PRODUCTION OF 3000 TPD OF BIO-ETHYLENE FROM ZEA-MAYS

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Supervised By

Engr. Usman Asghar

Group Members

Kashaf Tehreem	UW-19-CHE-BSc-008
Shehar Bano	UW-19-CHE-BSc-001
Abdullah Niaz	UW-19-CHE-BSc-020
Sami Jabbar	UW-19-CHE-BSc-010

Department of Chemical Engineering
Wah Engineering College
University of Wah, Wah Cantt

Production of 3000 Tons/Day of Bio-Ethylene from Zea-Mays

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Department of Chemical Engineering,
Wah Engineering College,
University of Wah
Wah Cantt.

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Abstract

The nonrenewable nature of fossil fuels and their connection to the buildup of greenhouse gases in the atmosphere have been recognized for a long time. As a result, renewable methods have been developed that utilize both non-biomass sources like wind, solar, geothermal, and hydroelectric power, and biomass sources that can be directly combusted or converted into value-added products using various thermochemical processes or using microorganisms. This combination of microorganisms and biomass has paved the way for the creation of a bio-economy, enabling the commercial production of biofuels, bio-chemicals, and other miscellaneous materials. This study focuses on the production of bio-ethylene from Zea Mays (a waste biomass of corn). Ethylene is a key feedstock for various downstream chemical products like PET, ethylene oxide etc. It is responsible for the production of about half of all plastics produced globally (a fast-growing industry all over the world). Annually, over 140 million tons of ethylene are produced, and demand for it is expected to rise, particularly in developing economies. With its resemblance to ethylene in terms of chemical makeup, bio-ethylene can also be used to make plastics and other downstream products with the current machinery and production capacity. This study provides techno-economic analysis for the production of bio-ethylene from Zea Mays (corn stover). The production of bio-ethylene from biomass-based pathways involves two primary steps: fermentation to produce bioethanol from biomass, followed by the catalytic dehydration of bioethanol to bio-ethylene. This research work discusses all the technical aspects with the economic evaluation of the production process. All indicators of economic analysis (Net Present Worth, IRR etc.) shows the Bio-Ethylene from Zea Mays is economically viable (27.8% rate of return) to produce value added product (bio-ethylene) with least environmental implications.

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CHAPTER # 01

INTRODUCTION

1.1 Introduction:

Ethylene, also referred to as ethene (CH2=CH2), is the first alkene. It is a colorless gas with a commonplace boiling point of -103.7 °C and is just marginally soluble in water and alcohol. Due to its high activity, this substance reacts readily when combined with various chemical reagents [1]. Ethylene due to its simple molecular formula and composition serves as the basic raw material for Plastic and Polymer industry. Ethylene is the important feed stock used in the production of various valuable products like plastics, PET bottles, PVC pipes, ethylene glycol etc. Almost 60% of Ethylene is used in Polyethylene manufacturing. At the present time, Ethylene is being produced from the cracking of fossil fuel based raw materials like Naphtha, Natural gas, shale gas etc. As we know, fossil fuels are depleting day by day and their prices are on the rise which is causing ethylene to get more expensive and eventually affecting the polymer industry. A number of people are now interested in making plastics using non-renewable feedstock. Utilizing biomass to create usable goods contributes to the reduction of greenhouse gas emissions. [2].

In this project, we have designed a complete process for the sustainable production of Ethylene (also known as Bio-Ethylene due to its production from biomass) from a cheap waste material, i.e., Corn stover (Zea-Mays) which is the remaining of the corn crop. Farmer used to get rid of this waste (corn stover) by land filling or by burning it. This action involves the serious environmental impacts (emission of GHG from its burning is major impact). Our project is a value addition project (converting waste into valuable product) which contains following Commercial advantages over the existing process:

i. Reducing Environmental Hazards
ii. Raw material is inexpensive which will obviously drastically affect the selling price of Ethylene. The low-price Ethylene will give the boost to the polymer industry of Pakistan (as ethylene is feed stock for polymer industry)
iii. This process is sustainable in term Ethylene production because the conventional production involves the fossil fuel based raw material which depleting day by day.
iv. This process is the first step towards bio-refinery in Pakistan.

