Spring 2023

WASTE PLASTICS PLANT DESIGN

Final Binder

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Executive Summary

The Aspen Plus design group was assigned to design a waste plastic recycling plant based on research done on pyrolysis of High-density polyethylene. The plant would convert 300,000 tons per year of HDPE into petroleum products, such as wax and liquid hydrocarbons. To test feasibility, we would need to find at what cost for raw feed would give us a 15% DCFROR.

After completing the FEL-2 study, it was found that a raw feed price of $0.59/lb of HDPE would provide a 15% after-tax DCFROR. Aside from economics, we were successful in designing a plant to convert 300,00 ton/yr of HDPE into petroleum products using Pyrolysis. This was accomplished under the conditions that product yield is identical to the smaller scale conducted in lab conditions and that the raw feed we purchase is prefiltered to eliminate potential contaminants that would cause unwanted side-reactions. We also have made the educated assumption that the wax will not contain tar, as was the case in the smaller scale model. This assumption is based on the fact that tar is created when a surface, such as metal or in a mineral fluidized bed, is present. Our reactor will utilize a steam fluidized bed and therefore, tar will have no place to form.

Before full scale implementation we suggest further studies on product yield in a pyrolysis reactor utilizing steam and potentially creating pilot plant to see the impact of scaling.

The binder attached provides a detailed summary of the economics, process design, related legislation, and other related information found during the FEL-2 stage.

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Memorandum

To: Project Management Office

From: The Aspen Minus (Dawson Grimm, Jimmy Hayes, Joseph Yoon, Joshua Rodrigues)

Date: 12-May-2023

Re: Design Basis Memorandum - Recycle of Waste Plastics Plant Feasibility Study

Introduction and Rationale

The economy of plastics is largely a one-way street, where the finished products remain permanently in that form. This means that waste plastics often end up discarded. In 2019, approximately 7 million tons of plastic waste were diverted to landfill in Texas. Pyrolytic decomposition of plastics can help to create a circular economy, where unused finished plastics can be processed back into raw materials for production of other petroleum-based products.

We were hired to develop a process design and FEL-2 study of the pyrolytic decomposition of polyolefin plastics. The client has requested a process design to convert polyethylene or propylene to their respective building blocks at feed rate of ~300,000 tons per year that generates a 15% after tax DCFROR. More specifically, we have designed a process involving the pyrolytic decomposition of HDPE plastics using superheated steam as a heat source. Wax is selected as the main product due to its high price relative to other possible products ($0.97/lbm) and easy separation from other products.

The metric that determines feasibility is the after tax DCFROR, meeting that minimum of 15% at the maximum possible feed rate will maximize revenue. We have delivered a total process design to convert HDPE to wax and liquid hydrocarbons. A complete economic analysis of our plant has been completed as well, including plant location, capital costs, operating costs, in order to determine a yearly revenue estimate. The capital cost is estimated at $81.1M and the operating costs at $36.M/yr. The price of waste HDPE needed to generate 15% after tax DCFROR with our process is estimated at $0.59/lbm. Current prices hover around $0.45/lbm. We recommend moving ahead with a FEL-3 study of this process.

Cost estimate for FEL2 study

Table 1. Cost estimate for FEL-2 Study

Category	Hourly Rate ($/hr)	Total Weekly Hours (hr)	Total for 15 Weeks ($)
Engineer (4)	100	50	75,000
Supervisors (3)	130	20	39,000
		Subtotal	114,000
		10% office supplies	11,400
		Total	125,400

QC: Checked by James Hayes

Raw Materials

We will be taking polyethylene (PE) plastics as feedstocks. The feed rate of PE to the plastics repurposing facility is expected to be about 300,000 tons/year.

PE plastics come in several different types, characterized by their density, molecular weight, and branching. Some include low density PE (LDPE) and high-density PE (HDPE).

There are several reasons why using a PE feedstock makes more sense than polypropylene (PP) plastics. First, more PE end use applications are single use than PP (56% vs 33%). Furthermore, more PE is produced than PP, so there is likely to be more available feedstock for a plastics repurposing facility[1]. For our process, we have chosen to operate with HDPE.

HDPE feedstock will be grinded up in a pre-treatment step before being put into the pyrolysis reactor[2]. Cleaned PE plastics will be used. We will try to buy waste PE plastic which has already been cleaned.

Natural HDPE waste plastic costs around $0.35-0.45 per pound. Prices have been as high as $1 per pound[3]. Color HDPE costs around $0.06-0.12 per pound. Prices have been as high as $0.50 per pound[4]. We are unsure of the effect of colorants on the product disposition; therefore, we will focus on natural HDPE feed and treat the effect of colorants as negligible.

Product and Co-product Disposition

Table 2. Products of pyrolysis of HDPE at 932 °F (500°C). Feedstock is 90-95% HDPE, 5-10% rubber[5].

Commented [JMH1]: Does it make a diff

Product	Yield (wt%)	Boiling Point (°F)	Production Rate (tons/yr)
Wax and Tar	97.58	~700	292,740
Ethylene	0.64	-154.7	1,920
Propylene	0.37	-53.7	1,110
Butylene	0.35	20.7	1,050
Toluene	0.3	231.1	900
Pentane	0.21	97	630
Propane	0.19	-43.6	570
Ethane	0.13	-128.2	390
Methane	0.09	-258.9	270
Benzene	0.09	176.2	270
Butane	0.04	30.2	120
Xylene	0	~280	0

The main products are waxes and tars. Until further research is completed (FEL3), we will assume 100% of the solids is wax. Running the reactor at a lower temperature gives more wax product. The lower temperature will also lower the metallurgical requirements. Furthermore, many of the individual components listed in Table 2 (ethylene, propylene, benzene, paraffin wax) are sold within a similar price range (~$0.50-1.00/lbm), but separating wax is much simpler due to its much higher boiling point than any other product. Therefore, wax products are more profitable. However, running the reactor at a lower temperature could present risks since the ratio of wax to tar in the solid products is still unknown.

The lab scale data is from experiments with a sand bed rather than steam fluidization. In reality, it does not seem likely that tar would form in our reactor since there would be no solid bed for it to accumulate on. Furthermore, any tar that did form would react with steam to form CO or CO_2. Therefore, we assume all solid products are wax.

Some liquid hydrocarbons are produced in the process. They may be sold off as a mixture of synthetic crude oil to nearby refineries.

The main gas product is ethylene, which can be used as a feedstock in producing new plastic. Some propylene and butylene are produced, which can also be used as plastic feedstock. However, given the energy needs of the facility and the difficulty in separating these products, we expect to instead use all of the vapor products for combustion in the boiler.

According to Al-Salem et al.[6], wax is classified as C_{19+}. The wax recovered from pyrolysis of HDPE results in two characteristically distinct fractions, one of light waxes (up to C_{40}) and a BP between 343 and 525 Celsius, and one of heavy waxes, with carbon point and boiling point both higher than the aforementioned. The lighter wax has MW between 100 and 1000 g/mol, and a high olefinic and aromatic content. Whether or not the heavier waxes have economic value is unclear at this time. As a result, we will be assuming that the waxes produced via pyrolysis will be all of one grade for now. Separating the waxes into different grades will be further investigated later in process development. The wax products will be stored as a liquid before being sold.

Key Assumptions

We will be making the following assumptions.

- All solid products are wax. No tar or char is formed in the reactor
- Lab scale data for product disposition at 930°F will scale up to a 300,000 tons/yr facility
- Colorants in PE have negligible effect on product yield
- For the purposes of analyzing the separation units, wax is treated as N-pentacostane ($C_{25}H_{52}$), although in reality there would be a distribution of wax molecular weight

Furthermore, this study is limited to HDPE as a feedstock, rather than LDPE. However, it should be noted that it could be possible for this process to run on either, but it is outside the scope of this study to investigate further.

Operating Philosophy

The facility will be continuously operated. There will be regular shut-off for maintenance and to potentially address buildup/fouling in small diameter pipes. Wax and tar products could potentially cause buildup if temperatures ever get too low in the process, but as long as they are continuously heated then there should be no issue.

Standard Units

The standard units for use in this process are shown in Table 3.

Table 3. Standard Units for use in this process.

Process Variable	Unit
Mass Flow Rate	lbm/hr
Volumetric Flow Rate	scf/hr
Heat Load	MMBTU/hr
Electricity	kW
Temperature	°F
Concentration	wt%
Pressure	psig
Density	lbm/scf
Specific Enthalpy	BTU/lbm
Land Area	Acres
Building and Process Unit Area	ft^2

Some exceptions are used. For example, gallons per minute (GPM) is standard for water flow rate so it is sometimes used instead. Additionally, yearly capacities are expressed in terms of tons/yr for feedstock and products.

Polyethylene Suppliers

We have identified several potential suppliers of HDPE. The specifications on the HDPE and the prices these companies have quoted us at are listed below:

Table 4. Waste HDPE distributors and typical prices.

Worldwide Distribution USA – Recycled plastics distribution located in Coldsprings, TX $.32/lb at 40,000 lbs for mixed, scrap HDPE $.45/lb at 20,000 lbs for refined scrap
De-Two Trading Inc. - Recycling center selling HDPE milk bottles in Ohio $.39/lb at 10,000 lbs
Tarquin Polymers & Colors- Polymers & chemical production in Sugar Land, TX $.46/lb at 30,000 lbs for scrap HDPE

These include transportation costs, handled by the suppliers. This aligns with our estimation of PE feed costing $0.40-0.45/lb.

Potential Customers & Finished Good Transport

Paraffin Wax:

Paraffin wax is commonly used in the manufacturing of candles, cosmetics, and coatings-related products. Paraffin wax is the main ingredient in some types of candles, as an emollient in lotions and creams, and as a coating on certain fruits, vegetables, paper, and cardboard to reduce moisture exposure. As a result, we will be primarily targeting these industries, with the further goal of targeting manufacturing plants near ours in the greater Houston area. Below is a list of potential customers, their location, and distance from our plant location:

- Packit Gourmet – Specialty foods producer (freeze-dried, camping foods). Packaging uses paraffin wax, located in Austin TX, 200 mi from our plant.
- Texas Food Pack – Food packaging manufacturing and distribution company. Make many cups, paper plates, wraps and boxes that use paraffin wax. Manufacturing plant in Dallas TX, 270 miles from our plant
- Mary Kay – Global cosmetics manufacturer, headquarters and manufacturing plant located in Dallas, TX. Produces skin lotions, lipsticks, and lip balms that contain paraffin wax as an emollient for its moisturizing/conditioning properties on the skin. 280 miles from our plant.
- Paddywax – Candle manufacturer based in Nashville, TN. Produce a large variety of candles made with paraffin wax. 800 mi from our plant.
- Globaltech Industries, Inc – One of largest candle manufacturers in the USA based in Cornelia, GA. 850 miles from our plant.

Synthetic Crude Oil:

Crude oil is used mostly commonly in oil refineries and olefins plants. There are many of these in the greater Houston area, however, two are extremely close to our proposed plant location in Baytown, TX. Exxon has an olefins plant just 5.2 mi from our plant, and Baytown Refinery is ~4 mi from our plant. Due to how close these companies are, it would likely be more cost effective to build a pipeline between our plant and these companies. This would also be more environmentally friendly as a pipeline would be less prone to spills. However, this requires much further investigation, as contracts with both companies would need to be created, the pipelines would need to follow local ordinance laws, etc. As a result, we will only be considering transportation through the traditional tanker truck method. The trucking companies listed above also have the required vehicles for transportation of our crude oil, we will use them for this product as well. Below is a list of refineries and plants in the greater Houston area we could sell our liquid oils to:

- Exxon Mobil Baytown Olefins Plant – Polyolefin manufacturing plant just 5 miles from our plant
- Exxon Mobil Baytown Refinery – Oil refinery somewhat adjacent to the olefins plant. 4 mi from our plant
- Petromax Refining Company LLC – wholesale refining and distribution of petroleum/petroleum-based products. 16 mi from our plant
- Valero Energy Houston Refinery – Network of oil refineries across Texas. Their Houston location is 25 mi from our plant. They also have a Port Arthur location, about 70 miles from our plant, as well as a Corpus Christi location that is 230 miles from our plant.

Transportation:

Transportation of paraffin wax is a trivial matter, dependent upon the distance necessary to reach the customer. For short- medium distances (<1000 miles), the wax is transport in tanker trucks that heated and insulated to retain the product as a liquid. For larger distances, the wax is transported in similarly heated tanker cars on trains. The wax is then unloaded in these tanker trucks to reach its destination. The same is true of crude oil, with slightly different specifications on the tanks themselves. Locomotive transportation is far more expensive, so we will only be considering potential customers within a tanker truck range from our plant. Below is a list of some trucking companies based in the Houston area that have the necessary trucks in their fleet:

- Quality Carriers – Tanker trucking company that specializes in transporting liquid bulk materials. They have a terminal in Houston that is 9 mi from our plant location.
- Trimac Transportation – Tanker trucking company servicing chemical, petroleum, and dry bulk product manufacturers. Their location is 11 mi from our plant location.
- Liquid Cargo – Tanker trucking company specializing in transporting food-grade and hazardous materials. They have a terminal in Houston, just outside of La Porte (13 miles from our plant location)
- Tankstar USA – Tanker trucking company that provides transportation services for the chemical, petroleum and food industries. They have a terminal in Pasadena (14 miles from our plant location)
- Schneider National – Nationwide logistics/transportation company. Provide trucking services for the chemical, food, and bulk products industries. Their location in Houston is 20 mi from our plant location.

Transportation via tanker trucks is relatively inexpensive over the proposed distances to our customers. A typical tanker truck could hold a load of 40,000-45,000 pounds of liquid wax. At our approximate production rate of 300,000 tons/year, this corresponds to 15,000 truckloads to safely transport the liquid wax out of our plant. To transport this over 100 miles costs between $300 and $600, and for 1000 miles, between $2000 and $4000. These figures are largely dependent on current gas prices and fluctuations in carrier rates. Given these figures, we estimate that our annual transportation costs for liquid wax will be ~$12M. A typical tanker truck can hold 55,000-60,000 pounds of liquid oil. At our approximate production rate of 2100 tons/year, this corresponds to 70 truckloads to safely transport the liquid oil out of our plant. To transport crude oil via tanker truck over 100 miles would cost between $2000 and $5000, and for 1000 miles, between $400 and $1000. Given these figures, we estimate our annual transportation costs of crude oil will be ~$300K. The negative relationship between distance and cost can be explained by the fact that crude oil is a hazardous material, and thus has high handling costs. The high fixed costs associated with loading and unloading are more evenly distributed across the longer distance. Crude oil transport is more expensive than paraffin wax due to supply and demand, where crude oil is much more desired product, and thus has a higher premium associated with its transportation to our customers.

Block Flow Diagram (BFD)

The BFD for our process is shown in Figure 1.

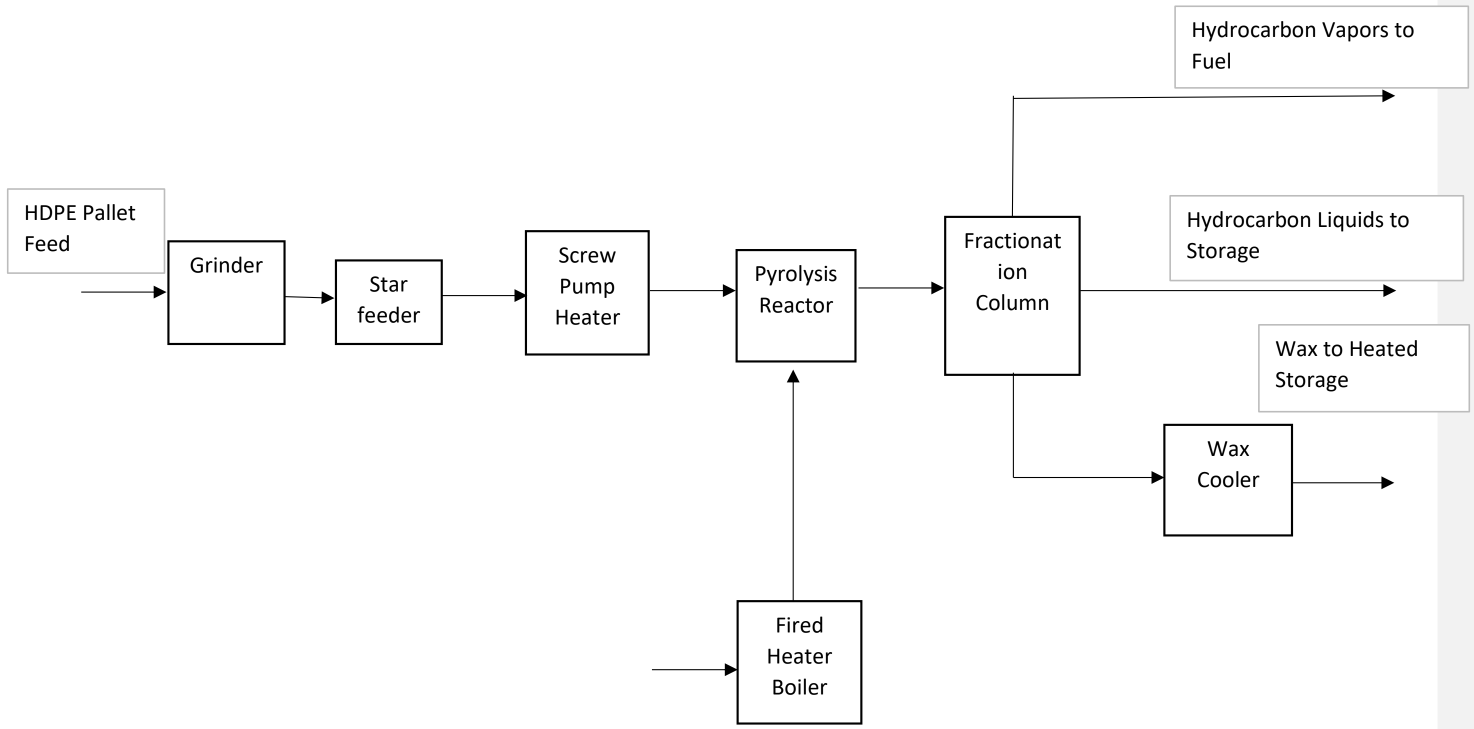


Figure 1. Block Flow Diagram of Pyrolysis Process

Pre-treatment, Grinder, Starfeeder, and Screw Pump Heater

Several steps will be taken to treat HDPE before it enters the reactor. We expect that the scarp HDPE we purchase to have been cleaned and treated to some degree.

In our process we will first grind scrap HDPE into a more workable form: pellets. We have identified models from Shini USA[7] and used a scaling rule of 0.6 to help design the unit for our process. The specifications of the pellet grinder are shown in Table 5

Table 5. Operating conditions and specifications of pellet grinder unit.

Parameter	Value
Operating Pressure	0 psig
Operating Temperature	100°F
Power	345 HP (254 kW)
Fly Knife Style	Scissor
Mass Flow Rate	19.0 lbm/s (300,000 tons/yr)
Rotor Speed	650 RPM

Next, an apron conveyer will transport the pellets to the starfeeder. The conveyer will have a length of 49 ft (15 m) and a width of 6 ft (2 m). The belt will move at a speed of 0.15 ft/s. Assuming a friction factor of 0.5, the power of the motor needed to power this conveyer is 9.02 HP.

$$Belt\ Pull = \mu H L_{belt} W_{belt} \rho_{HDPE} \ ,$$

$$Belt\ Speed\ = \frac{F_{HDPE}}{L_{belt} W_{belt} \rho_{HDPE}} \ ,$$

$$Conyevor\ Motor\ Power = Belt\ Speed * Belt\ Pull \ ,$$

Where the parameters are defined in Table 6.

Table 6. Apron conveyor and motor parameters.

Parameter	Value
Width of Belt, W_{belt}	6 ft
Length of Belt, L_{belt}	49 ft
Height of HDPE, H	3 ft
Density of HDPE, ρ_{HDPE}	61.2 lbm/ft^3
Friction Factor, μ	0.5
HDPE Feed Rate, F_{HDPE}	19.0 lbm/s (300,000 tons/yr)
$Belt\ Pull$	33,100 lbm
$Belt\ Speed$	0.15 ft/s
Conveyor Motor Power	9.02 HP (6.73 kW)

The starfeeder unit will exclude air from the PE pellets. Oxygen must be removed from the HDPE feed before it enters the reactor. This is due to the risk of combustion due to high temperatures in the reactor.

We have identified the ACS CI Series 16x16" Airlock Rotary Valve as a potential choice for this unit[8]. At a speed of ~14 RPM, it can process 1,176 ft^3/hr, which meets the feed flow rate in our process. An image of the unit is shown in Figure 2.

Figure 2. ACS CI Series 16x16" Airlock Rotary Valve.

The operating conditions and specifications of the starfeeder are shown below in Table 7.

Table 7. Specifications of starfeeder unit.

Parameter	Value
Operating Pressure	0 psig
Design Pressure	15 psig
Operating Temperature	100°F
Material of Construction	Cast Iron
Operating Mass Flow Rate	19.0 lbm/s (300,000 tons/yr)
Operating Volume Flow Rate	1,120 ft^3/hr
Design Volume Flow Rate	1,176 ft^3/hr
Rotor Speed	14 RPM
Valve Size	16 in
Power	0.16 HP (0.12 kW)

The screw pump heater (SPH) will heat the HDPE to a temperature of 350°F and melt it. A steam jacket will be present in the unit to provide the heat load. The SPH will deliver the HDPE to the reactor nozzle inlet. More typical centrifugal pumps cannot be used for this purpose because the feed is a solid. The operating conditions and specifications of the SPH are shown in Table 8. The model "XG090B01Z" from Hangzhou Xinglong Pump Co., Ltd meets the flow rate and required specifications for our process[9].

Table 8. Specifications of SPH unit.

Parameter	Value
Inlet Pressure	0 psig
Outlet Pressure	44 psig
Head	104 ft
Inlet Temperature	100°F
Outlet Temperature	350°F
Heat Load	15.0 MMBTU/hr
Operating Volumetric Flow Rate	1,120 ft^3/hr
Design Volumetric Flow Rate	1,480 ft^3/hr
Rotor Speed	238 RPM
Power	20.1 HP (15 kW)

Pyrolysis Reactor

The reactor will use high temperature steam to initiate the pyrolysis of a feed of liquid HDPE. Nozzles arranged on the bottom of the vessel will spray steam and droplets of liquid HDPE.

An injector nozzle, such as that typically used for fuel, will spray the steam. The velocity of the steam will be determined based on the calculated fluidization velocity. Atomizer nozzles arranged around the base of the reactor will inject liquid HDPE droplets into the reactor.

The high surface area to volume ratio of the droplets will speed up the pyrolysis reaction compared to using pellets.

Due to the elevated temperature in the reactor and of the steam inlet, a 25-12 Ni-Cr alloy (SS-309) will be used as the main construction material for the reactor. This alloy is capable of operating at very high temperatures (up to 1900°F).

The following equations from Khaghanikavkani and Farid [10] were used to determine residence time of the reactor.

$$-\ln\left(\frac{1-X}{1}\right) = k(t_{residence}) \quad (1)$$

$$k = k_0 e^{\frac{-E_a}{RT}} \quad (2)$$

The parameters (shown in Table 9) were based off experiments with 2-3mm diameter HDPE pellets, rather than liquid HDPE droplets. Therefore, the actual required residence time should be lower than calculated using these parameters. It is uncertain how much lower the actual required residence time would be, so this value is used for the reactor design instead.

Table 9. Parameters for integrated rate expression of pyrolysis reactor.

Parameter	Value
Conversion, X	99.999%
Rate constant, k	0.866 s^{-1}
Pre-exponential factor, k_0	22,500 s^{-1}
Activation Energy, E_a	65.33 kJ/mol
R	8.314
Temperature, T	930°F (500°C)
Residence Time, $t_{residence}$	**13.3s**

The following equations were used in determining the fluidization velocity of liquid HDPE particles within our pyrolysis reactor. This work was based on the Kunii-Levenspiel bubbling-bed model[11].

$$\varepsilon_{mf} = 0.586\psi^{-0.7}(\frac{\mu^2}{\rho_g \eta d_p^3})^{0.029}(\frac{\rho_g}{\rho_c})^{0.021} \qquad (3)$$

$$u_{mf} = \frac{\eta(\psi d_p)^2}{150\mu}\left(\frac{\varepsilon_{mf}^3}{1-\varepsilon_{mf}}\right) \qquad (4)$$

$$u_f = \frac{\eta d_p^2}{18\mu}\ for\ Re < 0.4 \qquad (5)$$

$$u_f = (\frac{0.0178\eta^2}{\rho_g \mu})^{1/3} d_p\ for\ 0.4 < Re < 500 \qquad (6)$$

If fluidization velocity is below the minimum, the HDPE droplets will not achieve lift and they will sink to the bottom of the reactor. If the fluidization velocity is above the maximum, the droplets may not react, and unreacted HDPE would enter the column.

We designed our reactor to maintain fluidization and operate close to but below the maximum fluidization velocity. It is assumed that all bubbles are of the same size ($d_p\ constant$) and that flow is smooth in emulsion phase.

The results of these calculations are summarized in Table 10.

Table 10. Parameters associated with fluidization velocity equations.

