SCALE UP FROM BENCH TO PLANT OF A WASTE TREATMENT PROCESS

FINAL REPORT

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CBE 488: PROCESS DESIGN & ECONOMIC ANALYSIS

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ABSTRACT

An industrial Avermectin fermentation process sporadically generates 200,000 L/day of aqueous waste containing several toxic compounds varying in composition. This project focuses on development of an efficient and reliable waste treatment scheme to neutralize avermectin and toluene in order to comply with environmental limits (Hazardous Air Pollutants, Volatile Organic Compounds, and Chemical Oxygen Demand). It is important to note that the incoming waste stream contains cell debris to which nearly all avermectin and ~95% of the toluene is bound, and that Avermectin is a large molecule with a high boiling point. With these parameters in mind, a treatment train consisting of **a steam stripper** to remove toluene, series of two **multiple effect evaporators** to separate avermectin, a high temperature incinerator to decompose the avermectin, and a chemical oxygen demand treatment tank was developed to meet the environmental COD limits. Process simulations made using Aspen Plus V14 indicate the proposed scheme can reach a final mass fraction of **1.62E-14** Avermectin, recover toluene at ~99.9% purity, and yield a processed stream virtually free of toluene and COD. Over a 10 year period, the total cost of this treatment system is estimated to cost **5.84%** of the revenue generated during the avermectin production process, after accounting for the recovered toluene. In other words the sale price of avermectin increases by ~6.0% due to waste treatment expenditures. Although this is above our proposed initial budget limitations, we believe that the safety factor resulting from the inherent thermodynamic separation associated with the multiple effect evaporators is worth the additional expenditure considering the toxicity of avermectin to aquatic life.

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SECTION 1: INTRODUCTION TO PROBLEM STATEMENT

Our team was tasked with treatment of a toxic waste stream stemming from an already-running chemical process designed to produce Avermectin. Avermectins are a group of 16-member, naturally occurring macrocyclic lactone endectocides that have a broad range of applications in agriculture, veterinary, and human medicine due to their anti-parasitic properties. Avermectins are however highly toxic to fish and aquatic organisms and are typically industrially produced via upstream fermentation of *Streptomyces avermitilis* and downstream extraction.[9] As such, waste streams from Avermectin production must be stringently monitored and treated to meet environmental safety requirements.

Our team was tasked with designing a robust, cost-effective, and environmentally-compliant wastewater treatment process to handle the waste stream resulting from an Avermectin production process. The waste stream is 200,000 L/day containing an average of ~7.96 kg/day of toluene, approximately 0.1 w/w% cell debris, and trace amounts of polyethylene glycol, palmitic acid glycerides, and fermentation residue.

The environmental stipulations require that the Volatile Organic Compound (VOC) toluene is a Hazardous Air Pollutant (HAP) and must be reduced to <100 ppm per the Clean Air Act, any residual Avermectin must be reduced to <1 ppb, the final pH of the stream must be between 6.5 and 7.5 and lastly, the stream must be also be treated to reduce Chemical Oxygen Demand (COD) prior to environmental discharge.[10, 11, 12] The environmental discharge criteria for our project is summarized below in Table 1.

Table 1.1: Summarized Environmental Discharge Criteria

Contaminant	Target Limit	Regulation Source
Toluene (VOC)	< 100 ppm	Clean Air Act (1990)
Avermectin	< 1 ppb	Environmental Safety Requirement
Final Stream pH	6.5 - 7.5	Wastewater Discharge Standard
COD (Chemical Oxygen Demand)	Significant Reduction Required	General Industrial Standard

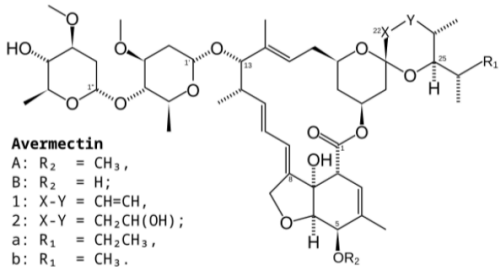

Figure 1.1: Avermectin Structure

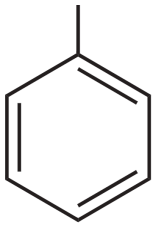

Figure 1.2: Toluene Structure

It should be noted that the toxic Avermectin compound has a strong tendency to bind to the cell debris, with nearly all the avermectin believed to be bound to the cell debris found in the waste stream, and also worth mentioning is its molecular weight (873.1 g/mol, per PubChem database) leading to it being very thermally stable.[13] The toluene on the other hand presents another dilemma, as only 5% of the toluene by mass is in aqueous phase while the other 95% by mass is once again bound to the cell debris in the waste stream being processed.

Over the course of the project, multiple treatment strategies–including ultrafiltration, centrifugation, caustic degradation, and multiple-effect evaporation (MEE)–were systematically evaluated through back-of-the-envelope calculations, decision matrices, and feedback from Dr. Raghavan.

Given the substantial scale of the waste stream (200,000 L/day), careful consideration was required to balance environmental compliance, energy consumption, capital cost, and operability. Taking a stream of this magnitude, with significant contamination hazards and persistent compounds required a robust and energy efficient solution to meet both economic and environmental objectives.

SECTION 2: METHOD SELECTION

Our team was presented with three major treatment operation choices and developed a design choice matrix to evaluate the three options and select one for full process design. The three major choices were based on the following three unit operations: centrifugation, membrane separation, and multiple effect evaporation (MEE). To quickly and efficiently decide on a design choice, the team conducted "back-of-the-envelope" calculations to evaluate the three potential design choices across various metrics. Each option was evaluated for its feasibility and goodness of fit (supported by the back-of-the-envelope calculations) for our process across 12 different metrics and assigned a numeric score with a value of 1 being an Excellent fit and a value of 5 being a Very Poor fit.[2] The matrix results are included below.

Table 2.1: Matrix to determine best cell debris removal

		Centrifuge	Membrane	Evaporator/MEE
1	Capital Cost	1	3	3
2	Utility Cost	2	1	4
3	Energy Consumption	1	2	4
4	Fouling	3	3	2
5	VOC Compliance	1	1	1
6	Particulate Size Handling	5	2	1
7	HAP Compliance	1	1	1
8	Solid Handling	5	3	1
9	Material of Construction	1	4	1
10	Requires Additional Equipment	4	2	4
11	Maintenance Requirements	3	3	1
12	Commercial Availability	1	1	1
	Total	**28**	**26**	**24**

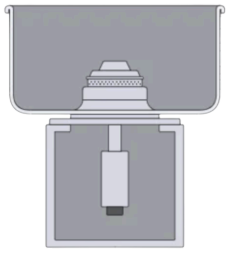

Figure 2.1 Centrifuge

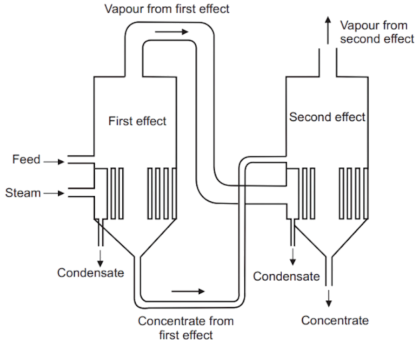

Figure 2.2 MEE

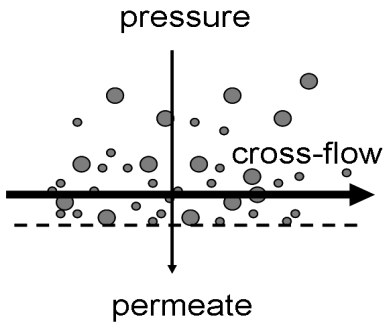

Figure 2.3 Membrane Filtration

When considering membrane filtration, we had to determine the type of membrane to use – ceramic or polymer. A ceramic membrane is what is usually used for wastewater treatment. Ceramic has a higher throughput, but it comes with a higher price tag. This is typically offset because ceramic membranes also have a longer operating life. Polymer membranes deal with a smaller throughput and a lower price tag, but having to replace them more often would not lend well for our process.[22] We decided if we were to move forward with membrane filtration, then we would use a ceramic membrane. Additionally, membrane filtration is a similar physical particulate separation method to centrifugation, but it comes with increased capital and utility costs. Additionally, although not explicitly stated in this chart, the pressure drop across the membrane would likely cause issues and require large amounts of pressure and energy. We do not believe that this option affords the team any additional advantages over centrifugation and will, therefore, be neglected in future process design.

When considering centrifugation, we had to evaluate the settling velocity (how fast the particles settle) using the Sigma Theory. This will help us determine if our particle can be separated from the liquid using centrifugation. The following are the calculations we did to determine the settling velocity[1].

$$\frac{Q}{\Sigma} = 2u_g$$

EQ: 2.1

$$u_g = \frac{\Delta p d_s^2 g}{18\mu}$$

EQ: 2.2

All of the numbers to solve for u_g were provided by lab data given to us in the problem statement.

$\Delta p = 1050\text{-}1000 \text{ kg/m}^3 = 50 \text{ kg/m}^3$

$d_s^2 = (.1 \text{ x}10\text{-}6 \text{ m})^2 = 1 \text{ x } 10\text{-}14 \text{ m}^2$

$g = 9.81 \text{ m/s}^2$

$\mu = 0.001093 \text{ Pa-s}$

$u_g = 2.49 \text{ E-}10 \text{ m/s}$

$\frac{Q}{\Sigma} = 4.99 \text{ E-}10 \text{ m/s}$

Table 2.2: Selection of sedimentation centrifuges

Type	Approximate efficiency (%)	Normal operating range Q, m^3/h at Q/Σ m/s
Tubular bowl	90	0.4 at 5×10^{-8} to 4 at 3.5×10^{-7}
Disc	45	0.1 at 7×10^{-8} to 110 at 4.5×10^{-7}
Solid bowl (scroll discharge)	60	0.7 at 1.5×10^{-6} to 15 at 1.5×10^{-5}
Solid bowl (basket)	75	0.4 at 5×10^{-6} to 4 at 1.5×10^{-4}

This Q/Σ value was then compared with Table 2.2, which contains the normal operating ranges for sedimentation centrifuges. Our settling velocity did not fall within the range of any of the centrifuges therefore centrifugation could not be used to remove the solids from our waste stream as is. If we wanted to continue with centrifugation, it would require additional equipment. We would need alum to flocculate (clump together) the organics. This would require empirical testing to determine the flocculant, which was outside the scope of our project as assigned to us.[1]

Other categories evaluated in the comparison matrix are evaluated below.

1. **Capital Cost:** The rough capital cost for each separation method was calculated: The centrifuge cost being $235,505, the filter being $1,521,700, and the MEE (calculation done for 4 evaporators) being $1,788,000.
2. **Utility Cost:** The utilities have been roughly costed for each separation method with membrane separation being the cheapest at $1800/year, centrifugation utilities costing $6400/year, and the evaporation utility cost $222,000/year. This does not include the incineration or COD costs, just the utilities that are used for each specific process.
3. **Energy Consumption[3]:** All three processes will result in high energy consumption due to the decision to utilize the incineration aspect rather than treating it with caustic. To get a full picture of what this looks like for each process would require a full design.
4. **Fouling[2]:** Both membrane separation and centrifugation have the potential for large amounts of fouling if not cleaned correctly. Evaporation must also be cleaned but the fouling potential is lower.
5. **VOC Compliance:** All three processes will be conducted to ensure they fall within the compliance guidelines.
6. **Particulate Size Handling:** The centrifuge can not handle the size of the particle we are dealing with. The ceramic membrane chosen would be able to handle the size of the particle.
7. **HAP Compliance:** All three processes will be conducted to ensure they fall within the compliance guidelines.
8. **Solid Handling:** The membrane and centrifuge will both be classified as solid handling steps requiring more people to operate the plan.
9. **Material of Construction:** All three techniques can be constructed with carbon steel. The membrane will be made of ceramic.
10. **Requires Additional Equipment:** The multiple-effect evaporation will require the most outside equipment, while the membrane and centrifuge will require only a pump and a tank to feed to each.
11. **Maintenance Requirements:** Both the centrifuge and membrane require at least monthly maintenance, while the evaporation process requires annual maintenance.
12. **Commercial Availability:** Both the membrane filters and the centrifuges would be bought from an industrial supplier. The evaporators can be easily purchased but require additional assembly to set up the process.

Ultimately, the Multiple Effect Evaporator design path was found to have the lowest matrix score (relatively best fit to our process) and selected as a unit operation.[2, 14]

Proof of Concept

After the above matrix was finished, a proof-of-concept Aspen simulation was also completed to ensure that avermectin could be separated with high fidelity from the wastewater stream via an evaporative process.[2]

Firstly, a single evaporator was used to model the Vapor-Liquid separation. The team notes that this is technically not considered an MEE setup, as there is only a single effect. Nevertheless, it is the simplest version of this separation and was modeled as a proof of concept–in other words, does evaporating the majority of the water off leave the avermectin in reasonable (environmentally safe) quantities? See the Aspen setup for the single evaporator below. The Aspen file will be attached, titled "IntialMEESimulation." Note that this is NOT the overall process file, but was simply a proof-of-concept for MEE operation specifically.

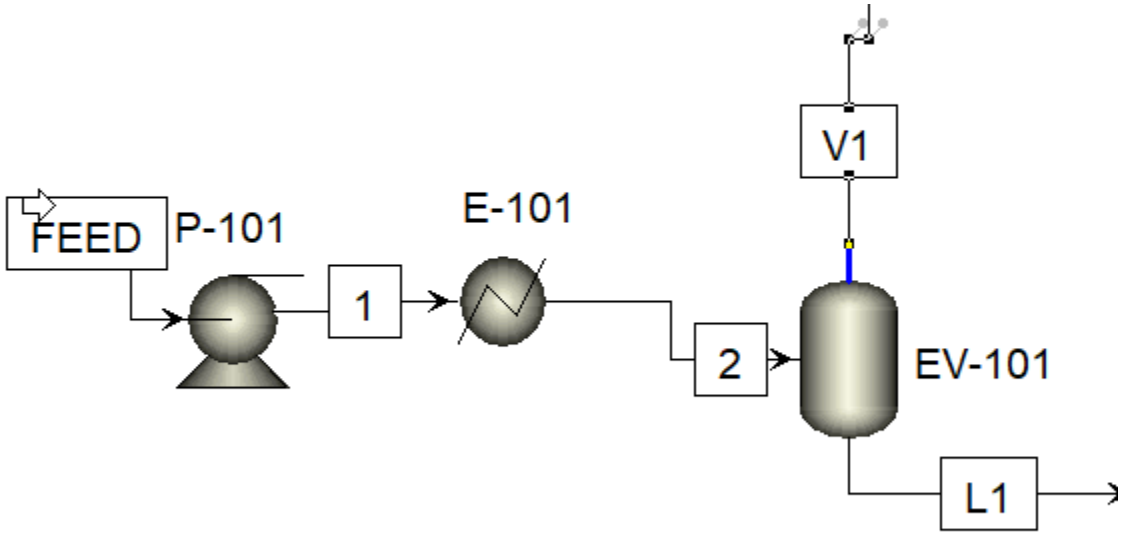

Figure 2.4: Single Effect Evaporator Aspen Setup

As specified above, the first effect (EV-101) was set to 240 ℃. The heat duty was iterated until 10,000 kg/day of water (95% concentration) was achieved in material stream L1. This required a heat duty of 4770 kW. Additionally, the pressure in EV-101 was 32.93 bar, which is relatively high, but reasonable for the high-pressure steam system in-use. Most importantly, the split between avermectin and water worked beautifully. In the initial waste water feed stream, 8 kg/day of avermectin are added to the system. V1 contains only 6.49E-06 kg/day, representing over a 1,000,000 X reduction in avermectin concentration. This is also below the 1 ppb limit specified in the problem statement. In fact, the only compound that appreciably vaporizes into the vapor stream (other than water) is glucose. This isn't a major concern, as the final output stream undergoes Chemical Oxygen Demand (COD) treatment that will remove any remaining toluene and reduce the COD that any remaining trace amounts of glucose would have.