1.2 Physical Properties:

Table 1.1: Physical Properties of Ethylene

Description	**Properties**
Chemical Formula	C2H4
Molecular weight	28.05 g/mol
Boiling Point	-103.7 °C
Melting Point	-169.2 °C
Colour	Colourless
Odour	Odorless
PH	2.5
Specific gravity	0.9740

1.3 Thermodynamic Data:

Table 1.2: Thermodynamic Properties of Ethylene

Description	Properties
Heat capacity, Cp (gas)	42.9 J/mol K
Thermal conductivity	0.020 W/m°C
Specific volume	0.0247 m3/mol
Flash Point	137 K

1.4 Reactions of Ethylene:

1.4.1 Hydrogenation of Ethylene:

At normal temperature, it is easily hydrogenated under pressure while a platinum or palladium chemical catalyst is present. Under 200 to 300°C, nickel catalyst is used. Under normal conditions of pressure and temperature, raney nickel works well as a catalyst.

$$C_2H_4 + H_2 \rightarrow C_2H_6 \qquad (1.1)$$

1.4.2 Addition Reaction of Ethylene:

Chlorine or bromine and $CH_2{=}CH_2$ combine to generate an addition compound. It joins with halogen acids to generate an addition complex. For instance, ethylene bromide is created when $CH_2{=}CH_2$ adds hydrogen bromide. The addition of halogen acids reacts in the following order: HI > HBr > HCl > HF. After absorbing concentrated sulfuric acid, it is hydrolyzed to produce ethanol.

$$C_2H_4 + Br_2 \rightarrow C_2H_4Br_2 \qquad (1.2)$$

1.4.3 Hydroxylation of Ethylene:

It is easily converted into glycols by adding hydroxyl groups to it. It is changed into cis-ethylene glycol by a cold, diluted alkaline permanganate solution. Osmium tetroxide quickly forms cyclic compounds like osmic ester when it reacts with $CH_2{=}CH_2$. 1,2-glycol is produced by refluxing osmic ester with ethanolic sodium hydrogen sulphate.

$$C_2H_4 + H_2O + KMnO_4 \rightarrow C_2H_6O_2 \qquad (1.3)$$

1.4.4 Ozonolysis Reaction:

To create an ozonide chemical, it adds ozone gas molecules. To create formic acid, the ozonide is oxidized using silver oxide, hydrogen peroxide, or peracids. Ozonide is reduced with zinc dust to produce formaldehyde. To provide the appropriate alcohol, reduction can alternatively be done using sodium borohydride or lithium aluminum hydride.

$$C_2H_4 + O_3 \rightarrow 2CH_2O \qquad (1.4)$$

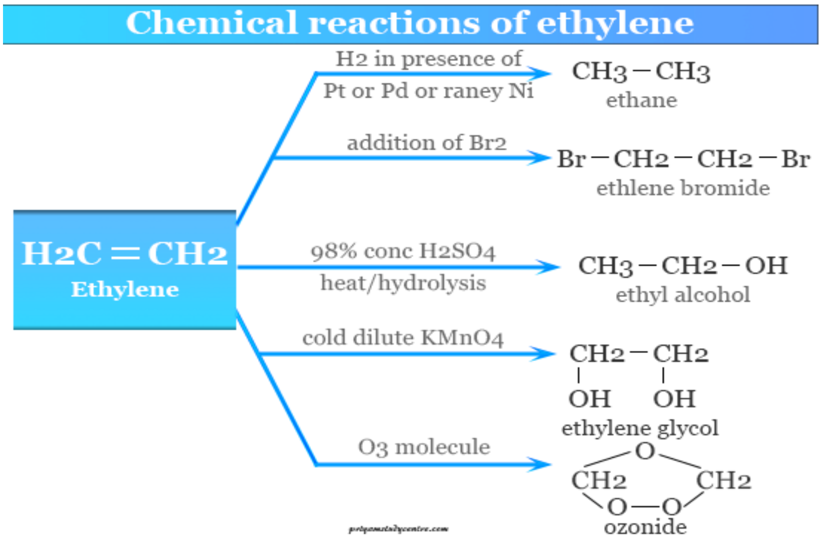


Figure 1.1: Chemical Reactions of Ethylene *[3]*

1.5 Industrial Applications:

There are many polymeric derivatives of ethylene that are employed in diverse contexts.

