handbook was used to calculate the total cost of the plant taking into account factors such as installation, piping, electronics, etc [20]. The Lang factor for the plant is 4.27, which is typical for solid and liquid processing plants.

Improvements to the proposed design can be made by looking into obtaining an alternative feed stream for the plant that could lower the cost for raw materials. The cost of raw materials accounts for the majority of the manufacturing costs of the proposed plant. An alternative microorganism with a higher productivity such as *Zymomonas mobilis* could replace the *S. cerevisiae* that was used in the proposed design. The DDGS evaporator system could be replaced by a double-effect evaporator to lower the plant's utility costs. Additionally, an alternative bioreactor such as a continuous stirred tank reactor (CSTR) could be used instead of multiple batch fermenters because CSTRs have a higher productivity and lower equipment costs but impellers in the CSTR may generate shear stress and may require additional separation equipment.

## Assumptions

Often, in designing processes such as a large-scale ethanol production plant, several assumptions must be made to make calculations easier that do not significantly impact the results.

Below lists the assumptions made for each group of processing equipment.

### *Fermenters*

- Conversion of 90 mole % glucose to alcohol
- 5 gallons of fusel oil produced for every 1000 gallons of ethanol produced
- Maximum survivable alcohol concentration of 16 vol % for yeast strain
- Negligible mass contribution from yeast and enzyme
- Corn is 60 wt. % starch
- Corn is 10 wt. % water
- Productivity of  $3 \text{ g L}^{-1} \text{ hr}^{-1}$  for yeast strain in batch reactor
- 5 hours of cleaning time required for each bank of fermenters
- Fermenters and harvest tanks are small field-erected storage vessels

### *Stripping Column*

- Bottoms must come out at 70 wt-% to 30 wt-% solid
- Solids all go to bottom
- Solids have same thermodynamic properties as glucose
- Pressure drop is negligible in column
- Solids dissolved in fluid are dilute enough to have negligible change in viscosity

### *Rectifying Column*

- Pressure is constant throughout the column
- Fusel oil is only made up of isoamyl alcohol
- 50 wt% isoamyl alcohol and water have a heterogeneous azeotrope at 95.1 °C
- The fraction of flooding is 0.80
- Each stage has a Murphree vapor efficiency of 0.70

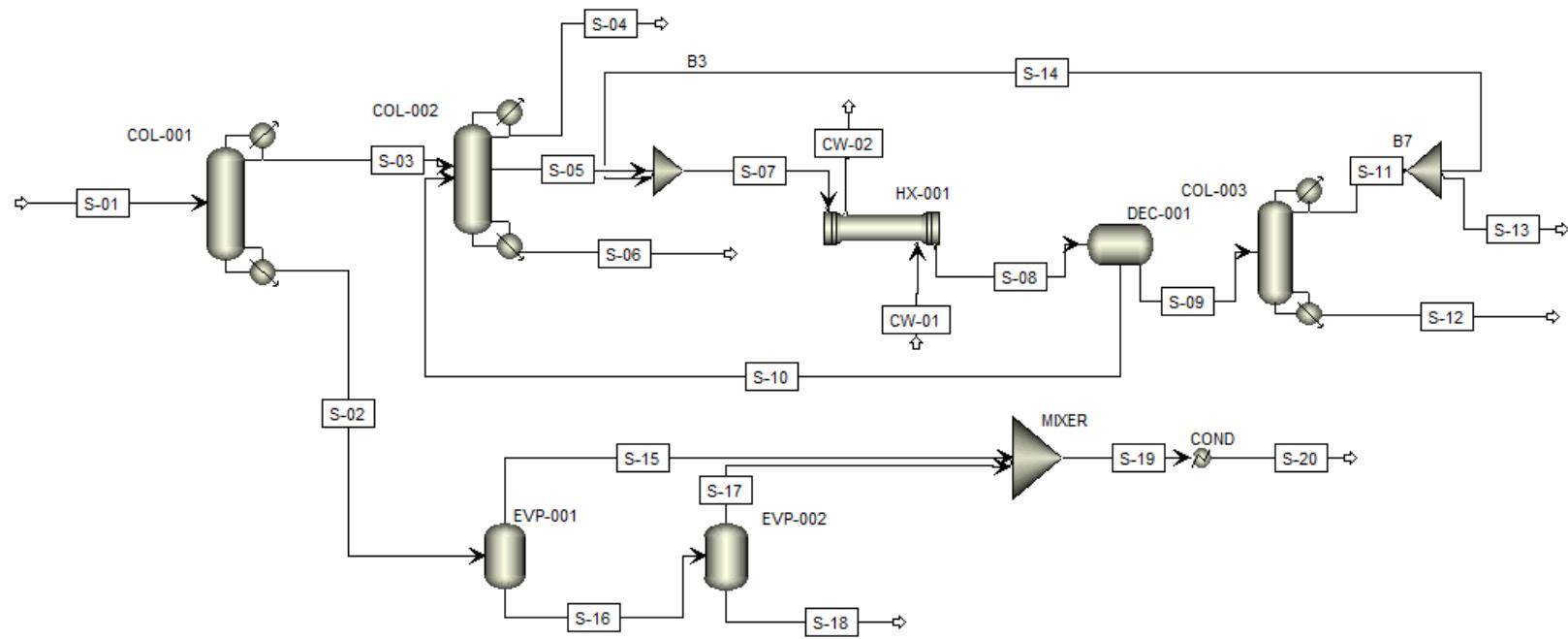
### *DDGS Production*

- DDGS moisture content of 15%
- no heat is lost to the environment
- No solids in the vapor stream
- Inlet steam is fully condensed to liquid water
- Stream used to evaporate liquid is saturated steam at 150 °C
- Solids have the same thermodynamic properties as glucose

**Appendix A:** Process Flow Diagram and Stream Table

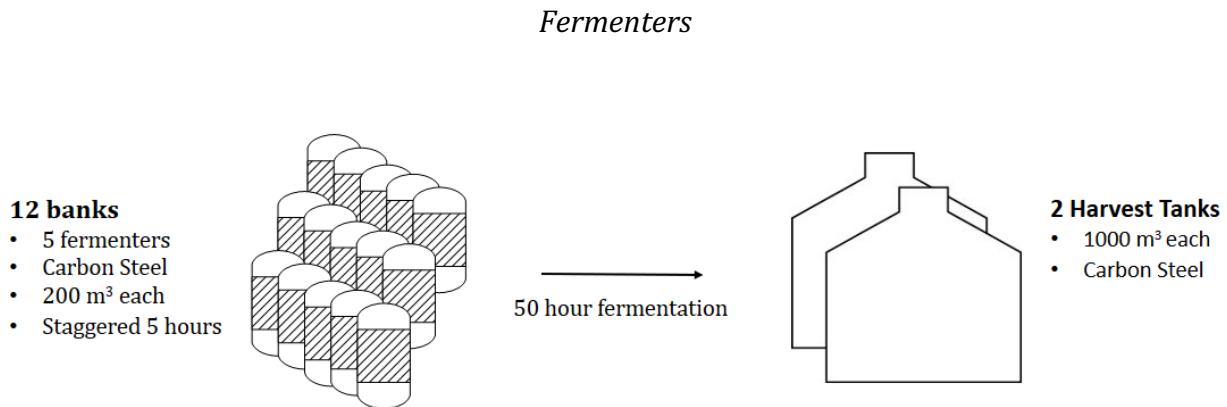
**Table A1:** Process Stream Specifications

Property	Units	Stream											
		S-01	S-02	S-03	S-04	S-05	S-06	S-07	S-08	S-09	S-10	S-11	
		Liquid Phase	Liquid Phase	Liquid Phase	Liquid Phase	Liquid Phase	Liquid Phase	Liquid Phase	Liquid Phase	Liquid Phase	Liquid Phase	Liquid Phase	
		C	78	120	100	78	86	98	85	5	5	5	95
		bar	1.7	1.7	1.7	1	1	1	1	1.7	1.7	1.7	1.7
		cal/gm	-3200	-3100	-3200	-1600	-3000	-3700	-2900	-3000	-1500	-3200	-1900
		kg/hr	230000	110000	110000	27000	1700	85000	1800	1800	230	1600	110
		<u>Mass Fractions</u>											
		Water	0.74	0.71	0.77	0.09	0.69	0.99	0.67	0.67	0.12	0.74	0.26
		Ethanol	0.11	0	0.23	0.91	0.23	0.01	0.25	0.25	0.33	0.24	0.66
		Isoamyl Alkohol	0	0	0	0	0.08	0	0.08	0.08	0.55	0.01	0.08
		Dextrose	0.03	0.05	0	0	0	0	0	0	0	0	0
		Solids	0.12	0.24	0	0	0	0	0	0	0	0	0
	Units	S-12	S-13	S-14	S-15	S-16	S-17	S-18	S-19	S-20	CW-01	CW-02	
	Units	Liquid Phase	Liquid Phase	Liquid Phase	Vapor Phase	Liquid Phase	Vapor Phase	Liquid Phase	Liquid Phase	Liquid Phase	-20	64	
		C	140	95	95	100	100	110	110	100	100	1	1
		bar	1.7	1.7	1.7	1	1	1	1	1	1	1	1
		cal/gm	-910	-1900	-1900	-3200	-2700	-3200	-2000	-3200	-3700	-2300	-2200
		kg/hr	120	5.6	110	43000	71000	31000	40000	74000	74000	2600	2600
		<u>Mass Fractions</u>											
		Water	0.01	0.26	0.26	1	0.53	1	0.16	1	1	0.22	0.22
		Ethanol	0.01	0.66	0.66	0	0	0	0	0	0	0	0
		Isoamyl Alkohol	0.98	0.08	0.08	0	0	0	0	0	0	0	0
		Dextrose	0	0	0	0	0.08	0	0.14	0	0	0	0
		Solids	0	0	0	0	0.39	0	0.7	0	0	0	0
		Ethylene Glycol	0	0	0	0	0	0	0	0	0.78	0.78	0.78



**Figure A1:** Process Flow Diagram

## **Appendix B:** Equipment Design & Hand Calculations



**Figure B1:** Process Schematic of the Fermentation Step

**Table B1:** Design Parameters for the Fermenters

<b><i>Fermenter Volume:</i></b>	200 m <sup>3</sup>
<b><i>No. of Vessels</i></b>	60
<b><i>Material of Construction:</i></b>	Carbon Steel
<b><i>Purchase Cost:</i></b>	\$ 3,223,506

**Table B2:** Design Parameters for the Harvest Tanks

<b><i>Harvest Tank Volume:</i></b>	1000 m <sup>3</sup>
<b><i>No. of Vessels</i></b>	2
<b><i>Material of Construction:</i></b>	Carbon Steel
<b><i>Purchase Cost:</i></b>	\$ 229,506

## Fermentation Hand Calculations

$$75,000,000 \frac{\text{gal EtOH}}{\text{year}} \cdot \frac{0.00379 \text{m}^3}{1 \text{gal}} \cdot \frac{789 \text{kg}}{\text{m}^3} \cdot \frac{1 \text{kcal}}{460 \text{kg}} \cdot \frac{1 \text{year}}{365 \text{days}} = 13321.1 \frac{\text{kcal EtOH}}{\text{day}}$$

Starting with  $\frac{5 \text{ gal Fuel oil}}{1000 \text{ gal EtOH}}$

$$V_{\text{Fuel}} = \frac{5 \text{ gal Fuel}}{1000 \text{ gal EtOH}} \cdot 75,000,000 \text{ gal EtOH} = 375,000 \text{ gal Fuel oil}$$

$$375,000 \frac{\text{gal Fuel}}{\text{year}} \cdot \frac{0.00379 \text{m}^3}{1 \text{gal}} \cdot \frac{810 \text{kg}}{\text{m}^3} \cdot \frac{1 \text{kcal}}{88.1 \text{kg}} \cdot \frac{1 \text{year}}{365 \text{days}} = 35.8 \frac{\text{kcal fuel}}{\text{day}}$$

• However, assumption that 1 mol of fuel for every mol of ethanol sacrificed, therefore need total of  $13321.1 \frac{\text{kcal}}{\text{day}}$  alcohol

1st iteration

$$\text{Total mol/s} = 13321.1 \frac{\text{kcal EtOH}}{\text{day}} + 35.8 \frac{\text{kcal fuel}}{\text{day}} = 13356.9 \frac{\text{kcal}}{\text{day alc.}}$$