SECTION 3: BFD/PFD

Once the methods of toluene separation and the separation of the organic were decided, a general block flow diagram (BFD) could be assembled as shown below.[2, 15]

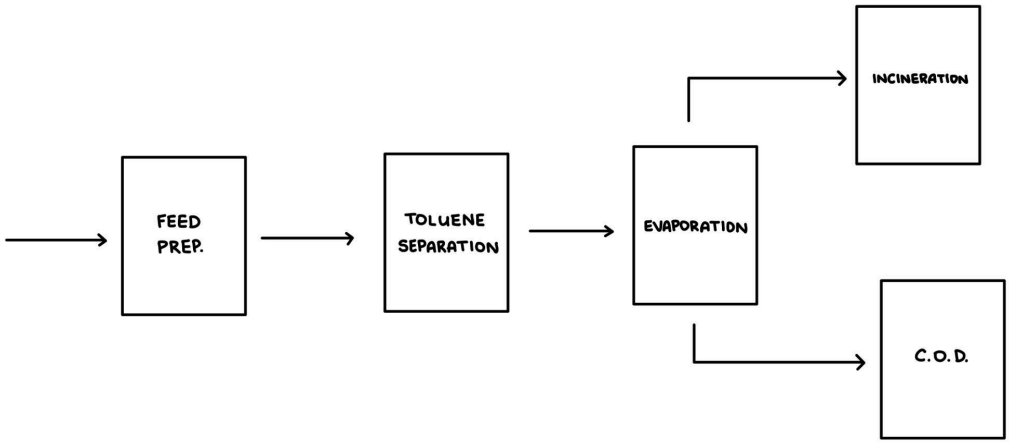

Figure 3.1: Avermectin Waste Treatment Block Flow Diagram

Then, we broke down the BFD into a process flow diagram (PFD) to determine all the equipment needed for our process.[2, 15, 16]

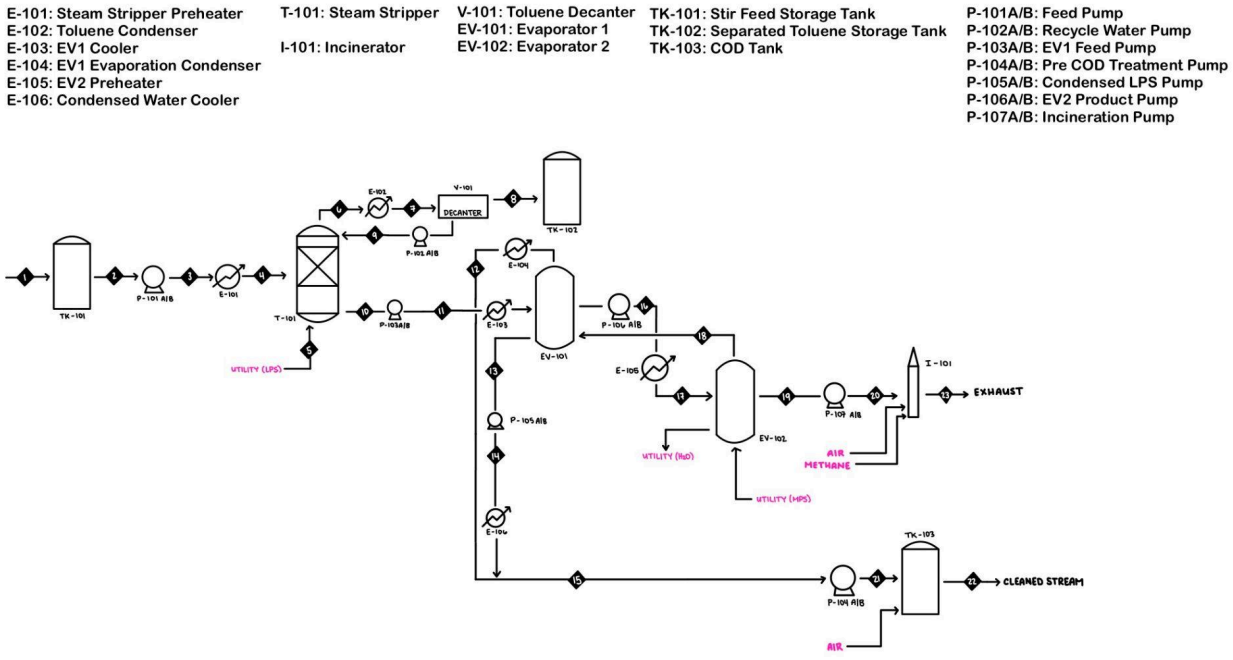

Figure 3.2: Avermectin Waste Treatment Process Flow Diagram

With the PFD designed, we could move to simulating the process in Aspen Plus V14.

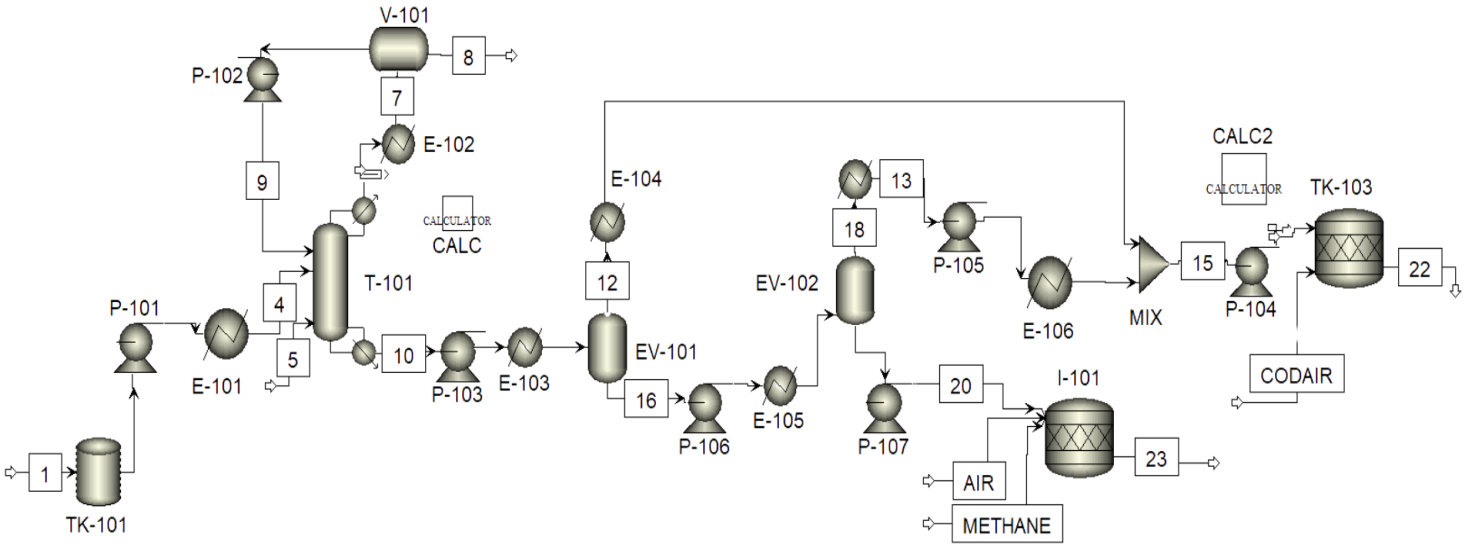

Figure 3.3: Aspen Plus V14 Simulation Process Flow Diagram

SECTION 4: MATERIAL AND ENERGY BALANCE

Material Balance

The stream tables are extremely lengthy and are not included directly in this report. They are listed in the Microsoft Excel file titled “Avermectin Material Balance.”

Energy Balance/Utility Considerations

In this process the Exchangers (E-101, E-102, E-103, E-104, E-105, E-106) and Pumps (P-101 A/B, P-102 A/B, P-103 A/B, P-104 A/B, P-105 A/B, P-106 A/B, P-107 A/B) utilized utilities of low pressure steam, mid pressure steam, cooling tower water, and electricity. All utility costs were provided or taken from Source 2 and on a basis of 340 operating days per year. Utilities for T-101, I-101, and TK-103 can be found in Section 5.

Table 4.1 Utility Summary 1

Equipment	Utility	Heat Duty (MMbtu/hr)	Flow (kg/hr)	Utility Cost ($/yr)
E-101	LPS	2.184	1,020.9	$78,724.27
E-102	CTW	0.563148	14,210.8	$1,826.06
E-103	CTW	0.188352	4,753.0	$610.75
E-104	CTW	10.040	253,355.7	$32,555.57
E-105	MPS	0.910116	476.9	$37,123.41
E-106	CTW	1.93	48,702.84	$6,258.19
T-101	MPS	-	179.921	$14,006.20
I-101	Compressed Air	-	366.125	$14,937.92
I-101	Methane	-	109.373	$99,869.37
TK-103	Compressed Air	-	8.62	$98.29
EV-102	MPS	0.014	4,541.2	$353,518.26

Table 4.2 Utility Summary 2

Equipment	Electric Duty (W)	Utility Cost ($/yr)
P-101 A/B	266.003	$243.83
P-102 A/B	7.768	$7.12
P-103 A/B	301.355	$276.23
P-104 A/B	275.092	$252.16
P-105 A/B	2522.294	$2,312.04
P-106 A/B	744.761	$682.68
P-107 A/B	6.852	$6.28

Note that the above utilities can be summed to find our overall C_{UT}, which will be used in the economic analysis section. Each of these values is summed in Microsoft Excel format, and found in the attached "Data Sheet." file.

Additionally, note that each pump, heater, etc. has a justification for specified temperature and pressure, either in the equipment design section (MEE, steam stripper, incinerator, COD tank-associated equipment) or in Appendix A of this report.

SECTION 5: SIMULATION AND EQUIPMENT DESIGN

Equipment Design

Multiple Effect Evaporators (MEEs)

In the above Methods Selection section, multiple effect evaporators were chosen as the optimal separation method for the avermectin treatment process. However, there are two common configurations for multiple effect evaporators, which are called co- and counter-current, or, equivalently, forward and backward feed. Each method will be briefly described, as well as associated pros and cons applicable to the avermectin process.[2]

In co-current (forward) feed, steam (heating medium) moves in the same direction as the wastewater flow. Thus, wastewater will first enter the hottest evaporator, and the temperature and pressure of each subsequent evaporator will be lower.

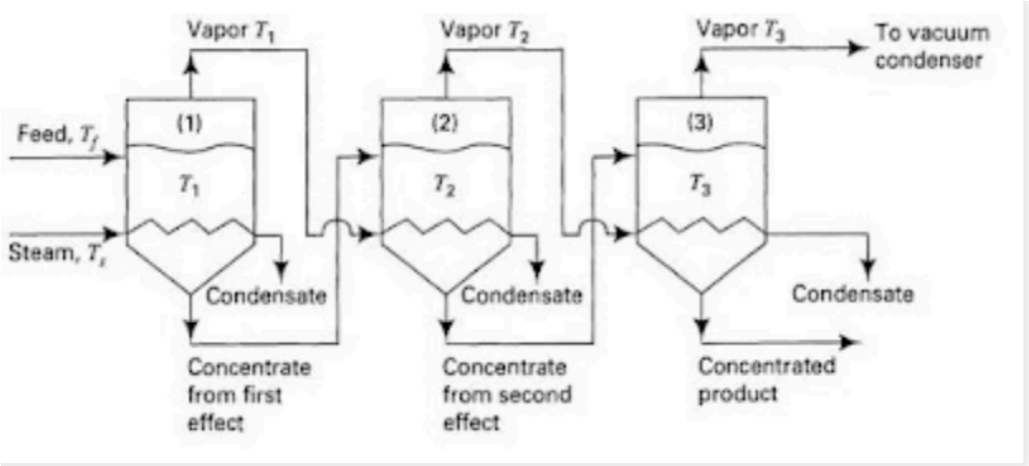

Figure 5.1: Co-Current Flow[4]

- Pros:
  - This configuration is good for evaporator trains that include heat-sensitive materials.[5] Note that, in our case, we don't have any specific components that we want to protect from heat denaturation. In fact, we want avermectin to degrade via caustic/thermal methods. Thus, this potential advantage does not apply to our process.
  - This configuration is easier operationally since pumps are not necessary between each effect as the liquid is flowing from higher pressure to lower pressure.
- Cons:
  - Relatively more expensive both for capital costing and utility purposes.[5,6] However, this may be partially offset by the fact that this operation does not require intermediate pumps.

In counter-current (backward) feed, steam and wastewater flow in opposite directions. That is, the steam is the coolest in the first stage entered by the wastewater, and the hottest in the most concentrated evaporator (last wastewater stage).

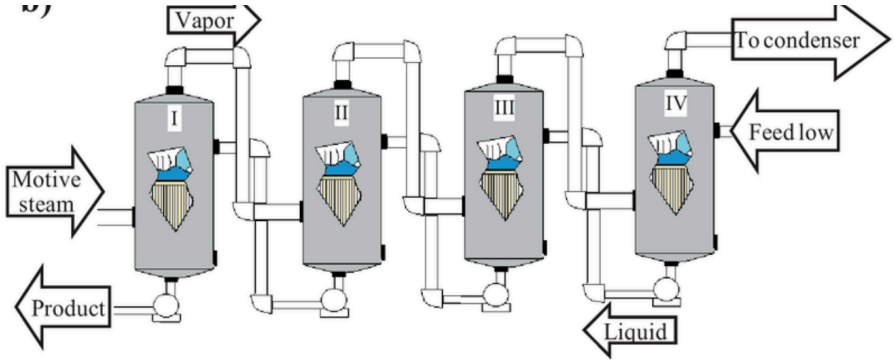

Figure 5.2: Counter-Current Flow

- Pros:
  - Our process requires incineration to 1200 ℃ after the multiple effect evaporator train, so this configuration (where the final product from the last effect is the hottest) decreases the heat input required for the incinerator to reach the necessary 1200 °C. This would reduce the amount of combustion fuel required (methane), which is one of our larger utility costs.
  - The concentrate is collected in the final, hottest effect. This results in a high viscosity that typically provides better heat transfer in practice, requiring smaller heat transfer area to achieve the same heat transfer rate.[5]
  - Additionally, the Log Mean Temperature Difference (LMTD) is typically much higher for counter-current flow vs. co-current flow at the same terminal temperature. This requires less area for heat exchange (to achieve the same heat transfer rate).[6]
  - Although caustic treatment was not used in this design iteration, it could potentially be implemented at a later date. This would require NaOH to be added prior to the first MEE in series, to be concentrated within the MEE train. Counter-current flow, in which the final MEE is the hottest, would allow the caustic treatment to be the most effective.
- Cons
  - Pumps are necessary to move liquid from one evaporator to the next because effects in series get progressively higher in pressure and temperature, potentially offset by the reduced necessary heat transfer area and reduced downstream fuel needs due to a higher terminal temperature outlet.