Parameter	Value
Dynamic Viscosity, μ	2.86E-6 Pa/s
Gas Density, ρ_g	2.824 kg/m^3
Liq. Particle Density, ρ_c	973 kg/m^3
Particle Diameter, d_p	0.00025 m
$\eta = g(\rho_c - \rho_g)$	9635.125 kg/m^{2}s^2
Sphericity, Ψ	1
Min Fluidization Velocity, u_{mf}	**0.87 ft/s**
Max Fluidization Velocity, u_f low Re	**3.84 ft/s**
Max Fluidization Velocity, u_f high Re	**2.24 ft/s**

The steam velocity was set based on the previously determined fluidization velocities.

The steam flow rate into the reactor was determined based on the heat load necessary to heat the PE feed from the inlet temperature (350°F) to the reactor temperature (930°F). This heat load is approximately 23.8 MMBTU/hr. The following parameters for HDPE (Table 11) were used in the calculation. These values were assumed to be constant with temperature. The heat of vaporization is assumed to be negligible. HDPE does not actually vaporize, but rather pyrolysis. The activation energy of the reaction is 65.3 kJ/mol, which, for an average of molecular weight 100,000 g/mol, corresponds the heat of reaction around 0.028 BTU/lbm of HDPE.

Table 11. Thermodynamic parameters for HDPE[12].

Parameter	Value
Heat Capacity, Cp	0.597 BTU/lbmF
Melting Point	266°F (130°C)
Heat of Melting	76.8 BTU/lbm
Heat of Vaporization/Reaction	0.028 BTU/lbm

The dimensions of the reactor were determined by modeling the shape of the reactor as a cylinder and cone, each section being equal in height.

$$V = \frac{\pi D^2 H}{6} \qquad (7)$$

To determine total volume (V), a density of 0.070 lbm/ft^3 was assumed for the steam and vapor hydrocarbons in the reactor. This density was determined based on steam tables[13] and the design pressure and temperature.

$$V = \frac{F_{steam} + F_{HDPE}}{\rho} t_{residence} \, , \qquad (8)$$

where ρ is density, F_{steam} is the mass flow rate of steam, and F_{HDPE} is the mass flow rate of HDPE.

The results for reactor sizing and operating conditions are shown in Table 12.

Table 12. Specifications of Pyrolysis Reactor

Process Variable	Value
Operating Pressure	44 psig
Operating Temperature	930°F (500 °C)
Volume, V	6,875 ft^3
Diameter, D	**17.3 ft**
Height, H	**43.3 ft**
Reactor Thickness	2.0 in
Steam Pressure	132 psig
Steam Temperature	1650 °F (900 °C)
Steam Flow Rate	265,000 tons/yr
Steam Velocity	2.0 ft/s

The reactor diagram based on these specifications is shown in Figure 3.

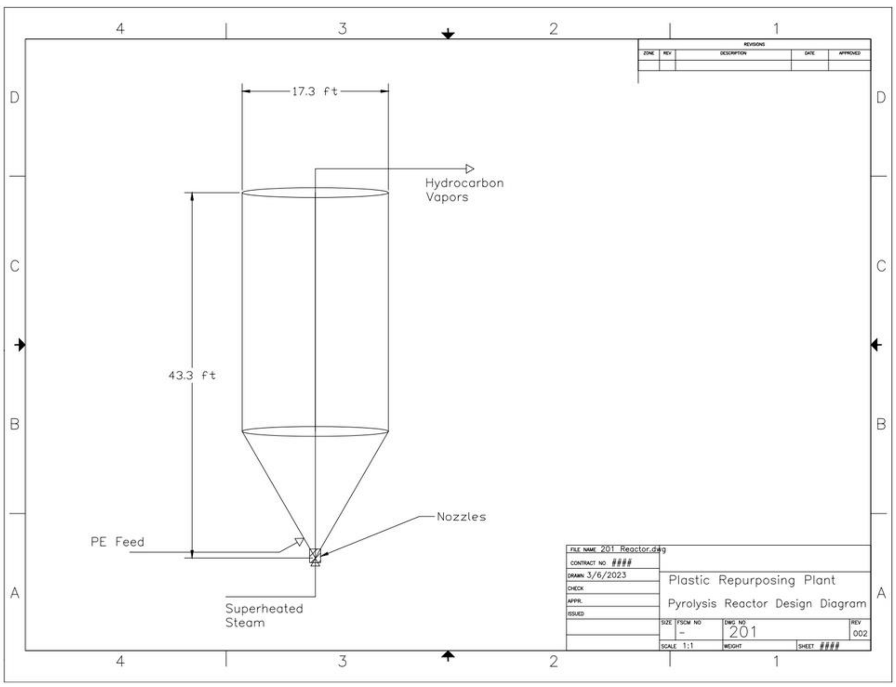


Figure 3. Pyrolysis Reactor Design Diagram.

Aspen Plus Simulation

The Aspen Plus V11 software was used to model the process. All units upstream of the column of the reactor could not be modeled in Aspen Plus because these units had purely solid HDPE feed. N-pentacostane ($C_{25}H_{52}$) used to represent waxes. PetroFrac unit operation was used to model the column. The fired heater boiler did not have a proper unit in Aspen Plus, so an RStoic unit with a HeatX unit were used as a substitute.

Additionally, although a hydrocyclone unit exists in Aspen Plus, it does not work for oil-water separations. Therefore, a generic separation unit was used to model the deoiler unit. Aspen does not have a membrane separator unit either, so the RO unit was modeled this way too.

The process layout of the simulation is shown below in Figure 4. Note that the reactor is modeled separately from the rest of the process. An empty vessel at 930°F was used to model the reactor since HDPE is not specified in the software.

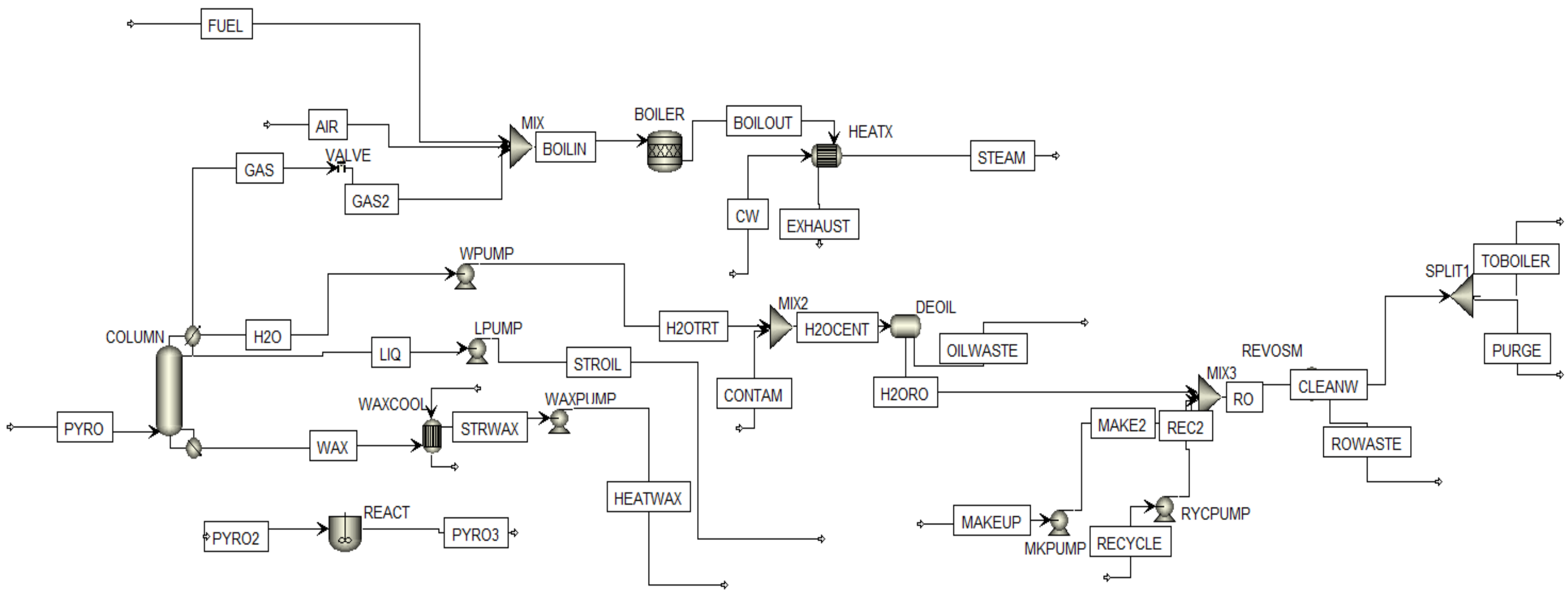


Figure 4. Process Flow Diagram in Aspen Plus V11.

We selected the Peng-Robinson equation of state model for the simulation based on the AspenTech manual (see Figure 5).

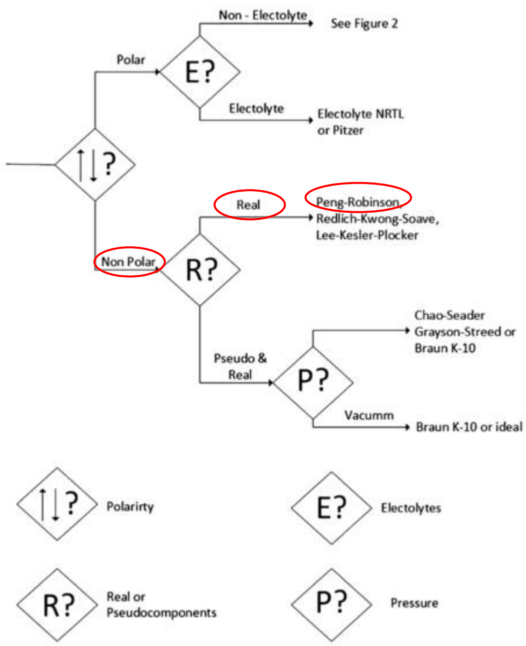


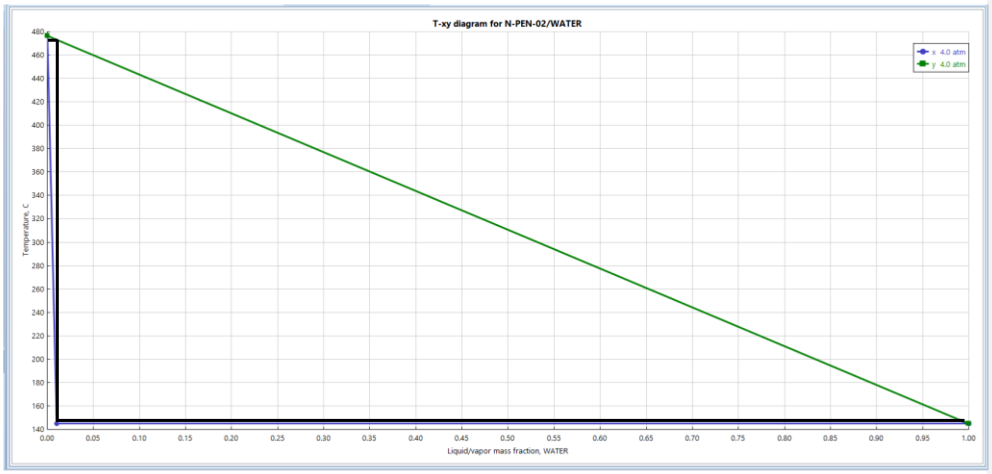
Figure 5. Modeling Methods in Aspen Plus V11.

Fractionation Column

The stream from the reactor will be separated into four different product streams in the column. This includes the bottoms wax product, and three distillate streams. The reflux drum will serve as a three-phase separator. The three phases separated in the drum are liquid water, liquid hydrocarbon, and vapor hydrocarbon streams. Both the column and reflux drum will operate at the same pressure as the reactor (44 psig).

The process stream from the reactor will be fed to the bottom stage. This column will not have a reboiler; instead, the liquid stream from the bottom stage will be cooled and a fraction will be pumped around back into the bottom stage. This will quench the superheated feed. All boil up in the column will be provided by the superheated feed.

The Aspen Plus V11 software was used to model the column. The simulation was used to determine the required number of stages, reflux ratio, and column sizing. Since wax and water are the main components in the column feed, a Txy and yx diagrams were used to estimate the minimum number of stages (Figure 6).



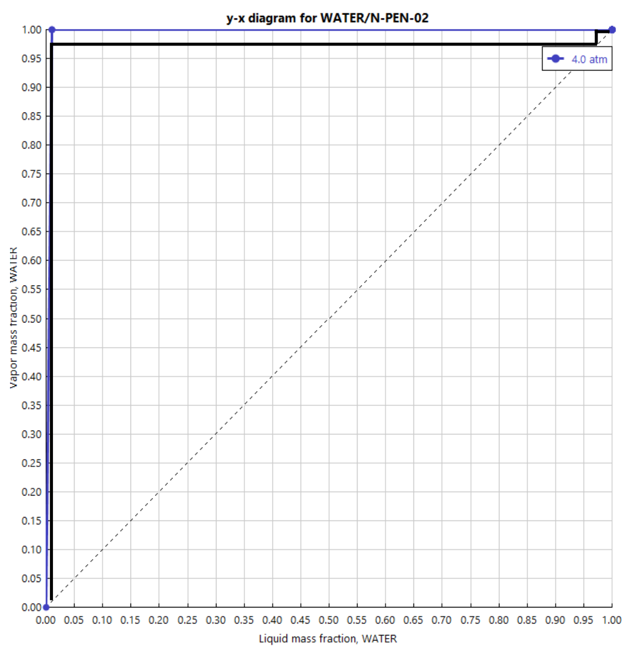


Figure 6. Txy and yx diagram for binary system of N-pentacostane and water at 44 psig (4 atm). Figures generated with Aspen Plus V11.

The diagrams indicate that a minimum of two stages will give product streams with 99wt%+ purity. We have designed the fractionation column to have three stages as a result.

The reflux drum is elevated to a height of 35 ft in order to maintain sufficient NPSHA for the pumps on the reflux, liquid water, and liquid hydrocarbon streams.

Table 13. Specifications of Fractionation Column.

Process Variable	Value
Operating Pressure	44 psig
Number of Trays	3
Feed Stage	3
Reflux Rate	27,900 tons/yr
Pump Around Rate	585,000 tons/yr
Bottoms Flow Rate	293,000 tons/yr
Vapor Hydrocarbon Flow Rate	6,330 tons/yr
Liquid Hydrocarbon Flow Rate	2,110 tons/yr
Water Flow Rate	264,000 tons/yr
Reflux Drum Temperature	120°F
Reflux Drum Diameter	3 ft
Reflux Drum Tangent to Tangent	9 ft
Reflux Drum Elevation	35 ft
Column Height	**16 ft**
Column Diameter	**20 ft**
Column Thickness	0.3125 in

High purity wax product can be achieved by this separation, with low toluene and benzene impurities. The stream results from Aspen Plus are summarized in Table 14. The liquid hydrocarbon and vapor hydrocarbon stream composition results are also summarized in Tables 15 and 16, respectively.

Table 14. Bottoms stream composition.

Bottoms Component	Weight Fraction
Wax	99.61%
Toluene	53.6 ppm
Benzene	9.93 ppm
Pentane	14.5 ppm
Butene	14.7 ppm
Water	0.376%

Table 15. Liquid hydrocarbon stream composition.

Liquid Hydrocarbon Component	Weight Fraction
Wax	53.9%
Toluene	28.8%
Benzene	5.16%
Pentane	5.52%
Butene	3.12%
Water	0.475%

Table 16. Vapor hydrocarbon fuel stream composition

Fuel Component	Weight Fraction
Methane	4.26%
Ethylene	30.1%
Ethane	6.10%
Propylene	17.1%
Propane	8.76%
Butene	15.5%
Butane	17.5%
Pentane	8.05%
Benzene	2.50%
Toluene	4.40%
Water	1.49%

Figure 7 shows the Fractionation Column and all associated units, including the reflux drum, air cooler condenser, bottoms cooler, reflux pump, and pumparound pump.

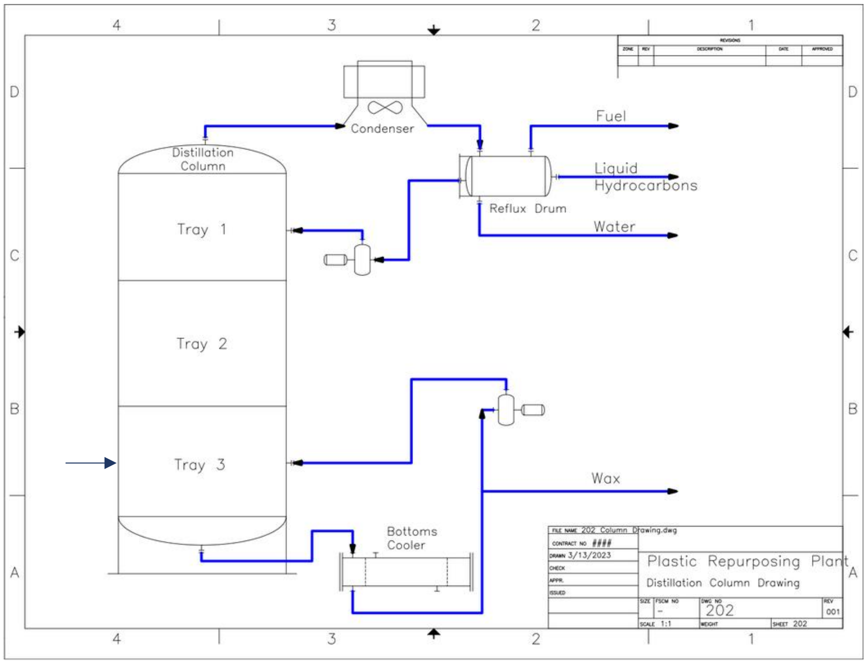


Figure 7. Process Flow Diagram of Fractionation Column unit.

<u>Aspen Plus Sensitivity Analysis on Column</u>

The effect of the number of trays on capital cost and revenue was investigated. In Aspen Plus, 8 trays or more gives “tray dried up” error. This error may have occurred due to the lack of significant amounts of material with intermediate boiling points (e.g., benzene and toluene) to fill the middle trays.

In this analysis, a price of $0.971/lb for wax and $0.22/lb for liquid hydrocarbon and off-specification wax was assumed.

It was found that increased stage count does not change the purity or flow rate of bottoms product, likely due to already very high separation factor between wax and all other components.

In fact, liquid hydrocarbon flow rate decreases slightly with increasing tray count, and vapor rate increases. Liquid hydrocarbon is more valuable than vapor, which is used for fuel anyway. Therefore, fewer stages is better. Table 17 summarizes the results of the stage count analysis.

Table 17. Sensitivity Analysis of Tray Count on Capital Cost and Product Value

Number of Stages	Equipment Cost	Installed Cost	Bottoms Purity (wt fraction)	Liquid Hydrocarbon Flow Rate (tons/yr)	Additional Product Value ($/yr)
3	596,700	1,787,800	0.9961	2,110	0
4	631,500	1,827,800	0.9961	2,103	-2,900
5	881,700	2,498,300	0.9961	2,103	-2,900
7	1,065,800	2,750,700	0.9961	2,103	-2,900

Next, an analysis of the effect of reflux ratio on capital cost and product value was conducted. We found that the utility cost was similar regardless of reflux ratio. This makes sense since the condenser cooling load wasn't changing significantly.

The "Tray dried up" error appeared again for reflux ratios above 20. It may be that there was not enough material to reflux in after that threshold.

The results of the reflux ratio sensitivity analysis are shown in Table 18. Note that the reflux ratio reported does not include the flow rate of water. In the Aspen Plus simulation, nearly all the water exits the reflux drum out the bottom and not through the reflux. By convention, Aspen reports reflux ratios in terms of hydrocarbon flow rates. The 1.13 ratio reflects the original design case reflected in Tables A1-A6 in the Appendix.

Table 18. Sensitivity Analysis of Reflux Ratio on Capital Cost and Product Value

Reflux Ratio (mol/mol)	Equipment Cost ($)	Installed Cost ($)	Bottoms Purity (wt fraction)	Bottoms Flow Rate (tons/yr)	Liquid Hydrocarbon Flow Rate (tons/yr)	Additional Product Value ($M/yr)
1.13	596,700	1,787,800	0.9961	292,740	2,110	0
0.25	623,900	1,817,700	0.9961	280,000	15,700	-18.76
0.37	623,800	1,817,600	0.9961	285,000	10,400	-11.38
0.61	608,800	1,825,400	0.9961	290,000	5,120	-4.00
1.96	611,800	1,827,800	0.9961	293,500	1,270	1.11
13.7	651,400	1,912,700	0.9957	294,000	715	1.83
16.9	644,300	1,913,400	0.9940	294,500	127	2.55

It was found that a reflux ratio of 2 gives an additional ~$1M/yr product value compared to the original case without decreasing wax product purity and only having ~$50K additional capital cost.

It is likely that our existing column design would be able to run at slightly higher reflux ratios and produce more product value than is considered in our economic analysis.

Flare

A flare will be used to handle overpressure scenarios and fugitive emissions. Sizing and design of the flare was conducted using nomograms provided by the National Air Oil group. The nominal tip diameter nomogram was designed using the equation,

$$d = \left[\frac{2.72*10^{-3}(W_1)\sqrt{\frac{t+460}{MW}}}{\sqrt{\Delta P_w}}\right]^{\frac{1}{2}} \quad (9)$$

The flare height nomogram was designed using the equations,

$$H = \sqrt{\frac{F * LHV *\epsilon}{12.56}} - 3.33d\sqrt{\frac{\Delta P}{55}}\cos\theta \quad (10)$$

$$\theta = \tan^{-1}\frac{1.47V_w}{550\sqrt{\frac{\Delta P}{55}}} \quad (11)$$

All process variables needed, along with flare height and nominal tip diameter, can be found in table 19.

Table 19. Process Variables for Flare Design

Process Variable	Value
Molecular Weight [MW]	44.0
Temperature [t]	130. °F
Pressure Drop [ΔP_w]	27.0 in. of H2O
Gas Flow Rate [F]	1,400 lbm/hr.
Nominal Tip Diameter	**2.0** in.
LHV	19,500 BTU/lb.
Emissivity [ϵ]	.13
Wind Speed [V_w]	15.0 mph
Flare Height	**21.0** ft

Fired Heater Boiler

All the heating load needed for the process will be met by the boiler. Information on the heat load and steam is given in Table 20.

Table 20. Boiler Unit specifications.

Process Variable	Value
Heat Load	115 MMBTU/hr
Steam Flow Rate	61,700 lbm/hr
Steam Temperature	1650°F
Steam Pressure	10 atm

Aspen Plus does not have a distinct unit for a fired heater boiler. A RSTOIC reactor with a HEATX unit were used in Aspen Plus. Therefore, Aspen HYSYS was used to model the boiler more accurately.

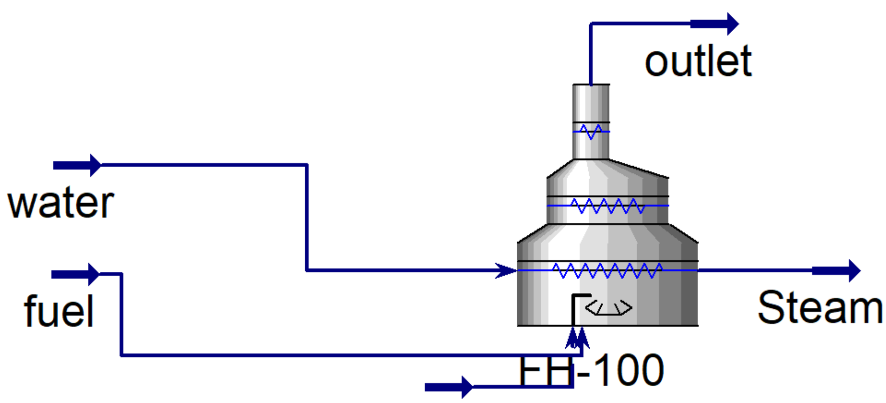


Figure 8. Aspen HYSYS simulation of the fired heater boiler. Fuel stream is composed of distillate vapor hydrocarbon from our process and additional natural gas (methane). The outlet is flue gas.

Table 21. Results of Aspen HYSYS simulation.

Process Variable	Value
Fuel Inlet Flow Rate	10,400 lbm/hr (45,500 tons/yr)
Inlet Air Flow Rate	215,000 lbm/hr (942,000 tons/yr)
Inlet Water Flow Rate	61,700 lbm/hr (270,300 tons/yr)
Steam Flow Rate	61,700 lbm/hr (270,300 tons/yr)
Flue Gas Flow Rate	226,000 lbm/hr (988,000 tons/yr)

The results indicate that an additional 8,940 lbm/hr (39,200 tons/yr) of natural gas is needed as fuel, on top of the 1,440 lbm/hr produced by the process. With a natural gas price of $0.07/lbm

(\$2.50/1,000 scf)[14], which is current as of mid-2023, this corresponds to a fuel cost of \$5.3M per year.

The flue gas is at a temperature of 1,760°F (compared to 1,690°F from Aspen Plus). The CO_2 mass fraction in this stream is 0.131, which corresponds to a CO_2 production rate of 29,400 lbm/hr (129,000 tons/yr).

Aspen HYSYS was also used to size the boiler unit. The results are shown in Table 22 and Figure 9.

Table 22. Sizing of Fired Heater Boiler.