• used goal seek in excel to change  $75,000,000 \frac{\text{gal EtOH}}{\text{year}}$   
Jntial total mol/s =  $13321.1 \frac{\text{kcal}}{\text{day}}$

Result

$$\text{EtOH} = 13285.4 \frac{\text{kcal}}{\text{day}} \cdot \frac{46.07 \text{kg}}{1 \text{kcal}} \cdot \frac{1 \text{day}}{24 \text{hrs}} \cdot \frac{1 \text{hr}}{3600 \text{s}} = \boxed{7.1 \frac{\text{kg EtOH}}{\text{s}}}$$

$$\text{Fuel} = 35.7 \frac{\text{kcal}}{\text{day}} \cdot \frac{88.15 \text{kg}}{1 \text{kcal}} \cdot \frac{1 \text{day}}{24 \text{hrs}} \cdot \frac{1 \text{hr}}{3600 \text{s}} = \boxed{0.0363 \frac{\text{kg Fuel}}{\text{s}}}$$

$$\text{Glucose} = \frac{\frac{1}{2} \text{ mol/c}}{0.9} = \frac{\frac{1}{2} \cdot 13321.1 \frac{\text{kcal}}{\text{day}}}{0.9} = 7400 \frac{\text{kcal Glucose}}{\text{day}}$$

$$\text{M glucose} = 7400 \frac{\text{kcal}}{\text{day}} \cdot \frac{180.16 \text{kg}}{1 \text{kcal}} = 1,333,294 \frac{\text{kg}}{\text{day}}$$

$$P_{\text{corn}} = \frac{1,333,294 \frac{\text{kg glucose}}{\text{day}}}{0.6} = 2,222,157 \frac{\text{kg}}{\text{day}} \cdot \frac{1 \text{day}}{24 \text{hrs}} \cdot \frac{1 \text{hr}}{3600 \text{s}} = \boxed{25.7 \frac{\text{kg}}{\text{s}}}$$

\* Assumption corn is 60 wt% glucose \*

$$\text{Corn} = 25.7 \frac{\text{kg}}{\text{s}}$$

$$\begin{aligned}\text{Glucose} &= 0.6 \cdot 25.7 \frac{\text{kg}}{\text{s}} = 15.4 \frac{\text{kg Glc}}{\text{s}} \\ * \text{Water} &= 0.1 \cdot 25.7 \frac{\text{kg}}{\text{s}} = 2.6 \frac{\text{kg H}_2\text{O}}{\text{s}} \\ \text{Solids} &= 25.7 - 15.4 - 2.6 = 7.7 \frac{\text{kg Solids}}{\text{s}}\end{aligned}$$

Volume of alcohol

$$V_{\text{Ethanol}} = \frac{7.1 \frac{\text{kg}}{\text{s}}}{789 \frac{\text{kg}}{\text{m}^3}} = 0.00898 \frac{\text{m}^3}{\text{s}}$$

$$V_{\text{Fuel}} = \frac{0.026 \frac{\text{kg}}{\text{s}}}{810 \frac{\text{kg}}{\text{m}^3}} = 4.49 \times 10^{-5} \frac{\text{m}^3}{\text{s}}$$

$$V_{\text{Alcohol}} = V_{\text{Ethanol}} + V_{\text{Fuel}} = 0.00898 + 4.49 \times 10^{-5} = 0.0090 \frac{\text{m}^3}{\text{s}}$$

$$V_{\text{Total Flow}} = \frac{V_{\text{Alcohol}}}{0.16} = \frac{0.009 \frac{\text{m}^3}{\text{s}}}{0.16} = 0.0563 \frac{\text{m}^3}{\text{s}}$$

$$V_{\text{Solids}} = \frac{7.7 \frac{\text{kg}}{\text{s}}}{500 \frac{\text{kg}}{\text{m}^3}} = 0.015 \frac{\text{m}^3}{\text{s}}$$

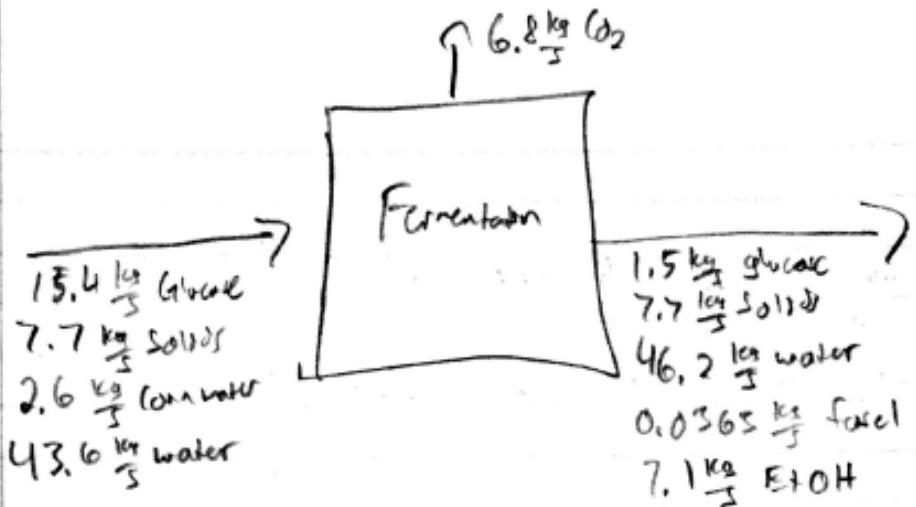
$$V_{\text{GV,out}} = \frac{0.1 \cdot 15.4 \frac{\text{kg}}{\text{s}}}{1500 \frac{\text{kg}}{\text{m}^3}} = 0.00103 \frac{\text{m}^3}{\text{s}}$$

$$\begin{aligned}V_{\text{water}} &= V_{\text{total}} - V_{\text{alcohol}} - V_{\text{solids}} - V_{\text{GV,out}} \\ &= 0.0563 \frac{\text{m}^3}{\text{s}} - 0.009 \frac{\text{m}^3}{\text{s}} - 0.015 \frac{\text{m}^3}{\text{s}} - 0.00103 \frac{\text{m}^3}{\text{s}} = 0.046 \frac{\text{m}^3}{\text{s}}\end{aligned}$$

$$\begin{aligned}M_{\text{water,middle}} &= V_{\text{water}} \cdot \rho_{\text{water}} - M_{\text{corn water}} \\ &= 0.046 \frac{\text{m}^3}{\text{s}} \cdot 1000 \frac{\text{kg}}{\text{m}^3} - 2.6 \frac{\text{kg}}{\text{s}} = 43.6 \frac{\text{kg water}}{\text{s}}\end{aligned}$$

$$M_{\text{CO}_2} = 13321.1 \frac{\text{kmol}}{\text{day}} \cdot \frac{44.01 \text{ kg}}{\text{kmol}} \frac{1 \text{ day}}{24 \text{ hrs}} \frac{1 \text{ hr}}{3600 \text{ s}} = 6.8 \frac{\text{kg}}{\text{s}}$$

\* Corn assumed to be 10 wt% water



### Sizing of Fermenters

$$C_{\text{Alcohol}} = \frac{M_{\text{Alcohol}}}{V_{\text{water}}} = \frac{7.1 \frac{\text{kg}}{\text{hr}} + 0.0365 \frac{\text{kg}}{\text{hr}}}{0.046 \frac{\text{m}^3}{\text{hr}}} = \frac{153 \frac{\text{kg}}{\text{hr}}}{\frac{\text{m}^3}{\text{hr}}} = 153 \frac{\text{g}}{\text{L}}$$

$$V_{\text{Total}} = \frac{V_{\text{alc}}}{0.16} = \frac{0.00125}{0.16} = 0.0078125 \frac{\text{m}^3}{\text{hr}} \cdot \frac{3600}{1\text{hr}} = 200 \frac{\text{m}^3}{\text{hr}}$$

$$\text{Fermentation Time} = \frac{C_{\text{AC}}}{\text{Productivity}} = \frac{153 \frac{\text{g}}{\text{L}}}{3 \frac{\text{g}}{\text{L/hr}}} = 50 \text{ hrs}$$

Assuming Storage Tank of 1000 m<sup>3</sup>

$$\text{Process Time} = \frac{1000 \frac{\text{m}^3}{\text{hr}}}{200 \frac{\text{m}^3/\text{hr}}{\text{hr}}} = 5 \text{ hrs}$$

$$\text{Total Time} = \text{Ferment} + \text{Processing} + \text{Cleaning} = 50 + 5 + 5 = 60 \text{ hrs}$$

$$\# \text{ of banks} = \frac{\text{Total Time}}{\text{Processing}} = \frac{60 \text{ hrs}}{5 \text{ hrs}} = 12 \text{ banks}$$

\*Assuming 200 m<sup>3</sup> fermenters \*

$$\# \text{ of fermenters in each bank} = \frac{\text{Tank Size}}{\text{Ferment size}} = \frac{1000 \frac{\text{m}^3}{\text{bank}}}{200 \frac{\text{m}^3}{\text{ferm}}} = 5 \text{ each}$$

$$\# \text{ of total Fermenters} = 5 \cdot 12 = 60 \text{ fermenters}$$

## Costing

\* Assumed to be small field erected storage tanks

$$200 \text{ m}^3 \text{ tank} = \$37570.00$$

$$1000 \text{ m}^3 \text{ tank} = \$80,246.90$$

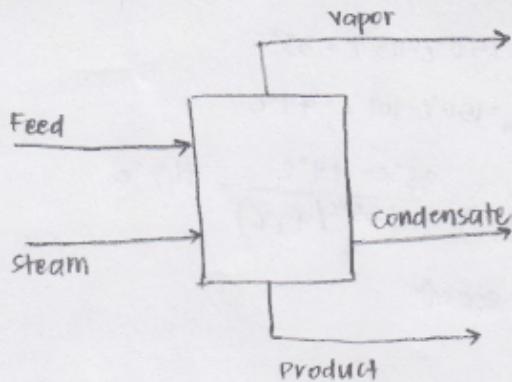
$$\begin{aligned}\text{Cost of Fermentors} &= \# \text{ Fermentors} \cdot \frac{\text{Cost}}{\text{Fermentor}} \cdot \text{Inflation} \\ &= 60 \cdot \$37,570.00 \cdot 1.43 \\ &\boxed{= \$13,223,506.00}\end{aligned}$$

$$\begin{aligned}\text{Cost of Storage Tanks} &= \# \text{ of tanks} \cdot \frac{\text{Cost}}{\text{Tank}} \cdot \text{inflation} \\ &= 2 \cdot \$80,246.90 \cdot 1.43\end{aligned}$$

$$\boxed{= \$229,506.00}$$

## Evaporators

Hand calculations: Waste Recovery & Before ASPEN optimization\*



Assume saturated steam  
at 150°C and fully  
condenses.