Although pumps are necessary in a counter-current configuration, the team believes the savings on methane and incinerator size lend themselves to choosing counter-current operation. Additionally, although this process doesn't involve caustic treatment, counter-current treatment would be significantly more efficient with this configuration. In future design streamlines and cost-saving initiatives, the team believes that it would be very feasible (and likely, result in increased overall cost savings) to implement NaOH caustic treatment for avermectin degradation.

To ensure the thoroughness of our design, the team tested each kind of evaporator that is manufactured in approximately the correct size for our process. This information was gathered from Turton et. al.[4] Early preliminary sizing estimates (from Weeks I-IV) indicated that our evaporators needed an area on the order of magnitude of 10-100 m^2. We found that there were four evaporator types that are manufactured with heat exchanger areas (A, m^2) that approximately match what we may need for our processes. These types are a.) agitated film, b.) short tube, c.) forced circulation and d.) falling film. See Figure 4.1 for the data that leads to this conclusion. Additionally, it was important to us to evaluate the economic repercussions of choosing a specific steam type (either mid-pressure steam or high-pressure steam), as well as the optimal number of evaporators in series.

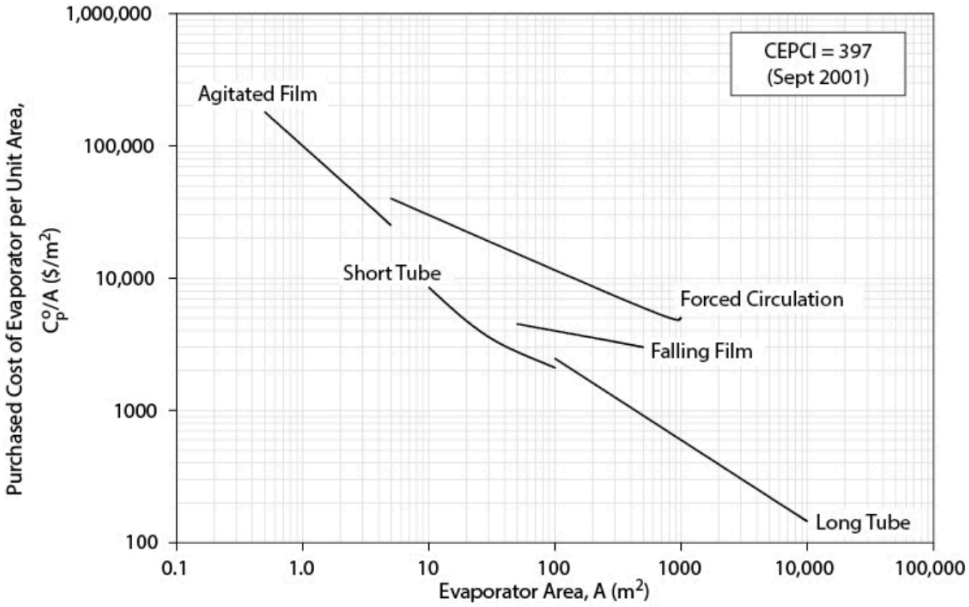

Figure 5.3: Costing/Production Curves for Evaporator Types

The first task was to perform initial sizing estimates for each type of evaporator. The following assumptions were made:

- ΔT remained constant between each multiple effect evaporator (from the initial steam temperature to 100 ℃), either high-pressure steam (HPS, 254 ℃) or mid-pressure steam (MPS, 184 ℃) is used,
- each evaporator in series has an equal heat duty (out of 4770 kW total), and the number of evaporators varied between 2 and 8 in series,
- And finally, each evaporator has an overall heat transfer coefficient of ~1700 W/m^2K.

The estimated overall heat transfer coefficient came from Perry's Chemical Engineering Handbook's estimation for U in evaporators with steam as the heating media.[7]

Note: In a single-effect simulation, the heat duty required to evaporate 190,000 kg/day (95%) of water of 4770 kW, so this total heat duty was chosen. See Weekly Report IV for further information on testing.

Using all of these stipulations, as well as the typical sizing equation (seen below), initial sizing estimates indicated that all evaporators would have between 15-35 m^2 of heat-transfer area. Referring to Figure 4.1 above, we note that this eliminates both agitated film (too small) or falling film (too big) from consideration.

$$A = \frac{Q}{U\Delta T}$$

EQN: 5.1

Where A represents heat transfer area, Q is the heat duty, and U is the overall heat transfer coefficient.

Note: To be clear, if the setup requires 2 evaporators in series and MPS is used as the initial heating media, the second evaporator will be held at 142 ℃ (ΔT, between heating media and liquid = 42℃) and the second evaporator will be held at 100 ℃ (ΔT still = 42℃). Because this setup requires ΔT to vary inversely with the number of evaporators (and thus Q), it causes all evaporators in series to be the same size. This was an intentional design choice for logistical and capital costing purposes.

Note: A maximum of 8 evaporators was chosen due to logistical constraints. Literature review (including Perry's Chemical Engineering handbook) indicates that, although more than 8 MEEs in series are *possible*, they are not often practical.

Even with the type of MEE train decided upon (counter-current flow being selected), there are still several MEE variables that can affect the overall price. The team was specifically interested in varying: a.) the evaporator type, within the realistic constraints described above, b.) the starting steam type (HPS vs. MPS), and c.) the number of evaporators in series. A JMP custom Design of Experiment (DoE) was created to analyze the effect significance of each of these variables on the total cost of the evaporators, necessary pumps, and utility costs, etc. See below for the JMP custom DoE setup.

	Starting Steam	Number of Evaporators	Evaporator Type	Cost
1	HPS	8	Forced Circulation	$18,684,633.54
2	LPS	8	Short Tube	$6,061,503.31
3	HPS	2	Forced Circulation	$7,282,845.37
4	LPS	2	Forced Circulation	$8,436,068.42
5	HPS	4	Short Tube	$3,859,665.45
6	LPS	3	Forced Circulation	$10,765,792.25
7	HPS	7	Short Tube	$5,033,459.56
8	HPS	8	Forced Circulation	$18,684,633.54
9	LPS	2	Short Tube	$3,667,967.31
10	HPS	2	Short Tube	$3,989,996.11
11	LPS	6	Forced Circulation	$19,207,663.04
12	LPS	7	Forced Circulation	$22,158,229.02
13	LPS	8	Short Tube	$6,061,503.31
14	HPS	4	Short Tube	$3,859,665.45

Figure 5.4: JMP Experimental Design

The table in Figure 4.2 has already been filled out with costing estimates for each design case. However, reaching this point required many calculations in Excel. The sheet will be attached to this report, titled "MEE Optimization." To make a true comparison between each case, capital costing needed to be completed for each evaporator in series and each pump required. Note that, because we have chosen a counter-current flow, pumps are needed to move the wastewater stream from lower to higher-pressure evaporators. The flow for each of these pumps, as well as the pressure they would need to operate at, was used to size and cost each pump for each evaporator in series (for each case). Evaporators were also sized and costed for each case, and utility costs were estimated based on the number of evaporators and the type of steam used. Note that, for this preliminary analysis, 2 evaporators were estimated to need 50% of the steam that 1 evaporator would need, etc.

Additionally, the method taught from CBE 480 was used to compare equipment alternatives via the use of a Net Present Value (NPV). The following equation was used.[4]

$$NPV =- (Capital\ Cost\ + (P/A) * Utility\ Cost)$$

EQ. 5.2

For this analysis a 10% interest rate over 10 years of operation was assumed (for the P/A factor). Using this factor for utilities allows us to account for the changing value of money over time when we do analysis on recurring operating costs. This factor is not needed for one-time costs that we can assume are occurring at the present (i.e., capital costs). Note that, in this equation, NPV is defined as negative. For clarity, we have defined NPV as a positive cash value, as seen in Figure 4.2 above in the "Cost" column.

The team wanted to see which variables had a significant impact on the total cost of the operation. To this end, JMP was used to calculate P-values, seen below in Figure 4.3.

Source	Logworth	PValue
Evaporator Type	4.041	0.00009
Number of Evaporators(2,8)	2.501	0.00316
Starting Steam	0.121	0.75679

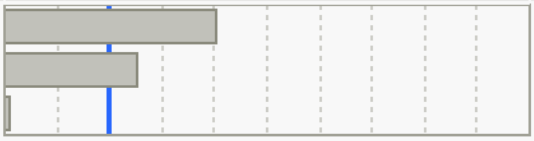

Figure 5.5: JMP Significance Results

As expected from the costing curves in Figure 4.1, short tube evaporators are significantly cheaper than forced circulation evaporators. Additionally, the number of evaporators has a significant impact on the price, although the directionality of this significance was not immediately apparent. Finally, HPS vs. LPS does not have a significant effect on costing.

Based on these results, the team was able to narrow down the search to determine which specific combination of steam and number of evaporators would be the most economical choice. A second, similar Excel sheet was created to find the NPV of 1-8 evaporators in series, each with a HPS and a MPS case. This came out to 16 trials. Note that all evaporators were assumed to be short tube evaporators, since we have already concluded that this is the cheapest option. The results of these trials are shown below. Note

that a single evaporator was tested as part of this plan, just to ensure that it wouldn't be more cost-effective than a multiple evaporator series.

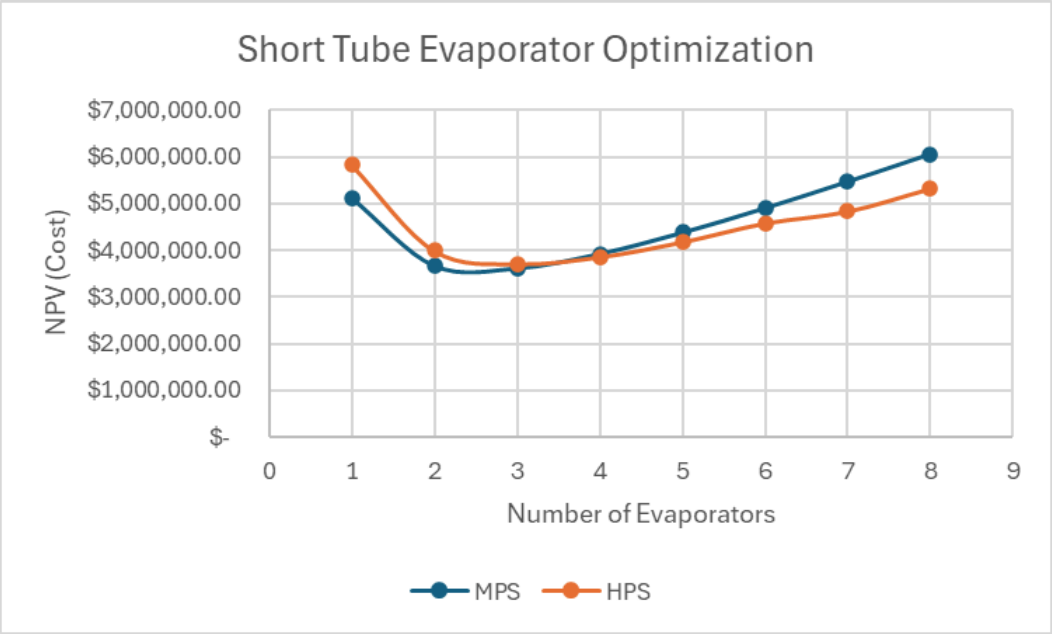

Figure 5.6: Short Tube Evaporator Case Costing

As indicated by the P-values from the JMP analysis, there is not much cost difference between using MPS vs. HPS. However, the minimum value on this graph lies when 2 evaporators in series are used with MPS. In this case, NPV = ~$3,600,000. This is beneficial not only because it's the cheapest option, but lower pressure (only 10 bar vs. 41 bar for HPS) results in significantly fewer safety concerns and accommodations that must be made to the process. In high pressure systems, which require special methods of construction, additional safety measures and equipment types are needed to avoid explosions. This results in higher capital and maintenance costs, and complicates the safety analysis associated with the process (e..g, HAZOP/FMEA analysis). Even if the HPS happened to be cheaper, the team would likely need to evaluate whether the decreased cost was worth the additional safety risk. Luckily, the NPV analysis supports our choice to use MPS as our heating media without the additional safety risk associated with HPS.

Note that although 10 bar is not "low" pressure, it is well within normal operating conditions for a chemical plant. The team will still analyze safety concerns associated with this operating pressure, but the risks associated will likely be much lower than if a higher pressure was necessary for the operation.

Following the above analysis, the team input two counter-current multiple effect evaporators in series into our process train (see attached Aspen file). As had been specified previously, the team still wanted the size (and thus, the heat duty, as all other variables remained the same in our previous analysis) of both evaporators to be the same. This was achieved using a Flowsheeting Design Specification, which will be discussed in further detail in the "Simulation" section (Section 5) of this report below.

Toluene Separation

A steam stripper was chosen as our method of choice for the actual separation of toluene out from the waste stream, being one of the most critical components responsible for removal of a key hazardous stream component. From early on in the project we recognized that it would be a challenging task–toluene is not only a Volatile Organic Compound (VOC) and HAP, but also only about 5% of it in our stream is actually in the aqueous phase. The other 95% is adsorbed to the surface of the cell debris in the stream. This meant that we couldn't simply rely on traditional liquid phase separation assumptions–we had to design the stripper to hit an aggressive <4 ppm toluene target in the bottoms product, in order to account for what the stripper doesn't readily strip (i.e., the toluene bound to the cell debris).

We started with some shortcut distillation tools in Aspen (DSTWU blocks), but they consistently gave us unrealistic results–sometimes even predicting fewer than one theoretical tray being needed to achieve the separation. That wasn't surprising in hindsight, given the dilute nature of the system and the fact that the striper is essentially removing trace organics from a mostly water feed (as much as ~98% water feed). So, instead we switched gears to using a RADFRAC distillation column block.and the NRTL property package, which gave us much better control over how the separation was simulated.