Process Variable	Value
Radiant Zone Height	14.0 ft
Radiant Zone Diameter	6.6 ft
Convective Zone Height	5.6 ft
Convective Zone Diameter	4.7 ft
Economizer Zone Height	38.1 ft
Economizer Zone Diameter	1.6 ft
Radiant/Convective Zone Thickness	1.0 ft
Economizer Zone Thickness	0.33 ft
Bottom Area	**137 ft^2**
Total Height	**57.7 ft**

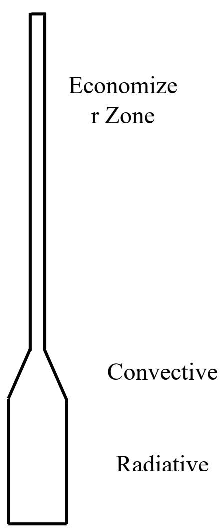


Figure 9. Diagram of relative sizes of each section of fired heater boiler.

There will be a backup boiler such that the process can continue to run while maintenance is being performed on the other, or some other disturbance occurs.

Heat Exchangers

There are three cooling units in the process. The first is a condenser on the top of the fractionation column. The second is a cooler for the column bottom pump around. The third is a cooler for the wax product stream before it is put into storage tanks. The three coolers are summarized in Table 23. The condenser is an air cooler and has a design air temperature of 100°F. The condenser outlet temperature was set at temperature of air coolant + 20°F. The bottoms and wax coolers will use process cooling water (CW) as the coolant. The heat load of the condenser, bottoms cooler, and wax cooler were calculated using the Aspen simulation.

The two heating units in the process, excluding the fired heater boiler (which is also a utility), are the screw pump heater and the heated wax storage tanks. The thermodynamic data in Table 11 was used to calculate the heat load of the screw pump heater.

There will be steam coils in the wax storage tanks that keep the material a liquid at 150°F. We assumed fiberglass insulation of 2 inches. The heat load was set equal to the rate of heat loss, calculated with the following equation.

$$Q = \frac{k\pi R^2 (T_{storage} - T_{air})}{L}, \qquad (12)$$

Where $L = 0.17\ ft$ is fiberglass thickness, $k = 0.045\ W/mK$, $T_{storage} = 150$°F, $T_{storage} =$ 100°F, and $R\ (radius\ of\ tank) = 25\ ft$.

Table 23. Heat exchanger specifications.

Unit	Heat Load (MMBTU/hr)	Pressure (psig)	Outlet Temperature (°F)	Working Fluid
Condenser	72.9	44	120	Air
Bottoms Cooler	41.1	44	487	CW
Wax Cooler	13.8	44	150	CW
Screw Pump Heater	15.0	44	350	Steam
Heated Wax Storage Tanks	0.27	0	150	Steam

Technology for Heat Exchange Network (HEN)

We have used pinch technology to analyze the heat exchange network of our facility.

The boiler was not considered in this analysis because it is both a process and utility unit. Furthermore, the heat load of the boiler is mostly used for superheating the reactor inlet steam, so there is little opportunity for heat recycling.

Additionally, the condenser on the column overhead was not included due to potential issues with operation during start-up. The same applies to wax storage tanks.

The pinch temperature difference was set at ΔT=20°F. Only the bottoms cooler, wax cooler, and screw pump heater were considered in the HEN analysis. The heating and cooling composite curves of the HEN are shown in Figure 10.

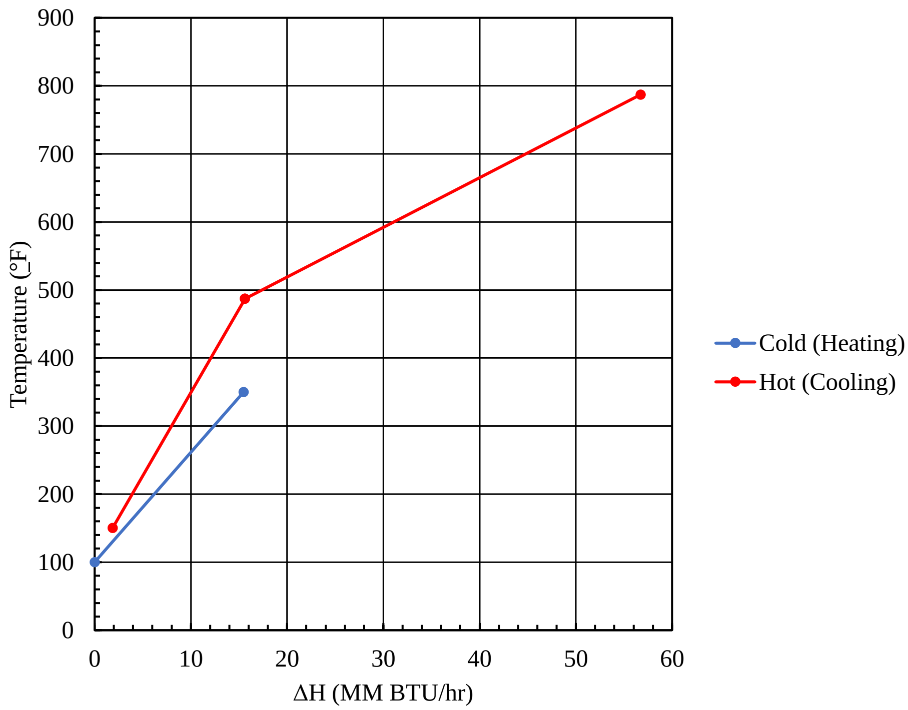


Figure 10. Hot and cold composite curves.

From Figure 10, we can determine the minimum cooling load given our specifications, $Q_{h,min}$ to be 41.2 MMBTU/hr. The minimum heating load, $Q_{c,min}$, is 1.86 MMBTU/hr. These values do not include the heating/cooling load for the boiler, condenser, or wax storage tanks.

We can interpret this to mean that most of the heat load of the screw pump cooler can be met by the heat recovered from the wax and bottoms cooler. Furthermore, we can see that the minimum cooling load is about the same load as the bottoms cooler. This demonstrates that significant external cooling is necessary regardless of heat recovery. This cooling will come from municipal cooling water.

Figure 11 shows how the screw pump heater, bottoms cooler, and wax cooler are connected in the HEN.

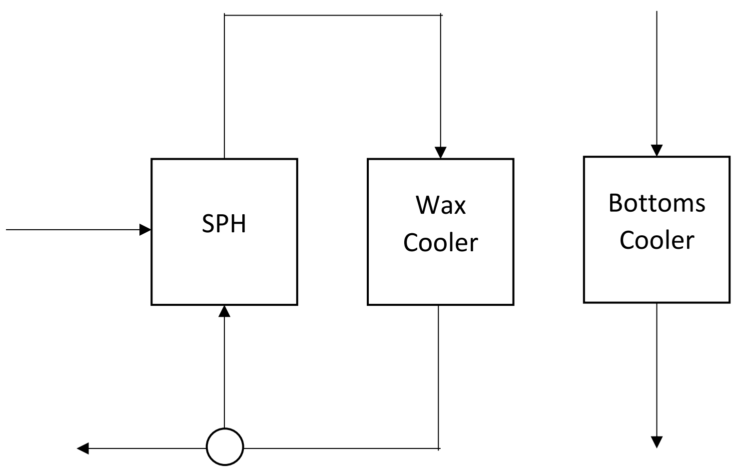


Figure 11. HEN Schematic of process water flow.

The heating load for the HEN will be met by 5,190 tons/yr of steam from the boiler. An outlet, at the same flow rate and but in liquid form, is sent to wastewater treatment. The steam through the wax storage tanks will also come directly from the boiler. The process water flow rate through each heat exchanger is given in Table 24.

The cooling load for the HEN will be met by 115,000 tons/yr of municipal water. This can provide all the cooling load for the bottoms cooler. Steam will be generated in the heat exchanger. It will be vented since there is no condenser unit to cool it or a turbine to generate electricity from it.

Table 24. Process water through heat exchanger.

Heat Exchanger	Process Water/Steam Flow Rate (tons/yr)
SPH	38,000
Bottoms Cooler	115,000
Wax Cooler	38,000
Wax Storage Tanks	320

Pumps

There will be a reflux pump and a pump-around pump. Additionally, each non-vapor product stream has an associated pump. The head of each pump was determined using the Aspen simulation. The efficiency of each pump is estimated to be 0.75. The Net Pressure Suction Head (NPSH) was calculated according to equations in the GPSA Engineering Handbook. The specific gravity and pressures were found from the Aspen Plus simulation.

$$NPSHA\ (ft) = \frac{2.31*(P_i - P_{vap})}{Specific\ Gravity} + z_i - z_{friction} \qquad (13)$$

$$NPSHR\ (ft) = NPSHA - 3ft \qquad (14)$$

A value of $z_{friction} = 3\ ft$ was assumed for the frictional losses. A margin of at least 3 ft is allowed between the NPSHA and NPSHR. P_{vap} was assumed negligible except for those pumps on streams coming out of the reflux drum. In those cases, $P_{vap} = 44\ psig\ (4\ atm)$. The specifications of the pumps are given in Tables 25a and b.

Table 25a. Specifications of pumps in process.

Unit	Flow Rate (tons/yr)	Flow Rate (GPM)	Pressure Change (psi)
Reflux	27,900	20.4	10
Pumparound	585,500	348.7	10
Column Water	263,600	121.6	44
Liquid Hydrocarbon	2,110	1.41	44
Wax	292,700	158.2	44
Makeup	6,700	3.42	10
Recycle	5,510	2.81	10

Table 25b. Specifications of pumps in process.

Unit	Pump Type	DH-SH (ft)	NPSHA (ft)	Power (HP)
Reflux	Centrifugal	34.1	32	1.50
Pumparound	Centrifugal	30.2	175	3.58
Column Water	Centrifugal	103	32	5.00
Liquid Hydrocarbon	Centrifugal	150	32	0.12
Wax	Centrifugal	133	175	7.50
Makeup	Centrifugal	24.2	31	0.12
Recycle	Centrifugal	24.1	31	0.12

Centrifugal was selected as the type for all pumps in this process due to their relatively low flow rates and head.

There are 7 unique pumps in total. Each will have a backup, such that the plant can still operate if one stops operating or needs maintenance.

Process Control

To ensure a consistent and safe process, process control will be used.

Reactor Control

The most important variable that need be controlled within the reactor is the temperature, as it determines reaction rate and product yield. Steam flow cannot vary as fluidization velocity must stay constant, therefore the flow of HDPE going into the reactor was varied to control heat transfer and thereby the overall temperature. The process control loop consists of a thermocouple sending any disturbances from the set point, 930°F, to a flow controller. A signal is then sent to the variable star feeder, which is able to vary its speed based on the current it is sent. In the case of the reactor's temperature being too high, more feed will be sent. This will give more volume/mass for heat transfer. The opposite will happen in the case of a temperature drop.

Fractionation Column Control

The variables to control in the column would be temperature, pressure and level, this would ensure the column is separating properly and safely. To control temperature, the column will utilize the pumparound and bottoms cooler. A thermocouple will be placed on or near the bottom stage and any disturbance from the set point will send a signal to a flow controller placed on the bottoms cooler cooling water stream. The flow controller will then modulate the control valve for the cooling water and decrease or increase the quenching of the feed stream.

For pressure, a pressure gauge will be placed at the vapor outlet for the column. In the case of deviation from set point pressure, the pressure controller will send a signal to the air cooler condenser and the cooling rate will increase (if overpressure) or decrease (if pressure drop).

To avoid flooding, a level controller will be placed within the column towards the bottom. Any deviation will send a signal to a flow controller on the bottoms product stream. This flow controller will then modulate a control valve on the same stream. In the case of a rise in level the valve will open increasing flow out of the column.

Reflux Drum (3-phase Separator)

The variables to control in the reflux drum would be temperature, pressure and level, this would ensure the separator is separating properly and safely. Temperature will be used to ensure cooling/condensing is properly taking place. To do this, a thermocouple will be placed within the drum and will send a signal to the condenser in the case of a deviation in temperature. If the temperature rises, the air cooler's fan speed will be increased.

Due to the reflux drum working with large vapor loads, it is important to control pressure in the drum for safety. To accomplish this, a pressure gauge will be placed at the vapor outlet, and this will control a control valve on the vapor outlet stream. If pressure rises too high the control valve will open more and increase flow rate out.

To ensure proper separation of the liquid hydrocarbons from the water, level control is necessary. The drum will utilize two level controllers, one at the liquid vapor interface, and one at the liquid-liquid interface. The liquid-liquid level controller will be placed on the left side of the weir and will ensure the water does not flood into the hydrocarbons. If the level rises too high, a signal will be sent to a flow controller on the water outlet stream and a control valve on the stream will then increase its opening and thereby increase outlet flow. If the liquid level of the hydrocarbon raises too high, a signal will be sent to a flow controller on the hydrocarbon outlet stream and a control valve on the stream will open more to increase hydrocarbon flow out.

Product and Feedstock Storage

A storage capability of 10 days of production is planned. Wax will be stored in heated vessels to keep it liquid. The storage tanks for wax and liquid hydrocarbons are summarized in Table 26.

Table 26. Specifications of Product Storage Tanks.

Specification	Wax Storage	Liquid Hydrocarbon Storage
Tank Type	Fixed Roof	Floating Roof
Blanketing	N_2	None
10-day production (ft^3)	286,000	2,373
Number of Tanks	3	3
Volume per Tank (ft^3)	95,200	791
Volume per Tank (gal)	712,000	5,920
Tank Height (ft)	48	16
Tank Diameter (ft)	50	7.9
Operating Temperature (°F)	150	100
Material of Construction	CS with Fiberglass	CS

A bund around the product storage tanks is necessary for spill prevention. The sizing of the bund is set to be,

$$A_{bund} = 1.1 * \frac{V_{Largest\ Tank}}{H_{bund}} \quad (15)$$

where the height of the bund, $H_{bund} = 2\ ft$ and the area of the bund $A_{bund} = 52{,}300\ ft^2$. A schematic of the bund surrounding the storage tanks is shown in Figure 12.

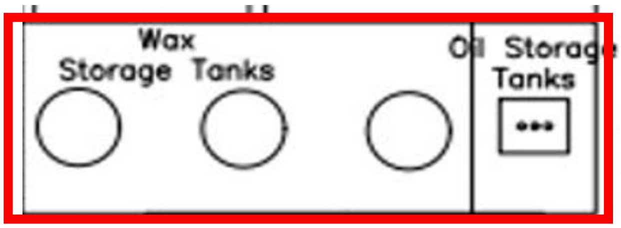


Figure 12. Bund around the product storage tanks.

The PE storage warehouse will also have 10 days of production. For a stack height of 30 ft, and assuming that only 60% of the warehouse area is used to stacking HDPE pallets, a total area of 15,900 ft^2 (0.37 acres) is required for the warehouse.

Materials of Construction

Some possible Corrosion Factors include water pH in the process. However, the Reverse Osmosis Unit will regulate pH of process water so this is unlikely to be an issue. Furthermore, there would be little influence of oxygen in the process since it would be excluded by the starfeeder. Only oxygen dissolved in water could have an effect.

The biggest corrosion and mechanical stress risk is temperature, since significantly high temperatures will be present in some of the process. This includes the Reactor, Boiler, and Fractionation Column most at risk.

Most process units will be constructed out of Carbon Steel (CS) whenever low corrosion risks and operating temperatures are present.

There will be several exceptions. First, the fired heater boiler can get up to 1760°F and contains oxygen; therefore, it will be made from SS-309. SS-309 is oxidation resistant up to its maximum temperature. The outlet piping from the boiler to reactor will also be made of SS-309. The reactor will be made of SS-309 as well, since it could reach a similar temperature during start up when the only inlet is steam at 1650 °F. Additionally, the column feed is at 930°F; therefore, it will be made of SS-304. The different alloys of steel and their allowable operating temperatures are shown in Table 27.

Table 27. Operating Temperatures of Steel Alloys[15,16,17].

Alloy	Upper Operating Temperature (°F)	Approximate Relative Cost
CS	650	1
SS-304	1600	4.5
SS-316	1700	6.5
SS-309	1900	10

Corrosion rates were estimated for various sections of the process (Figure 13). Experimental data on freshwater corrosion from Royani et al., 2019 was used. For stainless steel it is 0.02 mm/yr, and for carbon steel it is 0.50 mm/yr [18,19]. However, due to the reverse osmosis unit in our water treatment section, the actual process water will be purer than that in the experiment and corrosion rates will be lower.

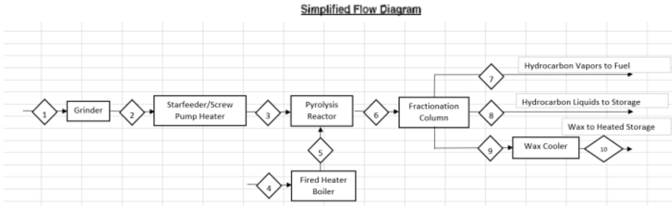
Requisition/Order No.

1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22	23
Equipment Stream or Line No.	Temperature °C		Pressure kPa(g)		Fluids & Corrosive Contaminants	Wt % Vapour	Vol. % H2S		H_2 Partial Press kPa(a)		Wt. % S in Liquid/Vapour		Wt. % Water	Material Required for H_2 Attack (Note 5)	Corrosion Rate mm/year			Mat'l for Col.18	Final Mat'l Selec'n (Note 9)	Corr'n All'nce mm(in)	Est'd Life Years	
	Oper. (Note 11)	Design	Oper.	Design			Oper.	Design	Oper.	Design	Oper.	Design			C.S. (Note 6)	Corr'n Mat'l Col 15 (Note 7)	Higher Alloy (Note 8)					
1	38	38	0	0															CS		20	
2	38	38	0	0															CS		20	
3	177	177	303	303															CS		20	
4	38	38	0	0									100		0.50				CS	3	6	
5	960	960	909	909									100			0.02			SS309	3	20	
6	500	500	303	303									46.9			0.02			SS304	3	20	
7	49	49	303	303									1.49		0.50				CS	3	6	
8	49	49	303	303									0.476		0.50				CS	3	6	
9	253	253	303	303									0.376		0.50				CS	3	6	
10	66	66	303	303									0.376		0.50				CS	3	6	

Notes

1. This form MUST be completed as required in GPS-A9.
2. Insert simplified flow diagram in space provided. Number all streams or lines. Name and number all process equipment.
3. For heat exchangers make a five line entry covering:
   a) Shell and shell cover.
   b) Channel and channel cover.
   c) Tubes - In Column 21 record minimum thickness in mm.
   d) Tube sheets and floating head.
   e) Baffles, tie rods, etc.
   Complete columns 2-21 for a) & b)
   Complete columns 16-21 for c)
   Complete columns 18-21 for d) & e).
4. If corrosion allowance exceeds 6mm on a side use more resistant material.
5. Where H_2attack is a design consideration enter the lowest alloy material required to resist this attack. When H_2attack is not a consideration enter a (-).
6. Where a dash (-) or C.S. has been entered in Col.15 enter the corrosion rate for C.S. in Col.16.
7. Where the material entered in Col.15 is other than C.S. enter the appropriate corrosion rate in Col.17.
8. Where the corrosion rate entered in Col's 16 or 17 is unacceptable enter in Col.18 the corrosion rate for the most economic higher alloy steel, and show this material in Col.19.
9. Selection should include consideration of solid vs. clad material for corrosion protection but backing material must be equal to or better than material listed or implied in Col.15. List both cladding and backing material in Col.20 if clad material is selected. Also for service at very high temperatures an alloy steel may be selected or high creep strength or to resist graphitisation or oxidation even where carbon steel or low CR.HO. steel may otherwise appear to meet the requirements for resistance to corrosion and hydrogen attack.
10. Col.22 is to be completed for all lines and equipment. Where life is 10 yrs or more, record 10 yrs. If less, record actual calculated life.
11. Enter both min. and max. operating temp's where stress corrosion, cracking or brittle fracture can influence material selection.

Simplified Flow Diagram



CES 98

Notes and References

Sources: GPS-A9

Issue Approval

Designer | Owner Approval

Completed by: James Hayes

Checked by:

Approved by:

CALTEX

Material Selection Basis For

Approvals Jan'95

Drawing No. | Sheet No. of | Rev.

STANDARD FORM No. SF-A-612

Figure 13. Estimated Corrosion rates for various process streams.

Utilities

A water treatment facility will be necessary for meeting the needs of the facility. A RO and de-oiler unit will be used for this purpose. Municipal water from the city of Houston will be used for our makeup water flow and external cooling water flow. Prices for municipal water shown in Table 28 reflect typical Houston values for our required flow rate.

Heat from burning hydrocarbon vapor products and natural gas will be used to vaporize the water into steam within the boiler. Out of the 10,400 lbm/hr of fuel demand, 8,940 lbm/hr will have to be purchased. The steam generated in the boiler will be used for transferring heat throughout the process.

Electrical utilities are also important for the operation of the plant. Pumps and fans in the facility may run on utility electricity, although they could also be powered by on-site generators if necessary. The office, laboratory, and plant operator buildings will also have an electrical demand.

The Aspen Plus Process Economics Analyzer indicates an electrical load of 215 kW for the process section. The simulation does not include the grinder, star feeder, conveyer, and screw pump heater. This adds an additional 278 kW of electrical load. The electrical load for the office, laboratories, and control building were estimated assuming an average value of 5 W/ft^2. This gives a load of 36.8 kW. The estimates cost of electricity based on typical Baytown, TX prices.

Nitrogen gas will be bought from a local facility, such as the Air Products facility in Baytown. The gas will be brought by trucks from the facility to ours.

The majority of this nitrogen will be used for wax storage blanketing at 0 psig (1 atm). The volume of nitrogen gas required is equal to the volume of product removed from the storage tanks. A pressure control system will be used to control the nitrogen blanket in the wax storage tank. As wax is added to the tank and the vapor space decreases, a valve opens to allow N_2 to vent to atmosphere. As wax is removed and the vapor space increases, another valve is opened to allow N_2 into the tank. Figure 14 depicts this process.

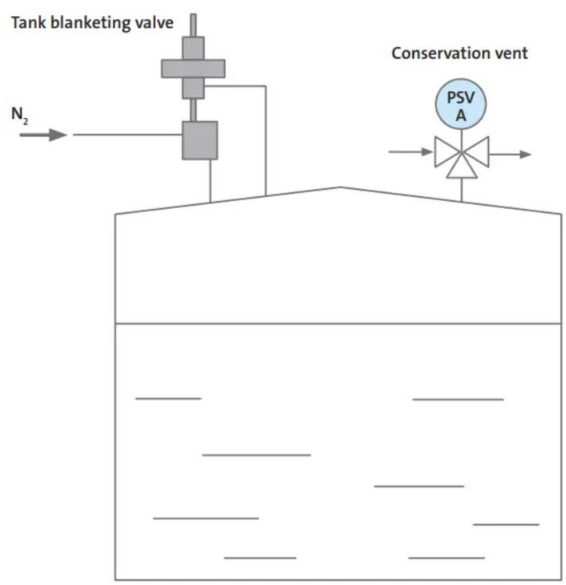


Figure 14. Pressure control system for nitrogen blanketing of wax storage tanks.

For process purging, nitrogen at 44 psig (4 atm) will be used. As a rule, 1.2 times the vessel volumes were used to estimate required nitrogen for purging. This corresponds to 14,400 ft^3 (407 m^3) per process purge. The nitrogen will be bought already pressurized from the vendor. Estimated utility costs shown in Table 28 are based on estimates from the Purity Gas company.

The utility demands are summarized in Table 28.

Table 28. Utilities for the Plastics Repurposing Plant and their costs [20,21,22,23,24]

Utility	Cost	Demand	Yearly Cost
Municipal	$0.0215/gal	121,000 tons/yr	$699,000
Electricity	$0.14/kWh	528 kW	$648,000
Steam	N/A	270,300 tons/yr	N/A
Natural Gas	$0.07/lbm	10,400 lbm/hr (45,500 tons/yr)	$5,270,000
N_2 Gas	$0.013/scf ($0.47/m^3)	10,400,000 scf/yr (295,000 m^3/yr)	$139,000

Water Treatment

To remove hydrocarbons and minerals from the process water, the water pre-treatment will include a de-oiling hydrocyclone typical of oil & gas plants to remove hydrocarbons. A de-oiler uses centrifugal force and differences in density to separate components[25].

Following this, the water will proceed through a reverse osmosis (RO) system to remove any additional components, especially dissolved metals and minerals like calcium. Both the water from the municipal and the water recycle will be sent through the water treatment section before being sent back to the fired heater boiler.

The de-oiler was modeled as a decanter in the Aspen Plus Simulation, and the RO unit as a generic separator. The sizing and specifications from the simulation were disregarded since they do not represent the actual unit.

A flow rate of 270,300 tons/yr of water will be sent through the water treatment process for the boiler needs. An additional 115,000 tons/yr of water will be sent through for cooling water needs. This corresponds to a total flow rate of 175.9 GPM (21.9 ft^3/hr) at the operating temperature and pressure.

Commercial De-oiler cyclone units that could meet our needs were investigated. Two possible candidates for our process include the FX200 by Henan Xingyang Mining Machinery Manufactory[26], which has a capacity of 25-40 m^3/hr (110-176 GPM), and the G800 MF-300 Series by Shanghai Saga Offshore Engineering Co., Ltd [27]. The specifications of the de-oiler unit are shown in Table 29, and a typical deoiler unit and tube are shown in Figure 15.