### Solid Balance:

$$x_{s,F} m_F = x_{s,P} m_P$$

$$x_{s,F} = 0.29$$

$$m_F = 81.16 \text{ kg/s}$$

$$x_{s,P} = 0.65 \leftarrow \text{chosen/assumed}$$

$$m_P = \frac{x_{s,F} m_F}{x_{s,P}} = \frac{0.29 (81.16 \text{ kg/s})}{0.65} = 36.46 \text{ kg/s}$$

### Liquid Balance:

$$x_L m_F = x_L m_P + x_V m_V \leftarrow \text{Assume vapor contains no solid}$$

$$x_L = 1$$

### Liquid in product stream

$$m_P = 36.46 \text{ kg/s}$$

$$m_{s,P} = 0.65 (36.46 \text{ kg/s}) = 23.70 \text{ kg/s}$$

$$m_{L,P} = m_P - m_{s,P} = 36.46 \text{ kg/s} - 23.70 \text{ kg/s} = 12.76 \text{ kg/s}$$

$$x_L m_F = x_L m_P + x_V m_V$$

$$(1 - 0.29) (81.16 \text{ kg/s}) = 12.76 \text{ kg/s} + m_V$$

$$m_V = 57.46 \text{ kg/s} - 12.76 \text{ kg/s} = 44.70 \text{ kg/s}$$

### Energy Balance: Assume all heat is transferred and no losses to the environment.

- Heat transferred is equal to  $\Delta H$  of vaporization of steam.

$$q = m_s \lambda \quad \text{where } \lambda \text{ is the specific enthalpy of vaporization at } 150^\circ\text{C}$$

$$q = 22500 \text{ kW} \leftarrow \text{modified in ASPEN to obtain desired heat duty}$$

$$\lambda = 2114.1 \text{ kJ/kg} \quad m_s = 10.6 \text{ kg/s steam} \rightarrow 22500 \text{ kW}$$

$$q = UA(LMTD) \rightarrow A = \frac{q}{U(LMTD)}$$

$$A = \frac{680 \text{ W}}{\text{m}^2 \text{K}} \leftarrow \text{obtained from online source for stainless steel.}$$

LMTD

$$T_{H,in} = 150^\circ C$$

$$T_{H,out} = 150^\circ C$$

$$T_{C,in} = 115^\circ C$$

$$T_{C,out} = 101^\circ C$$

co-current flow

$$\Delta T_1 = T_{H,in} - T_{C,in} = 150^\circ C - 115^\circ C = 35^\circ C$$

$$\Delta T_2 = T_{H,out} - T_{C,out} = 150^\circ C - 101^\circ C = 49^\circ C$$

$$LMTD = \frac{\Delta T_1 - \Delta T_2}{\ln(\Delta T_1 / \Delta T_2)} = \frac{35^\circ C - 49^\circ C}{\ln(35^\circ C / 49^\circ C)} = 41.7^\circ C$$

$$A = \frac{q}{U(LMTD)} = \frac{22500 \text{ kW}}{\frac{0.68 \text{ kW}}{\text{m}^2 \text{ K}} (41.7^\circ \text{ K})} \approx 800 \text{ m}^2$$

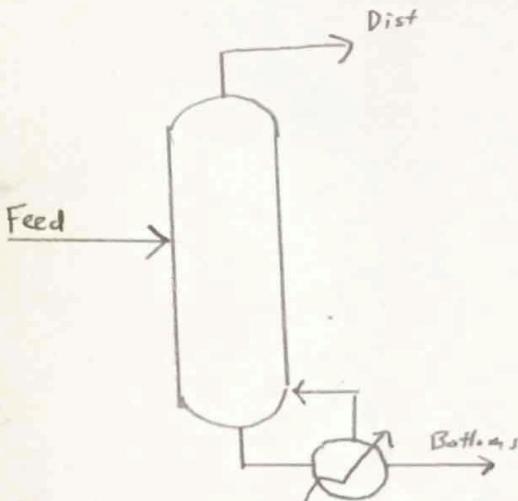
cost using Timmerhaus 14-33  $\rightarrow \$607,005$

$\rightarrow$  same balance on Evap 2

## Stripping Column

### Stripping Column

Component	Feed (kg/s)	Distillate (kg/s)	Bottoms (kg/s)
Water	46.210	24.603	21.607
Ethanol	7.100	7.100	0
Fusel	0.036	0.036	0
Glucose	1.540	0	1.540
Solids	7.720	0	7.720
Total	62.606	30.87	31.74
Temp (°C)	70	100	116



- Assume all Solids + glucose go to bottoms.
- Assume all Fusel + ethanol go to tops.

$$\text{wt-ratio of Solids/Liquids in bottoms} = 3/7$$

Bottoms

$$\frac{\text{Solids}}{\text{Liquids}} = \frac{3}{7} = \frac{9.26}{\text{Water}} \Rightarrow \text{Water in bottoms} = 21.607 \text{ kg/s}$$

$$\text{Dist} = \text{Feed} - \text{Bottoms} = 46.210 - 21.607 = 24.603 \text{ kg/s}$$

Fenske

$LK_B = \text{water}$

obtained from aspect

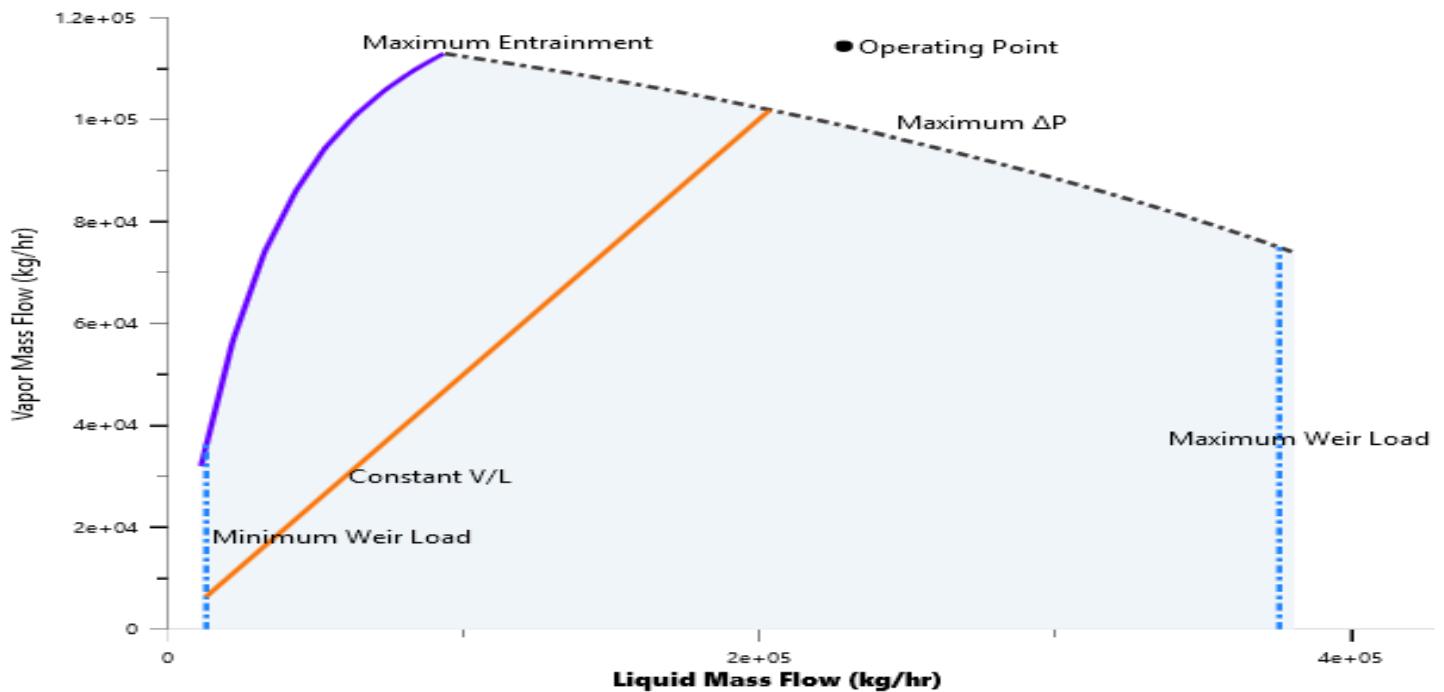
assume solids act same as glucose  
assume ethanol is LK

Component	Feed( $i, HK$ ) $\alpha$	Dist( $i, HK$ ) $\alpha$	Bottoms( $i, HK$ ) $\alpha$	$\alpha(i, r)_{H_2O}$	
Water	1	1	1	1	
Ethanol	7.183	6.749	2.971	4.478	$\leftarrow \alpha(LK, HK)_{H_2O}$
Fusel oil	12.748	12.152	1.717	4.567	
Glucose	$2.7 \times 10^{-9}$	$6.4 \times 10^{-8}$	$9.2 \times 10^{-8}$	$6 \times 10^{-8}$	
Solids	$2.7 \times 10^{-9}$	$6.4 \times 10^{-8}$	$9.2 \times 10^{-8}$	$6 \times 10^{-8}$	

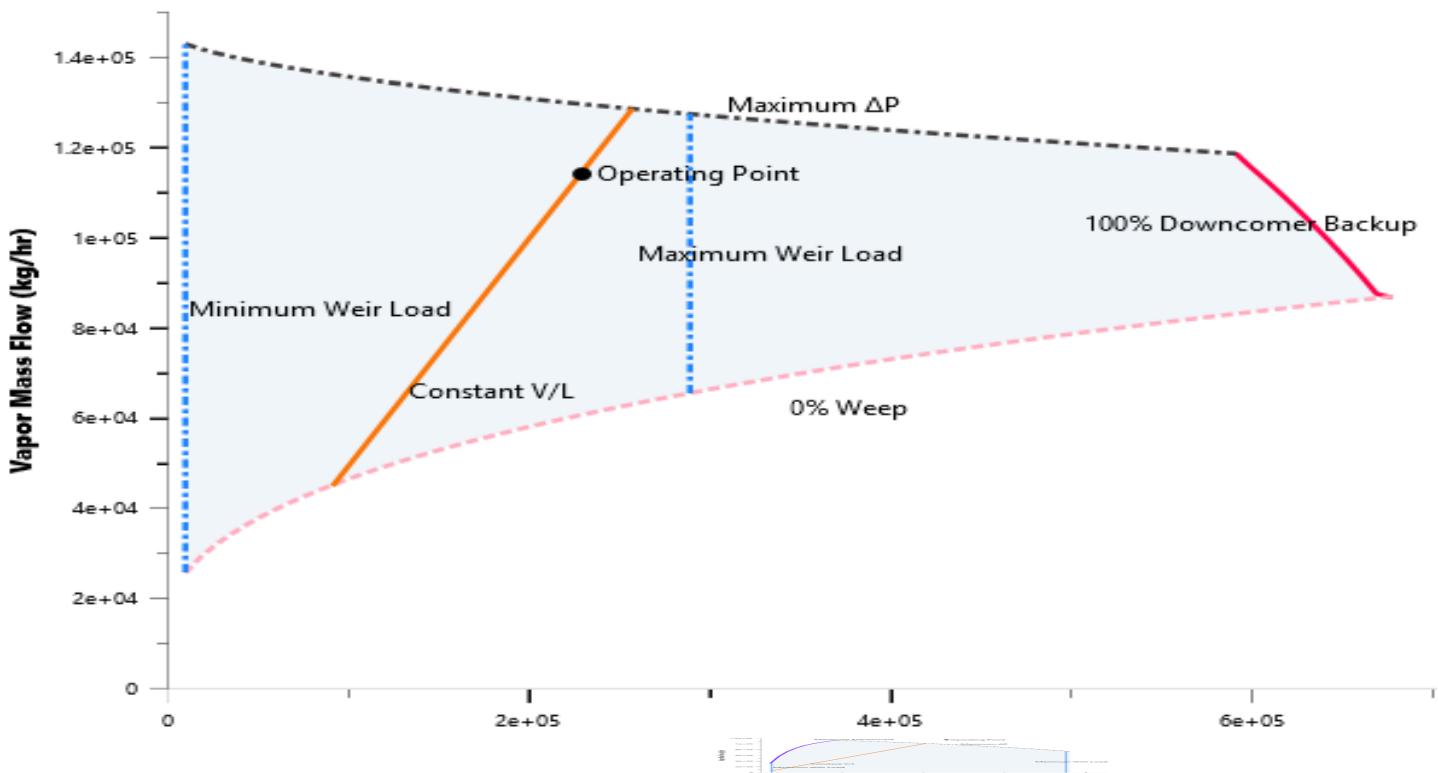
$$\alpha(i, r)_{H_2O} = (\alpha_{\text{dist}} \cdot \alpha_{B, H})^{0.5} = (6.749 \cdot 2.971)^{0.5} = 4.478$$

assume  $LK_{B, H} = 0.005$  to make possible calculation

$$N_{\min} = \frac{\log \left( \frac{LK_{B, H}}{LK_{\text{bottom}}} \cdot \frac{HK_{\text{dist}}}{HK_{\text{bottom}}} \right)}{\log(\alpha(LK, HK)_{H_2O})} = \frac{\log \left( \frac{7.1}{0.005} \cdot \frac{21.607}{24.603} \right)}{\log(4.478)} = 4.76 \text{ stages}$$