A brief mention on the difference between DWSTU and RADFRAC columns: DWSTU is what's considered a "shortcut model". Designing using a DWSTU block is a fast, simple method based on approximate correlations such as the Fenske-Underwood-Gilliland (FUG). It estimates the number of theoretical stages, minimum reflux, and recoveries all without performing full tray-by-tray calculations. DWSTU is an excellent tool for early-stage screening or for when just a rough idea of the column size or feasibility is satisfactory. However, it assumes constant molar overflow, ideal (or near ideal) behavior, binary-like separations, and no detailed energy or composition profiles. RADFRAC blocks on the other hand are considered Aspen Plus' rigorous distillation model. These block do full tray-by-tray mass and energy balances, support non-ideal mixtures (via NRTL, UNIQUAC, etc.) and do allow full customization of: Column internals (trays or packing), condenser/reboiler type, pressure and energy integration, side draws, heat duties and more options.

So, the team originally did try to make use of DWSTU blocks to get a quick handle on the feasibility of stripping toluene from the aqueous phase. But because the stream is extremely dilute and the system is highly non-ideal due to the organics and debris, DWSTU blocks would give nonphysical results such as predicting less than one theoretical stage. In order to develop realistic separation data and control over key parameters like column pressure, temperature profiles, and packing height, we switched to a RADFRAC block and the NRTL property method, which gave us a more accurate and reliable simulation of the steam stripping operation.

One of the major complications was how we would represent the toluene that was bound to the cell debris. Since Aspen cant' directly model this interaction, we were advised to use a calculator block as a workaround. The idea here was to scale the distillate flow to reflect the fact that, in the real world process, toluene bound to debris would slowly desorb and replenish the liquid-phase toluene over time. While this isn't a perfect physical model, it did allow us to capture the mass balance discrepancy in a way that accounted for the true toluene amount in the initial waste stream.

With the column framework in place, we established a flowsheeting design specification to hold the toluene composition in the bottoms stream at 4 ppm. From there we ran a sensitivity analysis on the steam flow rate to figure out the point where we achieved acceptable separation performance without sending a ton of excess water out the top in the distillate. Using the RADFRAC block, we imposed a flowsheeting Design Spec– 1) bottoms-phase toluene < 4 ppm–and allowed Aspen to vary the steam feed rate. It was also evident that once two theoretical ideal stages were present, additional stages would only marginally reduce the steam feed rate, so the tray count was then fixed at two while Aspen solve for the minimum steam flow rate. We ultimately landed on a steam feed rate of around 2.99 kg/min (997 L/min at 6 atm, 160 °C). Once the stripping was finished, we still had to collect the toluene from the column distillate. The vaporized toluene and steam in the distillate was condensed in an external condenser then sent to a liquid-liquid decanter, where we could readily separate the toluene rich phase from the water.

While RADFRAC blocks do have the ability to model the reboiler and condenser as integrated in their design, the team decided to design the condenser externally to allow us to setup a flowsheeting design spec to require the distillate be condensed fully to liquid before reaching the decanter. The reboiler was not the main source of energy/heat to separate the toluene, rather it was the steam stream that provided the energy to drive the separation as well as carry the vaporized toluene into the distillate.

Steam Stripper Design & Sizing

Having decided to use a steam stripping design to achieve the required toluene separation, the question of the column internals was next. The actual separation desired is not difficult to achieve in concept or in practice, the real difficulty was in the selection of appropriate column internal that best suited the waste stream being processed. Steam stripping operations typically benefit from low pressure and high temperature operations, so a desire to operate near atmospheric pressure was understood. The choice between trayed or packed internals is often the first decision made in separation tower design, and the team took into consideration the characteristics of the waste stream we were tasked with processing. A packed column internals design was selected over trays as they typically result in a lower pressure drop (ideal for near-atmospheric operation), higher interfacial area (beneficial for dilute toluene in water), lower liquid holdup requirements (not a problem since toluene volatilizes readily) and provides better handling of high flow rates.

With packing chosen over trays for internals, the next step was to select a packing material and two main types were investigated: Grid-Style packing and Random packing using Pall Rings. Grid-style packing–specifically Mellagrid packing manufactured by Sulzer–was ultimately selected for our final design. This decision was driven by the high-fouling potential of the waste stream, which not only contains organic residues from the fermentation process that produced the avermectin, but also approximately ~0.1 wt% cell debris. Grid-style packing is engineered to be resistant to fouling due to their open structure, mechanical robustness, and ease of cleaning or replacement in industrial service. Their relatively wide channels minimize the risk of clogging compared to conventional random packing or structured corrugated sheets.

To translate the two ideal theoretical stages into a packing height, we applied the Height Equivalent to a Theoretical Plate (HETP) approach as described in Turton. In the design, a conservative effective HETP of 0.1 plates per meter was assumed for Mellagrid under steam stripping conditions, influenced in large part by the HETP found for Mellapak products as shown below in Figure 5.7. Since simulation results (from RADFRAC and NRTL thermodynamics) indicated that 2 ideal stages were sufficient to meet the aggressive <4 ppm toluene spec, a total of 20 meters of packing was specified (each stage requiring a HETP of 10 meters of packing). Column diameter was found to be 0.396 meters based on flooding and vapor traffic constraints for the steam-to-waste ratio found through Aspen Plus' interactive sizing. Accounting for disengagement space, support structures, vapor-liquid distributors and flanging, the total column height was found to be 39.15 m. The random Pall Rings were also evaluated using Aspen hydraulic plots for benchmarking but were not selected due to their higher fouling risk, more liquid holdup, and reduced ease of cleaning.

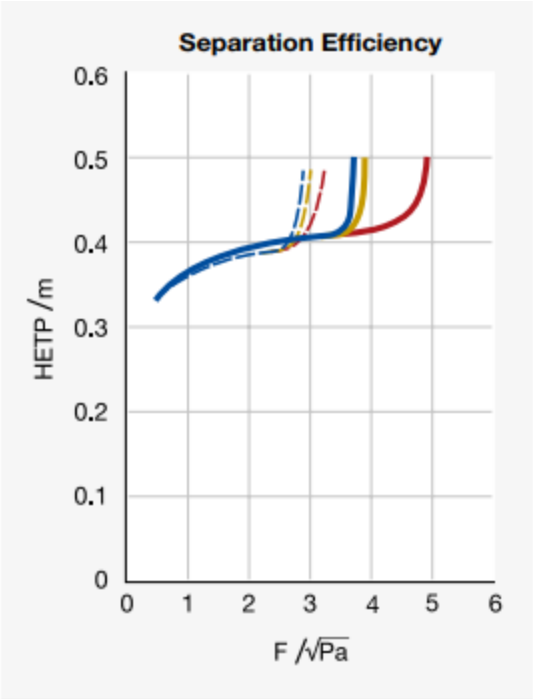

Figure 5.7: HETP Separation Efficiency of Mellapak

Metal was chosen as the packing material of construction for thermal durability, compatibility with steam and longevity under high-temperature services. The final bare module cost of the optimized steam stripper was estimated to be ~$59,103.00 including shell, Melagrid internals, support structures and distributors.

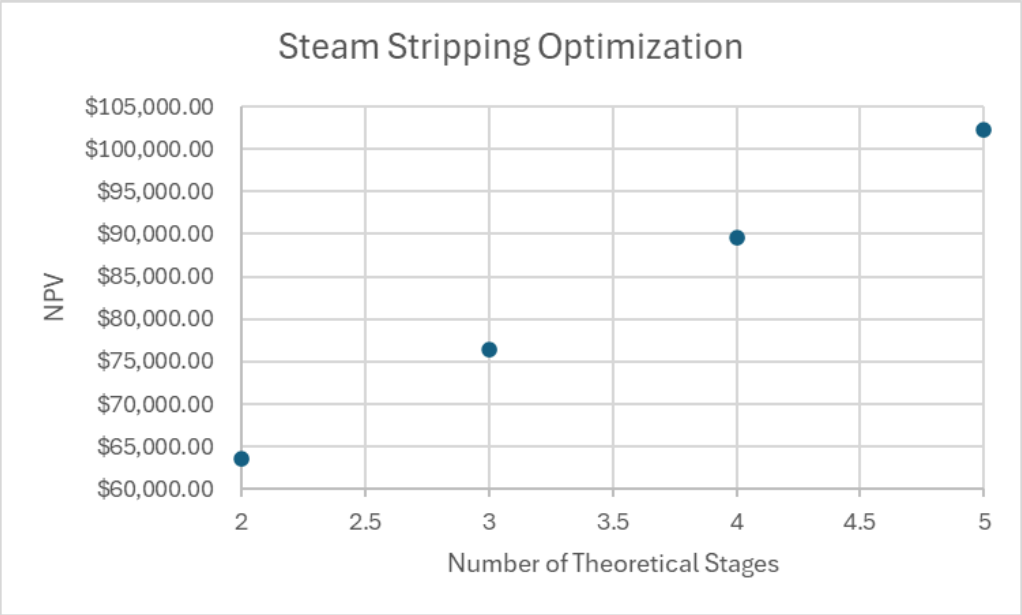

Figure 5.8: Steam Stripping Costing Optimization

Incineration

Background material for this project states that incineration infrastructure already exists within our industrial plant. However, it's necessary for the team to cost the utilities (fuel and air) to heat both the process stream and air utility to 1200 degrees Celsius. Additionally, enough air should be provided that the exhaust stream (Stream 23) is 6% v/v oxygen.

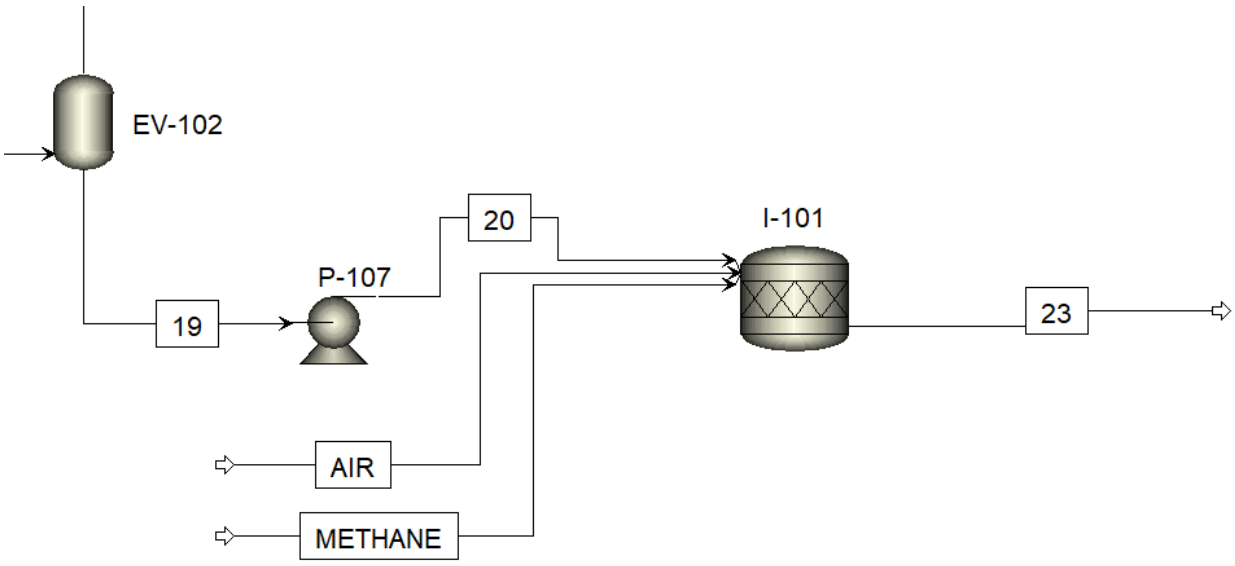

Figure 5.9: EV-102 Setup

In order to complete this analysis, several assumptions must be made. Firstly, it is assumed that the provided incinerator operates isobarically at atmospheric pressure. Secondly, assume that the incinerator

operates adiabatically (no heat loss) and all heat generated to heat the components to 1200 degrees Celsius will come from the combustion of natural gas, not external utilities.

The team has chosen to use natural gas (methane) as the fuel source to heat the incinerator. This is for a few reasons; firstly, it has high energy content, which allows us to use relatively small amounts of fuel to achieve large amounts of heat energy. Although it is more expensive than coal, the use of natural gas is also forward-looking. Many companies and countries highly regulate the burning of coal and are trying to transition to cleaner energy sources. As a waste treatment plant, we want to avoid the emissions of any additional pollutants, like sulfur oxides and nitrogen oxides. Additionally, per Turton, et. al., natural gas costs are around $3.16 / GJ, which is very reasonable. For reference, coal is listed as $2.04 / GJ.[4]

The combustion reaction for natural gas is as follows:

$$\mathbf{CH_4 + 2O_2 \rightarrow CO_2 + 2H_2O}$$
EQN 5.3

An Aspen simulation was used to vary air and methane feed flowrates to a stoichiometric reactor, as seen below in Figure 5.2. Methane was designated as a gaseous streams at standard conditions (25 degrees C and 1 atm), while air was assumed to be at the standard shop pressure of 3.3 barg, per Turton et. al.[4] Note that air was specified as a 79 mole percent N_2 and 21 mole percent O_2 stream, while other common components found in air were assumed to be negligible.

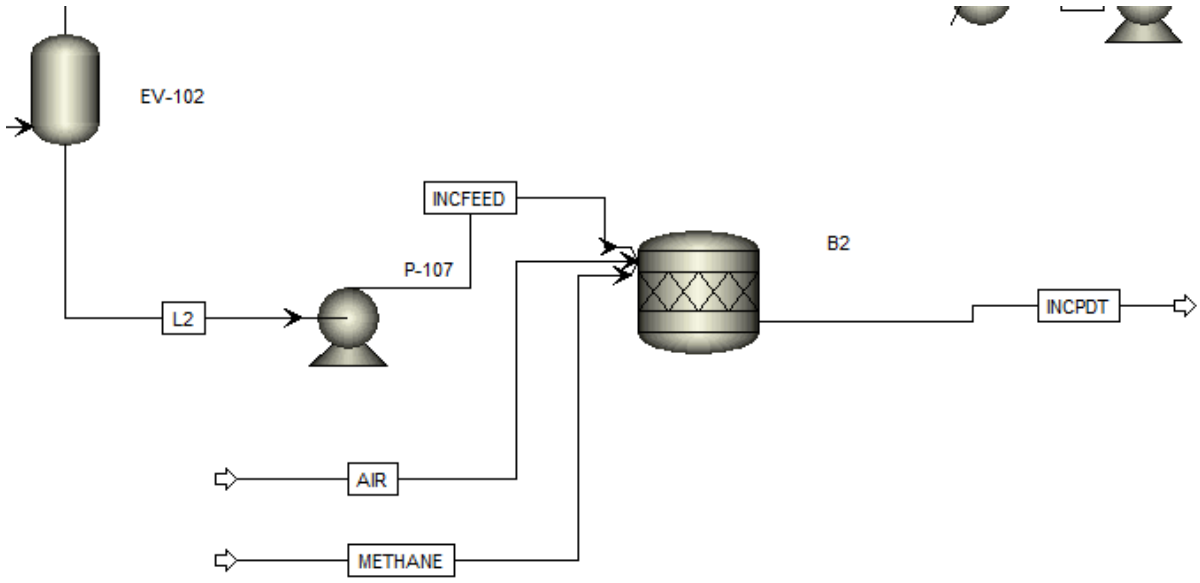

Figure 5.10: Incineration PFD

Using trial and error to acheive an outlet temperature of 1200 ℃ and 6% v/v oxygen, it was determined that the optimal air feed was 1839 kg/hr and methane feed was 72 kg/hr. The Aspen file for the overall process will be attached to this report, but these conditions result in an outlet temperature of 1200 ℃ and a volume percent of 6.0%.