- ✓ Polyethylene is used to make packaging, stretch films, containers, barrels, pallets, and other goods.
- ✓ The manufacturing of polyester films, resins, and fibers all uses ethylene oxide.
- ✓ Polypropylene (PP) is used to make a variety of products, including films, sheets, foamed goods, industrial goods, reinforced goods, containers, etc.
- ✓ The most common material used to make soda bottles and other packaging is polyethylene terephthalate (PET).
- ✓ PVC, a polymer of vinyl chloride, is used to produce bottles and packaging.
- ✓ The fabrication of electrical and electronic gadgets uses polystyrene (PS).

1.6 Storage and Handling of the Product:

Around -103°C is the typical storage temperature for liquid ethylene. The use of cryogenic tanks is prevalent. Transferring ethylene typically involves the use of pipelines. Trucks can also transport liquid ethylene. For low temperatures, these tanks are built of carbon steel or stainless steel. [4].

Before working with ethylene, one must receive training on how to handle and store it.

- ✓ An explosion could result from the combination of ethylene with trifluoro-methyl sub-fluoride, ozone, and nitrogen dioxide. .
- ✓ The OXIDENTS (perchlorates, peroxides, permanganate, chlorate, nitrate, chlorine, bromine, and fluorine), Nitro-methane, strong acid (hydrochloric acid, sulfuric acid, nitric acid, etc.), and a chlorine rink are incompatible with ethylene.
- ✓ Store in an airtight container in a cool, well-ventilated place.
- ✓ Metal container for transporting ethylene must be grounded
- ✓ While opening and closing the ethylene container, only non-sparking tools and equipment should be used.

- ✓ Use electrical equipment and fittings that are explosion-proof where ethylene is used, processed, or manufactured.

1.7 Safety Hazard:

- ✓ Inhaling ethylene gas can have an impact on you.
- ✓ Contact with liquid ethylene can cause frostbite on the skin..
- ✓ Ethylene exposure can result in headaches, vertigo, weariness, drowsiness, confusion, and unconsciousness.
- ✓ Ethylene is one of the volatile compounds that can cause deadly flames and explosions.

1.8 Motivation:

- ✓ A step towards bio refinery as we are using biomass for production of ethylene.
- ✓ We are converting a waste into useful product hence it is a value addition.
- ✓ Biorefinery based ethylene is sustainable alternative to oil-based ethylene because it can reduce dependence on fossil fuels.
- ✓ Compared to fossil fuel-based ethylene, bio-ethylene can cut GHG emissions by 40%.