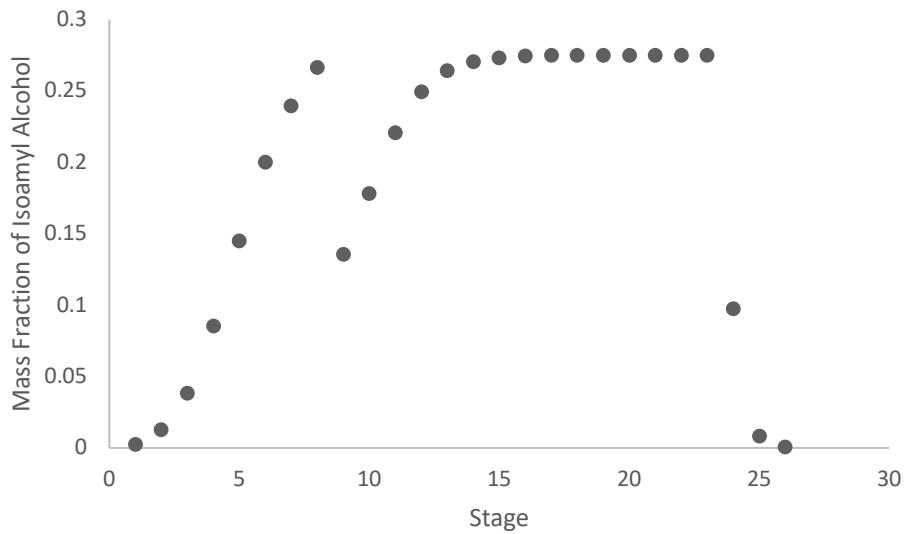


**Figure B2:** Hydraulic Plot Example with Bubble Cap Trays

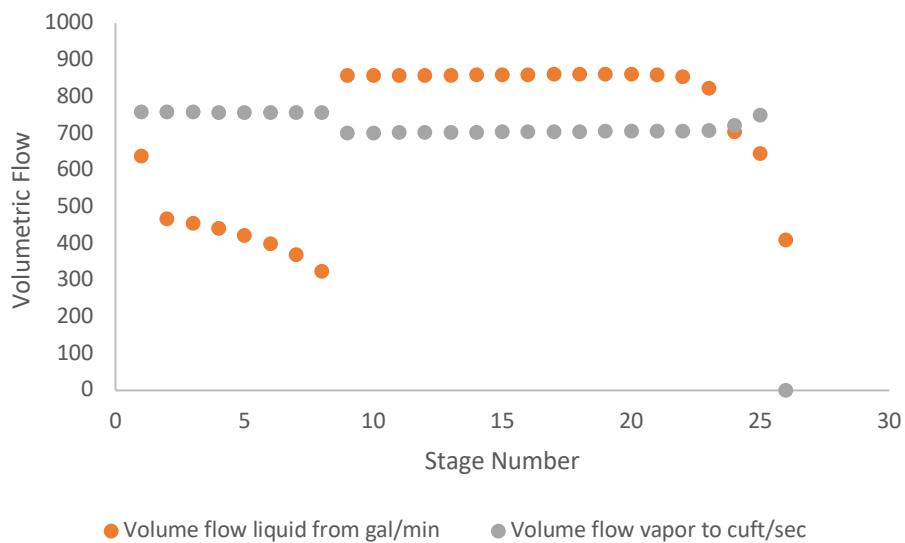


**Figure B3:** Hydraulic Plot Example with Nutter-BDP Trays

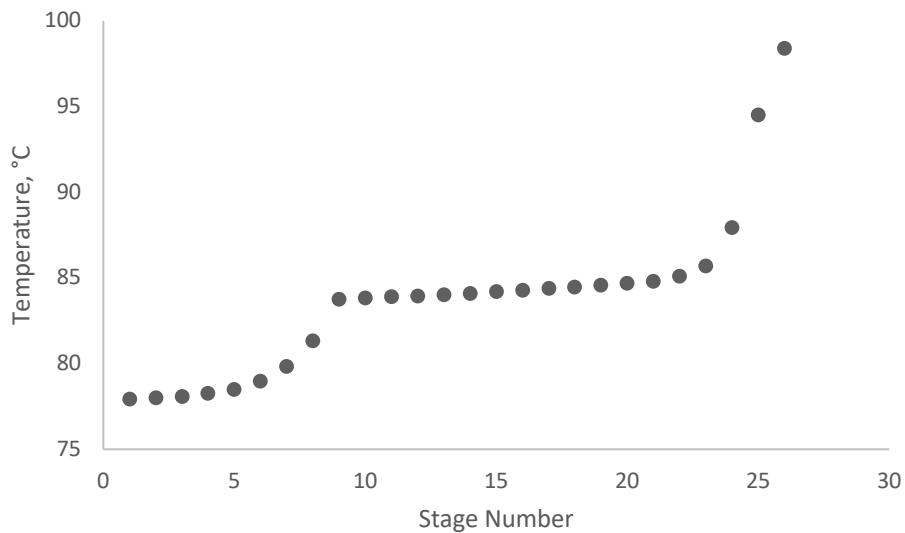
### *Rectifying Column*



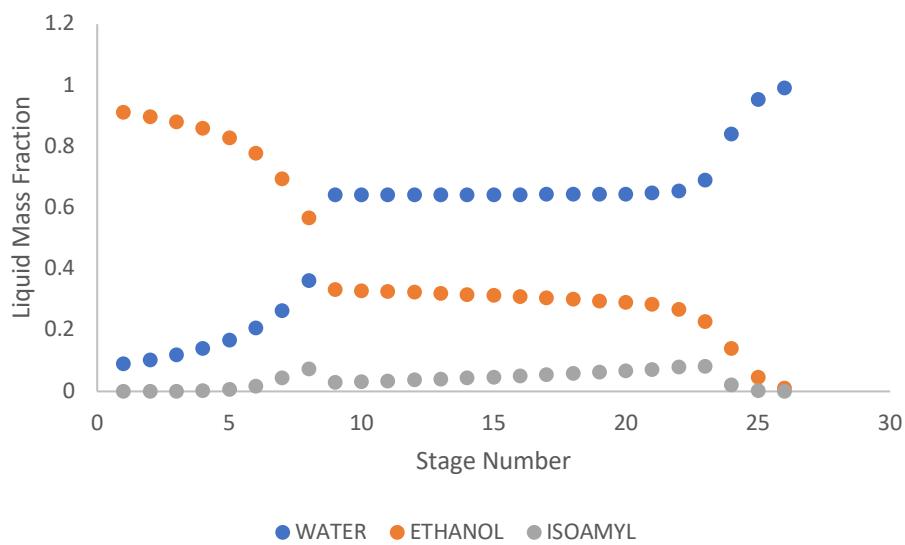
**Figure B4:** Composition Profile w/o Recovery System



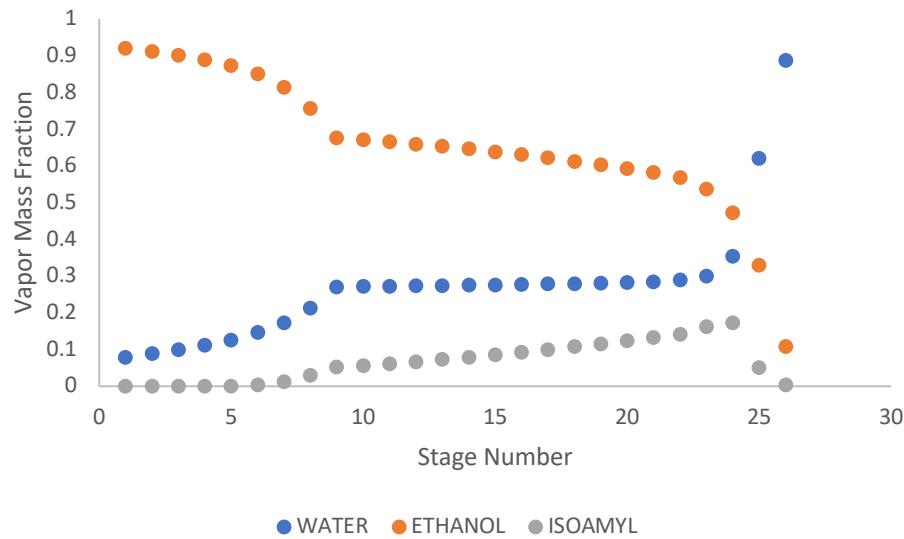
**Figure B5:** Volumetric Flow Profile w/ Recovery System



**Figure B6:** Temperature Profile w/ Recovery System



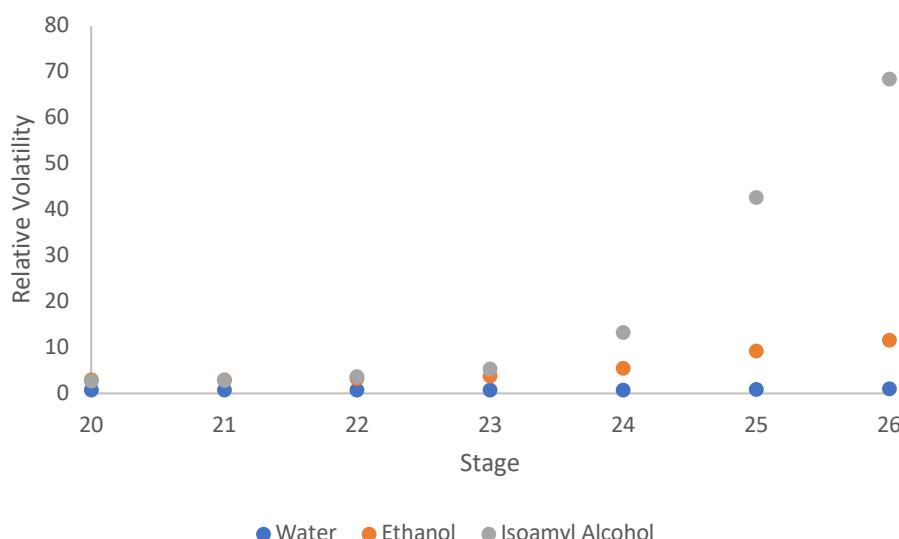
**Figure B7:** Liquid Composition Profile w/Recovery System



**Figure B8:** Vapor Composition Profile w/Recovery System

**Table B3:** Detailed Specifications of Distillation Column Design

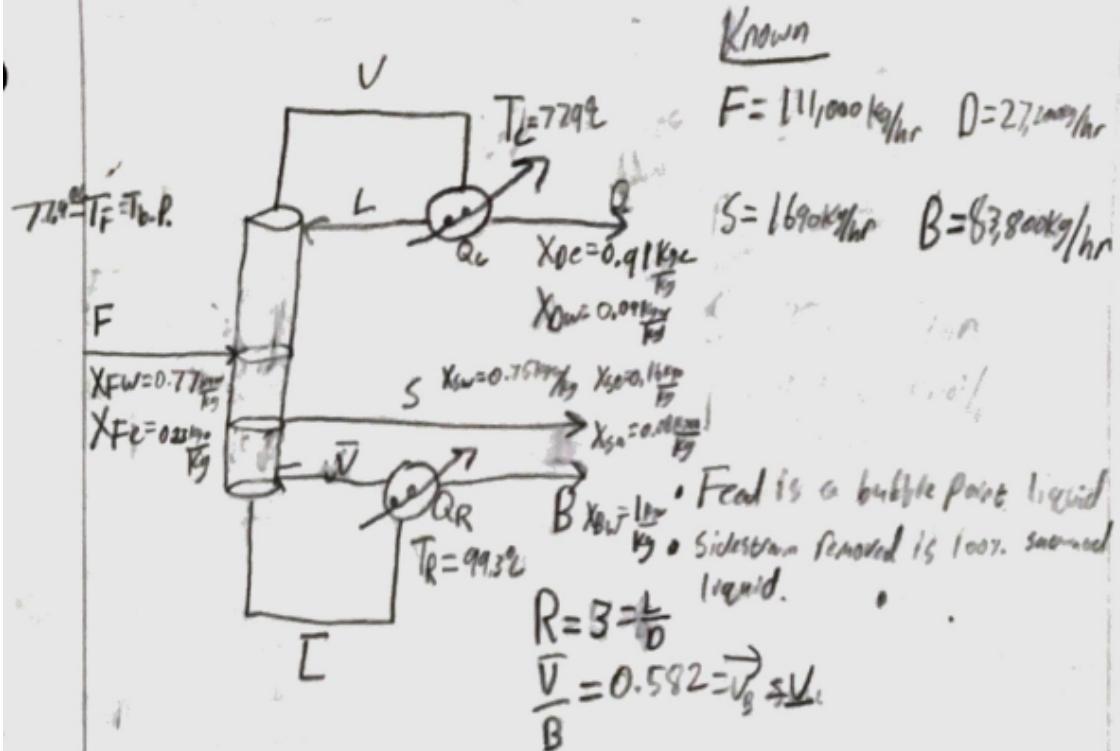
Specification	Correlation Value	Aspen Value
Number of Trays	25	25
Number of Sections	1	3
Column Pressure, psig	10	10
Height(s), m	11.65	4.26, 9.14, 1.22
Total Height, m	11.65	14.6
Diameter(s), m	3.23	3.92, 3.90, 3.54
Fraction of Flooding [14]	0.80	0.80
Tray Perforation	Bubble-Cap	Sieve
Tray Spacing, ft [12]	2	2
Tray Thickness	12 gauge; 2.77 mm	10 gauge; 3.40 mm
$A_h/A_a$	0.10	0.10
$A_d/A$	0.10	0.10
Reflux Ratio (Mass Basis)	3	3
Boilup Ratio (Mass Basis)	0.58	0.58
Tray and Column Material	Stainless Steel, 316, SA-240	Stainless Steel, 316, SA-240
Cost of Installed Column & Auxiliaries	\$1,072,500	\$524,810
Condenser Duty, kW	-29,400	-29,400
Condenser Temperature, °C	78	78
Reboiler Duty, kW	30,700	30,700
Reboiler Temperature, °C	99	99
Murphree Stage Efficiency [12]	0.70	0.70
Feed Stages	9, 24	9, 24
Total Pressure Drop, psi	3.0	3.0