Using the values from Turton, et. al and assuming 340 operational days/year, utility costs for both methane and air can be calculated. Note that the methane feed is already at standard conditions (1 atm, 25 degrees C).

Air:

$$0.50\ \frac{dollars}{100\ m3} * \frac{366.125\ m3}{hr} * \frac{24\ hr}{day} * \frac{340\ days}{yr} = \$14,932\ /year$$

Methane:

$$0.1119\ \frac{dollars}{std\ m^3} * \frac{109.373\ m^3}{hr} * \frac{24\ hr}{day} * \frac{340\ days}{yr} = \$99,868\ /year$$

This utility usage and costing seems perfectly reasonable for a wastewater process of our size. Note that the design specifications for air and methane flowrates will be discussed in the following section as well.

Chemical Oxygen Demand (COD) Treatment

The final stream leaving wastewater treatment (labeled Stream 15 below) is almost entirely pure. Using the steam stripper to remove toluene and the multiple effect evaporators (MEEs) to remove avermectin prove to be high-fidelity separation techniques. Avermectin is reduced to < 1 part per trillion, while toluene is reduced to ~85 ppm. Since we did not complete caustic treatment, we do not need to pH treat the wastewater stream before it is exhausted to the river.

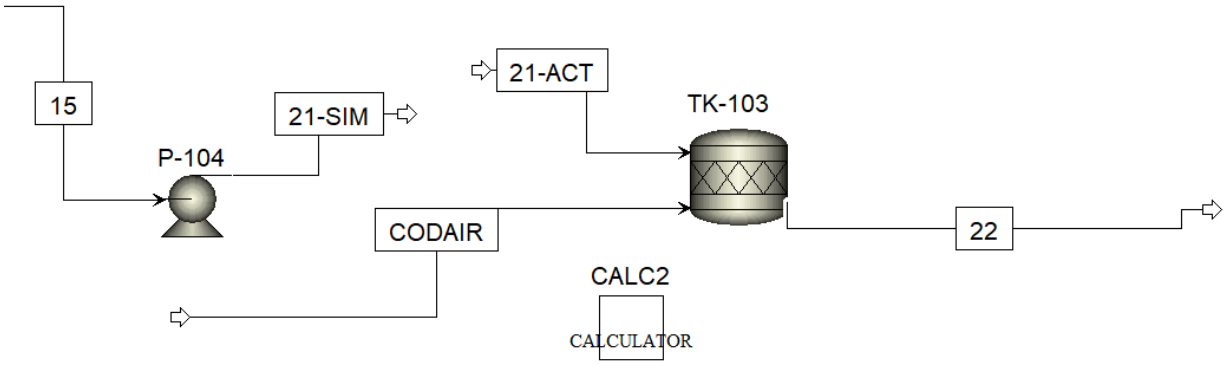

Figure 5.11: Chemical Oxygen Demand PFD

However, the team has also been tasked with treating the wastewater for chemical oxygen demand (COD). Luckily, a unique advantage of using the MEE approach is that many of the organic species, like glucose, polyethylene glycol, and triglyerides, preferentially flash into the liquid phase, which is then incinerated. Thus, we have significantly less COD treatment to complete than if we had employed another technique (such as membrane separation) that would not have preferentially separated these organic species. Note that, although we are currently hitting the toluene specification from the EPA (<100 ppm), the team was asked to complete COD treatment for toluene as well, since the microbes that work to

complete COD treatment cannot differentiate between glucose, toluene, or other biological molecules. Also note that, to maintain the fidelity of our overall mass balance, a calculator block was used immediately before COD treatment. The rest of our simulation models only aqueous toluene (5 ppm), but COD treatment will need to model the breakdown of *all* toluene, including that which is still adsorbed to cell debris, totaling 100 ppm. This change increases the total stream mass by less than 15 kg/day and will not significantly impact any upstream unit operations.

Per the Aspen simulation, our outlet stream (Stream 15) only contains 9.85 g/day of polyethylene glycol, 9.79 g/day of glucose, and 16.47 kg/day of toluene. All other organic species that originated from the avermectin industrial process are exhausted at < 1 mg/day and will be ignored in the following analysis. Despite the small quantities of polyethylene glycol, toluene, and glucose remaining, COD treatment for these species will still be costed for completeness. Again, note that although toluene is well below the EPA specification of 100 ppm, it is still necessary to provide sufficient oxygen to treat this component.

In order to fully reduce the chemical oxygen demand of a species, it must be reacted all the way to CO_2. Otherwise, as the organic species are reduced in nature, they will take up oxygen from the surrounding environment, reducing the dissolved oxygen percentage in the river effluent and harming wildlife. The balanced chemical reactions to treat glucose, toluene, and polyethylene glycol are shown below:

Glucose:

$$C_6H_{12}O_6 + 6O_2 \rightarrow 6CO_2 + 6H_2O$$

EQN: 5.4

Toluene:

$$C_7H_8 + 9O_2 \rightarrow 7CO_2 + 4H_2O$$

EQN: 5.5

Polyethylene glycol (PEG) 2000 refers to a molecule that demonstrates the following repeating structure: $H(OCH_2CH_2)_nOH$. However, the 2000 indicates the approximate molecular weight of the specific PEG. Using the known molecular weights for the atomic species H, O, and C, we find that n = 45. Therefore, the molecular species present is $H(OCH_2CH_2)_{45}OH$, which could equivalently be written as $C_{90}H_{182}O_{46}$.

Polyethylene glycol:

$$2C_{90}H_{182}O_{46}. + 225O_2 \rightarrow 180CO_2 + 182H_2O$$

EQN 5.6

Using Aspen, we can model this COD treatment as a RStoich reactor, using the stoichiometric relations described above, with PEG, glucose, and toluene all reacting to completion. See above for the PFD representation of this process. Note that a larger portion of this the PFD is shown for clarity of process flow, but the COD treatment is happening in TK-103. Air at 3.3 bar and 25 degrees C (standard shop conditions, per Turton et. al), is fed to the reactor to provide the necessary oxygen.

Using a Flowsheeting Design Specification, Aspen varied the oxygen flowrate of the CODAIR stream in order to result in negligible oxygen composition in the “CLEANWTR” stream. The resulting flowrate of

compressed air is 0.0236 cuft/sec, or 0.0401 m^3/min. The costing for this utility is worked out below, using the values from Turton et. al.

$$0.50\ \frac{dollars}{100\ m3} * \frac{2.409\ m3}{hr} * \frac{24\ hr}{day} * \frac{340\ days}{yr} = \$98.29\ /year$$

The design specification for air flowrate will be discussed in the following section as well.

Simulation

The Aspen simulation included several Flowsheeting Design Specifications to reach specified process goals. Specifically, Flowsheeting Design Specifications are used in our simulation to reach the following requirements/goals: fidelity of toluene separation, appropriate utility feed for COD reduction, efficient heat integration, necessary incinerator output temperature, and incinerator exhaust composition. Each of these specifications and the associated results will be discussed in detail in the following section.

Toluene Separation

The first Flowsheeting Design Specification sets the bottoms concentration of the toluene separation operation (steam stripper) at 4 ppm. The EPA regulation states that the toluene limit to be exhausted to wastewater is 100 ppm, but note that only 5% of the toluene in our system is in the aqueous phase, while the other 95% is bound to cell debris. Therefore, we need to decrease the toluene concentration *in aqueous phase* to a maximum of 5 ppm to account for the 95% of toluene not accounted for by the steam stripping operation. However, note that this mass-balance discrepancy is accounted for in subsequent process operations. The team chose to include a safety factor of 20%, which leads to the 4 ppm design specification. As noted above, the steam stripper height, width, and number of theoretical stages (and thus the packing height) was specified in a previous analysis. Therefore, the team chose to vary the steam flowrate (Stream 5) to change the toluene composition of the bottoms stream (Stream 10). As we know, increasing steam flowrate causes increased toluene separation (or, in other words, decreased toluene in the bottoms product). Therefore, we're able to set a design specification that sets the toluene bottoms composition at a desired value (4 ppm) by only varying the flowrate of the utility stream that feeds steam to the tower.

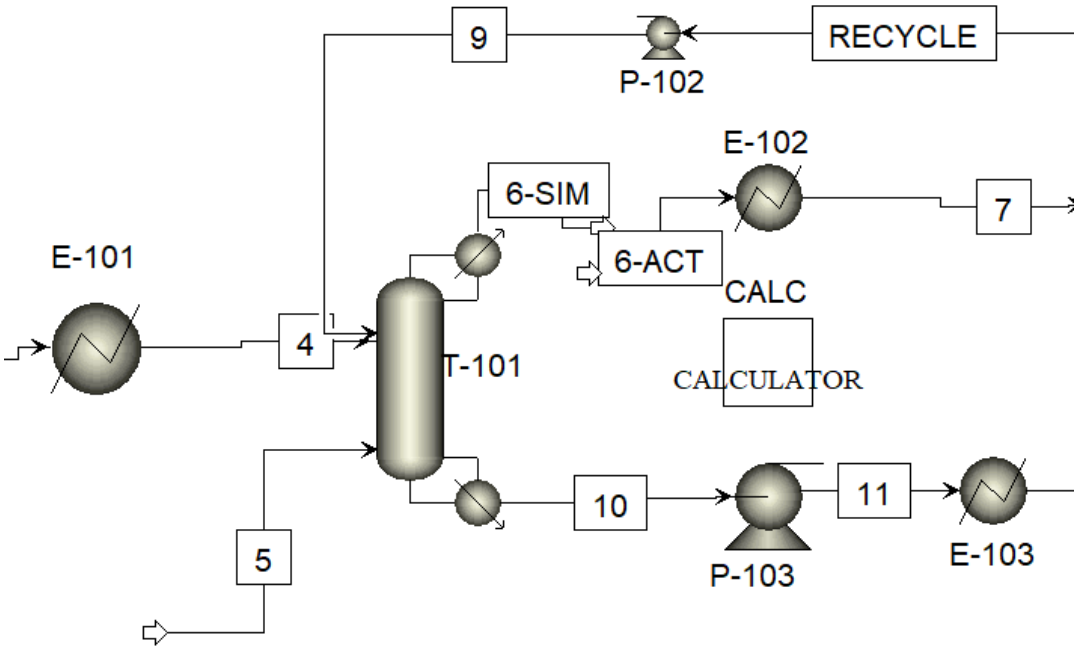

Figure 5.12: Steam Stripper PFD

Utility Feed for COD Reduction

The COD tanks (and necessary microbial feed) are already at the plant for which the avermectin waste treatment process was designed. However, the team was tasked with finding the necessary flowrate and cost of utility for this process. In COD treatment, microbes aerobically oxidize organic compounds to CO_2. This prevents the organics from being oxizided when exhausted to wastewater. To cost the necessary utility, which will be compressed air, the Aspen simulation includes an RStoich reactor that represents the pre-existing COD tanks. This reactor uses stoichiometric reaction equivalents to take all appreciable organics in the system (glucose, toluene, and polyethylene glycol) and react them with oxygen to produce carbon dioxide and water. Each reaction is specified to react the organic compound to completion. Note that thus process is described in-detail in the above "Chemical Oxygen Demand Treatment" section.

The design specification alters the flowrate of the compressed air utility stream to meet the oxygen demand exactly. This allows the team to visualize exactly how much compressed air is needed to push these reactions to completion, ensuring that organic compounds aren't being released to the environment to be oxizided, which would remove necessary oxygen from the aquatic ecosystem. As noted above, the resultant airflow is 0.0236 cuft/sec, at standard shop pressure (3.3 bar). This process is represented in Figure 5.11 above.

Efficient Heat Integration

Our process includes two counter-current multiple effect evaporators (MEEs) in series. In the real-world, the steam produced from the second (hotter, higher pressure) evaporator in series would be used to heat

the first (cooler, lower pressure) evaporator in series. This task is not easily accomplished in Aspen, so the team used a workaround method to ensure that the size and heat duty of both evaporators are balanced. First, a design specification was set on the second MEE: the flowrate of water coming out in the liquid stream (Stream 19) was specified to be 4000 kg/day. This allows the amount of expected suspended solids to stay below 5%, which is a rule-of-thumb limit on the maximum amount of suspended solids to maintain ease of pumping/transport. Then, a heat exchanger (labled COND-SIM in the simulation) was specified to condense the exiting steam from EV-102, but keep it at the same temperature. In the real world, this would be occurring in the first evaporator. However, we can then specify that the heat duty of evaporation for EV-101 be the same as the simulated heat duty of COND-SIM. In this manner, we can ensure that EV-101 is treated as if the steam from EV-102 is being used as its heat input. Via trial and error, the team was able to meet the 4000 kg/day effluent requirement while also making the size of both evaporators the same (a requirement from above MEE specifications).

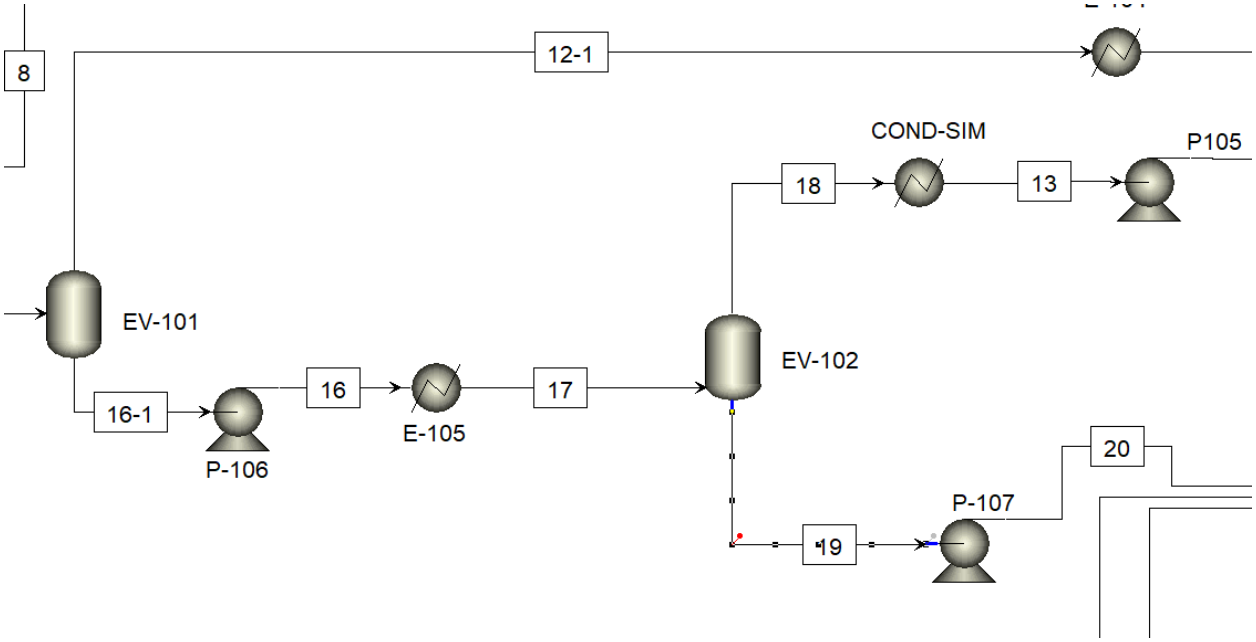

Figure 5.13: Efficient Heat Integration PFD

After this trial-and-error approach was completed, both evaporators had a heat duty of approximately 2550 kW. This value allows both the 4000 kg/day of water in effluent and allows consistently-sized reactors.