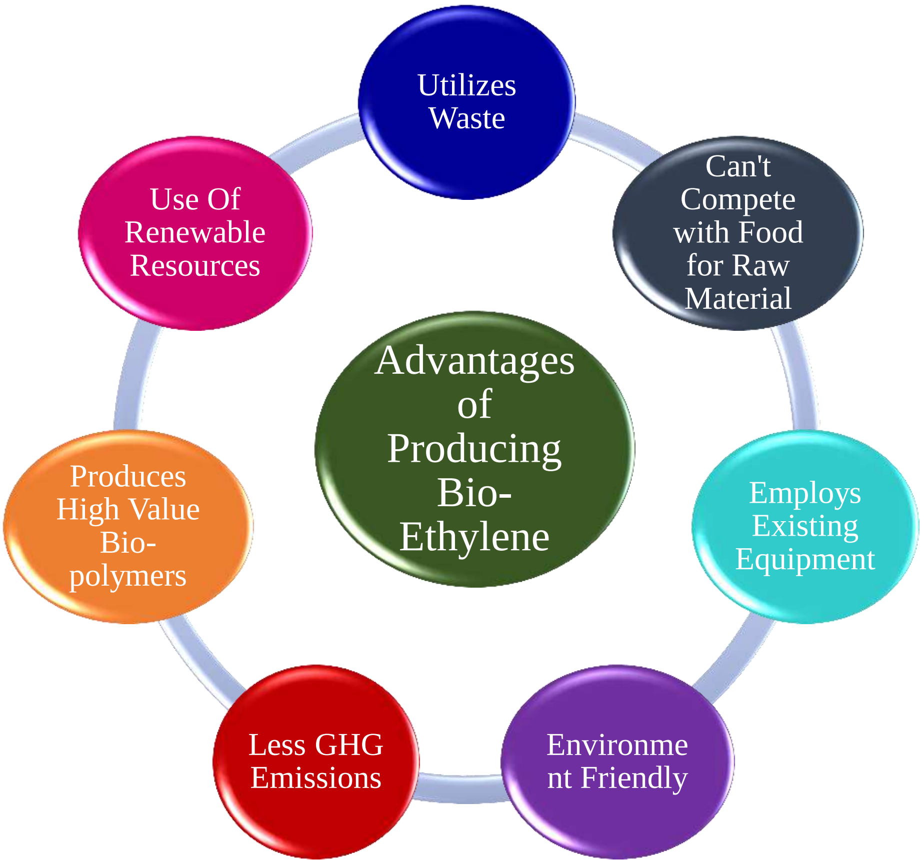


Figure 1.2: Advantages of Bio-Ethylene Production

CHAPTER # 02

PROCESS SELECTION

2.1 Global Ethylene Production and Consumption:

In 2021, the estimated value of the world's ethylene capacity was 216.35 mtpa. From 2021 to 2026, the market is anticipated to expand at an AAGR of more than 7%. The major nations in the globe, which together account for more than 50% of the world's ethylene capacity, are the US, China, Saudi Arabia, South Korea, and Iran. From 2020 to 2028, the demand for ethylene is anticipated to grow at a CAGR of 3.4%, reaching 233.9 million tons.

The expected increase in global ethylene usage from 2014 to 2020 is 4.3% per year. Additionally, estimates indicate that between 2020 and 2028, the growth rate of the world's ethylene consumption will climb by 3.4%.

The greatest user of ethylene worldwide, North America consumes 24% of the world's supply. 18% of the world's ethylene consumption is accounted for by China, the second-largest consumer. Third-largest user globally, the Middle East consumes 18% of the world's ethylene.

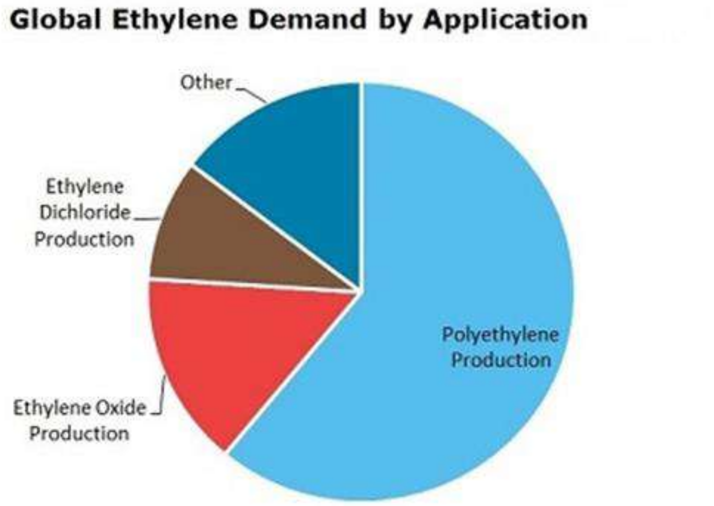


Figure 2.1: Ethylene Demand *[5]*

2.2 Ethylene Consumption and Production in Pakistan:

Almost 60% of ethylene is used in production of polyethylene. Pakistan's plastics sector is one of the nation's oldest, and evidence of its existence dates to 1947, the year Pakistan was founded. The years 1965 to 1975 marked a significant turning point in the usage of plastics in the country, when it became a powerful force with a sizable base.

Pakistan's plastics sector has made considerable progress toward success. Currently, plastics materials are the fourth most popular import, and this industry alone makes a considerable contribution to the national exchequer across multiple categories. The industry is expanding at a rate of 15% a year on average, and as it develops, it is outpacing all other industrial sectors. Over the past 15 years, the nation's per capita consumption has also increased.