**Figure B9:** Relative Volatility Profile of the Bottom Portion of Stripping Section

## Process Specifications

Flow rates inputted into Aspen Design



Overall Energy Balance:  $H_F = H_{ref} = (P = 1 \text{ atm}, T = 25^\circ\text{C})$

$$H_F F + Q_R = Q_L + D H_3 + B H_2 + Q_{loss} + S H_S \quad 307$$

$$\text{Condition} \\ Q_r = 2.94 \times 10^4 \text{ kW} = 1.06 \times 10^8 \text{ kJ/hr} \rightarrow Q_r = 3.02 \times 10^4 \text{ kW} = 1.11 \times 10^8 \text{ kJ/hr}$$

$$H_F = C_F(T_F - T_{Ff}) = 40485 \text{ J/kg K} (77.9 - 25) = 214 \text{ kJ/kg}$$

$$\rightarrow H_D = C_p (T_0 - T_{ref}) = 4220 \text{ J/kg K} (77.9^\circ\text{C} - 25^\circ\text{C}) = 223 \text{ kJ/kg}$$

$$\rightarrow H_B = C_{pw}(T_B - T_{Ref}) = 4186 \text{ J/kgK} (99.3 - 25)^\circ\text{C} = 311 \text{ kJ/kg}$$

$$H_S = (P_S(T_S - T_{Ref}) = 4082 \text{ J/kg K} (86.7 - 25)^\circ\text{C} = 25215 \text{ J/kg}$$

### Overall Energy Balance

$$(111,000 \text{ kg/hr}) (214 \text{ kJ/kg}) + 1,11 \times 10^8 \text{ kJ/hr} = 1,06 \times 10^8 \text{ kJ/hr} + 27,200 \text{ kg}_n (223 \text{ kJ/kg}) + 8300 \times 10^3 \text{ kg}_n (511 \text{ kJ/kg})$$

$$\dots + 1690 \text{ kg/hr} (252 \text{ kJ/kg}) + Q_{\text{loss}} \rightarrow Q_{\text{loss}} = -3.55 \times 10^6 \text{ kJ/hr}$$

Solving for  $S$  &  $B$  so E.B. balances.

2 eqns, 2 unk

$$① 111,000 = 27,200 + S + B$$

$$② 111,000 (214) + (1.11 - 1.06) \times 10^8 = 27,200 (223) + 311B + 252S$$

$$\rightarrow B = 2.66 \times 10^4 \text{ kg/hr} \quad S = 5.72 \times 10^4 \text{ kg/hr}$$

We know that  
fusel oil is not  
produced in this amount,  
So the flow rates inputted  
into Aspen are used

$$T_{\text{cold}} = 77.9^\circ \text{C}$$

$$T_{\text{ref}} = 28^\circ \text{C}$$

Distillation Column Design.  $\frac{A_h}{A_a} = 0.10$

Top Diameter

$$F_{LU} = \left[ \frac{R}{R+1} \right] \left[ \frac{P_L}{P_V} \right]^{\frac{1}{2}} \xrightarrow[T=75^\circ \text{C}]{y_e = 0.9189} F_{LU} = \frac{3}{4} \left[ \frac{7.4153}{807.4} \right]^{\frac{1}{2}} = 0.0719$$

$$\rightarrow C_F = 0.22 \frac{\text{ft}}{\text{s}} \text{ (from Fig. 6.23 P.226 Sieder 3rd)} = \text{Capacity Flooding Factor}$$

$$\rightarrow F_F = \text{Foaming Factor} = 1, \text{ not a foaming system}$$

$$\rightarrow F_{HA} = 1, \frac{A_h}{A_a} = 0.1 \rightarrow C = C_F \left( \frac{D}{20} \right)^{0.2} F_F F_{HA} \rightarrow \begin{array}{l} D = \text{Surface} \\ \text{tension of } F \\ \text{1 dyne/cm} \\ \text{and } L = 1 \text{ cm} \\ L + D = V \Rightarrow L = V - D \\ \frac{L}{V} = \frac{V - D}{V} \\ \frac{L}{V} = 1 - \frac{D}{V} \\ \frac{L}{V} = \frac{1}{2} \end{array}$$

$$\rightarrow C = (0.22) \left( \frac{2.2}{20} \right)^{0.2} = 0.224 \frac{\text{ft}}{\text{s}} \rightarrow U_f = C \left[ \frac{P_L - P_V}{P_V} \right]^{\frac{1}{2}}$$

$$\rightarrow U_f = (0.224 \frac{\text{ft}}{\text{s}}) \left[ \frac{807.4 - 7.4153}{7.4153} \right]^{\frac{1}{2}} = 2.33 \frac{\text{ft}}{\text{s}} = \text{flooding velocity}$$

$$\rightarrow f = 0.90 = \text{fraction of flooding} \rightarrow D_T = \left[ \frac{4V}{f u_f P_V \pi (1 - \frac{A_h}{A})} \right]^{\frac{1}{2}} = \text{Diameter of tray}$$

$$\rightarrow D_T = \left[ \frac{4(31.12 \text{ kg/s})}{(0.80)(2.33 \text{ ft/s})(\frac{1 \text{ m}}{3.28 \text{ ft}})(7.4153 \text{ kg/m}^3)(1 - 0.1)\pi} \right]^{\frac{1}{2}}$$

$$\rightarrow D_T = 3.23 \text{ m}$$

## Distillation Column Design

### Height

$$h_{tot} = h_{bottom} + h_{below} + (N_{trays} - 1)h_{tray} + N_{flooding thickness}$$

$$h_{tot} = \frac{4}{3.28} m + \frac{10}{3.28} m + (25-1)\frac{2}{3.28} m + (25)(\frac{2.77}{10^3}) m$$

$$\rightarrow h_{tot} = 11.65 m$$

### Costing

$$C_{tray} = \$30,000/tray \rightarrow C_{column} = 25(\$30,000) = \$750,000$$

From Fig 15-15, Peters, P. 796

### ASpen Distillation column costing

From

Fig 15-15  
Peters

Section 1:  $D_{T_1} = 3.92 m$ ,  $h_1 = 4.27 m$ , Sieve tray,  $N_{T_1} = 8$  trays,

$$\rightarrow C_{T_1} = \$15,000/tray \quad C_1 = 8(15,000) = \$120,000$$

Section 2:  $D_{T_2} = 3.90 m$ ,  $h_2 = 9.14 m$ , Sieve tray,  $N_{T_2} = 15$  trays

$$\rightarrow C_{T_2} = \$15,000/tray \rightarrow C_2 = 15(15,000) = \$225,000$$

Section 3:  $D_{T_3} = 3.54 m$ ,  $h_3 = 1.22 m$ ,  $N_{T_3} = 2$  trays

$$\rightarrow C_{T_3} = \$11,000/tray \rightarrow C_3 = 2(11,000) = \$22,000$$

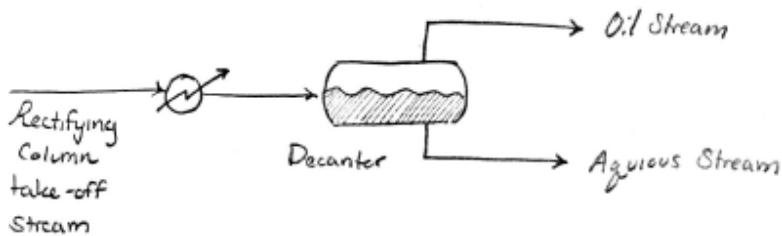
$$C_{Total} = \$367,000$$

*Decanter*

**Table B4:** Decanter equipment design specifications

<b><i>Decanter length:</i></b>	8.0 m
<b><i>Decanter Diameter:</i></b>	0.5 m
<b><i>Decanter Volume:</i></b>	1.57 m <sup>3</sup>
<b><i>Design Pressure:</i></b>	150 psi max
<b><i>Operating Volume:</i></b>	95% full
<b><i>Residence Time:</i></b>	43 minutes
<b><i>Material of construction:</i></b>	316 SS
<b><i>Purchase Cost:</i></b>	\$ 16,000

## DECANTER DESIGN CALCULATIONS



- Determine which phase separates:

$$\theta = \frac{V_L'}{V_H'} \left( \frac{\rho_L' \mu_H'}{\rho_H' \mu_L'} \right)$$

$V$  = volumetric flowrate  
 $\rho$  = density  
 $\mu$  = viscosity

\* subscript "L" denotes light phase  
 and subscript "H" denotes heavy phase

at  $10^\circ\text{C}$

$$\rho_H = 0.0013 \frac{\text{N}\cdot\text{s}}{\text{m}^2}, \rho_L = 55.50 \frac{\text{kmol}}{\text{m}^3}, \mu_L = 0.0060 \frac{\text{N}\cdot\text{s}}{\text{m}^2}, \mu_H = 9.28 \frac{\text{kmol}}{\text{m}^3}$$

$$V_L' = 1.88 \frac{\text{kmol}}{\text{h}} \left( \frac{1 \text{ m}^3}{9.28 \text{ kmol}} \right) \left( \frac{1 \text{ h}}{3600 \text{ s}} \right) = 5.63 \times 10^{-5} \frac{\text{m}^3}{\text{s}}$$

$$V_H' = 113 \frac{\text{kmol}}{\text{h}} \left( \frac{1 \text{ m}^3}{55.50 \text{ kmol}} \right) \left( \frac{1 \text{ h}}{3600 \text{ s}} \right) = 5.66 \times 10^{-4} \frac{\text{m}^3}{\text{s}}$$

$$\theta = \frac{(5.63 \times 10^{-5} \frac{\text{m}^3}{\text{s}})(9.28 \frac{\text{kmol}}{\text{m}^3})(0.0013 \frac{\text{N}\cdot\text{s}}{\text{m}^2})}{(5.66 \times 10^{-4} \frac{\text{m}^3}{\text{s}})(55.50 \frac{\text{kmol}}{\text{m}^3})(0.0060 \frac{\text{N}\cdot\text{s}}{\text{m}^2})} \rightarrow \theta = 0.0036$$

$\theta < 0.3 \therefore$  light phase always dispersed

[this is a check that a gravity decanter is a good design choice for this sep.]