Incinerator Outlet Temperature

The incinerator is an integral piece of the avermectin treatment process. In order to achieve high-fidelity avermectin degradation (without the use of caustic treatment), the avermectin-containing solution needs to be heated to 1200℃. This information was called out specifically in the background information. Therefore, the team needed a flowsheeting design specification to ensure that the incinerator outlet was heated to at least 1200℃. The incinerator itself doesn't provide heat, but rather the combustion of methane and oxygen to carbon dioxide and water (a highly exothermic reaction) provides the heat to achieve this temperature specification. The reactor is specified to react methane to completion. Therefore,

the amount of methane supplied to the reactor determines the outlet temperature of the incinerator. As specified above, the resultant methane flow is 72 kg/hr.

Note that the oxygen feed to the reactor also determines how much methane can react, but this will be discussed momentarily.

Incinerator Outlet Composition

The background information for this project specifies that the outlet composition of the incinerator exhaust must be 6% v/v oxygen. Because the gases in this system can be approximated as ideal (low pressure, high temperature conditions are met), the mole percent oxygen in the exhaust can be approximated as equal to the volume percent in the exhaust. In other words, oxygen needs to be supplied in excess to complete the necessary combustion reaction specified in the incinerator. To meet this specified process requirement, a flowsheeting design specification can be created to set the outlet composition of the incinerator to 6 mole percent oxygen by varying the oxygen flowrate of the compressed air stream. Note that by honing in on a good starting value, the compressed air is still composed of 21% oxygen and 79% oxygen even with variation from the design specification. Because the incinerator is specified to react methane to completion and the methane flowrate is varied by the aforementioned Incinerator Outlet Temperature design specification, inputting a variation on oxygen flowrate allows both the outlet composition *and* temperature incinerator specifications to be met. The resultant flow of compressed air is 1839 kg/hr.

Note on Design Specifications

Including flowsheeting design specifications allows the simulation to adjust automatically to any perturbations or stream changes made within the simulation, ensuring that critical output requirements are produced consistently. This ensures that our process is safe and that we're meeting the specified output stipulations (toluene concentration, stoichiometric utility feed, efficient heat integration, incinerator output temperature, exhaust composition)

SECTION 6: SIZING AND COSTING

SIZING

Exchangers

Equations 2.1 - 2.4 are used to size a heat exchanger.

$$Q = UA\Delta T_{lm}$$

EQ 6.1

$$\Delta T_{lm} = \frac{\Delta T_1 - \Delta T_2}{ln(\Delta T_1 / \Delta T_2)}$$

EQ 6.2

$$\Delta T_1 = T_{h,i} - T_{c,o}$$

EQ 6.3

$$\Delta T_2 = T_{h,o} - T_{c,i}$$

EQ 6.4

E-101[4]
Type: Floating Head
Q = 2184000 Btu/hr
U = 50 Btu/hr ft^2 F
$\Delta T_{lm} = 103.7$ F
A = 421.2 ft^2

E-102[4]
Type: Double Pipe
Q = 563148 Btu/hr
U = 150 Btu/hr ft^2 F
$\Delta T_{lm} = 99.6$ F
A = 37.7 ft^2

E-103[4]
Type: Floating Head
Q = 188352 Btu/hr
U = 50 Btu/hr ft^2 F
$\Delta T_{lm} = 33.8$ F
A = 111.6 ft^2

E-104[4]
Type: Floating Head
Q = 10040000 Btu/hr
U = 150 Btu/hr ft^2 F
ΔT_{lm} = 39.8 F
A = 1680.0 ft^2

E-105[4]
Type: Floating Head
Q = 910116 Btu/hr
U = 50 Btu/hr ft^2 F
ΔT_{lm} = 108.3 F
A = 168.0 ft^2

E-106[4]
Type: Floating Head
Q = 1930000 Btu/hr
U = 50 Btu/hr ft^2 F
ΔT_{lm} = 57.1 F
A = 676.0 ft^2

The area's will be used to cost all the heat exchangers.

Pumps
Equation 2.5 is used to size a pump.

$$Power = \frac{(Mass\ Flow\ Rate) * (\Delta Pressure)}{(Density) * (Efficiency)}$$

EQ. 6.5

P101-A/B
ΔP = 68948 Pa
Ɛ = 0.6
W_s = 0.266 kW

P102-A/B
ΔP = 68948 Pa
Ɛ = 0.6
W_s = 0.008 kW

P103-A/B
ΔP = 68947 Pa
Ɛ = 0.6
W_s = 0.301 kW

P104-A/B
ΔP = 68947 Pa
Ɛ = 0.6
W_s = 0.275 kW

P105-A/B
ΔP = 68947
Ɛ = 0.6
W_s = 2.522 kW

P106-A/B
ΔP = 328140 Pa
Ɛ = 0.6
W_s = 0.745 kW

P107-A/B
ΔP = 68948 Pa
Ɛ = 0.6
W_s = 0.007 kW

The wattage will be used to cost all the pumps

Vessels, Storage Tanks, and Evaporators
The Following Equations are used to size vessels and storage tanks

$$Volume = (Volumetric\ Flow\ Rate) * (Residence\ Time)$$
EQ. 6.6

$$Volume = \frac{(Mass\ Flow\ Rate) * (Residence\ Time)}{(Density)}$$
EQ. 6.7

$$Volume = (\frac{D}{2})^2 * (H) * (\pi)$$
EQ. 6.8

TK-101
Type: Floating Roof
Res Time: 24 hr
V: 35314.6 ft^3
% Full: 80%
Vact: 44143.3 ft3
H: 39 ft
D: 39 ft

TK-102
Type: Fixed Roof
Res Time: 168 hr
V: 1249.8 ft^3
% Full: 80%
V_{act}: 1562.2 ft^3
H: 13 ft
D: 13 ft

V-101
Orientation: Horizontal
Res Time: 0.083 hr
V: 1.17 ft^3
% Full: 60%
V_{act}: 1.96 ft^3
H: 3 ft
D: 1 ft

EV-101
Orientation: Vertical
Res Time: 0.083 hr
V: 27.4 ft^3
% Full: 60%
V_{act}: 45.6 ft^3
H: 9 ft
D: 3 ft

EV-102
Orientation: Vertical
Res Time: 0.083 hr
V: 15.0 ft^3
% Full: 60%
V_{act}: 25.0 ft^3
H: 7 ft

D: 2.3 ft
The height and diameter and orientation/type will be used to cost each type of vessel, tank, or evaporator

Sizing for the steam stripper was done in aspen and can be found in **Section 5**

Costing

After appropriately sizing each piece of equipment, the crucial design parameters (pressure, temperature, and material of construction) were used to find a bare module cost for each piece of equipment in the base case. This was done with a combination of the CAPCOST program and hand calculations done with the resources from Analysis, Synthesis, and Design of Chemical Processes (5th Edition). For the purpose of costing and after research into our process, the material of construction for all equipment was assumed to be carbon steel. Additionally, design pressure was used as a key parameter in all sizing calculations and was assumed to be (1.1)*(Operating Pressure). Similarly, operating temperatures were scaled by an additional 30 F to arrive at a suitable temperature rating.

Costing using the CAPCOST PROGRAM:
The CAPCOST Program was used to sized Exchangers (E-101, E-102, E-103, E-104, E-105), Pumps (P-101 A/B, P-102 A/B, P-103 A/B, P-104 A/B, P-105 A/B, P-106A/B, P-107A/B), and Vessels/Evaporators (V-101, EV-101, EV-102). For all calculations in the program, the CEPCI was set as 798.8.

Exchangers
To size exchangers, the following parameters were specified in CAPCOST:

a. Exchanger Type
b. Exchanger Area (m^2 or ft^2)
c. Pressure (Shell/Tube) = (1.1)*(Operating Pressure) (psig or barg)
d. Construction Material (All stainless steel)

Exchangers E-101, E-103, E-104, and E-105 were assumed to be Shell and tube (floating heat) exchangers. E-101 was assumed to be a double pipe exchangers due to its smaller size. All areas were taken from sizing calculations and may be further investigated in the previous section that addresses sizing. The bare module cost for each exchanger was found, totaling $372,730.

Table 6.1: Bare Module Cost of Exchangers

Exchangers	Type of Exchanger	Shell Pressure (barg)	Tube Pressure (barg)	MOC	Area (m^2)	Purchased Equipment Cost	Bare Module Cost
E-101	Floating Head	1.1	1.7	Carbon Steel / Carbon Steel	39.2	$ 19,200	$ 63,300
E-102	Double Pipe		5	Carbon Steel / Carbon Steel	3.5	$ 2,220	$ 9,930
E-103	Floating Head	5	0.781	Carbon Steel / Carbon Steel	10.4	$ 22,100	$ 64,900

E-104	Floating Head	5	1.09	Carbon Steel / Carbon Steel	156	$ 23,900	$ 103,000
E-105	Floating Head	5	3.51	Carbon Steel / Carbon Steel	15.6	$ 19,800	$ 61,100
E-106	Floating Head	5	4.84	Carbon Steel / Carbon Steel	62.8	$ 21,400	$ 70,500

Pumps

To size pumps the following parameters were specified into CAPCOST:

a. Pump Type
b. Shaft Power (kW)
c. Discharge Pressure = (1.1)*(Operating Pressure) (psig or barg)
d. Construction Material (All stainless steel)
e. Number of Spares (1 for all)

All pumps (P-101 A/B, P-102 A/B, P-103 A/B, P-104 A/B, P-105 A/B, P-106A/B, P-107A/B, P-108A/B) were considered cetrifugal pumps. All shaft powers were taken from sizing calculations and may be further investigated in the previous section that addresses sizing. It is important to note that the shaft power for P-101, P-102, P-103, P-104, P-106, and P-107 was considerably low, beneath the 1kW minimum used for costing. Thus, these pumps were costed as pumps with 1kW of capacity. Their true utilization is reflected in utility costs. The bare module cost for each pump was found, the total for all pumps being $138,700.

Table 6.2: Bare Module Cost of Pumps

Pumps (with drives)	Pump Type	Power (kilowatts)	# Spares	MOC	Discharge Pressure (barg)	Purchased Equipment Cost	Bare Module Cost
P-101	Centrifugal	1	1	Carbon Steel	0.71	$ 4,900	$ 19,500
P-102	Centrifugal	1	1	Carbon Steel	0.689	$ 4,900	$ 19,500
P-103	Centrifugal	1	1	Carbon Steel	0.71	$ 4,900	$ 19,500
P-104	Centrifugal	1	1	Carbon Steel	1.68	$ 4,900	$ 19,500
P-105	Centrifugal	2.52	1	Carbon Steel	3.28	$ 5,450	$ 21,700
P-106	Centrifugal	1	1	Carbon Steel	3.19	$ 4,900	$ 19,500
P-107	Centrifugal	1	1	Carbon Steel	3.22	$ 4,900	$ 19,500

Tank, Vessels, and Evaporators

To size the tanks, vessels, and evaporators (T-101, T-102, V-101, EV-101, EV-102) the following parameters were specified into CAPCOST:

a. Vessel Type (Vertical/Horizontal)
b. Length/Height (m or ft)
c. Diameter (m or ft)
d. Maximum Pressure = (1.1)*(Operating Pressure) (psig or barg)
e. Construction Material (All stainless steel)

The decanter was assumed to be a horizontal vessels. All lengths and diameters were taken from sizing calculations and may be further investigated in the previous section on sizing. The minimum volume for a vessel was 10 ft^3 therefore this was the volume assumed for our vessel since the actual size fell under this volume. The evaporators were also sized as if they were vessels but vertical in orientation. The bare module cost for the vessels, tanks, and evaporators was $299,880.

Table 6.3: Bare Module Cost of Evaporators, Vessels, Tank

Vessels	Orientation	Length/Height (meters)	Diameter (meters)	MOC	Pressure (barg)	Purchased Equipment Cost	Bare Module Cost
V-101	Horizontal	1.83	0.61	Carbon Steel	0.227	$ 2,890	$ 8,680
EV-101	Vertical	2.74	0.914	Carbon Steel	0.71	$ 4,160	$ 16,900
EV-102	Vertical	2.13	0.701	Carbon Steel	3.19	$ 2,890	$ 11,700
TK-101	Floating Roof	11.9	11.9	Carbon Steel	0	$ 196,000	$ 216,000
TK-102	Fixed Roof	3.96	3.96	Carbon Steel	0	$ 42,300	$ 46,600

Steam Stripper

To size the Steam Stripping tower (T-101), the following parameters were considered:

a. Stage Count (Ideal stages)
b. Height (meters)
c. Diameter (Sized using Interactive Sizing from Aspen)
d. Tower volume (calculated volume of a cylinder)
e. Steam Flow Rate (determined by Aspen's Flowsheeting Design Spec)
f. Steam Cost ($/kg)
g. Purchased Equipment Cost[4]

Verification was done with Aspen's packing hydraulics (the Billet-Schultes correlation, 50% flood) resulted in the 0.396 m Internal Diameter required for the two ideal trays.