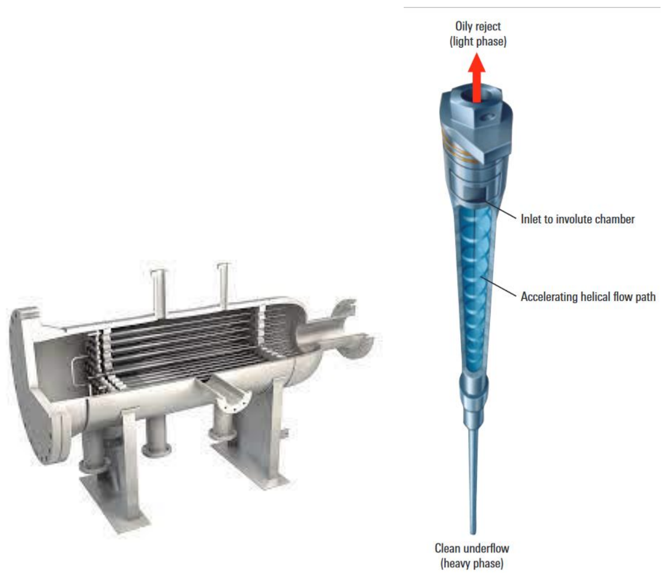


Figure 15. De-oiler tube and unit[28].

Table 29. Specifications of de-oiler unit.

Parameter	Value
Operating Pressure	10 psig
Operating Temperature	120°F (49°C)
Operating Flow Rate	176 GPM
Diameter of Tubes	0.656 ft (200 mm)
Number of Tubes	100
Hydrocarbons in Outlet	<10 ppm

Commercial reverse osmosis units that could meet our needs were investigated. A unit which could meet the needs of our process is the MW-190K-6580 by Shaanxi Aps Machinery[29]. The specifications of our RO unit are shown in Table 30.

Table 30. Specifications of RO unit.

Parameter	Value
Operating Inlet Pressure	10 psig
Operating Temperature	113°F (45°C)
Base Area	118 ft^2
Height	7.5 ft
Operating Flow Rate	132 GPM
Membrane Array	4:2
Membrane Quantity	30
Motor Power	30 HP (22.1 kW)
Dissolved Minerals in Outlet	<50 ppm

Duplicates of each water treatment unit will be present in the process.

Commissioning

Prior to start-up, safety teams will be commissioned to implement and test all control systems, instrumentation, and emergency functions throughout process equipment. Ensuring these systems are installed and functioning properly is imperative to a successful startup. Some of the activities will include instrumentation calibration (Sensors, Transmitters, Analyzers, etc.), inspecting fire protection, shut-down controls, and pressure relief controls.

Inspection will also be done on piping. Hydrostatic pressure tests for operating conditions prior to startup will be conducted. The pipes will also be flushed and cleaned with nitrogen gas after these tests, and again prior to start-up. Electrical Systems

With regards to electrical equipment, inspections will be performed after construction, on testing insulation resistance, continuity, and function of all components and wiring prior.

The product storage tanks must be tested for leaks after installation and prior to start-up. This will be accomplished by filling them with water. Since volume of water required is very great, only one tank will be filled at a time, and water will be reused to test the next tank. At typical municipal water price of $0.0215/gal and the largest tank volume of 712,000 gal, the cost will be $15,300 to perform this test.

Start-up

The water treatment facility will start up first since it needs to be treating water at all times during operation. Ideally, it would continue to operate even if the rest of the facility is shut down. A larger flow rate of water will also need to be purchased from the municipal during start-up and sent to the water treatment section.

A nitrogen purge will be used to flush oxygen out of the process. It will be used to pressurize the reactor to 44 psig.

The fired heater boiler will start up next since it provides all the heat load for the process. Steam will be run through the process before HDPE feed is introduced, and steam will also go to the Screw Pump Heater.

The Fractionation Column Start-up will begin by raising its pressure to the operating pressure (44 psig) with the steam. Then the fan on the air cooler will be started and cooling water will be circulated through the bottom cooler. The recycle pump can then be started. Next, a feed of wax and water will be introduced. Some wax to use for feed during start-up will need to be purchased. At this point, the reflux and pumparound pump can be turned on. The reflux ratio and distillate flow rates can be adjusted to meet the desired specifications.

Finally, HDPE feed can be introduced into the process and the pyrolysis reaction can begin.

Shut-down

Shut-down will start by stopping the HDPE feed to the reactor and the natural gas feed to the boiler. This will involve shutting off the motor of the star feeder and screw pump heater. The Fired heater boiler and flare will need to continue to operate with any remaining hydrocarbon vapor product, such that all flammable material is burned. The steam flow rate out of the boiler will be significantly diminished after the natural gas and HDPE feed is cut.

The Fractionation Column shut-down begins once the feed rate is diminished as HDPE feed into reactor stops. The air cooler fan and circulation of cooling water through bottoms cooler and wax cooler will be shut down. The recycle pump can be shut off at this point. The feed to the column will be completely cut off, and liquids will be drained from the trays and reflux drum. The column will next be vented to atmosphere and returned to atmosphere pressure.

Finally, wax, liquid hydrocarbon, column water, and water makeup pumps will all be shut down.

If necessary, the wastewater treatment section can then be shut down. However, it may not be necessary to do so. Continued treatment of water even after shut-down may be both useful and will speed up the next start-up.

Environmental Considerations

The EPA's New Source Performance Standards (NSPS)[30] establish emissions standards for new, modified, and reconstructed sources of air pollution. The relevant NSPS regulations to our plant include:

- 40 CFR Part 60 Subpart PPP - Standards of Performance for New Stationary Sources: Polyethylene Terephthalate Resin Plants: This regulation sets standards of performance for new stationary sources of VOC emissions from polyethylene terephthalate (PET)

resin plants, which includes polyethylene pyrolysis plants. The specific quantitative regulations include:
  - A limit of 23 parts per million by volume (ppmv) or less of total organic compounds (TOC) in the vent from the polymer production process.
- 40 CFR Part 60 Subpart F - Standards of Performance for New Stationary Sources: Incinerators: This regulation sets standards of performance for new stationary sources of emissions from incinerators, which may be relevant to polyethylene pyrolysis plants that use incineration as a method of waste disposal. The specific quantitative regulations include:
  - A limit of 180 milligrams per dry standard cubic meter (mg/dscm) or less of particulate matter (PM) emissions.
  - A limit of 150 parts per million by volume (ppmv) or less of carbon monoxide (CO) emissions.
- 40 CFR Part 60 Subpart OOOO - National Emission Standards for Hazardous Air Pollutants: Industrial, Commercial, and Institutional Boilers and Process Heaters: This regulation sets emission standards for hazardous air pollutants (HAPs) from industrial boilers and process heaters, which may apply to polyethylene pyrolysis plants that use boilers or process heaters. The specific quantitative regulations include:
  - A limit of 400 ppmv or less of nitrogen oxides (NO_x) emissions.
  - A limit of 9 ppmv or less of carbon monoxide (CO) emissions.

With respect to our plant, our waste output falls well below these EPA guidelines, assuming that no significant incomplete combustion occurs in the boiler.

Process Waste Management

Water that is utilized throughout the process will be recycled to our water treatment facility to remove any impurities and permit us to recycle it back to the boiler, where it will then be vaporized and used as a utility stream.

The water flow rate out of the process will be about 263,600 tons/yr. A makeup process water stream of 6,700 tons/yr will be combined with this stream and sent to our water treatment facility. Then 264,800 tons/yr of water will be sent to the boiler to be converted to superheated steam, from which it will go to the pyrolysis reactor. Furthermore, an additional 5,510 tons/yr of water will be recycled from the wax storage tank heater and the screw pump heater.

A water effluent steam in the form of a purge of 25 tons/yr will be sent to the Houston Ship Channel. The impurity in the steam is likely to be 10-40 ppm oil or less based on the capabilities of commercial de-oiler units, which will be used in our water treatment facility. Pentane, benzene, toluene, and wax may possibly be present in the effluent in the ppm level or lower.

The other waste stream in our process is the exhaust gas from the boiler. We are burning all of our 1,440 lbm/hr (6,330 tons/yr) of fuel product, in addition to another 8,940 lbm/hr (39,200 tons/yr). This combustion produces 29,400 lbm/hr (129,000 tons/yr) of CO_2 in flue gas, which is vented to the atmosphere. This is compared to ~50 million tons/yr from the Houston area[31]. The rest of the exhaust stream is mostly nitrogen, with some oxygen and water vapor. The exhaust gas temperature is 1760°F. Although it was not modeled in the Aspen Simulation, there could also be some hydrocarbons in the exhaust although they would likely be in the ppm level or lower. Furthermore, incomplete combustion could result in CO emissions or particulate matter in the form of elemental C, but are not present effects in our process.

Volatile Organic Compounds (VOC) emissions could be possible. The heavier VOCs like benzene and toluene are recovered and sold as oil product. Other VOCs will be recovered and sent to boiler as fuel. Some fugitive emissions of methane, ethylene, ethane, and other vapor hydrocarbon products are possible.

Nitrogen Oxides (NOx) and Sulfur Oxides (SOx) emissions are not likely to occur in any significant amounts. The HDPE feed is already highly processed, so unlikely that complete combustion of its pyrolysis products would release NOx or SOx. However, the natural gas feed could present issue if it is low quality. Furthermore, incomplete combustion could result in formation of NOx.

Our only liquid hazardous material is the synthetic crude oil, especially the waste from the de-oiler unit. There are many risks associated with our production and distribution of synthetic crude oil, namely that it is flammable and explosive. Following process safety protocols and proper maintenance is imperative to preventing spills/leaks or accidents. Any hydrocarbons present in the water stream will be removed via the de-oiler unit. Any of the synthetic crude oil hydrocarbons present with higher carbon chains will be separated and sold as off-spec. wax. Thus, the waste considerations are management by our internalized process safety protocols.

The vapor hydrocarbons, such as methane, ethylene, and propylene are also hazardous materials. The vapor hydrocarbon stream is recycled and combusted for heat energy in our process. The main byproduct of this specific process is carbon dioxide, which emissions management has previously been mentioned.

Up to this point, we have continuously worked to minimize any hazardous emissions in our design. When designing our separation and storage processes, our background goal was always to minimize environmental harm. We have maximizing materials and energy recycled. Future investigation into redirecting our CO_2 production can push this initiative further. Furthermore, our initial broad mission to minimize environmental harm now has the scaffolding of regulations. This allows us to set future goals to reduce emissions and improve our impact.

Fugitive Emissions

We will use a Leak Detection and Repair Program (LDAR). The purpose of this program is to reduce product losses, increase safety, decrease hazardous exposure to the surrounding community, and reduce emissions.

We will modify and replace leaking equipment with “leakless” components such as valves and pumps if necessary.

This includes: Bellows valves, diaphragm valves, Diaphragm pumps, canned motor pumps, magnetic drive pumps. Pumps can also include dual seals with or without barrier fluid.

The process for leak detection and response will include:

- Identifying components
- Leak definition
- Monitoring components
- Repairing components
- Recordkeeping

The program will also include a Training Program and Third-Party Audits.

Table 31. Sources of Equipment Leaks

Equipment	Common Leak Source
Pump	Seal
Valve	Stem, gland, O-Ring
Connector	Gasket failure, flanges
Sampling Connection	Outlet at purges
Compressor	Seal
Pressure Relief Devices	Worn seals, ruptured disks
Open-Ended Lines	Point open to the atmosphere

Personnel Protection

To protect our plant personnel from any minor hazards and from further damage in the event of a major hazard, we will require a variety of personal protective equipment (PPE) to be worn in the plant. All personnel on the plant floor (near equipment) will be required to wear non-conductive, fire-resistant clothing & boots. When working near or on equipment, similarly protective gloves must be worn also. Ear plugs will be provided at all entrances to the plant floor to reduce noise exposure to our personnel. All clothing and boots will be provided by plant management and is factored in our labor costs. In addition to training all personnel on the use of each piece of equipment that is relevant to them, we have a medical facility on-site with a doctor and two nurses to carry out annual health screenings and treat any issues arising from plant work.

Controlled Space

The will be a required video for visitors to our facility, particularly for entering the control building or ISBL area. There will be a separate video for office visits. All visitors must be accompanied by trained operator or engineer.

All vehicles going into the site must check in at security booth, and present ID and parking pass.

Positive pressure should be maintained in the control building. Specifically, pressure in the building should be kept 0.1 inches of water above air pressure. For reference, the highest recorded air pressure in Houston 31.06 inHg (422.7 inches of water). Inlet air louvres to the control building will be shut in a release event.

Our plant is a Class I Hazardous facility, i.e. flammable gases/vapors present. There are 4 types of equipment: explosion-proof, intrinsically-safe, purged & pressurized, non-incendive. We will use intrinsically-safe equipment, where the circuit is specifically designed to withhold the available energy below the level which could cause ignition.

Fire and Explosion Index

Dow Fire & Explosion Index[32] is calculated.

Table 32a. DFEI: Calculation of Material Factors.

Component	Flash Point (°F)	Boiling Point (°F)	N_R	N_F	MF
Water	-	212	0	0	1
Polyethylene	649	-	1	1	14
Wax and Tar	450	700	1	1	14
Ethylene	-213	-155	1	4	21
Propylene	-162	-53.7	1	4	21
Butylene	-110	20.7	1	4	21
Toluene	40	231	1	3	16
Pentane	57	97	1	4	21
Propane	-156	-44	1	4	21
Ethane	-211	-128	1	4	21
Methane	-306	-259	1	4	21
Benzene	12.2	176	1	3	16
Butane	-76	30.2	1	4	21
Xylene	90	280	1	3	16

Table 32b. DFEI: Calculation of Index.

Material Factor	21	
1. General Process Hazards	Penalty Factor Range	Penalty Factor Used
Base Factor	1	1
A. Exothermic Chemical Reactions	0.3 to 1.25	0.3
B. Endothermic Processes	0.20 to 0.40	0.20
C. Material Handling and Transfer	0.25 to 1.05	0.25
D. Enclosed or Indoor Process Units	0.25 to 0.90	-
E. Access	0.20 to 0.35	-
F. Drainage and Spill Control	0.25 to 0.50	-
General Process Hazards Factor (F1)		1.75
2. Special Process Hazards		
Base Factor	1	1
A. Toxic Materials	0.20 to 0.80	0.20
B. Sub-Atmospheric Pressure (<500 mm Hg)	0.5	-
C. Operation In or Near Flammable Range _X_Inerted __Not Inerted		
1. Tank Farms Storage Flammable Liquids	0.5	-
2. Process Upset or Purge Failure	0.3	-
3. Always in Flammable Range	0.8	-
D. Dust Explosion	0.25 to 2.00	-
E. Pressure Operating Pressure = 59 atm Relief Setting = 15 psig		0.35
F. Low Temperature	0.20 to 0.30	-
G. Quantity of Flammable/Unstable Material Total Combustible Energy: ~6000 MM BTU		2
1. Liquids or Gases in Process		
2. Liquids or Gases in Storage		
3. Combustible Solids in Storage, Dust in Process		-
H. Corrosion and Erosion	0.10 to 0.75	-
I. Leakage -- Joints and Packing	0.10 to 1.50	-
J. Use of Fired Equipment		-
K. Hot Oil Heat Exchange System	0.15 to 1.15	-
L. Rotating Equipment	0.5	0.5
Special Process Hazards Factor (F2)		4.05
Process Unit Hazards Factor (F1 x F2 = F3)		7.09
Fire and Explosion Index (F3 x MF = F&EI)		**149**

F&EI of the plant was determined to be 149. 128-158 is considered to be a "Heavy" degree of hazard.

Based on the F&EI, we can estimate the radius of exposure in Figure 16.

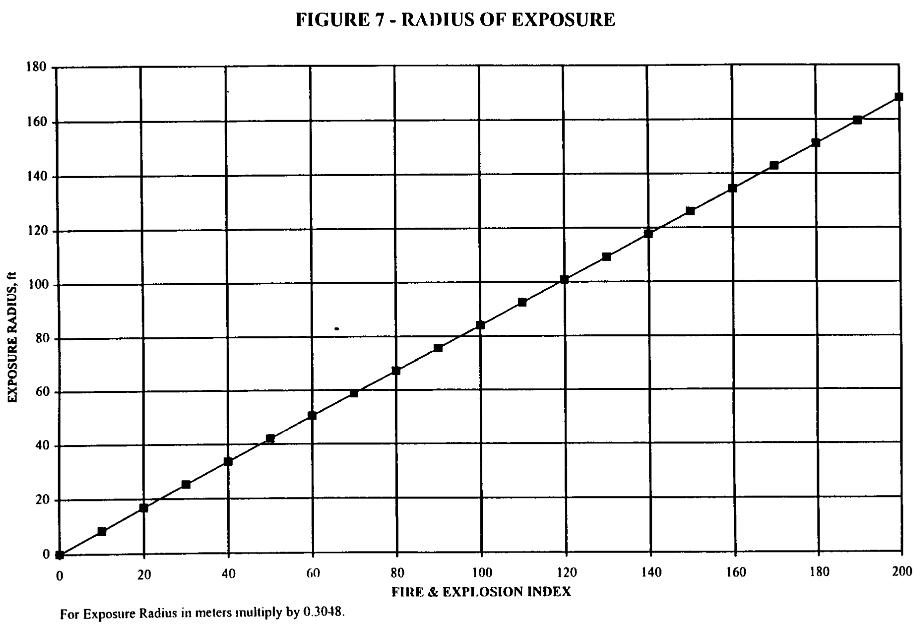


Figure 16. Radius of Exposure as a function of F&EI.

This indicates a radius of 125 ft, which corresponds to an area of ~1 Acre.

Fire Prevention

Our plant will manage fire prevention through a comprehensive seven-part program detailed below:

1. Conduct a fire risk assessment: identify possible fire risks, including all hot surfaces, electrical components, and any hot chemicals to be handles
2. Good housekeeping: Maintaining a clean plant and free of clutter to minimize the risk of a fire starting. Removing combustibles and keeping work areas clear promotes safety in all aspects of process.

3. Maintaining Equipment: Regular inspections and maintenance of all equipment and components identified in the risk assessment.
4. Fire Detection & Suppression Systems: Install detection systems throughout the plant to quickly detect a fire, and alert personnel. Install sprinklers or foam systems in especially high-risk areas.
5. Personnel Training: Train all personnel on prevention and response procedures, i.e., handling procedures, safe use of electrical equipment, and fire extinguishers
6. Emergency Response Strategy Implementation: Implement the plan detailed below for evacuation, equipment shutdown, and contact/coordination with emergency services.
7. Conduct regular fire drills: conducting scheduled and unannounced fire drills allows the plant to test the effectiveness of our prevention/response strategy and identifying areas for improvement.

Through this program, we can identify all fire hazards and how to prevent them. Through conducting regular drills of our response strategy, we can quantify what areas need improvement and how to make the plant a safer operating environment.

Critical Alarms

In analyzing potential emergencies or sudden changes to operating conditions, it is necessary to have critical alarms in place to initialize emergency protocols within the plant, critical to preventing severe accidents. Below are some scenarios, planned alarms, and operator intervention procedures:

Unique Weather Events:

- Alarms: Wind speed/direction, atmospheric pressure and temperature
- Operator Intervention: Shut down non-critical equipment, secure loose items, activate emergency procedures, evacuate if necessary

Loss of Cooling:

- Alarms: Temperature, pressure and flow rate of cooling water
- Operator Intervention: Shut down affected equipment, isolate cooling system, determine the cause and repair system

Loss of Electric Power:

- Alarms: Voltage, frequency, and current of all critical components
- Operator Intervention: Start emergency generators/back-up power, shut down non-critical equipment, determine the cause and repair the system.

Loss of Steam:

- Alarms: Steam pressure, temperature, and flow rate (particularly in the boiler and reactor)

- Operator Intervention: Shut down affected equipment, isolate steam system from process, determine the cause and repair the system.

Loss of Instrument Air:

- Alarms: Pressure and flow rate of instrument air
- Operator Intervention: Shut down non-critical equipment, isolate the instrument air system, investigate the cause and make necessary repairs.

Minor Fire:

- Alarms: Smoke and heat detectors, manual fire alarms
- Operator Intervention: Activate appropriate fire suppression system (sprinklers, CO_2), evacuate, if necessary, investigate the cause once fire is extinguished

Major Fire:

- Alarms: Smoke and heat detectors, manual fire alarms
- Operator Intervention: Activate fire suppression system, immediately evacuate the area and call the fire department. Follow emergency/evacuation procedures on where to go in order to account for all personnel.

Our greatest risk is a potential fire, as it poses the greatest risk to equipment and personnel. However, it is a uniquely unlikely event, the losing of cooling or electric power pose a lower severity but greater frequency, posing a similar threat to our plant. Thus defines the needs for these critical measures to maintain the health of our plant.

Emergency Response Strategies

We will be considering response strategies for the following emergency scenarios: Chemical Release, Fire, Explosion, & Natural Disaster. We have listed each of these scenarios along with the steps, in order, of how to respond to each.

1. Chemical release
   a. Identify the source and isolated the area.
   b. Evacuate personnel from affected area.
   c. Notify emergency response team & local emergency services.
   d. Monitor the area and establish a perimeter to prevent public exposure.
2. Fire:
   a. Activate the alarm and immediately notify emergency response team and local fire dept. Evacuate the area and if necessary, the entire plant.
   b. Ensure that all doors and windows are closed.
   c. Shut down equipment and attempt to isolate the source.
   d. Engage suppression systems, using extinguishers if possible.
   e. Establish a perimeter and monitor the area for hazardous chemical exposure.
3. Explosion:
   a. Activate the alarm and immediately notify emergency response team and local fire dept.

b. Evacuate the entire plant.
c. If possible, ensure that all doors and windows are closed.
d. Shut down all plant equipment and engage fire suppression systems to control any resulting fires.
e. Assess harm to personnel, and containment of hazardous chemicals.
f. Once all hazards are contained, begin cleanup procedures and assess plant/equipment damage.

4. Natural Disaster:
    a. Monitor weather alerts and take appropriate protection measures, such as shutting down equipment, or the entire plant depending on the reported severity of the event.
    b. Activate the alarm and notify the emergency response team to secure outdoor equipment and materials.
    c. Coordinate with local emergency services.
    d. Evacuate all personnel prior to the event.
    e. Following the event, assess all damage and initiate cleanup procedures.

Muster points will be established closest to the highest concentrations of personnel across the plant. This will include the parking lot for our offices and laboratory. Our response strategies are only for the immediate time frame after the event and intended to provide a safe transition to the procedures that local/state emergency services already have in place for incidents like this.

Site Security

Resources from the Cybersecurity and Infrastructure Security Agency (CISA) were used to plan our site security[33].

From these governing bodies, we found potential security issues included release situations, theft, and sabotage.

A release risk is described as, Toxic, flammable, or explosive chemicals or materials that can be released at a facility.

A theft risk/scenario is described as, chemicals or materials that, if stolen or diverted, can be converted into weapons using simple chemistry, equipment, or techniques.

A sabotage risk/scenario is described as, chemicals or materials that can be mixed with readily available materials.

All these risks can create dangerous scenarios and therefore, our plant must have security measures in place to prevent them. This could include prescreening of handlers and surveillance systems.

Cyber Security

Cyber security is governed by two groups, each of which having differing roles and responsibilities. These groups include,

- Cybersecurity & Infrastructure Security Agency (CISA)
    - Chemical Facility Anti-Terrorism Facility Standards (CFATS)
        - Focused on high-risk chemical facilities.
        - If the facility possesses any chemicals that exceed the screening threshold quantities (STQ) and concentration, it must be reported to the CISA.
    - ChemLock
        - Voluntary program which provides no-cost services to help the facility better understand the risks and improve chemical security to work with their business model.
        - Services include on-site assessments, documentation, exercises, training, and other resources.
- Occupational Safety and Health Administration (OSHA)
    - Chemical Facility Security and Safety Working Group
        - Has oversight of chemical facility security and safety and reduce risks with hazardous chemicals to workers and communities.
        - Resources
            - Community planning and preparedness
            - Enhancing federal operational coordination
            - Improving data management
                - Facility Registry Service (FRS)
                - Substance Registry System (SRS)
            - Modernizing Policies and Regulations
                - Joint EPA/OSHA/CISA Safety Advisory
                - OSHA Process Safety Management
                - Chemical Facility Anti-Terrorism Standard

Notable Systems to be protected are,

- Unit Process Control and Data Analytics
- Information Technology (IT)
    - Office software
    - Payroll information
    - Business process control

Programs and software that should be implemented within the plant are,

- Access Control Policies/Procedures Consultation
- Analysis and Detection
- Anti-Phishing – Cybercrime prevention
- Audit Log Monitoring
- Automated Indicator Sharing (AIS)

Location of Facility and Plot Plan

Our proposed facility will be near the Houston ship canal, on the east bank (Figure 17). The coordinates are 29°42'36.4"N 94°58'11.5"W. The address is Evergreen Rd, Baytown, TX 77523.

Figure 17. Location of Proposed Facility. Source: Google Maps

Up to about 0.5 mi^2 (320 acres) of open area could be available on this plot of land. Ultimately, only around 25 acres were used in the plot plan (Figure 18). The cost of land around the channel is typically around $100-130K/acre or greater.

A draft plot plan was designed in this plot. Sections were designated for the process itself, in addition to facilities to store the product and the PE feedstock, as well as for offloading of material or equipment. Space was also left aside for office/administration buildings. A laboratory building for testing samples of product and process streams was included. Sections for water and electricity utilities are also present.