- Velocity of dispersed phase ( $u_d$ ):

$$u_d = \frac{d_d^2 g (\rho_d - \rho_c)}{18 \mu_c}$$

subscript "d" denotes dispersed phase,  
 subscript "c" denotes continuous phase

$d_d$  = diameter of dispersed phase

$$u_d = \frac{\left(300 \mu\text{m} \frac{10^{-6} \text{m}}{1 \mu\text{m}}\right)^2 (9.81 \frac{\text{m}}{\text{s}^2})(999.8 \frac{\text{kg}}{\text{m}^3} - 818.3 \frac{\text{kg}}{\text{m}^3})}{18 (0.0060 \frac{\text{N}\cdot\text{s}}{\text{m}^2})} \quad \text{(assumption)}$$

$$= 6.84 \times 10^{-3} \frac{\text{m}}{\text{s}}$$

- Volumetric flow rate of continuous phase :  $5.66 \times 10^{-4} \frac{m^3}{s}$

- time of separation :  $t = \frac{100 \mu_c}{\rho_d - \rho_c} = \frac{100 (1.30 \text{ cp})}{999.8 \frac{\text{kg}}{m^3} - 812.3 \frac{\text{kg}}{m^3}} = 43 \text{ min}$

- Area of liquid interface :  $A_i = \frac{L_c}{u_c} = \frac{5.66 \times 10^{-4} \frac{m^3}{s}}{6.84 \times 10^{-3} \frac{m}{s}} = 0.082 m^2$

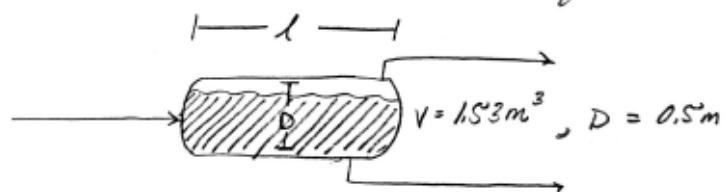
- radius of decanter :  $r = \sqrt{\frac{A_i}{\pi}} = \sqrt{\frac{0.082 m^2}{\pi}} = 0.16 m$

$$D = 2 \cdot r = 0.32 \rightarrow \text{round up to } 0.5 m$$

- Common  $\frac{L}{D}$  ratio for gravity decanters is 2-5 (5 is chosen to allow for maximum separation time)  
length =  $5 \times 0.5 = 2.5 m$

- hold-up volume:  $V_h = 43 \text{ min} (5.66 \times 10^{-4} \frac{m^3}{s}) (\frac{60 s}{1 \text{ min}}) = 1.46 m^3$

- assume vessel operates at 98% of capacity:  $\frac{1.46 m^3}{0.98} = 1.53 m^3$   
(length is adjusted to meet this required volume)



- find length of vessel,  $V = \frac{\pi D^2}{4} l \rightarrow \frac{4V}{\pi D^2} = l$

$$\rightarrow l = \frac{4(1.53 m^3)}{\pi (0.5)^2} \rightarrow l = 7.79 m \rightarrow l = 8.0 m$$

new  $V = \frac{\pi (0.5)^2}{4} (8.0) \rightarrow V = 1.57 m^3$

- Decanter is sized as a horizontal storage tank for operation at atmospheric pressure (minimum pressure rating is used)

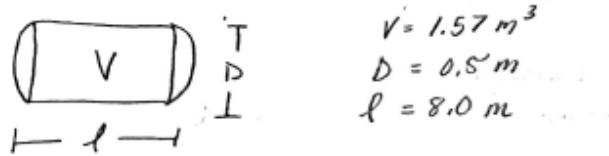


Fig 12.54 (Peters and Timmerhaus) is used for purchase cost of tank:

$$\text{for } l = 8.0 \text{ m} \quad \left. \begin{matrix} \\ D = 0.5 \text{ m} \end{matrix} \right\} \text{ purchase cost} = \$11,532 \quad * \text{operating max pressure of } 15.0 \text{ psia}$$

adjusting price from Jan 2002 to Nov. 2018  $\rightarrow$  multiply by 1.43

$$\left[ \begin{matrix} \text{Purchase} \\ \text{Cost} \end{matrix} \right] = 1.43 * \$11,532 \longrightarrow \boxed{\$16,491}$$

*Decanter Feed Chiller*

**Table B5:** Summary of Design Specifications for the Decanter Feed Stream Chiller

<b>Tube-side Fluid:</b>	Water/Alcohol	<b>Shell Side Fluid:</b>	Coolant (11 gpm)
<b>Tube-side Pressure Drop (psi):</b>	0.50	<b>Shell-side Pressure Drop (psi):</b>	0.20
<b>Heat Duty (kW):</b>	163	<b>Corrected LMTD (°C):</b>	24.0
<b>Exchanger Area (m³):</b>	50.1	<b>Exchanger Length (m):</b>	6
<b>UA (kW/m³-K):</b>	0.1357	<b>Shell Diameter (m):</b>	0.36
<b>Head Type:</b>	Fixed	<b>Pitch:</b>	Square
<b>Tube OD (in)</b>	0.75	<b>No. of Tube/Shell Passes:</b>	1/1
<b>Material of Construction</b>	316 SS	<b>Max Operating Pressure (psia):</b>	150
<b>Exchanger Cost (USD): \$17,116</b>			

**Table B6:** Stream Summaries for Shell and Tube Side Liquids

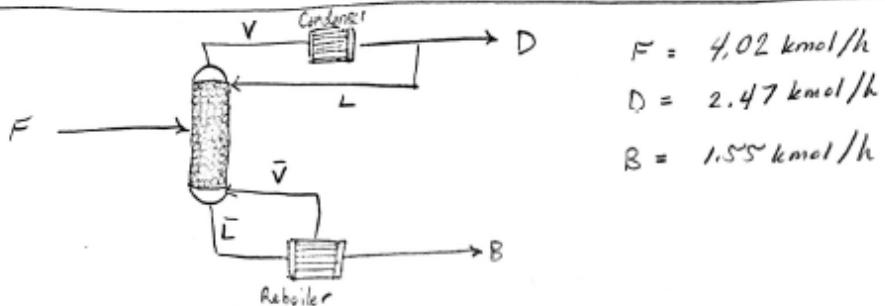
<b>Parameter</b>	<b>Tube-Side</b>	<b>Shell Side</b>
<b>Fluid:</b>	Water/Alcohol	Coolant (50% ethylene glycol)
<b>No. of Passes:</b>	1	1
<b>Inlet Temp (°C):</b>	86.0	-20.0
<b>Outlet Temp (°C):</b>	5.0	62.9
<b>Pressure Drop (psi):</b>	0.50	0.18
<b>No. of Tubes/Baffles:</b>	149	60

*Packed Column*

**Table B7:** Design Specifications for the Packed Distillation Column

<b>Total Column Height (m):</b>	6.1	<b>Operating Pressure (psig):</b>	10
<b>Packed Height (m):</b>	3.5	<b>Column Pressure Drop (psi):</b>	0.015
<b>Column Diameter (m):</b>	0.50	<b>Top Stage Temperature (°C):</b>	95
<b>Packing Type:</b>	½" Raschig Super-Ring	<b>Bottom Stage Temperature (°C):</b>	138
<b>HETP (ft)</b>	1.5	<b>Boil-up Ratio</b>	9
<b>Material of Construction</b>	316 SS	<b>Reflux Ratio:</b>	3
<b>Column Cost (USD):</b> \$64,000			

## PACKED COLUMN DESIGN CALCULATIONS



### Internal Mass Balances

- Feed is all liquid:  $\bar{L} = F + L$   
 - from reflux ratio:  $\frac{L}{D} = R \rightarrow R = \frac{L}{D} \rightarrow 3D = L$   
 $\rightarrow L = 3(2.47 \text{ kmol/h}) \rightarrow \boxed{L = 7.41 \frac{\text{kmol}}{\text{h}}}$   
 $\rightarrow \bar{L} = 4.02 \frac{\text{kmol}}{\text{h}} + 7.41 \frac{\text{kmol}}{\text{h}} \rightarrow \boxed{\bar{L} = 11.43 \frac{\text{kmol}}{\text{h}}}$

- $V = L + D \rightarrow V = 7.41 \frac{\text{kmol}}{\text{h}} + 2.47 \frac{\text{kmol}}{\text{h}} \rightarrow \boxed{V = 9.88 \frac{\text{kmol}}{\text{h}}}$
- $\bar{V} = \bar{L} - B \rightarrow \bar{V} = (11.43 \frac{\text{kmol}}{\text{h}} - 1.55 \frac{\text{kmol}}{\text{h}}) \rightarrow \boxed{\bar{V} = 9.88 \frac{\text{kmol}}{\text{h}}}$

### Column Sizing

$$F_{LV} = \frac{LM_L}{GM_G} \left( \frac{\rho_G}{\rho_L} \right)^{0.5} \quad * G \text{ is mostly water}$$

\* L is mostly isoamyl alcohol

- $F_{LV} = \frac{(7.41 \frac{\text{kmol}}{\text{h}}) \times 80.77 \frac{\text{kg}}{\text{kmol}}}{(9.88 \frac{\text{kmol}}{\text{h}}) \times 30.10 \frac{\text{kg}}{\text{kmol}}} \left[ \frac{0.578 \frac{\text{kg}}{\text{m}^2}}{731.00 \frac{\text{kg}}{\text{m}^2}} \right]^{0.5} \rightarrow F_{LV} = 5.66 \times 10^{-2}$

- $Y = \frac{u_r^2 F_p}{g} \left[ \frac{\rho_v}{\rho_{H_2O(l)}} \right] \underbrace{f(\rho_v)}_{\text{liquid density correction}} f(\mu_L) \quad \begin{matrix} \text{liquid viscosity correction} \\ \text{using } F_p \text{ chart: } Y @ \text{flooding} = 0.2 \end{matrix}$

• Density correction factor:

$$\left[ \frac{\text{density ratio of water to liquid}}{\rho_{\text{water}} / \rho_{\text{liquid}}} \right] = \frac{999 \frac{\text{kg}}{\text{m}^3}}{731 \frac{\text{kg}}{\text{m}^3}} = 1.4$$

$$\text{correction factor} = 1.6 \cdot f(\rho_L)$$

• Viscosity correction factor:

$$\mu_L = 0.462 \rightarrow \text{correction: } f(\mu_L) = 0.8$$

• @ flooding:  $0.2 = \frac{u_v^2 F_p}{g} \left[ \frac{\rho_v}{\rho_{H_2O(L)}} \right] f(\rho_L) f(\mu_L)$

$$\rightarrow u_v = \left[ \frac{(0.2) g}{F_p} \left[ \frac{\rho_{H_2O(L)}}{\rho_v} \right] \frac{1}{f(\rho_L) f(\mu_L)} \right]^{0.5}$$

$$\rightarrow u_v = \left[ \left[ \frac{(0.2)(9.81 \frac{\text{m}}{\text{s}^2})}{(96 \frac{\text{ft}^2}{\text{min}^2}) \times \frac{3.28 \text{ ft}}{\text{m}}} \right] \left[ \frac{999 \frac{\text{kg}}{\text{m}^3}}{0.578 \frac{\text{kg}}{\text{m}^3}} \right] \left[ \frac{1}{(0.8)(1.6)} \right] \right]^{0.5}$$

$$\rightarrow u_{vf} = 1.129 \frac{\text{m}}{\text{s}}$$

• operate @ 75% of flooding:  $u_v = 0.75 u_{vf} = (0.75)(1.129 \frac{\text{m}}{\text{s}}) = 0.84 \frac{\text{m}}{\text{s}}$

• Diameter of column:  $D_T = \sqrt{\frac{4GM_g}{u_v \pi \rho_v}} \rightarrow D_T = \sqrt{\frac{4(9.88 \frac{\text{kmol}}{\text{m}} \times 30.10 \frac{\text{kg}}{\text{kmol}}) \frac{1\text{h}}{3600\text{s}}}{(0.84 \frac{\text{m}}{\text{s}}) \pi (0.578 \frac{\text{kg}}{\text{m}^3})}}$

$$\boxed{D_T = 0.47 \text{ m}}$$

Aspen value of  $0.44 \text{ m}$  ✓

\* small difference due to the fact that Aspen operates at 80% flooding velocity.