Table 6.4: Bare Module Cost of Tower

Towers	Description	Height (m)	Diameter (meters)	MOC	Pressure (barg)	Purchased Equipment Cost	Packing Type	Bare Module Cost
T-101	Packed Tower	39.15	0.396	Carbon Steel	1	$ 2,890	Stainless Steel Grid	$ 59,103

$$Z_{tot} = N_{TP} \times HETP + h_{distrib} + h_{collector/redistrib} + h_{VL\ diseng.} + h_{sump} + h_{nozzle} + h_{contingency}$$

EQ 6.9

Table 6.5: Equation 6.9 Symbol Definitions

		Value Used	Source
NTP	Ideal Stages from Aspen Design Spec	2	Aspen Plus V14
HETP	Height per stage for Mellagrid	10 m per stage^{-1}	Source
$h_{distrib}$	Top Liquid Distributor	0.90 m	Turton
$h_{collector/redistrib}$	Mid-bed redistributor+bottom collector	1.80 m	Turton
$h_{VL\ diseng.}$	Vapor-Liquid Disengagement Zones (Top + Bottom)	8.00 m	Turton
h_{sump}	Liquid holdup for level control	1.00 m	Turton
h_{nozzle}	Inlet/outlet nozzle clearance (0.25 m each x 4)	1.00 m	Turton
$h_{contingency}$	Packing support, instrument skirts, fouling margin (~25 % of bed)	6.50 m	Towler & Sinnott
Total Z_{tot}	20 m + 19.15 m = 39.15 m	-	Calculated total

Result: A 0.396 m ID × 39.15 m packed steam-stripper meets the <4 ppm toluene bottoms target.

SECTION 7: HSE

Determining Tank Flammability

When determining if a floating head tank is needed, the following equations can be used to solve for the mole fractions in the vapor phase above the liquid.

$$log_{10}(P) = A - \frac{B}{T+C}$$

EQ. 7.1

It is important to note that the feed stream entering TK-101 is a non–ideal solution. Therefore a *modified* Raoult's law must be employed. This can be seen in equation 7.2, where gamma is an activity coefficient representing a deviation from the ideal.

$$y_i = x_i \frac{\gamma_i P^{SAT}}{P_i}$$

EQ. 7.2

The team used Aspen to determine the Activity Coefficient by conducting a P-x-y analysis on a binary system of toluene and water. For flammability purposes, none of the other stream components were flammable, therefore the stream is modeled just as water and toluene. The results for this analysis can be found in the attached data sheet under "Determining Activity Coefficient".

The feed stream was given in terms of mass fraction, so this was converted to mol fraction for a case in which toluene was 2 percent by mass of the total feed stream, and also for a case in which only the unbound toluene was accounted for (5% of the 2% by mass).

Table 7.1 Mole Fraction Calculation

Mol Frac. Calc (in feed stream)										
Production (L/day)	Mass Frac. Toluene	Density of Stream (kg/L)	Production (kg/day)	Production of Toluene (kg/day)	Molecular Weight of toluene (g/mol)	Mols of Toluene (mol/day)	Mass of Water (kg/day)	Molecular Weight of Water (g/mol)	Mols of Water (mols/day)	Mol Fraction
200000	0.02	1	200000	4000	92.14	43412.2	196000	18.02	10876803.55	**0.00398**
200000	0.001	1	200000	200	92.14	2170.61	199800	18.02	11087680.36	**0.00020**

Using the constants in Table 7.1 and Equation 7.2, we solved for the saturated pressure of the vapor phase. This pressure was converted from bar into atm, and then Equation 7.2 was used to get the vapor fraction (yi). Using MSDSs found online, the flammability limits were used to determine if a floating head is needed. If the vapor fraction falls within the limits, a floating head tank is required. If it is above or below the limits, then a standard tank can be used. As can be seen TK-101 needs a floating head, but TK-102 does not.

Table 7.2 Flammability Limits TK-101

Flammability Limits														
Tank	Substance	Temp (K)	A	B	C	P_sat (bar)	P_sat (atm)	Activity Coefficient (aspen)	P (atm)	yi	yi (%)	LFL	UFL	Tank Type
TK-101 (.004)	Toluene	298.15	4.23679	1426.448	-45.957	0.03807	0.0376	8379.87	1	1.252	125.17%	1.10%	7.10%	fixed roof
TK-101 (.004)	Toluene	348.15	4.07827	1343.943	-53.773	0.3258	0.3215	5889.75	1	7.527	752.73%	1.10%	7.10%	fixed roof
TK-101 (.0002)	Toluene	298.15	4.23679	1426.448	-45.957	0.0381	0.0376	8379.87	1	0.062	**6.16%**	1.10%	7.10%	**floating roof**
TK-101 (.0002)	Toluene	348.15	4.07827	1343.943	-53.773	0.3258	0.3215	5889.75	1	0.371	37.06%	1.10%	7.10%	fixed roof
TK-101 (.004)	Toluene	323.17	4.14157	1377.578	-50.507	0.12282	0.12121	7105.73	1	3.424	342.39%	1.10%	7.10%	fixed roof
TK-101 (.0002)	Toluene	323.17	4.14157	1377.578	-50.507	0.12282	0.12121	7105.73	1	0.169	16.86%	1.10%	7.10%	fixed roof

Table 7.3 Flammability Limits TK-102

Flammability Limits														
Tank	Substance	Temp (K)	A	B	C	P_sat (bar)	P_sat (atm)	Activity Coefficient (aspen)	P (atm)	yi	yi (%)	LFL	UFL	Tank Type
TK-102	Toluene	368.15	4.07827	1343.943	-53.773	0.6358129998	0.6274838495	1	1	0.626	62.62%	1.10%	7.10%	fixed roof

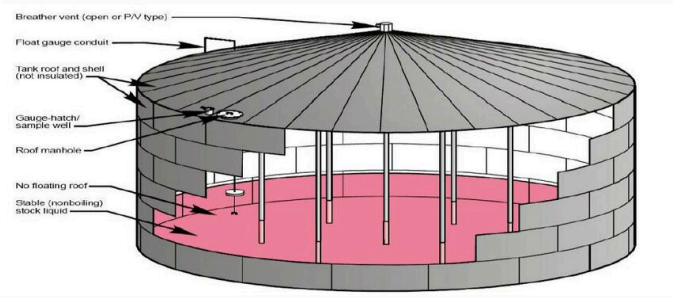

Figure 7.1 Fixed Roof Tank

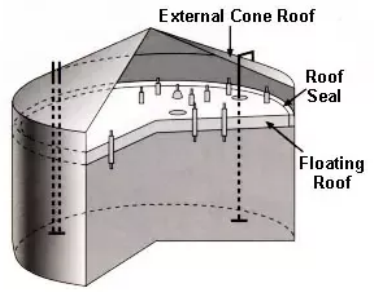

Figure 7.2 Floating Roof Tank

FMEA

Failure Modes and Effect Analysis (FMEA)

Failure Mode and Effect Analysis (FMEA) is a systematic, team-based risk assessment tool meant to identify ways that a process can fail (the failure modes), evaluate the effects of each failure, and then allow for prioritization of the risks identified so that limited resources can be directed towards the most consequential problems identified. Each failure mode is assigned a Severity (S), Occurrence (O), and Detection (D) score (1 = best, 10 = worst). The product of the three (S×O×D) results in a Risk Priority Number (RPN), providing an objective rating of the failure mode evaluated. Conventional hazard reviews (checklists, "what-if" analyses) would flag obvious issues but do not offer much guidance on where to focus resources / where to act first. An FMEA on the other hand converts qualitative concerns into quantitative RPNs, letting the team couple safety with economics. The same risk ranking logic used in the Center for Chemical Process Safety (CCPS) and Department Of Energy (DOE) guidance recommended for large chemical facilities. [17, 18, 19, 20, 21]

The process train was divided into five logically independent subsystems for several key reasons including capturing subsystem interactions, as well as ensuring that the values determined for each subsystem were not adversely impacted by values of other subsystems (i.e., some items may not seem as severe if compared to other items, etc):

1. S-1: Feed Conditioning & Toluene Removal

   TK-101, P-101 A/B, E-101, and the Steam Stripper T-101 all see the variable composition wastewater and the highest vapor-phase toluene loading. Operate under the same low pressure/steam conditions, and share common control loops for temperature and flow. Grouping them together isolates the flammability dominated hazards and keeps the high VOC (toluene) zone separate from downstream clean-in-place utilities, making the FMEA severity rankings comparable within the subsystem.

2. S-2: Distillate Recovery & Recycle Loop

   The condenser/decanter train (E-102, V-101) recycle pumps (P-102 A/B), P-103 A/B, and the fixed roof toluene tank TK-102 form a closed section where the vapor distillate from T-101 is cooled, phase split and then either recycled or stored. These units operate within the same pressure envelope, handle essentially pure organic vapor/liquid, and are physically remote relative to the aqueous line, so separating them keeps solvent handling risks from interfering with the water handling side for RPN calculations.

3. S-3: High Solids Concentration - Multiple Effect Evaporators

   EV-101, EV-102, the inter-effect heat exchangers (E-103 through E-106), and the transfer pumps P-104 A/B and P-105 A/B are all dedicated to flashing water and concentrating the non-volatile solids (products from the fermentation production of avermectin) before disposal. They operate at elevated temperature/pressure, have tight heat integration ties, and their dominant failure modes (tube side scale/fouling, steam balance upsets) are distinct from either the VOC or combustion sections, thus justifying a standalone subsystem.

4. S-4: Thermal Decomposition - Incineration

The existing incinerator (I-101), combustion air blower, and methane feed controls convert the MEE bottoms to CO_2 + H_2O at 1200 °C. Because combustion risks and utility costs dominate here, and because the hardware is effectively a "black-box" auxiliary to our systems design, it is analyzed separately from other systems to keep the FMEA focused only on the high temperature, high O_2 hazards only.

5. S-5: Final Polishing & Environmental Compliance (COD/Biological-Oxidation)
   The aerobic COD tank (TK-103), its duty/stand-by blowers, and the effluent probes sit in one subsystem because they all hinge on the aerobic bugs getting enough air. If the blower fails or the microbes crash, our polishing step fails. This subsystem exists at the very end of our process train, right where the treated stream is released to the environment. So, the focus is on meeting the <100 ppm toluene/COD limit, without dragging in stripping or incinerator potential concerns.

The team met and worked to develop ~15-20 plausible failures per subsystem. Historical data from analogous solvent-recovery units and manufacturer reliability figures supported the development of the Occurrence estimates with the Severity and Detection values were developed with input from CCPS guidance/literature. Scores were assigned by consensus in a workshop, disagreements wherever they were encountered were resolved by taking the conservative (higher risk) value. RPNs ranged across the different subsystems, and the entire table is found in the separate Microsoft Excel file titled "Avermectin Waste Treatment FMEA". Applying the Pareto Principle, we found that ~20% of the failure modes (worst two per subsystem) accounted for ~80% of the cumulated RPN. Given schedule (two weeks remaining) and budget constraints (~ 6% of plant CAPEX reserved for safeguards), we concentrated mitigation design efforts on those top two RPNs per subsystem. Lower ranked items were either Inherently Safe by Design (ISD) or covered by standard operating procedures and did not justify additional hardware at this stage. The selected mitigations bring every subsystem into a region considered ALARP (as low as reasonably possible) and meet Occupational Safety Health Administration (OSHA) Process Safety Management (PSM) and EPA Risk Management Policies (RMP) expectations.

<u>FMEA RPN Mitigation</u>

Once the top ranked failure modes were found through the RPN sort, we ran them through a fast "hierarchy of controls" filter, preference was given to those inherently safe first, then engineered controls/hardware, then administrative/procedural controls. This approach mirrors the CCPS playbook/recommended practices. Wherever an inherent change (e.g., specifying a fail-closed control valve in place of a gate valve) could significantly reduce the Severity or Occurence score, it was adopted first. When inherent fixes were unavailable, we selected engineered safeguards (e.g., redundant instrumentation, hard-wired trips, duty/stand-by equipment) that raised the detectability and would drive the RPN down, a practice in line with ALARP guidance. Administrative measures were reserved for residual, low-consequence cases to avoid over reliance on operator vigilance alone.

Potential safeguards were then screened through a brief cost-benefit analysis–potential Capital Cost divided by projected RPN reduction–then checked against OSHA PSM and EPA RMP "shall" clauses to ensure regulatory compliance. The final count resulted in ten actions, the highest RPN from each

subsystem, to capture ~80% of the total risk reduction in an effort to apply the Pareto Principle to focus resources on the vital few rather than the trivial many.

For brevity, the key results and mitigation strategies were explained in this section, for the complete FMEA worksheet including all ratings, RPN calculations and actions items please see the separate Microsoft Excel file titled “Avermectin Waste Treatment FMEA”.

SECTION 8: FINANCIAL ANALYSIS

After all equipment, utilities, and manufacturing costs were settled, the team was able to complete a full profitability / financial analysis on the avermectin waste treatment process. Per Dr. Raghavan, we decided to center the analysis on the *increase in sale price of avermectin* due to the waste treatment process.

Note that the majority of this financial analysis process will come directly from CBE 480 instruction and the Turton et. al design textbook, but with edits where necessary to fit our specific process. One difficult consideration for our process is that there isn't a singular main product that generates a majority of revenue. Rather, we're treating waste to meet environmental regulations, which is cost-intensive but does not directly make money for the company. Other than a small percentage of COMd that is reclaimed via the sale of purified toluene, the process doesn't produce revenue from product sale.

Note that the team was asked specifically *not* to use discounted methods to avoid complicating the sale price increase analysis, and therefore only non-discounted methods will be employed. From here, several assumptions must be made. Firstly, a MACRS depreciation method over six fiscal years (the current federal policy for chemical plants) will be employed. Additionally, the assumption is made that the analysis commences in year zero and ends in year 12, allowing for 2 years of process construction/startup and 10 years of plant operation. Because a wastewater treatment process for an existing production plant is being evaluated, this process is considered a Brownfield expansion to an existing plant.

The process of evaluating profitability first involves the calculation of several starting values, namely FCI (total Fixed Capital Investment), land costs, working capital, revenue, CoM_d (cost of manufacturing, without depreciation), and salvage value. The calculation and/or assumptions for each of these values will be laid out below:

FCI_L: Equations 8.1-8.6 are used to calculate FCI for the process. C_{BM}, or the bare module cost for each piece of equipment, is calculated using the extensive charts and best-fit equations from Turton et. al. The sum of all C_{BM} values in the project is the total bare module cost, or C_{TBM}. Note that Contingency (C_c) is assumed to be 15% of C_{TBM}, while the Fee (C_F) is assumed to be 3% of C_{TBM}, per Turton heuristics. Additionally, because the process was assumed to be Brownfield, Outside Battery Limits Costs (OSBL) costs were assumed to be 30% of Inside Battery Limits Costs (ISBL) costs, and cost of land was assumed to be zero.

$$C_{TBM} = \sum C_{BM}$$

EQ. 8.1

$$C_C = 0.15 * C_{TBM}$$

EQ. 8.2

$$C_F = 0.03 * C_{TBM}$$

EQ. 8.3

$$ISBL = C_{TBM} + C_C + C_F$$

EQ. 8.4

$$OSBL = 0.3 * ISBL$$

EQ. 8.5

$$FCI = OSBL + ISBL$$

EQ. 8.6

Table 8.1: Capital Costing Information

C_{TBM}	$972,813.65
C_C	$145,922.05
C_F	$29,184.41
ISBL	$1,147,920.11
OSBL	$344,376.03
FCI	$1,492,296.14

Since a Brownfield project is assumed, land cost is assumed to be zero. Thus, FCI = FCI_L, where FCI_L is defined by Turton et. al as total fixed capital investment minus land costs. Working capital (WC) is assumed to be 20% of FCI_L, while salvage value is assumed to be zero. Plant revenue was calculated using the given flow rate and cost of avermectin from the industrial process (400 kg / day, $200 / kg). Ideally, our yearly expenditures for wastewater treatment to reflect only a small percentage (ideally 1-2% or less) of the revenue generated from avermectin production. However, an interesting caveat is that our process purifies dilute toluene in the initial feed solution to high-purity (0.999 mass percent) toluene recovered from the steam-stripping column. Given background information states that toluene can be sold for $800/MT. At this point, our simulation produces 4082 kg/day of toluene at >0.999% by mass. Note that the plant is assumed to operate for 24 hours a day, for 340 days out of any given year.