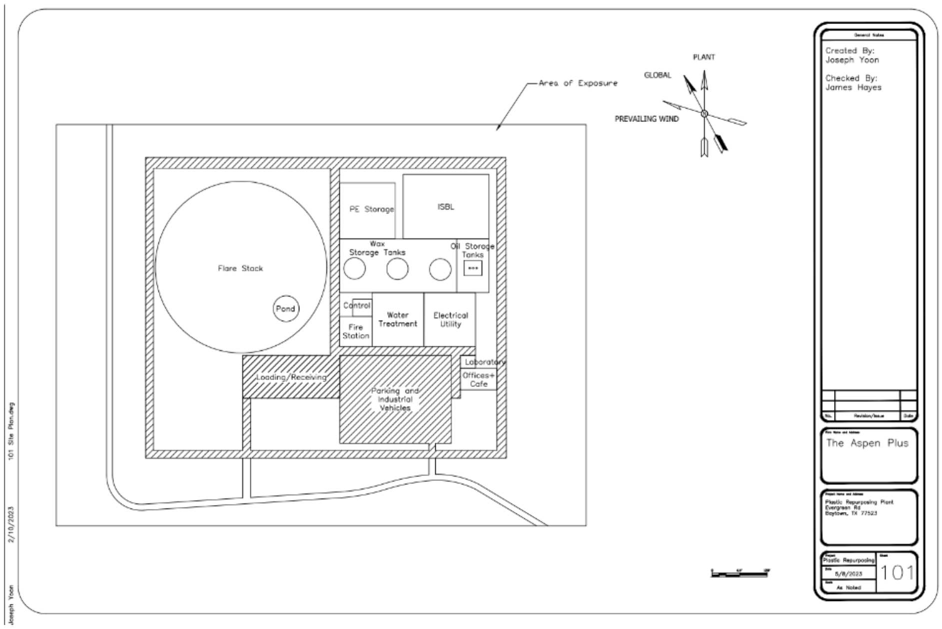


Figure 18. Preliminary Plot Plan for Plastics Repurposing Facility.

The layout of the ISBL section is shown in Figure 19, with equipment IDs shown.

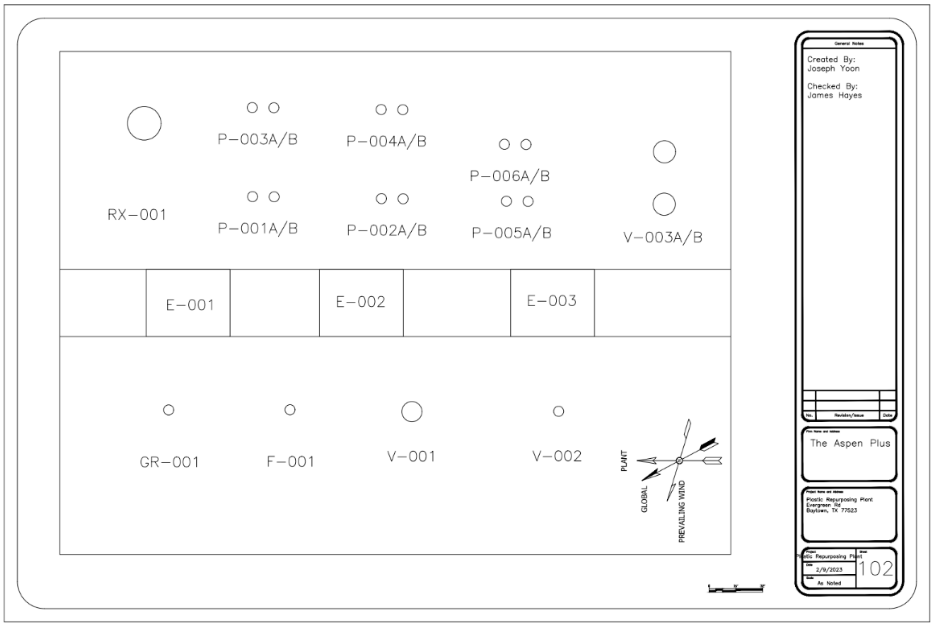


Figure 19. Plant ISBL Layout.

Table 33. Equipment IDs.

ID	Unit
RX-001	Pyrolysis Reactor
P-001A/B	Screw Pump Heater
P-002A/B	Pump Around Pump
P-003A/B	Reflux Pump
P-004A/B	Liquid Hydrocarbon Pump
P-005A/B	Column Water Pump
P-006A/B	Wax pump
E-001	Bottoms Cooler
E-002	Condenser
E-003	Wax Cooler
GR-001	Grinder
F-001	Starfeeder
V-001	Fractionation Column
V-002	Reflux Drum

The required distance between different buildings and unit operations was estimated based on the figure in "Spacing for Fire and Safety Considerations .pdf" (Caltex, 1995) and summarized in Table 34.

Table 34. Distances between buildings and/or unit operations. P-M refers to Personal or Maintenance access (2.5-4 ft).

Locations	Distance (ft)
Office Building to Plant Road	25
Office Building to Product Storage Tanks	200
Office Building to Tank Truck Loading Racks	200
Office Building to Warehouse	200
Office Building to Flares	300
Office Building to Reactor	200
Office Building to Heat Exchangers	200
Office Building to Column	200
Office Building to Pump	100
Office Building to Compressor	100
Office Building to Utility Boiler	200
Parking Lot to Product Storage Tanks	100
Parking Lot to Main Electrical Substation	50
Plant Road to Utility Boiler	25
Plant Road to Flares	100
Plant Road to Reactor	25
Plant Road to Column	25
Product Storage Tanks to Column	100
Product Storage Tanks to Reactor	100
Product Storage Tanks to Tank Truck Loading Racks	100
Column to Reactor	P-M
Column to Column	10

The Material and Personnel Flow within the facility is demonstrated in Figure 20.

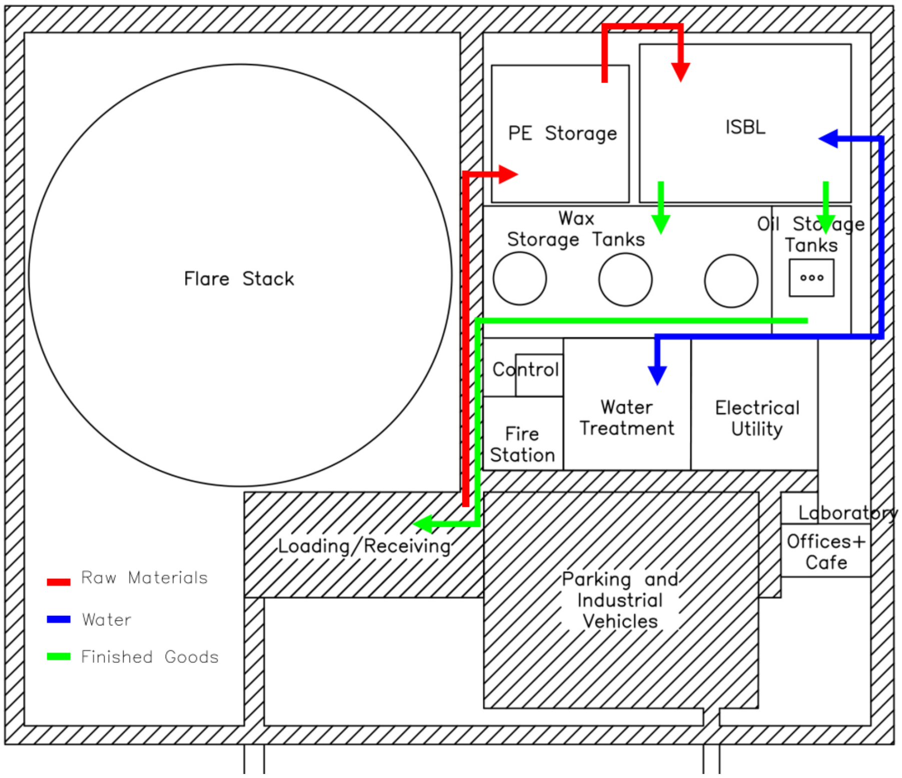


Figure 20. Material and Personnel Flow.

Constructability

The Offices, lab, facilities, flare, and ISBL will all be stick built. All of the storage facilities, including the Wax Storage Tanks, Oil Storage Tanks, and PE Storage Warehouse will also be built on site.

The heaviest and largest pieces of equipment, namely the fired heater boiler and reactor, will be transported to the site on a barge. Smaller equipment, like pumps, the column, and exchangers, will be transported to the site by truck. All parts of the site are accessible by road, and the site is right alongside the Houston ship channel for barge delivery.

The proposed laydown area is suggested to be around the flare and is shown in Figure 21. The parking lot could serve as an additional laydown area if needed.

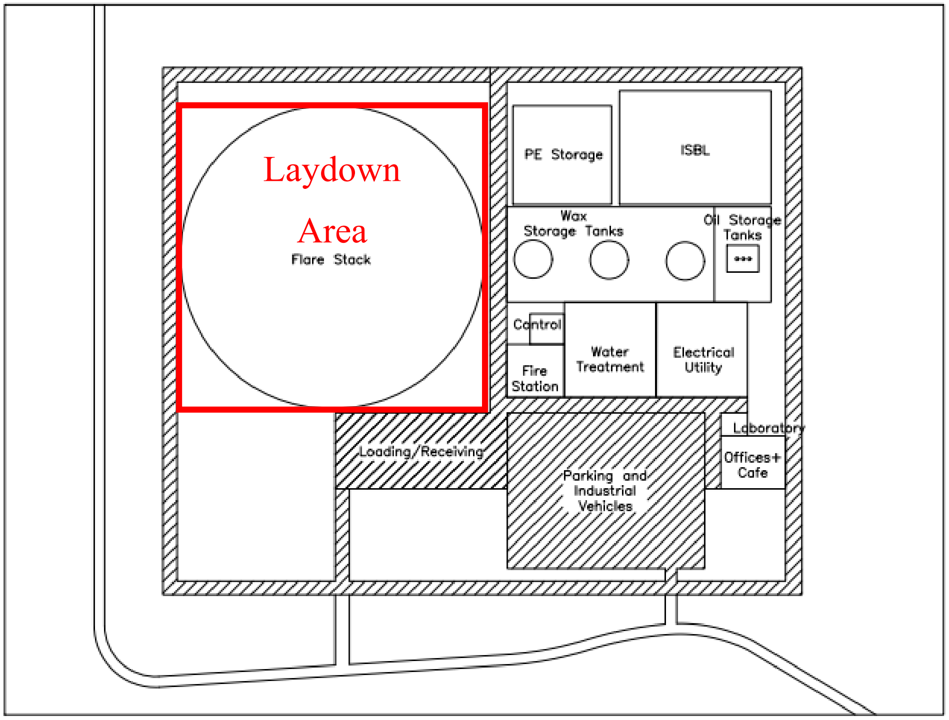


Figure 21. Proposed Laydown area during construction.

Meteorological Conditions

Houston has a hot, humid climate. Since it is on the Gulf Coast, there is a significant risk of Hurricanes in this area. In Houston, winter lows are generally around 46°F, but can get as low as 34°F (all time min 5°F)[34,35]. Therefore, freezing conditions are possible but not common. The summer highs are generally around 92°F but can get as high as 97°F (all time max 109°F). The air-cooling temperature used for this process design will use the maximum air temperature. Houston ocean temperatures are around 45-90°F, depending on the time of year[36]. They are typically lower than the air temperature.

The 2018 International Plumbing Code indicates that 4.6 inches/hour is the expected maximum hourly rainfall rate in Houston[37]. The maximum total rainfall in Houston for a single event was 60.58 inches during Hurricane Harvey in 2017[38].

Wind speeds can reach significant extremes in the Houston area. Hurricane Ike had winds up to 110 mph[39]. The all-time highest wind speeds in the Houston area were 145 mph during a

Hurricane in 1915[40]. The prevailing winds in Houston are typically from the Southeast (SE), but in January wind comes from the North[41]. The average wind speed is 12.8 mph.

All these meteorological conditions are summarized in Table 36.

Table 36. Meteorological Conditions in the Houston area.

Condition	Value
Winter Low Temperature	34.0°F
All Time Winter Low	5.0°F
Summer High Temperature	97.0°F
All Time Summer High	109.°F
Ocean Temperature Low	45.0°F
Ocean Temperature High	90.0°F
Maximum Hourly Rainfall Rate	4.6 in/hr
Maximum Total Rainfall Event	60.6 in
All Time Max Wind Speed	145. mph
Prevailing Winds Direction	SE

Stormwater Run-off

To determine the volume of run off of stormwater per duration of storm, the rational method was utilized to calculate intensity of rainfall per minute length of storm, and then the peak flow of rain for different length storms.

$$\mathrm{I} = \frac{b}{(\mathrm{d} + \mathrm{Tc})^{\mathrm{e}}} \qquad (13)$$

where b,d,e are coefficients dependent upon recurrence (b = 54.68 in; d = 6.96 min; e = 0.6623)[42,43]. Q = C*I*A = "Peak Rain flow (cfs)", where C = runoff coefficient; I = Intensity (in / hr) A = Area in acres. Run off Coefficients: Concrete/Asphalt areas – 0.95; Unpaved areas – 0.4; Undeveloped Areas – 0.2; Roofed Areas – 1.0. This was used to calculate Figure 22.

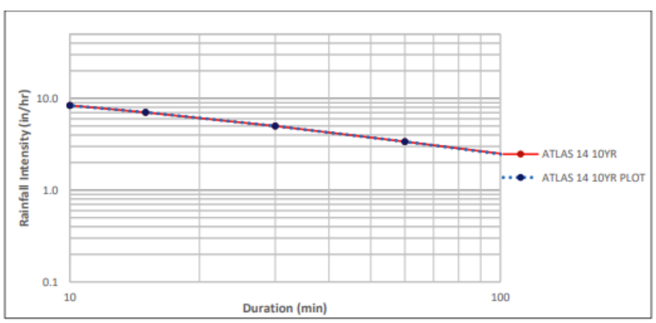


Figure 22. 10-year storm recurrence rainfall intensity (in/hr) vs storm duration (min) for Houston, TX

From these values, the volume of stormwater run-off could be determined. V = QD = "Run-off volume", where Q = Peak Flow Rate; D = Storm Duration. This was used to calculate table 37. The relevant table is below:

Table 37. Storm Duration and Total Volume of Run-off on Plant Property

Storm Duration	Total Vol. (ft^3)	Volume (acre-ft)	Ft of water per acre
15 min	65200	1.50	0.115
30 min	82400	2.02	0.154
60 min	61300	1.33	0.101
2 hours	40800	0.94	0.072
3 hours	31600	0.73	0.055
4 hours	26300	0.60	0.046
5 hours	22700	0.52	0.040
6 hours	20200	0.46	0.035
12 hours	12800	0.29	0.022

We will design drainage system to collect stormwater run-off. Drains will channel the water to the treatment facility, collect the first 30 mins of rainfall, and test for quality. If it's clean, it will be recycled to be used in our process, the water fails, discharge through further drain system to the municipal storm drains.

Design Codes, Standards, and Practices

Our plant falls under the Heavy Industrial (HI) Development Regulations for Baytown, TX[44].

Table 38. Property Development Standards for HI uses in Baytown[44]:

Ordinance Restriction	Amount
Minimum Lot Size (Acres)	5
Minimum Front Setback (ft)	50
Minimum Rear Setback (ft)	20
Minimum Interior Side Setback (ft)	10
Minimum Street Side Setback (ft)	50
Maximum Height (ft)	60/75
Maximum Lot Coverage %	90
Minimum Lot Frontage (ft)	60

Buildings in the heavy industrial (HI) zoning district have a maximum height of 60 feet unless the entire building is equipped with a National Fire Protection Association ("NFPA") 13 fire sprinkler system, then the maximum height shall be 75 feet.

See the ordinance code (reference 14) for more specifics on building codes and site design regulations.

Scheduling and Employees

The shift schedule that will be used at our facility is 8/5 plus OT. When coverage is needed 24/7 there will be four separate shift teams and they will work shifts five days in a row, with alternating 1 or 2day breaks (Figure 23). There will be three shifts per day for operations, maintenance, quality, medical, safety, etc., but not for HR, administration, and engineers.

We will assume a utilization factor of 85% with a yearly working period of 2,080 hours for non-shift schedule roles.

The burdened labor rates will be assumed to be 50% of the unburdened rate.

The German company Revalyu runs a PET repurposing plant, which has a capacity around 80,000 tons per year and employs 70 people[45].

The total number of employees which will be employed at the proposed facility is estimated to be 108, with 53 on site at one time at maximum.

We will assume that contact, process engineers, doctors, and nurses will do an additional 10% hours of overtime, and their overtime wage will be 150% their normal working wage. This will account for the fact that they will always be on call.

Table 39. Number of employees and their wage for given roles.

Role	Number (per shift)	Unburdened Wage ($/hr)	Burdened Wage ($/hr)	Cost ($/yr)
HR	2	60	90	374,000
Administrator	2	60	90	374,000
Secretary	2	30	45	187,000
Janitor	2	20	30	125,000
Senior Engineer/Project Leads	3	90	135	842,000
Process Engineer	4	60	90	749,000
Contact Engineer	4	60	90	861,000
Mechanical Engineer	1	60	90	187,000
Senior Process Designer	1	90	135	280,000
Process Designer	2	60	90	374,000
Product Development	2	60	90	374,000
Quality	3*	40	60	1,576,800
Marketing	3	50	75	468,000
Accounting	3	50	75	468,000
Operator	4*	60	90	3,153,600
Maintenance	3*	40	60	1,576,800
Laborers	4*	20	30	1,051,200
Fire and Safety	3*	30	45	1,182,600
Doctor - Medical	1	90	135	323,000
Nurse - Medical	2	50	75	343,000
Cafeteria	2	20	30	374,000
Total	**104**			**15,244,000**
At once	**53**	*shift schedule		

The total yearly wages for all employees will be around $15.2M.

The Shift schedule for the facility is shown below.

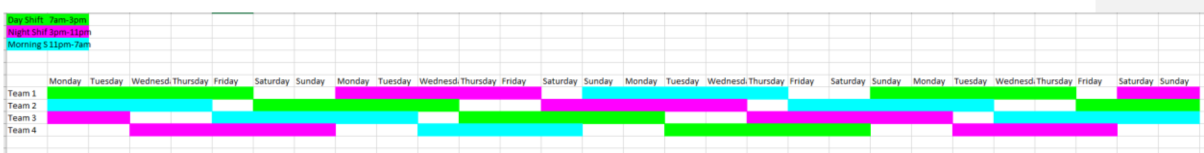


Figure 23. Shift schedule for the facility.

Table 40. Area per employee for various site sections. Minimum medical area is 108ft^2.

Site Area	Area per employee (ft^2)	Number of employees	Area (ft^2)
Office/Administration	108	36 (in office)	3,800
Laboratory	215	5	1,080
Canteen	11	53 (total at once)	583
Kitchen	38	2	76
Medical	1.6	53 (total at once)	108
Control Building	430	4	1,720
Garage	1,080	53 (total at once)	52,700
Fire Station	n/a	n/a	5,380

The building containing the office, canteen, medical, and kitchen will be 4,570 ft^2. The laboratory building will be 1,080 ft^2. The control building will be 1,720 ft^2. The parking lot and garage will be 52,700 ft^2. The fire station will be 5,380 ft^2.

Product Quality Control and Management

We expect to produce several mixtures of products. The gas products will be one such mixture. However, we expect most of this product will be used for the energy demands of the facility. Aromatics (benzene, toluene, etc.) and other liquids are expected to be sold as a mixture, without much regard to quality, to a refinery, as liquid hydrocarbon/synthetic crude oil.

The solids will also be sold as a mixture. We assume that the solids will consist of waxes with varying carbon numbers. The mixture would have specifications based on a range of density, melting/boiling point, etc. Off specification mixtures can be sold at a downgrade. Any specifications may not be important for the liquids since they are going to a refinery anyway. The price of crude oil (~0.22 $/lbm) is chosen for the price of liquids and off-specification wax.

Table 41. Product Mixtures, Yields, and Prices [46,47,48].

Product Mixture	Yield (%)	On-Spec Price ($/lbm)	Off-Spec Price ($/lbm)
Fuel Gas	2.11	0.070	N/A
Liquids	0.703	0.22	0.22
Wax and Tars	97.6	0.971	0.22

Our lab will need to meet ISO Certification and Training.

According to ISO 9001:2015 – Our quality management systems:

a) needs to demonstrate its ability to consistently provide products and services that meet customer and applicable statutory and regulatory requirements, and

b) aims to enhance customer satisfaction through the effective application of the system, including processes for improvement of the system and the assurance of conformity to customer and applicable statutory and regulatory requirements."

We will also use the American Society of Testing and Materials (ASTM) Standards for our quality testing procedures. This organization sets testing standards for use in manufacturing.

We will take samples of our liquid hydrocarbon product tank weekly and take samples from each shipment.

The exact composition of the liquid hydrocarbons is not significant since it is planned to be sent to a refinery. However, boiling point, density, and molecular weight may be of interest to a refinery. The specifications for our product are given below in Table 42.

Table 42. Specifications for liquid hydrocarbon/synthetic crude oil product.

Property	Specification
Molecular Weight	130
Density	42.7 lb/ft^3 (@120°F)
Bubble Point	188.4°F
Dew Point	754.6°F

There are some properties of Polyethylene that may be of interest to us, but no extensive testing is necessary since the HDPE is a raw material. They include:

- Mechanical testing including tensile, flexural, shear and compressive properties
- Thermal testing including DSC, DMTA,TMA, TGA, HDT and Vicat Softening Points
- Rheological testing including capillary, rotational, Melt Flow Rate/Index and DSV
- Weathering and Chemical Resistance testing including UV, temperature, humidity, repeated sterilization and environmental impact
- Optical testing including color, haze and gloss

The most important product to test is our main product: wax. It will be tested in every shipment and weekly in our storage tanks.

Any off-specification product should be sent back into the column feed as a first option. Off-spec wax could also be sent to the wax storage tanks, such that higher purity wax produced later can be mixed with this material to produce on-specification product. If the product is still off-spec after mixing, it would be sold at a downgrade, but we would like to avoid this. If the wax purity is very low (~50% or less), it should be sent to the liquid hydrocarbon storage tanks instead and sold at a downgrade.

During upsets, start-up, and shut-down, quality tests would be taken at increased frequency during. It should be multiple times per day and as soon as possible after start-up or upset.

The specifications for our wax product are shown in Table 43.

Table 43. Specifications of wax product[49].

Property	Specifications	Test Method	Typical Values for Wax
Melt Point	100°F	ASTM D87	100-160°F (43-71°C) for paraffin waxes
Congealing Point (Temperature at which molten wax ceases to flow)	125°F-130°F	ASTM D938	Varies widely
Oil Content	0.5%	ASTM D721	Fully Refined <0.5%, Semi-refined 0.5-1.0%, Scale 1.0-3.0%
Kinematic Viscosity	4 @ 212°C	ASTM D445	2.9-7.5 for paraffin
Mottling	None	In-House	pass or fail
Saybolt Color	White	ASTM D6045	Lightest color is highest, +30 is maximum
ASTM Color	White	ASTM D6045	Darkest color is highest, 8 is black
Odor Test	1	ASTM D1833	A value of 1 or less is acceptable for paraffin

Information from Blended Wax Inc was used to determine the required lab equipment[50]. The list of equipment that would be in our lab is given below:

- Differential Scanning Calorimeter (DSC)
- Infrared Spectrometer (FTIR)
- Autoviscometer
- Brookfield Viscometer
- Inductively Coupled Plasma (ICP) Machine
- Karl Fischer Water Analyzer
- Needle Penetrometer
- Autotitrator
- Particle Size Analyzer
- Densitometer
- Gas Chromatograph (GC)
- Lovibond Colorimeter
- Automated Instrument
- Inductively Coupled Plasma (ICP) Machine
- Karl Fischer Water Analyzer
- Needle Penetrometer
- Autotitrator
- Particle Size Analyzer

- Densitometer
- Gas Chromatograph (GC
- Lovibond Colorimeter
- Automated Instrument

Available literature was used to help design the lab[51]. It was determined that 25 ELF (equivalent linear feet) per researcher is needed, and 2 researchers at minimum should be working. This is reflected in Table 39.

Other quality system needs for our facility include gowning procedures, to ensure accurate testing and safety of lab workers. Regular Sanitation and Cleaning should be conducted in the lab. Workers should wear all appropriate PPE. This includes Face masks, lab coats, and goggles.

Regular calibration should be conducted on all instrumentation. Workers should follow all procedures on calibrating instruments such that all measurements are accurate. Training should stress the importance of this.

SWOT Analysis

Strengths (Internal factors that are positive for the project) • Lean production • Structured workforce • Technological skills • Purity of wax product	**Weaknesses** (Internal factors that are negative for the project) • Employee health problems • Inability of product diversification • High carbon footprint and pollution

Opportunities (External factors that have the potential to benefit the project)
 	
• Decrease in waste PE costs
• Close proximity of other petrochemical plants
• Abundant water resources
• Strong local market potential
	Threats (External factors that are a potential detriment to the project)
	
	• Natural disasters
	• Fluctuating market prices
	    ○ Increase in waste PE costs
	    ○ Decrease in wax prices
	• Tax increases
	• New laws restricting production

Figure 24. SWOT Analysis.

Project Schedule

The chosen start date is 2/1/2023, which marks the beginning of the FEL2 study. The end of FEL2 is set as 5/1/2023.