• For high-efficiency random packings w/ low viscosity liquids:  
 $HETP(\text{ft}) = 1.5 D_p(\text{in}) \rightarrow HETP = 1.5(0.5 \text{ in}) = 0.75 \text{ ft}$

• height of column ( $l_T$ )  $\rightarrow l_T = (0.75 \text{ ft})(8 \text{ stages})$   
 $\uparrow$  0.5 inch Rasching rings

$$l_T = 6 \text{ ft} + (4 \text{ ft for reboiler}) + (10 \text{ ft for head-space})$$

$$\boxed{l_T = 20 \text{ ft}} \rightarrow \boxed{6.10 \text{ m}}$$

• packed Column Costing:

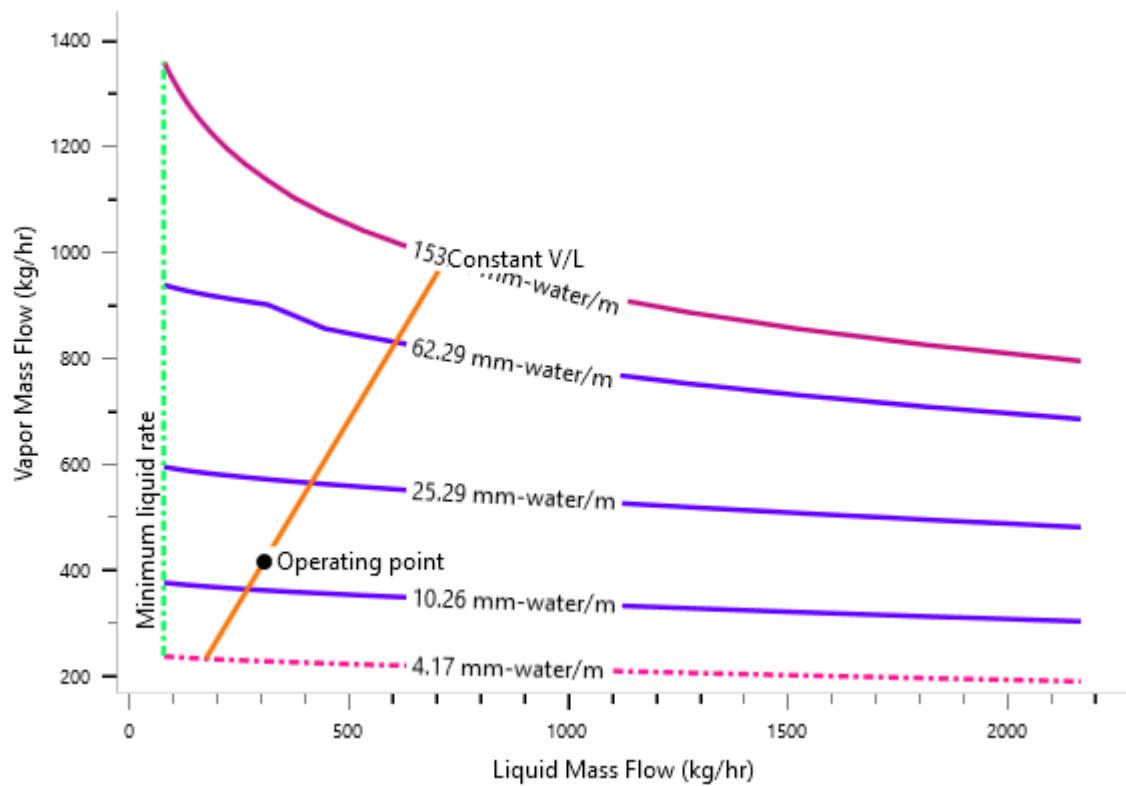
Diameter = 0.5m, Stainless Steel, 6.10 m Column height  
using fig 15-16 Timmerhaus and Peters

cost of packed column (including installation & aux.) = \$ 43,927

cost of porcelain packing = \$ 622 (0.5" Rasching Rings)  
(fig 15-14)

Adjust cost to 2018:

$$\text{Cost} = [\$622 + \$43,927] 1.43 \rightarrow \boxed{\text{Cost} = \$63,705}$$



**Figure B10:** Example Hydraulic Plot for a Middle Stage in the Packed Distillation Column  
Verifying Stable Operation

## **Appendix C: Economics and Capital Investment**

**Table C1:** Summary of Equipment Purchase Price for all Ethanol Production and Fusel Recovery Equipment

<b>Fusel Recovery Plant Addon</b>		
<b>Major Equipment</b>	<b>Purchase Cost</b>	
<i>Fusel Take-off stream heat exchanger</i>	\$	17,116
<i>Fusel Oil Decanter</i>	\$	16,491
<i>Packed Distillation Column</i>	\$	63,705
<i>Fusel Oil Product Storage Tank</i>	\$	108,306
<b>Total Major Equipment Cost:</b>	\$	205,618

<b>Ethanol Production Plant</b>		
<b>Major Equipment</b>	<b>Purchase Cost</b>	
<i>Fermenters</i>	\$	3,223,506
<i>Harvest Tanks</i>	\$	229,506
<i>Stripping Column</i>	\$	581,071
<i>Double-effect Evaporator</i>	\$	1,568,902
<i>Rectifying Column</i>	\$	367,000
<i>Ethanol Product Storage Tanks</i>	\$	339,373
<b>Total Major Equipment Cost:</b>	\$	6,309,358

**Table C2:** Calculation of Total Capital Investment from Equipment Purchase Cost Based on the Factorial Method of Cost Estimation

<b>Ethanol Plant</b>		<b>Fusel Recovery Plant Addon</b>	
<i>Equipment Purchase Cost:</i>	\$ 6,309,358	<i>Equipment Purchase Cost:</i>	\$ 205,618
<i>Installation Cost:</i>	\$ 2,586,837	<i>Installation Cost:</i>	\$ 84,303
<i>Piping Cost:</i>	\$ 2,302,916	<i>Piping Cost:</i>	\$ 75,051
<i>Electrical Costs:</i>	\$ 788,670	<i>Electrical Costs:</i>	\$ 25,702
<i>Instrument Costs:</i>	\$ 820,217	<i>Instrument Costs:</i>	\$ 26,730
<i>Building and Service Cost:</i>	\$ 1,924,354	<i>Building and Service Cost:</i>	\$ 62,713
<i>Excavation and site prep cost:</i>	\$ 946,404	<i>Excavation and site prep cost:</i>	\$ 30,843
<i>Auxiliaries cost:</i>	\$ 3,249,319	<i>Auxiliaries cost:</i>	\$ 105,893
<b>Total Physical Plant Investment:</b>	\$ 18,928,074	<b>Total Physical Plant Investment:</b>	\$ 616,854
<i>Field Expense:</i>	\$ 2,460,650	<i>Field Expense:</i>	\$ 80,191
<i>Engineering Cost:</i>	\$ 2,460,650	<i>Engineering Cost:</i>	\$ 80,191
<i>Direct Plant costs:</i>	\$ 4,921,299	<i>Direct Plant costs:</i>	\$ 160,382
<i>Contractor's fees:</i>	\$ 820,217	<i>Contractor's fees:</i>	\$ 26,730
<i>Contingency cost:</i>	\$ 2,460,650	<i>Contingency cost:</i>	\$ 80,191
<i>One-time labor fees:</i>	\$ 3,280,866	<i>One-time labor fees:</i>	\$ 106,921
<b>Total Capital Cost:</b>	\$ 27,130,239	<b>Total Capital Cost:</b>	\$ 884,157

**Table C3:** Utilities Cost Calculations for Both Cases of Ethanol & Fusel Recovery [7]

	Fusel Recovery Plant Add-on	Ethanol Production Plant	Total Plant
<i>Steam @ 100 psi (lb/hr):</i>	500	374859	375359
<i>Cooling Water (lb/h):</i>	54857	11190000	11244857
<i>Electricity (kW):</i>	52.507	74.599	127.106
	Unit	Cost Per Unit	
<i>Steam @ 100 psi (lbm):</i>	1000	\$ 13.87	
<i>Cooling Water (lbm):</i>	1000	\$ 0.25	
<i>Electricity (kWh):</i>	1	\$ 0.06	
	Fusel Recovery Plant Add-on	Ethanol Production Plant	Total Plant
<i>Steam @ 100 psi (Unit/Year):</i>	3942	2955388.356	2959330.356
<i>Cooling Water (Unit/Year):</i>	432492.588	88221960	88654452.59
<i>Electricity (Unit/Year):</i>	413965.188	588138.516	1002103.704
<i>Steam @ 100 psi (USD/Year):</i> \$	54,668	\$ 40,985,326	\$ 41,039,993
<i>Cooling Water (USD/Year):</i> \$	108,988	\$ 22,231,934	\$ 22,340,922
<i>Electricity (USD/Year):</i> \$	26,639	\$ 37,847	\$ 64,485
<i>Total Utilities Cost (USD/Year):</i> \$	<b>190,294.45</b>	\$ <b>63,255,106.35</b>	\$ <b>63,445,401</b>

**Table C4:** Summary of Raw Annual Raw Material Costs for the Ethanol Production Facility  
*These are shared by both cases analyzed in this design report.*

Pricing for Raw Materials	Cost per year
Corn	\$ 113,675,475.77
Yeast	\$ 4,639,124.53
Process Water	\$ 1,107,937.00
Total	<b>\$ 119,422,537.30</b>

**Table C4:** Calculation of Fixed Cost for the Ethanol Plant Running w/o Fusel Recovery

	Without Fusel Recovery				
	hrs/day	USD/hour	USD/day	USD/year	
<i>Operators:</i>	300	\$ 40	\$ 12,000	\$ 4,380,000	
<i>Supervisory:</i>		\$ 150	\$ 3,600	\$ 1,314,000	
<i>Benefits:</i>		\$ 26	\$ 624	\$ 227,760	
<i>Plant overhead:</i>		\$ 54	\$ 1,298	\$ 473,741	
<i>Maintenance Materials:</i>		\$ 7	\$ 164	\$ 59,687	
<i>Operating Materials:</i>		\$ 50	\$ 1,200	\$ 438,000	
<i>Local Taxes and Insurance:</i>		\$ 6	\$ 149	\$ 54,260	
<i>Total Fixed Cost:</i>			\$ 19,034	\$ 6,947,448	
<i>Capital Investment (10 year fixed straight line):</i>			\$ 7,433	\$ 2,713,024	

**Table C5:** Calculation of Fixed Cost for the Ethanol Plant Running w/ Fusel Recovery

	<b>With Fusel Recovery</b>			
	hrs/day	USD/hour	USD/day	USD/year
<i>Operators:</i>	350	\$40	\$ 14,000	\$ 5,110,000
<i>Supervisory:</i>		\$ 175	\$ 4,200	\$ 1,533,000
<i>Benefits:</i>		\$ 30	\$ 728	\$ 265,720
<i>Plant overhead:</i>		\$ 63	\$ 1,514	\$ 552,698
<i>Maintenance Materials:</i>		\$ 7	\$ 169	\$ 61,632
<i>Operating Materials:</i>		\$ 58	\$ 1,400	\$ 511,000
<i>Local Taxes and Insurance:</i>		\$ 6	\$ 154	\$ 56,029
<b>Total Fixed Cost:</b>			\$ 22,165	\$ 8,090,078
<i>Depreciation (10 year fixed straight line):</i>			\$ 7,675	\$ 2,801,440

**Table C6:** Ethanol Production Plant Cash Flow (no fusel oil recovery) for Years 1-5 of Hypothetical Operation

Ethanol Production Plant					
<b>Total Depreciable Capital</b>	\$ 27,130,239.40	Ethanol Selling Price (USD/gal)	\$ 1.64		
DCFRR	0.1	Fusel Selling Price (USD/gal)	\$ 10.28		
<b>Year:</b>	<b>1</b>	<b>2</b>	<b>3</b>	<b>4</b>	<b>5</b>
<b>total Production Volume (gal/year):</b>	75523832	75523832	75523832	75523832	75523832
<b>isel Production Volume (gal/year):</b>	0	0	0	0	0
<b>Sales Revenue:</b>	\$ 163,613,203.51	\$ 163,613,203.51	\$ 163,613,203.51	\$ 163,613,203.51	\$ 163,613,203.51
<b>Variable Costs (USD/year):</b>	\$ 182,677,643.65	\$ 182,677,643.65	\$ 182,677,643.65	\$ 182,677,643.65	\$ 182,677,643.65
<b>Fixed Costs (USD/year):</b>	\$ 6,947,447.81	\$ 6,947,447.81	\$ 6,947,447.81	\$ 6,947,447.81	\$ 6,947,447.81
<b>Depreciation (USD/year):</b>	\$ 2,713,023.94	\$ 2,713,023.94	\$ 2,713,023.94	\$ 2,713,023.94	\$ 2,713,023.94
<b>Gross Profit (USD/year):</b>	\$ (28,724,911.89)	\$ (28,724,911.89)	\$ (28,724,911.89)	\$ (28,724,911.89)	\$ (28,724,911.89)
<b>Income Tax (21%) (USD/year):</b>	0	0	0	0	0
<b>Net Profit (USD/year):</b>	\$ (28,724,911.89)	\$ (28,724,911.89)	\$ (28,724,911.89)	\$ (28,724,911.89)	\$ (28,724,911.89)
<b>Cash Flow (USD/year):</b>	\$ (26,011,887.95)	\$ (26,011,887.95)	\$ (26,011,887.95)	\$ (26,011,887.95)	\$ (26,011,887.95)
<b>NPW:</b>	\$ (287,249,118.88)				