This presents a unique opportunity when it comes to profitability evaluation, because in addition to purifying wastewater to the limits specified by the EPA, we are also able to produce nearly pure toluene, which could either be sent back to the avermectin industrial process (for a credit) or sold externally (for a profit). In either case, the team has decided to treat this toluene “production” as a profit–in other words, we don’t start “cutting into” the 1-2% avermectin revenue budget until all profits from toluene have been exhausted. For example, if the toluene produced can be sold for 1 million dollars / year, and our process costs 1.2 million dollars / year to run, we will only assume that our “deficit” (what’s cutting into our 2% budget from avermectin production) is $200,000 / year.

Finally, CoM_d (cost of manufacturing, minus depreciation) is calculated using the equation given in CBE 480 (Equation 8.11). C_{UT}, or utility costs, have been calculated in the attached Excel sheet, using values obtained from Turton, et. al.[4] C_{RM} (cost of raw material) is zero in our case, as we receive a wastewater

stream from a production plant. Though we use compressed air, steam, etc. throughout our process, we treat these as utilities, not raw materials. Finally, C_{OL} (cost of operating labor) can be calculated using equations from CBE 480, written as Equation 8.10 here, where P is the total number of processing steps handling particulate solid and N_{np} is the total number of nonparticulate processing steps. Note that although there are suspended solids in our system, there are no manual "particulate processing" steps, as all streams with suspended solids contain >5% solid particulate, and thus can be pumped and processed like normal. Also, note that a salary of $50,000 / year is assumed for each operator. Waste treatment cost (ironically) was assumed to be zero for this process, as although we are waste-treating treating the whole stream, these costs are wrapped up within our utility and operating costs. Incineration might be called a "waste treatment" process, but as it's an integral part of our process, we'll cost the utlities as-normal.

$$FCI_L = FCI - L = FCI$$

EQ. 8.7

$$WC = 0.20 * FCI_L$$

EQ. 8.8

$$Revenue = mass\ flow\ rate\ pdt\ (kg/hr) * \frac{24\ hrs}{day} * \frac{340\ days}{year} * \frac{\$\ (sale\ price)}{kg}$$

EQ. 8.9

$$C_{OL} = 4.5 * 50,000 * (6.29 + 31.7P^2 + 0.23N_{np})^{0.5}$$

EQ. 8.10

$$COM_d = .18FCI_L + 2.73C_{OL} + 1.23(C_{WT}+C_{UT}+C_{RM})$$

EQ. 8.11

Table 8.2: Manufacturing Costing Information

Working Capital (WC)	$298,459.23
Revenue (Avermcetin)	$27,200,000 / year
Revenue (Toluene)	$1110288.77 / year
C_{OL}	$700,000.00 / year
C_{RM}	$0.00 / year
C_{WT}	$0.00 / year
COM_d	$2,970,882.92 / year

Using information from Equations 8.7-8.11, the team has provided an Excel sheet (attached as "Profitability" worksheet in "Data Sheet for 488" workbook) including investment costs, depreciation (MACRS method), revenue, expenses, income tax, after-tax profit, and after-tax cashflow. In this manner, the team has determined how well proposed budget expectations have been met via the evaluation of net-postive revenue from toluene and net-negative costs for capital expenditures, manufacturing costs, etc.

Dr. Raghavann requested that we evaluate process profitability via analyzing the percent increase in sale price (of avermectin) that would be required to make the same profit as without waste treatment. The team was given a goal of 1-2% sale price increase. In the 10 years of active waste treatment, we don't get quite to this level of cost efficiency. See Figure 8.1 below for the necessary sale price increase over time.

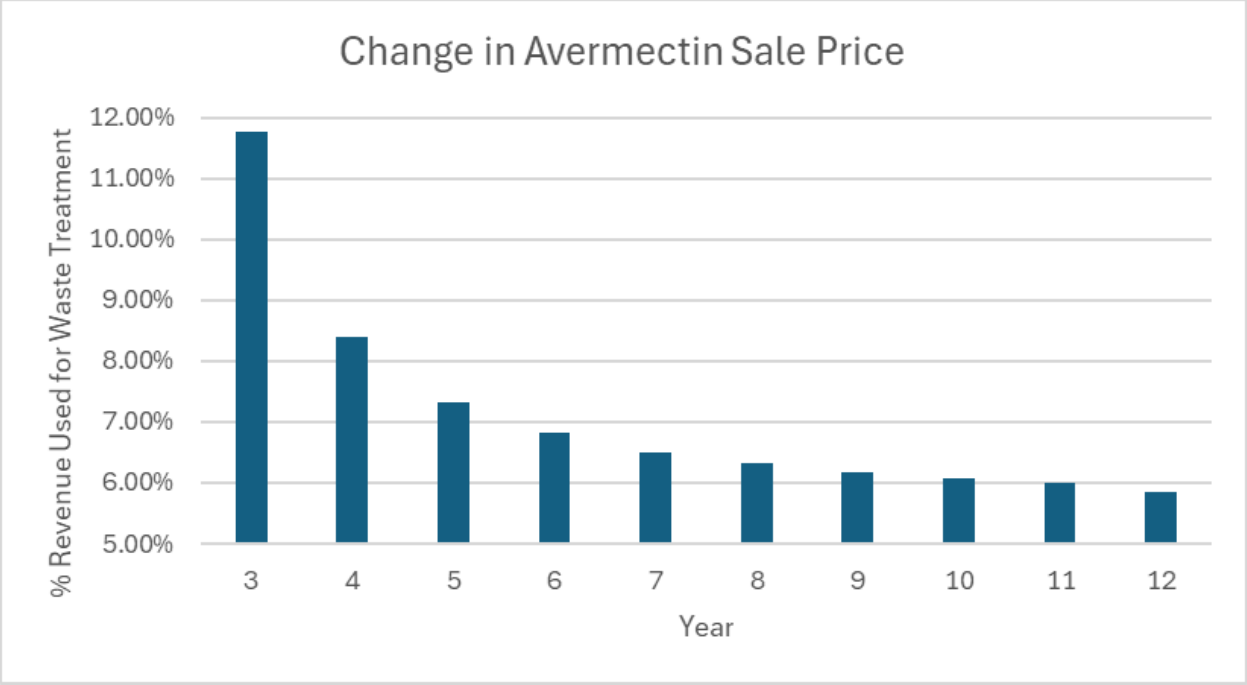

Figure 8.1: Necessary Sale Price Increase

Note that this value is artificially inflated in year 3 (the first year of avermectin production) because we've only produced and sold one year's worth of avermectin, but we've spent all up-front capital cost and working capital. Then, as we continue to produce avermectin (but not spend nearly as much in up-front costs), our necessary sale price increase continues to decrease over the years of operation. In fact, if we were to extend the plant's operation out to 20, 30, or 40 years, the necessary sale price increase value would continue to decrease over time. However, for the purposes of this project, a 10 year operation (with two years of build phase) are assumed. By Year 12 (on Figure 8.1 above), we see that the overall increase in sale price is 5.84%. Note that all other specific values are included in the attached profitability spreadsheet.

Although the goal was to be within 1-2% of avermectin production revenue, we did not quite reach that level of cost efficiency. However, with additional iterations, we could likely decrease utility costs to get

closer to this desired value. Unfortunately, time constraints did not permit us to iterate further through design choices.

Nevertheless, the team believes that slightly increased cost for multiple effect evaporative separation (as compared to membrane separation) is worth a relatively small additional price–using a method that leverages thermodynamic principles to separate a highly toxic compound is a much higher-fidelity method than one which requires the function of a physical barrier. For example, if a tear or lesion were to occur in a membrane that the operators/engineers were unaware of, it would be relatively easy to "accidentally" exhaust avermectin to an effluent stream. By leveraging different boiling points of water and pollutants, however, we can ensure that as long as water is being boiled off at known temperatures and pressures, only *extremely* negligible amounts of avermectin will come with it.

Note that the team has previously ruled out centrifugation as an option–the particles in our system are simply too small to settle effectively without the use of an additional flocculant.

SECTION 9: CONCLUSION AND RECOMMENDATIONS

In conclusion, the team was able to effectively separate toluene, avermectin, and organic pollutants from the given wastewater stream. To accomplish this, a steam stripper operation was used to remove toluene, which was then decanted off to ~99.9% purity to be sold for a sale credit. Additionally, two counter-current multiple effect evaporators in series were used to leverage the high boiling point of avermectin, separating it from water to a purity of approximately 16 parts per quadrillion (mass fraction of 1.6e-14). This a significant improvement over the EPA-required separation to 1 ppb. Concentrated avermectin is then sent to an incinerator, where it is burned until decomposition occurs (1200 degrees Celsius). Using this method to purify the industrial waste resulted in an overall sale price increase of avermectin of >6%. This increase is reasonable, but not quite within the ideal 1-2% range.Although this method it is slightly more costly than prefered, the team believes that the safety factor associated with this method is worth the slight additional expenditure. Leveraging invariant thermodynamic properties (such as boiling temperature and flash behavior) is an extremely high-fidelity separation method that will not fail as long as the evaporators continue operating at the specified heat duty (and the feed stream avermectin concentration does not increase by several orders of magnitude).

The team recommends that the company move forward with the proposed process. However, if the company would like to iterate further to result in more potential cost savings, the team recommends that a study be completed which evaluates if a better overall heat transfer coeffient (U) can be achieved in the multiple effect evaporators. Much of the utility costing for this process is tied up in the steam necessary to heat the wastewater stream to flash temperatures; if this could be reduced, the total expenditures per year would decrease for the lifetime of the plant. Additionally, the team would recommend a study be done on the advantages and disadvantages of adding caustic (NaOH) to the process. If NaOH could be added to the concentrated avermectin stream and held at an increased temperature for several minutes, the avermectin could likely be degraded without the use of the incinerator (which incurs additional utility costs). However, this would require in-depth analysis on the cost of equipment needed up-front, the cost of NaOH, and whether this would be cheaper overall than the current incineration process. Overall, the proposed process provides a reliable and high-purity solution for wastewater treatment, leveraging proven thermodynamic principles for consistent performance. With continued optimization, the company can potentially reduce operational costs while maintaining the rigorous environmental standards.

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APPENDIX A: EQUIPMENT CONDITIONS

TK-101: Conditions were set at atmospheric pressure and 25 C which were the given conditions of the waste stream.

P-101A/B: Pressure was set to be an increase in 10 psi from atmospheric pressure. This is so the liquid can enter the tower which was set at atmospheric pressure.

E-101: The temperature was set at 90 C so the toluene in the stream would remain liquid, but we could reduce the heat duty of the stream stripper. The liquid at 90 C is closer to the operating temperature of the stripper.

T-101: Justification in Section 5

E-102: This is set at 90 C so the toluene and water condense

V-101: This is set at atmospheric pressure and 85 C so the toluene and water can be separated and decanted.

P-102 A/B: This pressure is set at 0.69 bar so the condensed water can fall back into the tower

P-103 A/B: This pressure is also set at 0.69 bar so the toluene stripped waste will move from the steam stripper to the first evaporator.

E-103: The temperature was set at 105 C so we could reduce the heat duty of the evaporator. This was done by heating the liquid at 105 C is closer to the operating temperature of the first evaporator.

EV-101: Justification in Section 5

E-104: This was set at 35 C in order to condense the steam from the first evaporator into liquid using cooling tower water. Additionally we cooled it to 35 C because this stream is going to the COD tank and microbes grow at 37 C.

P-104 A/B: The pressure was set at 0.69 bar to move the liquid from two streams into one stream and sending it into the COD tanks.

P-105 A/B: This pump was set at a pressure of 3.28 bar to move the solid waste out of the first evaporator to the stream that goes to the COD tanks.

P-106 A/B: This pump's pressure was set at 3.28 bar to take the waste stream from the first evaporator to the second evaporator.

E-105: The temperature was set at 145 C so we could reduce the heat duty of the evaporator. This was done by heating the liquid from the first evaporator to the operating temperature of the second evaporator.

EV-102: Justification in Section 5

P-107 A/B: This pump's pressure was set at 3.23 bar to take the waste stream from the second evaporator to the incinerator

E-106: The temperature of this evaporator was specified at 35 C to cool the condensed liquid coming out of the first evaporator. We cooled it to 35 C because this stream is going to the COD tank and microbes grow at 37 C.

TK-103 (COD Tanks): Justification in Section 5

I-101: Justification in Section 5

APPENDIX B: CONTRIBUTIONS

Frank: FMEA work, defining subsystem boundaries and interactions, mitigation strategies as well (a team effort for FMEA work, but I was heavily involved in compiling our results). Worked with Calre on the introduction statement to start this report, summarizing the environmental limits in the table found in Section 1. Helped find supporting CCPS sources to develop FMEA choices. Wrote out the sizing/costing of the Steam Stripper. Wrote out the steam stripper section in the Simulation section explaining the HETP and how we iterated to find tower sizing specs.

Abigail: I worked on researching and drafting the introduction, laying the groundwork for a clear understanding of the project's goals. Additionally, I developed the method selection matrix, carefully evaluating potential approaches to identify the most efficient and cost-effective design. My contributions extended to the energy balance and utility calculations, where I analyzed the energy requirements and utility usage to optimize process efficiency. I also handled the sizing and costing of equipment, ensuring accurate estimates for capital and operational expenses. I also conducted the gamma calculations for determining flammability limits, a critical safety component in the design process. Finally, I helped with the conclusion, integrating all findings into a cohesive summary that highlighted the project's successes and potential areas for further improvement.

Clare: I worked on the research for the introduction, providing a comprehensive overview of the project's background. In addition to this work, I did the method selection, carefully evaluating the various approaches to ensure the most efficient and cost-effective process design. My contributions also included costing and sizing calculations for the major equipment, such as tanks, where I conducted thorough flammability assessments to ensure compliance with safety standards. I was responsible for creating all the block flow diagrams (BFDs) and process flow diagrams (PFDs), and I worked Appendix A.

Morgan: I put together the material balance (stream table) Excel sheet that is called out in Section 4 and is attached to this report. I also wrote the MEE, COD, and incineration sections under "Equipment Design" in Section 5, and wrote all sections under "Simulation" in Section 5. I also completed Section 8 (Financial Analysis) and cowrote the conclusion.