The specialized process units will cause procurement and construction periods to be long (over 1 year). This will especially be the case for the pyrolysis reactor. It will take particularly long as it must be made unique for our process, and out of the uncommon SS-309 alloy.

The schedule indicates a 3-year duration from FEL2 to start-up. The estimated start-up date is February 2025.

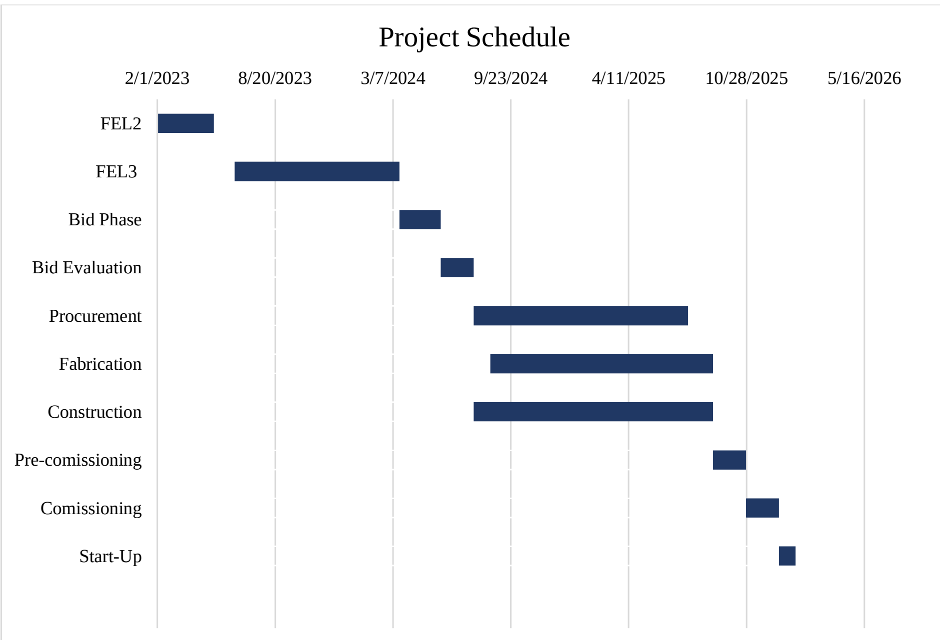


Figure 25. Project Schedule.

Capital and Equipment Costs (CAPEX)

Before estimating our facility's capital cost, we looked into the capital costs of existing plants and papers analyzing pyrolysis plant economics. There is an existing plastics pyrolysis plant in Indiana, owned and operated by Brightmark. Construction on this project started in 2019 and the plant began creating products in 2022. These products include naphtha, diesel, and waxes.

There is a similar pyrolysis plant in development in Böhlen, Germany by Dow and Mura. It will have a capacity of 120,000 tons per year[52]. There are no capital cost estimates for this facility currently, so it is not reflected in Table 44.

Many journal articles have modeled the costs of plastics pyrolysis plants. A few are summarized in Table 44.

Table 44. Capital Costs (CAPEX) of Plastic Pyrolysis Plants. Scaling Factor of 0.6 and inflation to 2023 dollars are used to project the capital cost of the proposed facility.

Plant or Article	Capacity	CAPEX ($M)	Year	Projected CAPEX of Proposed Facility ($M)
Brightmark, Indiana	100,000 tons/yr	260	2019	562
Wright et al, 2010[53]	28 million gallons/yr 37,000 tons/yr	48.2	2010	225
Riedewald et al, 2021[54]	40,000	23-30 (avg: 26.5)	2021	94.2
Badger et al, 2011[55]	200 wet-tons/day (73,000 tons/yr)	6.03	2011	17.8
Yadav et al., 2022[56]	240 tons/day (88,000 tons/yr)	56	2022	124

All of the projections based on the journal articles give a lower estimated capital cost of this proposed facility than the one based on the Brightmark facility. Badger et al.'s estimation is very low compared to the others so it was disregarded. The average of the other CAPEX estimates based on this approximation is $251.3M.

Aspen Plus Process Economic Analyzer was used to estimate equipment costs. Where costs could not be found using Aspen, or were found to be inaccurate, either the FACT method (Olin Engineering's cost estimation method) or commercially available prices were used instead. For example, the de-oiler and reverse osmosis unit cost estimates from Aspen were significantly lower than commercially available prices, so the latter prices were used instead. Equipment Costs are shown below in Table 45.

Table 45. Aspen Plus Estimations of Equipment Costs. A number in parentheses next to the equipment name is used to indicate the number of copies of that piece of equipment. Inflation adjustment has been applied.

Equipment	Equipment Cost ($)	Installed Cost ($)
Grinder**	223,000	892,000
Auger Conveyer*	48,900	196,000
Motor*	1,200	4,800
Star feeder**	10,000	40,000
Screw Pump Heater**	50,000	200,000
Pyrolysis Reactor	2,830,000	3,550,000
Fractionation Column	448,000	1,380,000
Reflux Drum	41,600	155,000
Boiler (2)*	4,950,000	19,800,000
Air Cooler (Condenser)	69,600	255,000
Bottoms Cooler	60,600	254,000
Wax Product Cooler	27,600	201,000
Reflux Pump (2)	13,920	99,500
Pumparound Pump (2)	21,750	145,000
Column Water Pump (2)	16,820	123,000
Liquid Hydrocarbon Pump (2)	13,050	84,40
Wax Pump (2)	19,140	126,000
Makeup Pump (2)	12,470	87,000
Recycle Pump (2)	12,470	83,800
Flare**	21,000	21,000
De-oiler (2)**	152,000	608,000
Reverse Osmosis Unit (2)**	200,000	800,000

*Costs determined based on the FACT Method

**Costs determined based on commercially available units

Aspen Plus gives costs in 2013 Dollars, and FACT gives costs in 2006 dollars. Therefore, the CEPCI Index was used to account for inflation. Since 2013, this factor is 1.45. Since 2006, this factor is 1.73 [57,58].

The FACT Method was also used to estimate the tankage costs for wax and liquid hydrocarbons.

The PE storage warehouse cost was estimated using a cost of $30/ft^2, which is typical for pre-engineered metal building (PEMB) construction[59].

The breakdown of capital costs of the project is shown in Table 46.

Table 46. Overall project capital costs.

Component	Cost
Total ISBL	$27.7M
Wastewater Treatment Facility	$1.4M
Product Storage Tanks	$18.7M
PE Storage Warehouse	$0.5M
Other OSBL	$9.7M
Total Project Cost	$57.9M
Total Project Cost w/ Contingency	$81.1M

Fixed and Operating Costs

We have used the total equipment costs from the Aspen and FACT estimates from Table 47.

Table 47. Fixed and Operating Costs of the Pyrolysis Plant[53].

Category	Subcategory	Quantity ($/year)	Subtotal ($/year)
Staff Salaries & Benefits	Salaries Benefits	$10.2M $5.1M	$15.2M
Insurance	Property Damage (PD)	(0.2% of replacement value)	$1.1M
	Business Interruption (BI)	(0.45% of expected gross profit)	
Maintenance	Labor Parts	(2% of total equip. cost)	$0.96M
3	Property Tax	(0.02% of prop. Value)	$0.08M
Utilities	Electric	($0.14/kWh, 252 kW)	$0.65M
	Municipal Water	($0.0215/gal, 121,000 tons/yr)	$0.70M
	Natural Gas	($0.07/lbm, 8,940 lbm/hr)	$5.3M
	Nitrogen Gas	($0.013/scf, 10,400,000 scf/yr)	$0.14M
Transportation			$12.3M
		Total ($/year)	$36.4M

The annual property tax in Baytown, TX for commercial property is 0.02% of property value. Based on an average value of $300K/acre in the area, and our property area of 13 acres, this corresponds to a property value of $3.9M, which corresponds to annual property tax of ~$80K.

Overall Economic Analysis

A cost for PE feedstock of $0.45/lbm, and the product prices listed in Table 48 were utilized to estimate the NPV for the project. A time period of 20 years and a discount rate of 15% were used.

$$NPV = \sum_{i}^{t}\left(\frac{R_{product} - R_{feedstock} - R_{OPEX}}{(1+i)^{t}}\right) - R_{CAPEX} \qquad (17)$$

Table 48. Preliminary Economic Analysis of the Pyrolysis plant.

Category	Cost
ISBL Cost	$27.7M
OSBL Cost	$9.7M
CAPEX Total	**$81.1M**
OPEX Total	**$36.4M/yr**
Feedstock	$260M/yr
Products	$482M/yr
Net Profit	**$186M/yr**
NPV	**$1,086M**

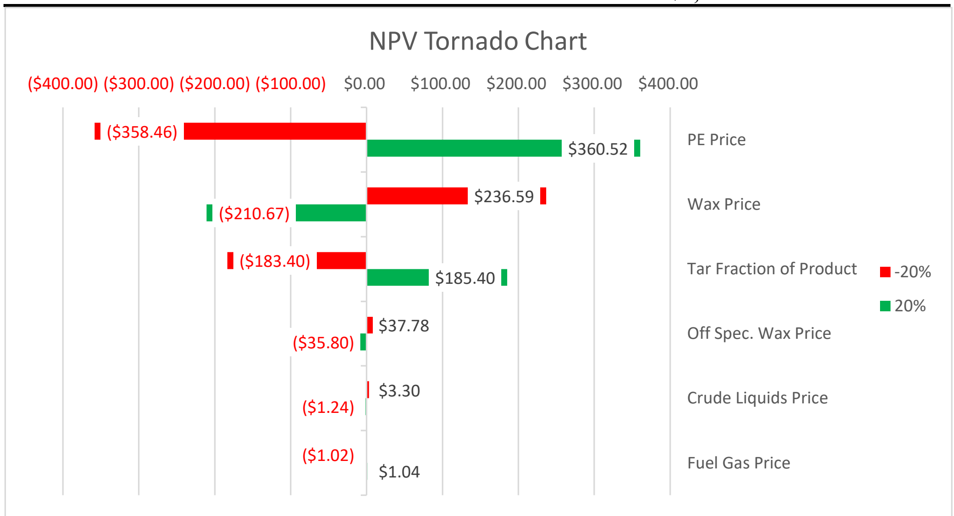


Figure 26. Net Present Value Tornado Chart

Following further sensitivity analysis for a 15% discounted cash flow rate of return, the following cases were plotted for market variations in raw materials and product prices. We assumed a 2% YoY growth rate and a 2% inflation rate. Working capital infusion was assumed to be 5% of capital costs. The corporate tax rate in the USA is currently 21%. For the following cases below, the annual cash flow and net present value is plotted over a 10-year period. The variations in these cases are summarized in the DCFROR Tornado Chart.

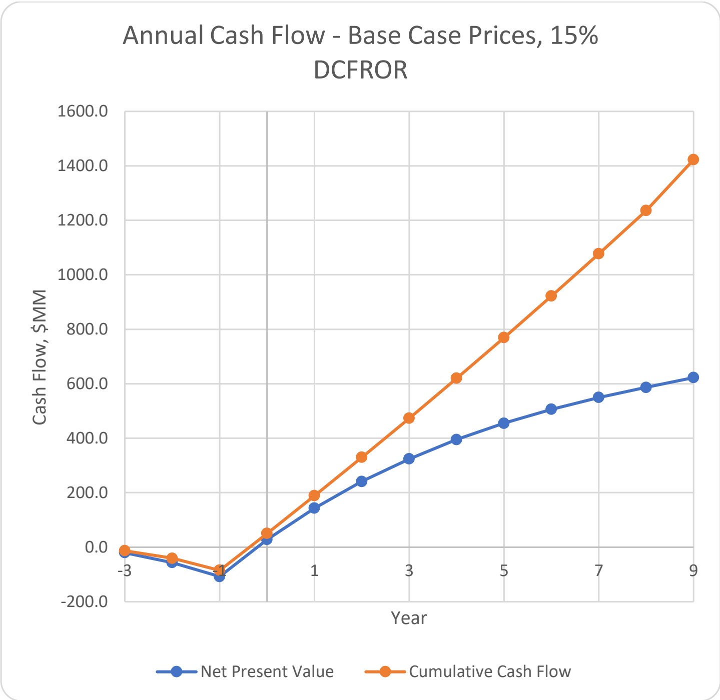


Figure 27. Annual Cash Flow at 15% DCFROR Base Case

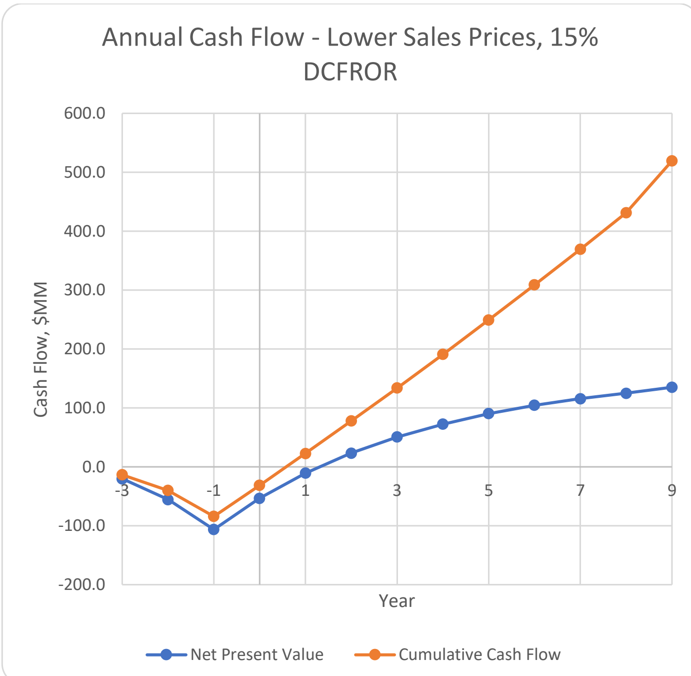


Figure 28. Annual Cash Flow at 15% DCFROR Lower Sales

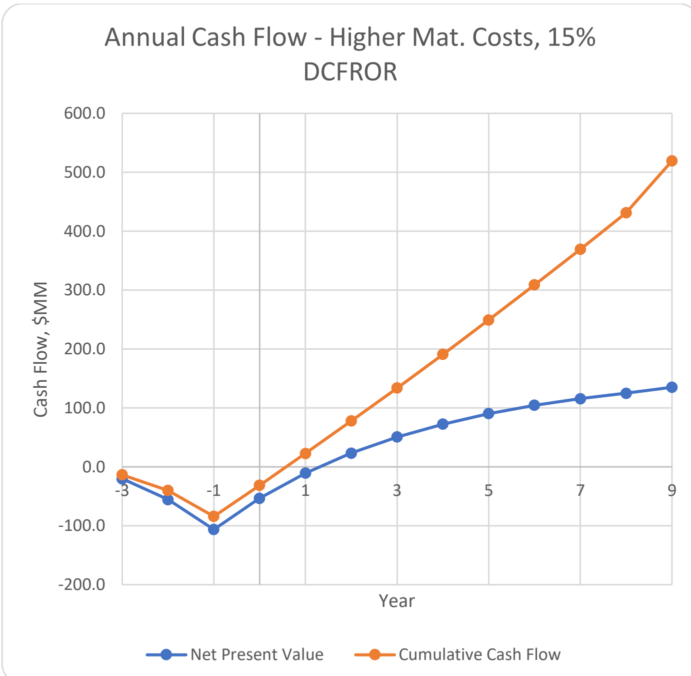


Figure 30. Annual Cash Flow at 15% DCFROR Higher Capital Costs

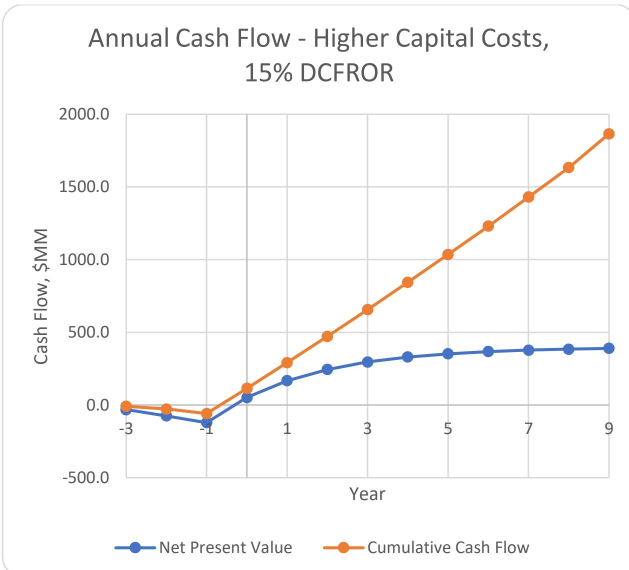


Figure 29. Annual Cash Flow at 15% DCFROR Higher Materials Costs

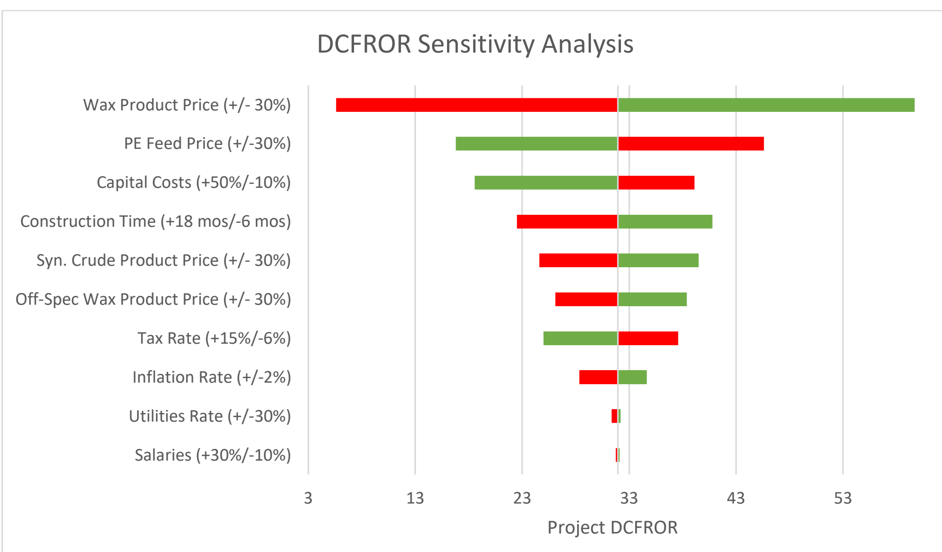


Figure 31. DCFROR Tornado Chart

From the tornado chart, it is clear to see that our project's DCFROR has the highest dependency on our largest product, paraffin wax, and our polyethylene feedstock. This follows intuition as they are the largest factors in our cash flow. The market conditions of product and feed prices can be relatively managed through a diverse and complex supply chain, where one distributor can't strain the economics of our plant. Doing this mitigates large price fluctuations, allowing the prices to trend with the economy itself.

A mere 10% reduction in capital costs has the same collateral effect as increasing capex 50%; i.e., it's better to spend more than be short on investment funds. Construction time sees a similar phenomenon occurring, where the opportunity cost of extending time is much higher than shortening the deadline. Utilities and salaries are predictably manageable, as they are relatively overhead in determining our annual cash flow and growth.

Following this analysis, the price of waste plastic that results in a 15% after-tax DCFROR is $0.585/lb. Our base case DCFROR (current prices) is 31.95%.

Conclusion

We recommend moving ahead with a FEL-3 study of a pyrolysis process for waste HDPE.

The current price of waste HDPE is significantly lower than that required to generate an after tax DCFROR of 15%. The NPV for a 20 life project period is $1,086M, compared to $81.1M capital investment and $36.4M/yr operating costs.

Focusing on wax products presents an opportunity to generate high profits from the pyrolysis of HDPE. Wax prices are significantly higher than waste HDPE ($0.97/lbm vs $0.45/lbm) and separation of wax from other products is relatively simple. Only a 3-tray column was needed to accomplish this. This resulted in the relatively low capital cost of the project compared to the NPV and cumulative net cash flow.

It should be noted that this study is based on lab scale data of product yield. The data is also from experiments using a sand bed rather than steam. Further lab scale experiments, or potentially a pilot plant study, should be done to confirm the desired product disposition.

Ultimately, the pyrolysis of polyolefin plastics like HDPE will allow us to improve the circularity of the plastics economy by converting them to other petroleum products. This can be accomplished with lower carbon intensity than making new plastics.

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Appendix A: Aspen Plus and Aspen HYSYS Stream Results

Table A1. Aspen Plus Stream Results for Column feed, distillates, and bottoms.

	Units	PYRO	GAS	H2O	LIQ	WAX
Description						
From			COLUMN	COLUMN	COLUMN	COLUMN
To		COLUMN	VALVE	WPUMP	LPUMP	WAXCOOL
Stream Class		CONVEN	CONVEN	CONVEN	CONVEN	CONVEN
Maximum Relative Error						
Cost Flow	$/hr					
MIXED Substream						
Phase		Vapor Phase	Vapor Phase	Liquid Phase	Liquid Phase	Liquid Phase
Temperature	F	932	120.000014322818	120.000014322818	120.000014322818	487.467500527899
Pressure	psig	44.0878463265404	44.0878463265404	44.0878463265404	44.0878463265404	44.0878463265404
Molar Vapor Fraction		1	1	0	0	0
Molar Liquid Fraction		0	0	1	1	1
Molar Solid Fraction		0	0	0	0	0
Mass Vapor Fraction		1	1	0	0	0
Mass Liquid Fraction		0	0	1	1	1

Mass Solid Fraction		0	0	0	0	0
Molar Enthalpy	Btu/lbmol	-93347.9485557899	-3961.96672795989	-122109.28779429	-63959.9131131837	-197329.150127673
Mass Enthalpy	Btu/lb	-2596.33109956451	-108.030905834859	-6778.09547197102	-490.718681634699	-598.910794935486
Molar Entropy	Btu/lbmol-R	-24.8447418615596	-30.7650249831223	-37.6006284643799	-178.489003014663	-471.017546811411
Mass Entropy	Btu/lb-R	-0.691018677472783	-0.838869618339823	-2.0871520433976	-1.36941849953191	-1.42957841356366
Molar Density	lbmol/cuft	0.00395773567580937	0.00982310081837193	3.42616858854793	0.32777847416291	0.122435415413623
Mass Density	lb/cuft	0.142295605643223	0.360256153616614	61.7233864498957	42.7224059577136	40.3400250478057
Enthalpy Flow	MMBtu/hr	-334.533974559084	-0.155942033452288	-407.597005958274	-0.235998762115896	-40.0011740463358
Average MW		35.9537920920207	36.6743821811169	18.01528	130.339266685585	329.480035752117
Mole Flows	kmol/hr	1625.55322011265	17.8532838342802	1514.07722767966	1.67366140156857	91.9490471971355
Mole Fractions						
Mass Flows	tons/year	564744.000000001	6326.83700409023	263569.270232721	2107.89320453055	292739.999558659
Mass Fractions						
METHA-01		0.00047809272874081	0.0425582340216125	0	0.000225732224851862	9.05823395922276E-07
ETHYL-01		0.0033997705154902	0.301167020908436	0	0.00551407542951108	1.00507808369071E-05
ETHAN-01		0.000690578385958948	0.0610061259657314	0	0.00158081267653485	2.36387397131114E-06
PROPY-01		0.00196549232926777	0.171048207970432	0	0.011806098521872	9.97428942913353E-06

PROPA-01		0.00100930687178615	0.0875628831016408	0	0.00683673828093995	5.44099489843365E-06
N-BUT-01		0.000212485657218138	0.0174957021309506	0	0.00416571888898093	1.79904969627876E-06
N-PEN-01		0.00111554970039522	0.0805095359777401	0	0.055216349450011	1.44808390231283E-05
BENZE-01		0.00047809272874081	0.0250191251297286	0	0.0516160342402237	9.93026407005864E-06
TOLUE-01		0.00159364242913603	0.0439711073396902	0	0.287539129083004	5.36318738997044E-05
1-BUT-01		0.00185924950065871	0.154768373100773	0	0.0315550776596072	1.46580455097989E-05
WATER		0.46883897836896	0.0148936688832149	1	0.00475379169240976	0.00375923846524954
O-XYL-01		0	0	0	0	0
N-PEN-02		0.518358760783647	1.54700500877058E-08	0	0.539190441852054	0.99611752570002
AIR		0	0	0	0	0
OXYGE-01		0	0	0	0	0
NITRO-01		0	0	0	0	0
CARBO-01		0	0	0	0	0
SODIU-01		0	0	0	0	0
Volume Flow	gal/min	112893.561546917	499.556110329295	121.465814225653	1.40346613739655	206.421505660372

Table A2. Aspen Plus Stream Results for Boiler Inputs and Outlet

	Units	GAS2	FUEL	AIR	BOILIN	BOILOUT
Description						
From		VALVE			MIX	BOILER
To		MIX	MIX	MIX	BOILER	HEATX
Stream Class		CONVEN	CONVEN	CONVEN	CONVEN	CONVEN

Maximum Relative Error						
Cost Flow	$/hr					
MIXED Substream						
Phase		Vapor Phase	Vapor Phase	Vapor Phase	Vapor Phase	Vapor Phase
Temperature	F	113.955662602566	100.000000000001	100.000000000001	100.100818822271	3215.90219768296
Pressure	psig	2.70857975211166	2.70857975211166	-1.77635683940025E-15	-1.77635683940025E-15	-0.192175002492528
Molar Vapor Fraction		1	1	1	1	1
Molar Liquid Fraction		0	0	0	0	0
Molar Solid Fraction		0	0	0	0	0
Mass Vapor Fraction		1	1	1	0.999999999999997	1
Mass Liquid Fraction		0	0	0	3.33066907387547E-15	0
Mass Solid Fraction		0	0	0	0	0
Molar Enthalpy	Btu/lbmol	-3961.96672795989	-31849.208534119	157.084302469182	-2072.95752161602	-2069.07999018241
Mass Enthalpy	Btu/lb	-108.030905834859	-1985.26989957582	5.46079310996737	-74.2305656510305	-74.2305656510305