**Table C6:** Ethanol Production Plant Cash Flow (no fusel oil recovery) for Years 6-10 of Speculative Operation

<b>Year:</b>	<b>6</b>	<b>7</b>	<b>8</b>	<b>9</b>	<b>10</b>
<b>total Production Volume (gal/year):</b>	75523832	75523832	75523832	75523832	75523832
<b>isel Production Volume (gal/year):</b>	0	0	0	0	0
<b>Sales Revenue:</b>	\$ 163,613,203.51	\$ 163,613,203.51	\$ 163,613,203.51	\$ 163,613,203.51	\$ 163,613,203.51
<b>Variable Costs (USD/year):</b>	\$ 182,677,643.65	\$ 182,677,643.65	\$ 182,677,643.65	\$ 182,677,643.65	\$ 182,677,643.65
<b>Fixed Costs (USD/year):</b>	\$ 6,947,447.81	\$ 6,947,447.81	\$ 6,947,447.81	\$ 6,947,447.81	\$ 6,947,447.81
<b>Depreciation (USD/year):</b>	\$ 2,713,023.94	\$ 2,713,023.94	\$ 2,713,023.94	\$ 2,713,023.94	\$ 2,713,023.94
<b>Gross Profit (USD/year):</b>	\$ (28,724,911.89)	\$ (28,724,911.89)	\$ (28,724,911.89)	\$ (28,724,911.89)	\$ (28,724,911.89)
<b>Income Tax (21%) (USD/year):</b>	0	0	0	0	0
<b>Net Profit (USD/year):</b>	\$ (28,724,911.89)	\$ (28,724,911.89)	\$ (28,724,911.89)	\$ (28,724,911.89)	\$ (28,724,911.89)
<b>Cash Flow (USD/year):</b>	\$ (26,011,887.95)	\$ (26,011,887.95)	\$ (26,011,887.95)	\$ (26,011,887.95)	\$ (26,011,887.95)
<b>NPW:</b>					

**Table C7:** Ethanol Production Plant Cash Flow (with added fusel recovery) for Years 1-5 of Hypothetical Operation

Ethanol Production Plant					
<b>Total Depreciable Capital</b>	\$ 28,014,396.80	Ethanol Selling Price (USD/gal)	\$ 1.64		
DCFRR	0.1	Fusel Selling Price (USD/gal)	\$ 10.28		
<b>Year:</b>	<b>1</b>	<b>2</b>	<b>3</b>	<b>4</b>	<b>5</b>
<b>total Production Volume (gal/year):</b>	75523832	75523832	75523832	75523832	75523832
<b>isel Production Volume (gal/year):</b>	356357	356357	356357	356357	356357
<b>Sales Revenue:</b>	\$ 167,201,029.64	\$ 167,201,029.64	\$ 167,201,029.64	\$ 167,201,029.64	\$ 167,201,029.64
<b>Variable Costs (USD/year):</b>	\$ 182,867,938.10	\$ 182,867,938.10	\$ 182,867,938.10	\$ 182,867,938.10	\$ 182,867,938.10
<b>Fixed Costs (USD/year):</b>	\$ 8,090,078.07	\$ 8,090,078.07	\$ 8,090,078.07	\$ 8,090,078.07	\$ 8,090,078.07
<b>Depreciation (USD/year):</b>	\$ 2,801,439.68	\$ 2,801,439.68	\$ 2,801,439.68	\$ 2,801,439.68	\$ 2,801,439.68
<b>Gross Profit (USD/year):</b>	\$ (26,558,426.21)	\$ (26,558,426.21)	\$ (26,558,426.21)	\$ (26,558,426.21)	\$ (26,558,426.21)
<b>Income Tax (21%) (USD/year):</b>	0	0	0	0	0
<b>Net Profit (USD/year):</b>	\$ (26,558,426.21)	\$ (26,558,426.21)	\$ (26,558,426.21)	\$ (26,558,426.21)	\$ (26,558,426.21)
<b>Cash Flow (USD/year):</b>	\$ (23,756,986.53)	\$ (23,756,986.53)	\$ (23,756,986.53)	\$ (23,756,986.53)	\$ (23,756,986.53)
<b>NPW:</b>	\$ (265,584,262.09)				

**Table C8:** Ethanol Production Plant Cash Flow (with added fusel recovery) for Years 6-10 of Hypothetical Operation

Year:	6	7	8	9	10
<i>total Production Volume (gal/year):</i>	75523832	75523832	75523832	75523832	75523832
<i>total Production Volume (gal/year):</i>	356357	356357	356357	356357	356357
<i>Sales Revenue:</i> \$	167,201,029.64	\$ 167,201,029.64	\$ 167,201,029.64	\$ 167,201,029.64	\$ 167,201,029.64
<i>Variable Costs (USD/year):</i> \$	182,867,938.10	\$ 182,867,938.10	\$ 182,867,938.10	\$ 182,867,938.10	\$ 182,867,938.10
<i>Fixed Costs (USD/year):</i> \$	8,090,078.07	\$ 8,090,078.07	\$ 8,090,078.07	\$ 8,090,078.07	\$ 8,090,078.07
<i>Depreciation (USD/year):</i> \$	2,801,439.68	\$ 2,801,439.68	\$ 2,801,439.68	\$ 2,801,439.68	\$ 2,801,439.68
<i>Gross Profit (USD/year):</i> \$	(26,558,426.21)	\$ (26,558,426.21)	\$ (26,558,426.21)	\$ (26,558,426.21)	\$ (26,558,426.21)
<i>Income Tax (21%) (USD/year):</i> \$	0	0	0	0	0
<i>Net Profit (USD/year):</i> \$	(26,558,426.21)	\$ (26,558,426.21)	\$ (26,558,426.21)	\$ (26,558,426.21)	\$ (26,558,426.21)
<i>Cash Flow (USD/year):</i> \$	(23,756,986.53)	\$ (23,756,986.53)	\$ (23,756,986.53)	\$ (23,756,986.53)	\$ (23,756,986.53)
<i>NPW:</i>					

**Table D1:** Process Equipment Hazards**a.)** Fermentation Considerations

<b>Issue</b>	<b>Causes</b>	<b>Consequences</b>	<b>Safeguards</b>
Fermenter pressure exceeds upper limit	<ul style="list-style-type: none"> <li>• Plug in pressure relief valve</li> <li>• Excess CO<sub>2</sub> being evolved in reaction</li> <li>• Pressure gauge malfunction</li> <li>• Clog in reactor effluent line</li> </ul>	<ul style="list-style-type: none"> <li>• Fermenter shell tensile strength compromised; potential rupture</li> <li>• Fermenter effluent composition compromised <ul style="list-style-type: none"> <li>• Stripping/rectifying column malfunction</li> </ul> </li> </ul>	<ul style="list-style-type: none"> <li>• Manual pressure relief valve that can be turned</li> <li>• Addition of an automatic back up pressure relief system</li> <li>• Installation of different pressure gauges from different manufacturers</li> <li>• High pressure alarm</li> </ul>
Fermenter pressure is below lower limit	<ul style="list-style-type: none"> <li>• Under production of CO<sub>2</sub> in fermenter</li> <li>• Pressure gauge malfunction <ul style="list-style-type: none"> <li>• Leak in fermentation vessel</li> </ul> </li> </ul>	<ul style="list-style-type: none"> <li>• Fermenter effluent composition compromised <ul style="list-style-type: none"> <li>• Stripping/rectifying column malfunction</li> </ul> </li> <li>• Inhalation exposure limits exceeded</li> </ul>	<ul style="list-style-type: none"> <li>• Installation of CO<sub>2</sub> monitoring apparatus with alarm <ul style="list-style-type: none"> <li>• Low pressure alarm</li> </ul> </li> <li>• Installation of different pressure gauges from different manufacturers</li> </ul>

## b.) Distillation, Stripping, and Packed Column Considerations

Issue	Causes	Consequences	Safeguards
Condenser exceeds/fails to operate within temperature limit	<ul style="list-style-type: none"> <li>Insufficient water flow through condenser</li> <li>Thermocouple malfunction</li> <li>Too much water flow through condenser</li> <li>Changes in column feed flowrates <ul style="list-style-type: none"> <li>Changes in reboiler operating conditions</li> </ul> </li> </ul>	<ul style="list-style-type: none"> <li>Not enough reflux back into column, causes distillate/bottoms composition to have higher impurity</li> <li>Too much reflux back into column, causes distillate/bottoms composition to have higher impurity</li> </ul>	<ul style="list-style-type: none"> <li>Control loop installation of reflux valve and flow meter for water flow.</li> <li>Thermocouples from different manufacturers installed before and after condenser</li> <li>Control loop implementation that regulates the flows, temperatures, and pressures of internal and external streams</li> </ul>
Reboiler exceeds/fails to operate within temperature limit	<ul style="list-style-type: none"> <li>Insufficient steam flow through reboiler</li> <li>Too much steam flow through reboiler</li> <li>Changes in column feed flowrates <ul style="list-style-type: none"> <li>Changes in condenser operating conditions</li> </ul> </li> </ul>	<ul style="list-style-type: none"> <li>Not enough boilup back into column, causes distillate/bottoms composition to have higher impurity</li> <li>Too much boilup back into column, causes distillate/bottoms composition to have higher impurity</li> </ul>	<ul style="list-style-type: none"> <li>Control loop installation of pressure sensors for steam flow.</li> <li>Thermocouples from different manufacturers installed before and after reboiler</li> <li>Control loop implementation that regulates the flows, temperatures, and pressures of internal and external streams</li> </ul>
Pressure inside the column not maintained	<ul style="list-style-type: none"> <li>Leakage in piping and fittings</li> </ul>	<ul style="list-style-type: none"> <li>• Explosion</li> <li>• Inhalation exposure limits exceeded</li> </ul>	<ul style="list-style-type: none"> <li>Pressure sensor and control loop implementation of the inlet and outlet streams of the column</li> </ul>

c.) Heat Exchanger and Evaporator Considerations

<b>Issue</b>	<b>Causes</b>	<b>Consequences</b>	<b>Safeguards</b>
Heat Exchanger Exceed Upper Temperature Limit	<ul style="list-style-type: none"> <li>• Cooling water flow is insufficient</li> <li>• Thermocouple malfunction</li> </ul>	<ul style="list-style-type: none"> <li>• Tensile strength of drum is compromised, could crack</li> <li>• Design separation not achieved, different product compositions</li> <li>• Autoignition temperature of anhydrous by products reached; explosion</li> </ul>	<ul style="list-style-type: none"> <li>• Flow sensor and control loop implementation</li> <li>• Thermocouples from different manufacturers installed at all exiting and entering streams</li> </ul>
Heat Exchanger Drops Below Lower Temperature Limit	<ul style="list-style-type: none"> <li>• Too much cooling water flow in coil</li> <li>• Thermocouple malfunction</li> </ul>	<ul style="list-style-type: none"> <li>• Design separation not achieved, different product compositions           <ul style="list-style-type: none"> <li>• DDGS not dried up to specifications</li> </ul> </li> </ul>	<ul style="list-style-type: none"> <li>• Flow sensor and control loop implementation</li> <li>• Thermocouples from different manufacturers installed at all exiting and entering streams</li> <li>• DDGS quality testing</li> </ul>
Pressure inside the apparatus not maintained	<ul style="list-style-type: none"> <li>• Leakage in piping and fittings</li> <li>• Leakage in vessel</li> </ul>	<ul style="list-style-type: none"> <li>• Explosion</li> <li>• Inhalation exposure limits exceeded</li> </ul>	<ul style="list-style-type: none"> <li>• Pressure sensor and control loop implementation of the inlet and outlet streams of the column</li> </ul>

**d.) Decanter Considerations**

<b>Issue</b>	<b>Causes</b>	<b>Consequences</b>	<b>Safeguards</b>
Decanter below lower temperature limit	<ul style="list-style-type: none"> <li>• Heat exchanger over cooling feed</li> <li>• Thermocouple malfunction</li> </ul>	<ul style="list-style-type: none"> <li>• Packed column malfunction</li> <li>• Fusel oil composition compromised</li> </ul>	<ul style="list-style-type: none"> <li>• Installation of cooling water flow meter on heat exchanger</li> <li>• Thermocouples from different manufacturers installed at all exiting and entering streams</li> </ul>
Decanter above higher temperature limit	<ul style="list-style-type: none"> <li>• Heat exchanger under cooling feed</li> <li>• Thermocouple malfunction</li> </ul>	<ul style="list-style-type: none"> <li>• Packed column malfunction</li> <li>• Fusel oil composition compromised</li> </ul>	<ul style="list-style-type: none"> <li>• Installation of cooling water flow meter on heat exchanger</li> <li>• Thermocouples from different manufacturers installed at all exiting and entering streams</li> </ul>

## References

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