Molar Entropy	Btu/lbmol-R	-28.4015274627322	-19.2371374278933	1.24957299450987	0.268719273582159	15.9205279805673
Mass Entropy	Btu/lb-R	-0.774424155871826	-1.19911645052929	0.0434394748014971	0.00962257232545096	0.571166800253277
Molar Density	lbmol/cuft	0.00286001716742176	0.00290454091357966	0.00244784290494028	0.00244764257437359	0.00036765483404193
Mass Density	lb/cuft	0.104889362642581	0.0465968527867392	0.07041425806351	0.068352693264232	0.0102478979344734
Enthalpy Flow	MMBtu/hr	-0.155942033452288	-17.7419625750365	1.17364068208744	-16.7242639264014	-16.7242639264014
Average MW		36.6743821811169	16.04276	28.7658402920378	27.9259292103648	27.8736928923521
Mole Flows	kmol/hr	17.8532838342802	252.678770470574	3388.97299188059	3659.50504618545	3666.36309223266
Mole Fractions						
Mass Flows	tons/year	6326.83700409023	39170.0000000001	942000	987496.837004093	987496.837004093
Mass Fractions						
METHA-01		0.0425582340216125	1	0	0.0399386180612883	0
ETHYL-01		0.301167020908436	0	0	0.00192956025872031	0
ETHAN-01		0.0610061259657314	0	0	0.000390862836996185	0
PROPY-01		0.171048207970432	0	0	0.00109589630175814	0
PROPA-01		0.0875628831016408	0	0	0.000561010494649303	0
N-BUT-01		0.0174957021309506	0	0	0.000112093985020207	0
N-PEN-01		0.0805095359777401	0	0	0.000515820094119439	0
BENZE-01		0.0250191251297286	0	0	0.000160296135389114	0.000160296135389114

TOLUE-01		0.0439711073396902	0	0	0.000281720425425952	0.000281720425425952
1-BUT-01		0.154768373100773	0	0	0.000991592310277704	0
WATER		0.0148936688832149	0	0	9.5422903533412E-05	0.0975189287905183
O-XYL-01		0	0	0	0	0
N-PEN-02		1.54700500877058E-08	0	0	9.91157456736448E-11	9.91157456736448E-11
AIR		0	0	0	0	0
OXYGE-01		0	0	0.21	0.200324692279678	0.0215336244734295
NITRO-01		0	0	0.79	0.753602413814028	0.753602413814028
CARBO-01		0	0	0	0	0.126903016262094
SODIU-01		0	0	0	0	0
Volume Flow	gal/min	1715.7904127625	23911.45612239	380539.140041745	410950.126943117	2741005.82904405

Table A3. Stream results for Boiler steam output, wax stream to storage

	Units	CW	STEAM	EXHAUST	STRWAX	HEATWAX
Description						
From			HEATX	HEATX	WAXCOOL	WAXPUMP
To		HEATX			WAXPUMP	
Stream Class		CONVEN	CONVEN	CONVEN	CONVEN	CONVEN
Maximum Relative Error						
Cost Flow	$/hr					
MIXED Substream						

Phase		Liquid Phase	Vapor Phase	Vapor Phase	Liquid Phase	Liquid Phase
Temperature	F	100.000000000001	1652	1689.21319942139	150.000000000001	150.066986778376
Pressure	psig	-1.77635683940025E-15	132.263538979621	-0.192175002492528	44.0878463265404	88.1756926530807
Molar Vapor Fraction		0	1	1	0	0
Molar Liquid Fraction		1	0	0	1	1
Molar Solid Fraction		0	0	0	0	0
Mass Vapor Fraction		0	1	1	0	0
Mass Liquid Fraction		1	0	0	1	1
Mass Solid Fraction		0	0	0	0	0
Molar Enthalpy	Btu/lbmol	-123258.90273291	-89672.7054837918	-16292.5582792691	-265221.01509632	-265146.070412708
Mass Enthalpy	Btu/lb	-6841.90879813749	-4977.59154916225	-584.513804546488	-804.968393580783	-804.74093007623
Molar Entropy	Btu/lbmol-R	-39.3031560743171	-3.10104208470734	10.9385581934569	-558.705905205334	-558.700660455633
Mass Entropy	Btu/lb-R	-2.18165668667471	-0.172133993182862	0.392433045585366	-1.69572005760517	-1.69570413934266
Molar Density	lbmol/cuft	3.40132610955506	0.0064922578684175	0.0006288041465426	0.145146046264697	0.145141915193242

Mass Density	lb/cuft	61.2758422349451	0.116959843331744	0.017527093670166	47.8227245125707	47.821363407
Enthalpy Flow	MMBtu/hr	-421.92281408168	-306.955192729759	-131.691885278322	-53.7637342417459	-53.7485419838809
Average MW		18.01528	18.01528	27.8736928923521	329.480035752117	329.480035752117
Mole Flows	kmol/hr	1552.67461378495	1552.67461378495	3666.36309223266	91.9490471971355	91.9490471971355
Mole Fractions						
Mass Flows	tons/year	270288.27023	270288.27023	987496.837004093	292739.999558659	292739.999558659
Mass Fractions						
METHA-01		0	0	0	9.05823395922276E-07	9.05823395922276E-07
ETHYL-01		0	0	0	1.00507808369071E-05	1.00507808369071E-05
ETHAN-01		0	0	0	2.36387397131114E-06	2.36387397131114E-06
PROPY-01		0	0	0	9.97428942913353E-06	9.97428942913353E-06
PROPA-01		0	0	0	5.44099489843365E-06	5.44099489843365E-06
N-BUT-01		0	0	0	1.79904969627876E-06	1.79904969627876E-06
N-PEN-01		0	0	0	1.44808390231283E-05	1.44808390231283E-05
BENZE-01		0	0	0.000160296135389114	9.93026407005864E-06	9.93026407005864E-06
TOLUE-01		0	0	0.000281720425425952	5.36318738997044E-05	5.36318738997044E-05
1-BUT-01		0	0	0	1.46580455097989E-05	1.46580455097989E-05
WATER		1	1	0.0975189287905183	0.00375923846524954	0.00375923846524954
O-XYL-01		0	0	0	0	0

N-PEN-02		0	0	9.91157456736448E-11	0.99611752570002	0.99611752570002
AIR		0	0	0	0	0
OXYGE-01		0	0	0.0215336244734295	0	0
NITRO-01		0	0	0.753602413814028	0	0
CARBO-01		0	0	0.126903016262094	0	0
SODIU-01		0	0	0	0	0
Volume Flow	gal/min	125.472036454249	65735.4224463221	1602635.81073075	174.123260303924	174.128216250863

Table A4. Stream results for oil to storage, water to treatment section.

	Units	STROIL	H2OTRT	CONTAM	H2OCENT	OILWASTE
Description						
From		LPUMP	WPUMP		MIX2	DEOIL
To			MIX2	MIX2	DEOIL	
Stream Class		CONVEN	CONVEN	CONVEN	CONVEN	CONVEN
Maximum Relative Error						
Cost Flow	$/hr					
MIXED Substream						
Phase		Liquid Phase	Liquid Phase			Liquid Phase
Temperature	F	120.224893604593	158.917501567411	100.000000000001	158.923593836971	158
Pressure	psig	88.1756926530807	88.1756926530807	-1.77635683940025E-15	-1.77635683940025E-15	-1.77635683940025E-15
Molar Vapor		0	0	0.580263636661975	0.000534994677245049	0

Fraction						
Molar Liquid Fraction		1	1	0.419736363338025	0.999465005322755	1
Molar Solid Fraction		0	0	0	0	0
Mass Vapor Fraction		0	0	0.666666666666667	0.00195408970378337	0
Mass Liquid Fraction		1	1	0.333333333333333	0.998045910296217	1
Mass Solid Fraction		0	0	0	0	0
Molar Enthalpy	Btu/lbmol	-63926.7262953411	-122105.979746525	-60579.8155963813	-122063.150689983	2998.7506714986
Mass Enthalpy	Btu/lb	-490.464062910146	-6777.91184741647	-823.192745163809	-6761.01546555636	38.4714230706228
Molar Entropy	Btu/lbmol-R	-178.476923755997	-37.3541960876287	-37.07349550437	-37.3454965334425	-59.7622389149934
Mass Entropy	Btu/lb-R	-1.3693258240169	-2.07347296781558	-0.503775593184267	-2.06854794591343	-0.766698745180228
Molar Density	lbmol/cuft	0.327739655808151	3.28819809783799	0.00443175824957849	1.85660798060502	0.667625972594807
Mass Density	lb/cuft	42.7173464018204	59.2378094280187	0.326138804191942	33.5191393753405	52.0397654631438
Enthalpy Flow	MMBtu/hr	-0.235876309667976	-407.585963797689	-0.140861181581761	-407.72682498015	0.00431058358400835
Average MW		130.339266685585	18.01528	73.5912894668749	18.0539670870193	77.9474849680953

Mole Flows	kmol/hr	1.67366140156857	1514.07722767966	1.05470042398888	1515.13192810365	0.652020804042476
Mole Fractions						
Mass Flows	tons/year	2107.89320453055	263569.270232721	750.000000000001	264319.270232721	491.09927163406
Mass Fractions						
METHA-01		0.000225732224851862	0	0	0	0
ETHYL-01		0.00551407542951108	0	0	0	0
ETHAN-01		0.00158081267653485	0	0	0	0
PROPY-01		0.011806098521872	0	0	0	0
PROPA-01		0.00683673828093995	0	0	0	0
N-BUT-01		0.00416571888898093	0	0	0	0
N-PEN-01		0.0552163494500111	0	0	0	0
BENZE-01		0.0516160342402238	0	0.333333333333333	0.000945825855904818	0.467893829779177
TOLUE-01		0.287539129083004	0	0.333333333333333	0.000945825855904818	0.508272657108085
1-BUT-01		0.0315550776596072	0	0	0	0
WATER		0.00475379169240976	1	0	0.997162522432286	0.0238335131127385
O-XYL-01		0	0	0	0	0
N-PEN-02		0.539190441852054	0	0	0	0
AIR		0	0	0	0	0
OXYGE-01		0	0	0	0	0
NITRO-01		0	0	0	0	0

CARBO-01		0	0	0	0	0
SODIU-01		0	0	0.333333333333333	0.000945825855904818	6.91581730746842E-17
Volume Flow	gal/min	1.40363236765111	126.562434774218	65.4135661355938	224.308124679057	0.268437383262643

Table A5. Stream Results for Makeup and Recycle Streams

	Units	H2ORO	MAKEUP	MAKE2	RECYCLE	REC2
Description						
From		DEOIL		MKPUMP		RYCPUMP
To		MIX3	MKPUMP	MIX3	RYCPUMP	MIX3
Stream Class		CONVEN	CONVEN	CONVEN	CONVEN	CONVEN
Maximum Relative Error						
Cost Flow	$/hr					
MIXED Substream						
Phase		Liquid Phase	Liquid Phase	Liquid Phase	Liquid Phase	Liquid Phase
Temperature	F	158	100.000000000001	100.012052509227	100.000000000001	100.012052509227
Pressure	psig	-1.77635683940025E-15	-1.77635683940025E-15	10.2871641428594	-1.77635683940025E-15	10.2871641428594
Molar Vapor Fraction		0	0	0	0	0
Molar Liquid Fraction		1	1	1	1	1
Molar Solid Fraction		0	0	0	0	0

Mass Vapor Fraction		0	0	0	0	0
Mass Liquid Fraction		1	1	1	1	1
Mass Solid Fraction		0	0	0	0	0
Molar Enthalpy	Btu/lbmol	-122143.092048213	-123258.90273291	-123258.156499734	-123258.90273291	-123258.156499734
Mass Enthalpy	Btu/lb	-6775.11996847485	-6841.90879813749	-6841.86737590168	-6841.90879813749	-6841.86737590168
Molar Entropy	Btu/lbmol-R	-37.3776439040357	-39.3031560743171	-39.3029864264846	-39.3031560743171	-39.3029864264846
Mass Entropy	Btu/lb-R	-2.07328975664719	-2.18165668667471	-2.18164726978901	-2.18165668667471	-2.18164726978901
Molar Density	lbmol/cuft	3.29146551347696	3.40132610955506	3.40130341013392	3.40132610955506	3.40130341013392
Mass Density	lb/cuft	59.3391374701569	61.2758422349451	61.2754332985174	61.2758422349451	61.2754332985174
Enthalpy Flow	MMBtu/hr	-407.818276842852	-10.4587700085606	-10.4587066891493	-8.60741161600049	-8.60735950507002
Average MW		18.0281814368681	18.01528	18.01528	18.01528	18.01528
Mole Flows	kmol/hr	1514.47990729961	38.4882403646552	38.4882403646552	31.6752473687626	31.6752473687626
Mole Fractions						
Mass Flows	tons/year	263828.170677338	6700	6700	5514.00000000001	5514.00000000001
Mass Fractions						
METHA-01		0	0	0	0	0

ETHYL-01		0	0	0	0	0
ETHAN-01		0	0	0	0	0
PROPY-01		0	0	0	0	0
PROPA-01		0	0	0	0	0
N-BUT-01		0	0	0	0	0
N-PEN-01		0	0	0	0	0
BENZE-01		7.66313477367703E-05	0	0	0	0
TOLUE-01		1.4686858008174E-06	0	0	0	0
1-BUT-01		0	0	0	0	0
WATER		0.998974313514875	1	1	1	1
O-XYL-01		0	0	0	0	0
N-PEN-02		0	0	0	0	0
AIR		0	0	0	0	0
OXYGE-01		0	0	0	0	0
NITRO-01		0	0	0	0	0
CARBO-01		0	0	0	0	0
SODIU-01		0.000947586451587268	0	0	0	0
Volume Flow	gal/min	126.470423864436	3.11024464187113	3.11026539884116	2.55968491869811	2.55970200137465

Table A6. Stream results for water treatment outlets

	Units	RO	ROWASTE	CLEANW	TOBOILER	PURGE
Description						

From		MIX3	REVOSM	REVOSM	SPLIT1	SPLIT1
To		REVOSM		SPLIT1		
Stream Class		CONVEN	CONVEN	CONVEN	CONVEN	CONVEN
Maximum Relative Error						
Cost Flow	$/hr					
MIXED Substream						
Phase		Liquid Phase	Liquid Phase	Liquid Phase	Liquid Phase	Liquid Phase
Temperature	F	155.440628520096	155.440628520096	155.440628520096	155.440628520096	155.440628520096
Pressure	psig	-1.77635683940025E-15	-1.77635683940025E-15	-1.77635683940025E-15	-1.77635683940025E-15	-1.77635683940025E-15
Molar Vapor Fraction		0	0	0	0	0
Molar Liquid Fraction		1	1	1	1	1
Molar Solid Fraction		0	0	0	0	0
Mass Vapor Fraction		0	0	0	0	0
Mass Liquid Fraction		1	1	1	1	1
Mass Solid Fraction		0	0	0	0	0

Molar Enthalpy	Btu/lbmol	-122192.463920277	-183979.370395593	-122175.250867298	-122175.250867298	-122175.250867298
Mass Enthalpy	Btu/lb	-6778.07333184071	-3148.04234652629	-6781.34981196991	-6781.34981196991	-6781.34981196991
Molar Entropy	Btu/lbmol-R	-37.4589296775497	-24.6480876835423	-37.4614125984095	-37.4614125984095	-37.4614125984095
Mass Entropy	Btu/lb-R	-2.07786441275422	-0.421749588673139	-2.07929954288597	-2.07929954288597	-2.07929954288597
Molar Density	lbmol/cuft	3.29638365716675	1.77021231390519	3.29480107539238	3.29480107539238	3.29480107539238
Mass Density	lb/cuft	59.425919634651	103.455580049035	59.360327826382	59.360327826382	59.360327826382
Enthalpy Flow	MMBtu/hr	-426.884343037072	-0.178841557022389	-426.705445155154	-418.195880024916	-8.50956513023833
Average MW		18.027610198058	58.44247	18.0163616764975	18.0163616764975	18.0163616764975
Mole Flows	kmol/hr	1584.64339503303	0.440925335975703	1584.20246969705	1552.60954242501	31.5929272720429
Mole Fractions						
Mass Flows	tons/year	276042.170677338	249.000000045769	275793.170677292	270293.170677292	5500.00000000002
Mass Fractions						
METHA-01		0	0	0	0	0
ETHYL-01		0	0	0	0	0
ETHAN-01		0	0	0	0	0
PROPY-01		0	0	0	0	0
PROPA-01		0	0	0	0	0
N-BUT-01		0	0	0	0	0

N-PEN-01		0	0	0	0	0
BENZE-01		7.32406510219885E-05	0	7.33067763798537E-05	7.33067763798537E-05	7.33067763798537E-05
TOLUE-01		1.40370106197417E-06	0	1.40496839416966E-06	1.40496839416966E-06	1.40496839416966E-06
1-BUT-01		0	0	0	0	0
WATER		0.999019696923862	0	0.999921662349506	0.999921662349506	0.999921662349506
O-XYL-01		0	0	0	0	0
N-PEN-02		0	0	0	0	0
AIR		0	0	0	0	0
OXYGE-01		0	0	0	0	0
NITRO-01		0	0	0	0	0
CARBO-01		0	0	0	0	0
SODIU-01		0.000905658724054595	1	3.62590572012316E-06	3.62590572012316E-06	3.62590572012316E-06
Volume Flow	gal/min	132.13216766542	0.0684627696327189	132.158851013062	129.523275669457	2.63557534360546

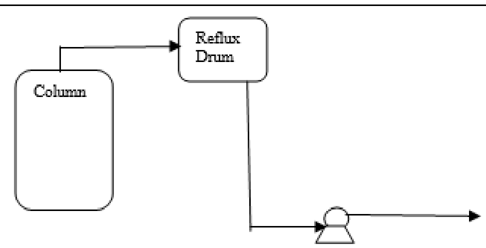
Table A7. Aspen HYSYS Stream Results.

Stream Name	Steam	air	fuel	outlet	water
Vapour / Phase Fraction	1.00000000000000	1.00000000000000	1.00000000000000	1.00000000000000	0.000000000000000
Temperature [C]	900.000000000000	49.0000000000000	49.0000000000000	957.861861819117	49.0000000000000
Pressure [kPa]	100.000000000000	100.000000000000	100.000000000000	100.000000000000	100.000000000000
Molar Flow [kgmole/h]	1710.89347205876	3723.11924421586	274.335102205379	4014.49618143469	1710.89347205876

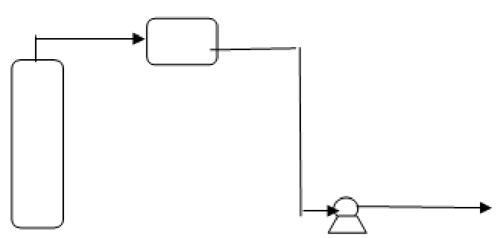
Mass Flow [kg/h]	30821.9178082192	107412.996873988	5188.02490062013	112601.021774608	30821.9178082192
Std Ideal Liq Vol Flow [m3/h]	30.8841177125259	124.169500618667	15.6558994414509	134.913226090559	30.8841177125259
Molar Enthalpy [kJ/kgmole]	-208424.544935958	694.404474475227	-64627.1459436831	-36148.5950826550	-284354.825310974
Molar Entropy [kJ/kgmole-C]	224.564314467856	154.067331129845	185.577976861852	207.956943606845	59.7227685525714
Heat Flow [kJ/h]	-356592193.347747	2585350.66218832	-17729494.6877023	-145118396.923548	-486500814.372953
Liq Vol Flow @Std Cond [m3/h]	30.3721925207187	87971.7520616840	6464.48623755004	94627.5287334173	30.3721925207187

Appendix B: Equipment Sheets

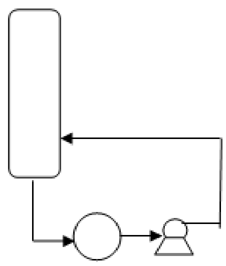
Olin Engineering Process Pump Load Sheet	
Project Name: Plastics Repurposing Plant	Item Name: Column Water Pump
Prepared by: James Hayes	Checked by: Joseph Yoon
OPERATING CONDITIONS	
Fluid type:	Liquid Water
Suction Temperature °F	120
Vapor Pressure @ Suction Temperature PSIA	
Viscosity @ Suction Temperature	
FLOWRATE	
Volumetric Flow Rate @ op. conditions GPM	121.6 GPM
I Mass Flow Rate Lb/Hour	60,200 lb/hr (263,600 tons/yr)
Specific Gravity @ Flow Temperature	0.989
Specific Gravity @ 60 °F	1.00
EXTREME OPERATING CONDITIONS	
Maximum Temperature °F	200
Vapor Pressure @ Maximum T PSIA	7.5
Minimum Temperature °F	100
Viscosity at Minimum T c stokes	
SUCTION CONDITIONS	
Estimated Suction Pressure PSIA	66
Estimated Discharge Pressure PSIA	110
Estimated NPSH Available Feet	32
HYDRAULIC CIRCUIT SKETCH	



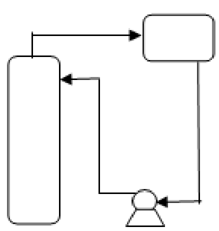
Olin Engineering Process Pump Load Sheet	
Project Name: Plastics Repurposing Plant	Item Name: Liquid Hydrocarbon Product Pump
Prepared by: James Hayes	Checked by:
OPERATING CONDITIONS	
Fluid type:	Liquid Hydrocarbon
Suction Temperature °F	120
Vapor Pressure @ Suction Temperature PSIA	
Viscosity @ Suction Temperature	
FLOWRATE	
Volumetric Flow Rate @ op. conditions GPM	1.41 GPM
I Mass Flow Rate Lb/Hour	482 lb/hr (2,110 tons/yr)
Specific Gravity @ Flow Temperature	0.681
Specific Gravity @ 60 °F	0.789
EXTREME OPERATING CONDITIONS	
Maximum Temperature °F	200
Vapor Pressure @ Maximum T PSIA	
Minimum Temperature °F	100
Viscosity at Minimum T c stokes	
SUCTION CONDITIONS	
Estimated Suction Pressure PSIA	66
Estimated Discharge Pressure PSIA	110
Estimated NPSH Available Feet	32
HYDRAULIC CIRCUIT SKETCH	



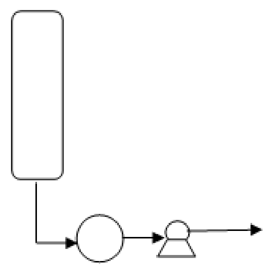
Olin Engineering Process Pump Load Sheet	
Project Name: Plastics Repurposing Plant	Item Name: Pumparound
Prepared by: James Hayes	Checked by:
OPERATING CONDITIONS	
Fluid type:	Liquid Wax
Suction Temperature °F	487
Vapor Pressure @ Suction Temperature PSIA	
Viscosity @ Suction Temperature	
FLOWRATE	
Volumetric Flow Rate @ op. conditions GPM	348.7 GPM
I Mass Flow Rate Lb/Hour	134,000 lb/hr (585,000 tons/yr)
Specific Gravity @ Flow Temperature	0.776
Specific Gravity @ 60 °F	0.805
EXTREME OPERATING CONDITIONS	
Maximum Temperature °F	600
Vapor Pressure @ Maximum T PSIA	
Minimum Temperature °F	400
Viscosity at Minimum T c stokes	
SUCTION CONDITIONS	
Estimated Suction Pressure PSIA	60
Estimated Discharge Pressure PSIA	70
Estimated NPSH Available Feet	175
HYDRAULIC CIRCUIT SKETCH	



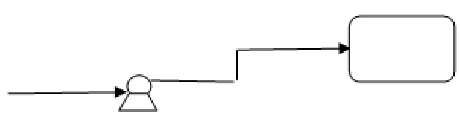
Olin Engineering Process Pump Load Sheet	
Project Name: Plastics Repurposing Plant	Item Name: Reflux Pump
Prepared by: James Hayes	Checked by:
OPERATING CONDITIONS	
Fluid type:	Liquid Hydrocarbon
Suction Temperature °F	120
Vapor Pressure @ Suction Temperature PSIA	
Viscosity @ Suction Temperature	
FLOWRATE	
Volumetric Flow Rate @ op. conditions GPM	20.4 GPM
I Mass Flow Rate Lb/Hour	6,781 lb/hr (29,700 tons/yr)
Specific Gravity @ Flow Temperature	0.664
Specific Gravity @ 60 °F	0.789
EXTREME OPERATING CONDITIONS	
Maximum Temperature °F	200
Vapor Pressure @ Maximum T PSIA	
Minimum Temperature °F	100
Viscosity at Minimum T c stokes	
SUCTION CONDITIONS	
Estimated Suction Pressure PSIA	60
Estimated Discharge Pressure PSIA	70
Estimated NPSH Available Feet	32
HYDRAULIC CIRCUIT SKETCH	



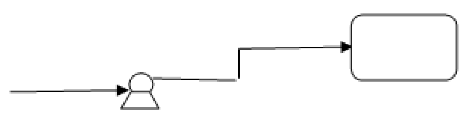
Olin Engineering Process Pump Load Sheet	
Project Name: Plastics Repurposing Plant	Item Name: Pumparound
Prepared by: James Hayes	Checked by:
OPERATING CONDITIONS	
Fluid type:	Liquid Wax
Suction Temperature °F	487
Vapor Pressure @ Suction Temperature PSIA	
Viscosity @ Suction Temperature	
FLOWRATE	
Volumetric Flow Rate @ op. conditions GPM	348.7 GPM
I Mass Flow Rate Lb/Hour	66,900 lb/hr (293,000 tons/yr)
Specific Gravity @ Flow Temperature	0.776
Specific Gravity @ 60 °F	0.805
EXTREME OPERATING CONDITIONS	
Maximum Temperature °F	600
Vapor Pressure @ Maximum T PSIA	
Minimum Temperature °F	400
Viscosity at Minimum T c stokes	
SUCTION CONDITIONS	
Estimated Suction Pressure PSIA	60
Estimated Discharge Pressure PSIA	110
Estimated NPSH Available Feet	175
HYDRAULIC CIRCUIT SKETCH	



Olin Engineering Process Pump Load Sheet	
Project Name: Plastics Repurposing Plant	Item Name: Recycle Water Pump
Prepared by: James Hayes	Checked by:
OPERATING CONDITIONS	
Fluid type:	Liquid Water
Suction Temperature °F	100
Vapor Pressure @ Suction Temperature PSIA	
Viscosity @ Suction Temperature	
FLOWRATE	
Volumetric Flow Rate @ op. conditions GPM	2.81 GPM
I Mass Flow Rate Lb/Hour	1,260 lb/hr (5,500 tons/yr)
Specific Gravity @ Flow Temperature	0.984
Specific Gravity @ 60 °F	1.00
EXTREME OPERATING CONDITIONS	
Maximum Temperature °F	
Vapor Pressure @ Maximum T PSIA	
Minimum Temperature °F	
Viscosity at Minimum T c stokes	
SUCTION CONDITIONS	
Estimated Suction Pressure PSIA	15
Estimated Discharge Pressure PSIA	25
Estimated NPSH Available Feet	31
HYDRAULIC CIRCUIT SKETCH	

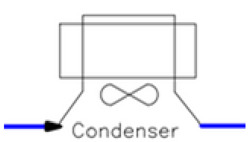


Olin Engineering Process Pump Load Sheet	
Project Name: Plastics Repurposing Plant	Item Name: Makeup Water Pump
Prepared by: James Hayes	Checked by: Joseph Yoon
OPERATING CONDITIONS	
Fluid type:	Liquid Water
Suction Temperature °F	100
Vapor Pressure @ Suction Temperature PSIA	
Viscosity @ Suction Temperature	
FLOWRATE	
Volumetric Flow Rate @ op. conditions GPM	3.42 GPM
I Mass Flow Rate Lb/Hour	1,530 lb/hr (6,700 tons/yr)
Specific Gravity @ Flow Temperature	0.983
Specific Gravity @ 60 °F	1.00
EXTREME OPERATING CONDITIONS	
Maximum Temperature °F	
Vapor Pressure @ Maximum T PSIA	
Minimum Temperature °F	
Viscosity at Minimum T c stokes	
SUCTION CONDITIONS	
Estimated Suction Pressure PSIA	15
Estimated Discharge Pressure PSIA	25
Estimated NPSH Available Feet	31
HYDRAULIC CIRCUIT SKETCH	



Olin Engineering Heat Exchanger Load Sheet					
Prepared by: Joseph Yoon					
		COLD FLUID		HOT FLUID	
FLUID ALLOCATION					
Fluid Name		Air		Water, Hydrocarbons	
Fluid Quantity; Total	Lb/Hr	460,000 (2e6 tons/yr)		68,900 lbm/hr (301,700 Tons/yr)	
		In	Out	In	Out
Vapor	Lb/Hr	460,000	460,000	8,700	1,450
Liquid	Lb/Hr				7,260
Steam	Lb/Hr			60,200	
Water	Lb/Hr				60,200
FLUID PROPERTIES					
Temperature	°F	100	120	300	120
Density	Lbs/Cubic Foot	0.8	0.8		62.4
Viscosity	Centistokes				
Molecular Weight		29	29	18	18
Specific Heat	BTU/Lb/°F	0.24	0.24	0.445	1
Thermal Conductivity					
Phase Change Enthalpy	BTU/Lb				
DESIGN BASIS					
Heat Transferred	MM BTU/Hr	72.9			
Pressure Drop Allowed	PSI				
Fouling Resistances		-		-	
Expected Individual Coefficients	BTU/Hr/SqFt//°F				
LMTD Correction Factor		-			
Overall U Clean	BTU/Hr/SqFt//°F	-			
Overall U Fouled	BTU/Hr/SqFt//°F	-			
Expected Surface Area	Square Feet	4560			
Recommended Tube Diameter	Inches				
Recommended Tube Length	Feet				
Expected Number of Tubes					
Expected Bundle Diameter	Inches				

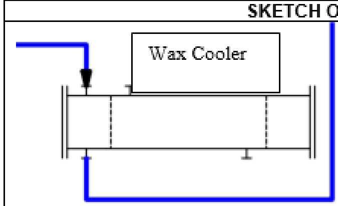
SKETCH OF HEAT EXCHANGER CONFIGURATION



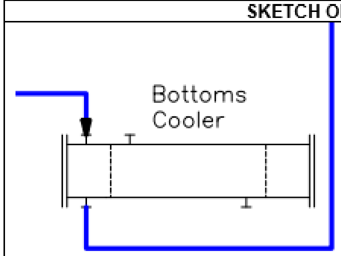
Olin Engineering Heat Exchanger Load Sheet

Prepared by: Joseph Yoon					
		COLD FLUID		HOT FLUID	
FLUID ALLOCATION					
Fluid Name		Water (to Superheated Steam)		Wax	
Fluid Quantity; Total	Lb/Hr	8,700		66,700	
		In	Out	In	Out
Vapor	Lb/Hr				
Liquid	Lb/Hr			66,700	66,700
Steam	Lb/Hr		8,700		
Water	Lb/Hr	8,700			
FLUID PROPERTIES					
Temperature	°F	120	350	487	150
Density	Lbs/Cubic Foot	62.4			40.3
Viscosity	Centistokes				
Molecular Weight		18	18	329	329
Specific Heat	BTU/Lb/°F	1	1	0.696	0.520
Thermal Conductivity					
Phase Change Enthalpy	BTU/Lb	970			
DESIGN BASIS					
Heat Transferred	MM BTU/Hr	13.8			
Pressure Drop Allowed	PSI				
Fouling Resistances		-		-	
Expected Individual Coefficients	BTU/Hr/SqFt//°F				
LMTD Correction Factor		-			
Overall U Clean	BTU/Hr/SqFt//°F	-			
Overall U Fouled	BTU/Hr/SqFt//°F	-			
Expected Surface Area	Square Feet	4560			
Recommended Tube Diameter	Inches				
Recommended Tube Length	Feet				
Expected Number of Tubes					
Expected Bundle Diameter	Inches				

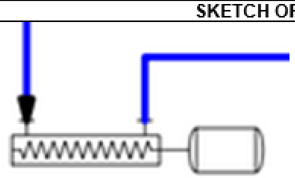
SKETCH OF HEAT EXCHANGER CONFIGURATION



Olin Engineering Heat Exchanger Load Sheet					
Prepared by: Joseph Yoon					
		COLD FLUID		HOT FLUID	
FLUID ALLOCATION					
Fluid Name		Water (Superheated Steam)		Wax	
Fluid Quantity; Total	Lb/Hr	26,000		66,800	
		In	Out	In	Out
Vapor	Lb/Hr				
Liquid	Lb/Hr			66.800	66,800
Steam	Lb/Hr		26,000		
Water	Lb/Hr	26,000			
FLUID PROPERTIES					
Temperature	°F	100	350	787	487
Density	Lbs/Cubic Foot	62.4			40.3
Viscosity	Centistokes				
Molecular Weight		18	18	329	329
Specific Heat	BTU/Lb/°F	1	1	0.696	0.696
Thermal Conductivity					
Phase Change Enthalpy	BTU/Lb	970			
DESIGN BASIS					
Heat Transferred	MM BTU/Hr	41.1			
Pressure Drop Allowed	PSI				
Fouling Resistances		-		-	
Expected Individual Coefficients	BTU/Hr/SqFt//°F				
LMTD Correction Factor		-			
Overall U Clean	BTU/Hr/SqFt//°F	-			
Overall U Fouled	BTU/Hr/SqFt//°F	-			
Expected Surface Area	Square Feet	2,140			
Recommended Tube Diameter	Inches	1			
Recommended Tube Length	Feet	20			
Expected Number of Tubes					
Expected Bundle Diameter	Inches				
SKETCH OF HEAT EXCHANGER CONFIGURATION					



Olin Engineering Heat Exchanger Load Sheet					
Prepared by: Joseph Yoon					
		COLD FLUID		HOT FLUID	
FLUID ALLOCATION					
Fluid Name		Water (Superheated Steam)		HDPE	
Fluid Quantity; Total	Lb/Hr	8,700		68,500	
		In	Out	In	Out
Vapor	Lb/Hr				
Liquid	Lb/Hr			68,500	68,500
Steam	Lb/Hr		8,700		
Water	Lb/Hr	8,700			
FLUID PROPERTIES					
Temperature	°F	400	120	100	350
Density	Lbs/Cubic Foot	62.4			40.3
Viscosity	Centistokes				
Molecular Weight		18	18	329	329
Specific Heat	BTU/Lb/°F	0.445	1	0.597	0.597
Thermal Conductivity					
Phase Change Enthalpy	BTU/Lb	970			
DESIGN BASIS					
Heat Transferred	MM BTU/Hr	15.0			
Pressure Drop Allowed	PSI				
Fouling Resistances		-		-	
Expected Individual Coefficients	BTU/Hr/SqFt/°F				
LMTD Correction Factor		-			
Overall U Clean	BTU/Hr/SqFt/°F	-			
Overall U Fouled	BTU/Hr/SqFt/°F	-			
Expected Surface Area	Square Feet	4560			
Recommended Tube Diameter	Inches				
Recommended Tube Length	Feet				
Expected Number of Tubes					
Expected Bundle Diameter	Inches				
SKETCH OF HEAT EXCHANGER CONFIGURATION					

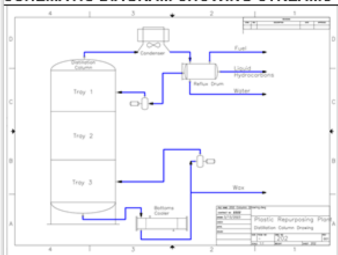


P-001

OLIN ENGINEERING FRACTIONATOR COLUMN LOAD SHEET

Prepared By: James Hayes	Checked By:

SCHEMATIC DIAGRAM SHOWING STREAMS



PROCESS NOTES

Design Pressure	44 psig
Top Temperature	120°F
Bottoms Temperature	487°F
Design Temperature	
Wall thickness	0.3125 in

Distillation Tray Load Sheet

Tray Number		Tray	VAPOR TO TRAY								LIQUID TO TRAY				
			T	P	Mass	Volume	Density				Mass	Volume	Density	Vis	Surface Tension
		Type				Flow	@					GPM@	@		
Theor.	Actual		°F	PSIA	Lb/Hr	ACFS	P & T				Lb/Hr	P & T	P & T	CP	Dynes/cm
	1		120	59	1443.49	0.01435					6836.87	17.3741	0.684		
	2		454	59	68414.8	0.32115					7236.67	18.0265			
	3		787	59	69295.5	0.32533					200370	499.175	0.646		

Final Presentation Questions:

Q1: To insert the N_2 blanket and purge the system, you will need to bring the N_2 up to process pressure (4 atm). How will you accomplish this?

The N_2 is shipped in pressurized tanks as a liquid and placed in pressurized tanks on arrival. For this reason, we will assume that expanders are needed to decrease pressure. We have found that it is an industry norm for the provider of the N_2 to also provide necessary equipment for it. This would include expanders/compressors and refrigerated storage tanks.

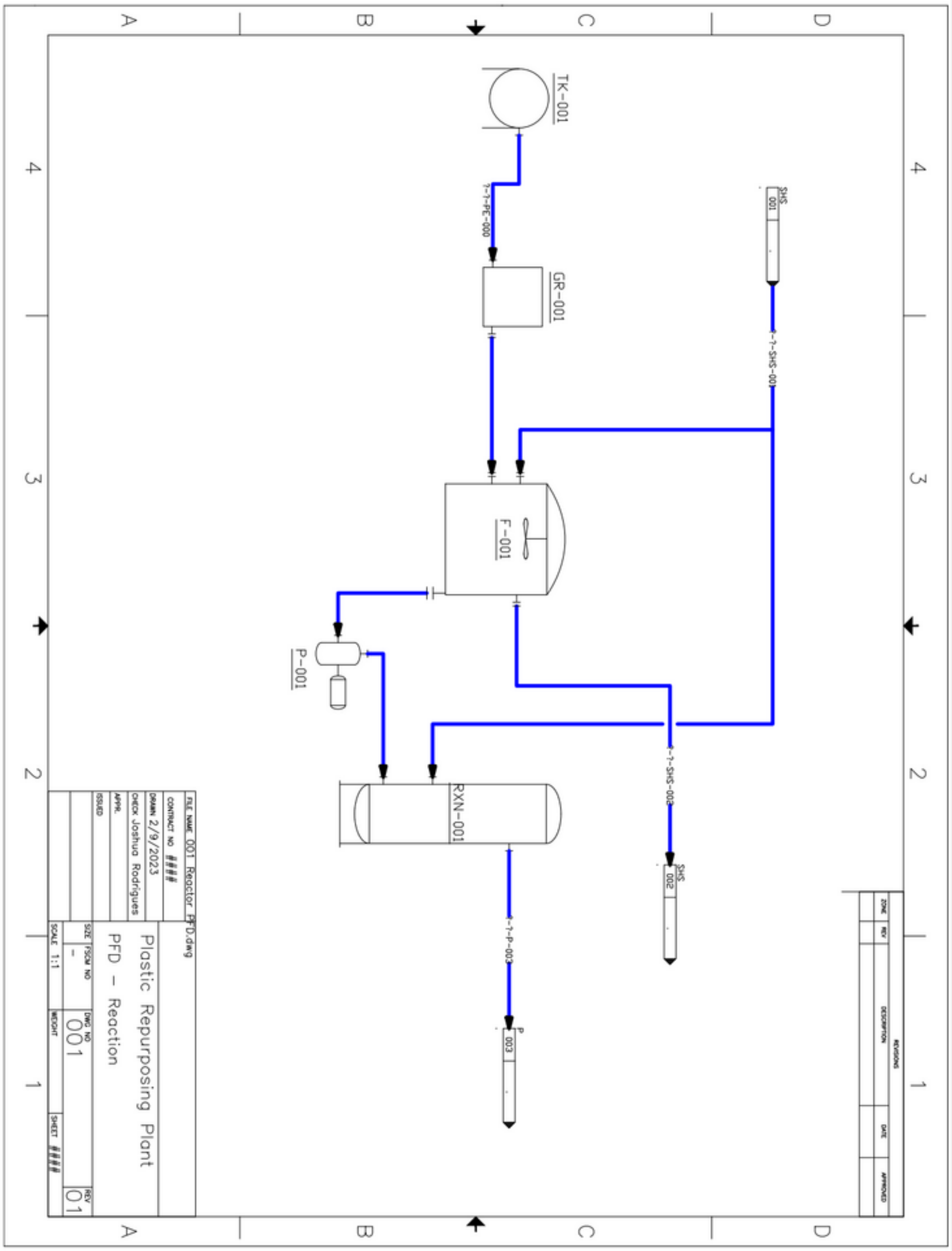
Q2: What would happen if unreacted HDPE found its way into the distillation column?

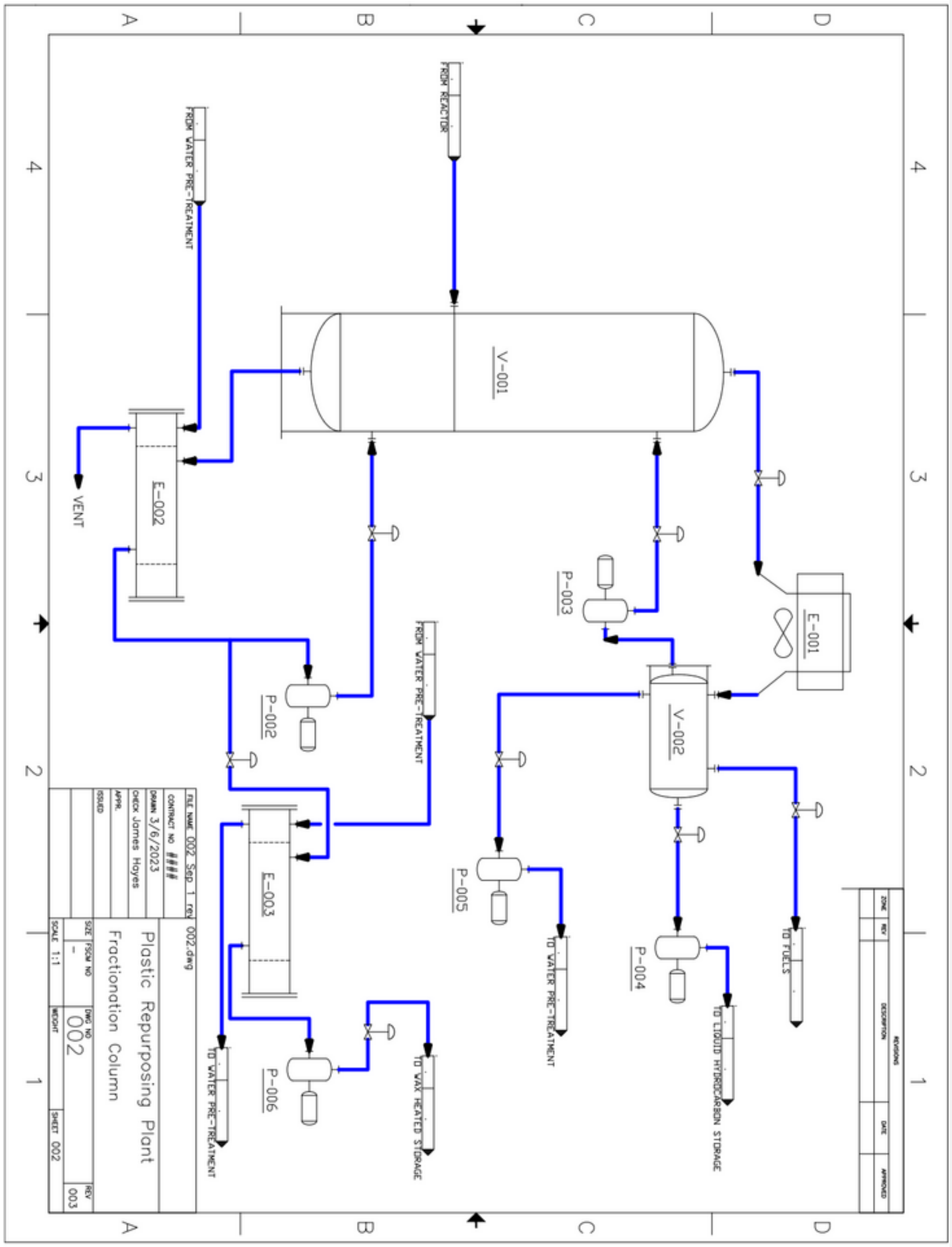
At the operating temperature (930 °F), unreacted HDPE will not boil and leave as a vapor. This means any unreacted HDPE would leave as droplets, most likely due to fluidization velocity being too high. To avoid these unreacted droplets from leaving a mesh will be placed at the outlet of the reactor to collect any droplets before they leave the reactor.

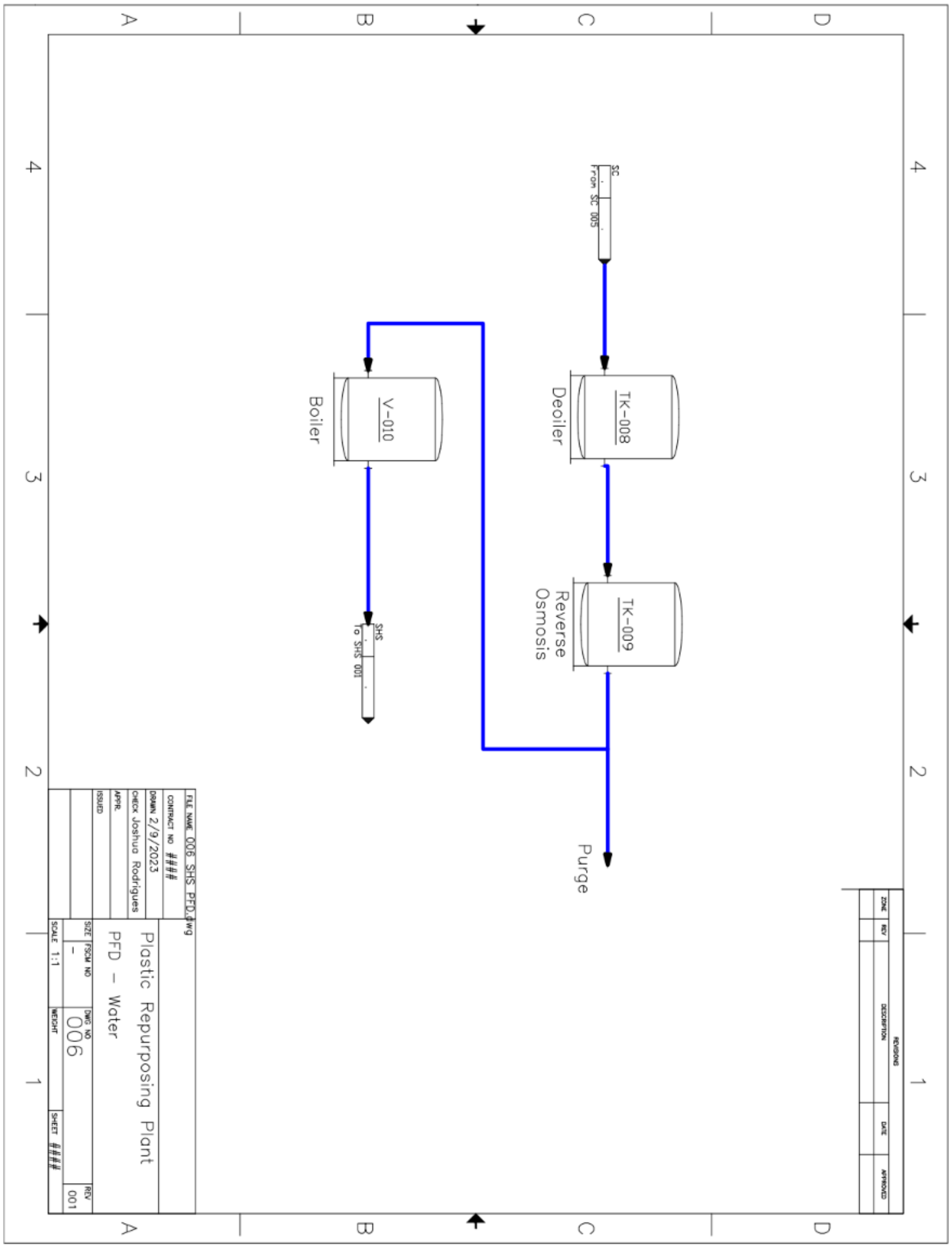
Q3: What if the system goes outside the range of required fluidization velocity?

In the case of the fluidization velocity being too high, the unreacted droplets will be pushed up to the outlet and will be collected on the droplet mesh. In the case the fluidization velocity is too low, the force of gravity on the droplets will exceed its upward acceleration, causing the droplets to accumulate at the bottom.

Appendix C: Process Flow Diagram (PFD)




FROM REACTOR
FROM WATER PRE-TREATMENT
V-001
E-001
E-002
E-003
V-002
P-002
P-003
P-004
P-005
P-006
VENT
TO FUELS
TO LIQUID HYDROCARBON STORAGE
TO WATER PRE-TREATMENT
TO WAX HEATED STORAGE
FROM WATER PRE-TREATMENT
TO WATER PRE-TREATMENT
FILE NAME 002 Sep 1 rev 002.dwg
CONTRACT NO ####
DRAWN 3/6/2023
CHECK James Hoyes
APPR.
ISSUED
Plastic Repurposing Plant
Fractionation Column
SIZE FSCM NO
DWG NO 002
REV 003
SCALE 1:1
WEIGHT
SHEET 002
REVISIONS
ZONE
REV
DESCRIPTION
DATE
APPROVED
A
B
C
D
1
2
3
4



SC
From SC 005
TK-008
Deoiler
TK-009
Reverse Osmosis
V-010
Boiler
SHS
To SHS 001
Purge
REVISIONS
ZONE
REV
DESCRIPTION
DATE
APPROVED
FILE NAME 006_SHS_PFD.dwg
CONTRACT NO ####
DRAWN 2/9/2023
CHECK Joshua Rodrigues
APPR.
ISSUED
Plastic Repurposing Plant
PFD - Water
SIZE -
FSCM NO
DWG NO 006
REV 001
SCALE 1:1
WEIGHT
SHEET ####