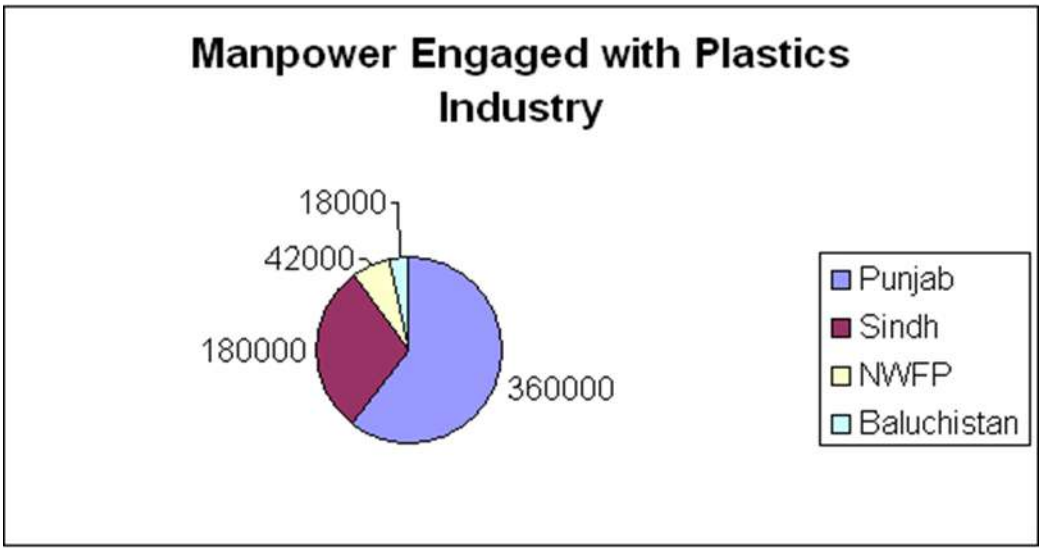


Figure 2.2: Manpower Engaged with Plastic Sector *[2]*

There are 6000 plastics processors operating in Pakistan now. The entire sector is a SME that is self-financing. Through customs duties, sales taxes, and income taxes, the sector gives the national exchequer about 8 billion rupees a year. There is a significant concentration of plastics processors in and around the industrial centres of Karachi, Lahore, Gujranwala, Peshawar, Faisalabad, Hyderabad, Rawalpindi, Gadoon, and Hattar, whether they be producers of woven, extruded, injection-molded, blow-molded, or tubular films.

Both the structured and un-organized sectors of the industry can be separated. The organized sector, which has between 600 and 700 units, is capable of manufacturing goods of high quality. Products of poor quality and low price are produced by the un-organized sector. Despite this, over the past 15 years, the un-organized sector has expanded more quickly than the structured sector.

2.3 Market Assessment:

The growth of the ethylene market is primarily being driven by the rising demand for polyethylene products across a variety of industries, while variables like the volatility of raw material costs could restrain market expansion. The going up demand for polyethylene products from a variety of industries, including consumer electronics, construction, and automotive, is one of the significant drivers fueling the expansion of the worldwide ethylene market. The market for polyethylene was estimated at USD 66.24 billion in 2021, and it is anticipated to expand at a CAGR of 3.7% over the next five years. The worldwide polyethylene market is also projected to be significantly impacted by the growing demand for plastic. [5].

2.3.1 Corn Production in Pakistan:

In Pakistan, maize has overtaken rice and wheat as the third-most important cereal crop, covering 1.3 million hectares. Pakistan's corn (maize) production increased by 7.79 percent between 2019 and 2020, reaching 7,800,000 tons. In 2020, maize production soared by 32.16 percent following a decline of 3.78 percent in 2017.

Presently, Pakistan's two most important provinces for corn production are Punjab and KPK. Nearly all of the country's maize (corn) production comes from these two provinces. In which KPK contributes 21% and Punjab contributes 76% of nation's grain production. In the provinces of Sindh and Baluchistan, maize (corn) grains are only produced by 2 to 3 percent. Azad Kashmir's 0.122 million hectares of land are used for maize (corn) production, which is also gaining prominence [6].

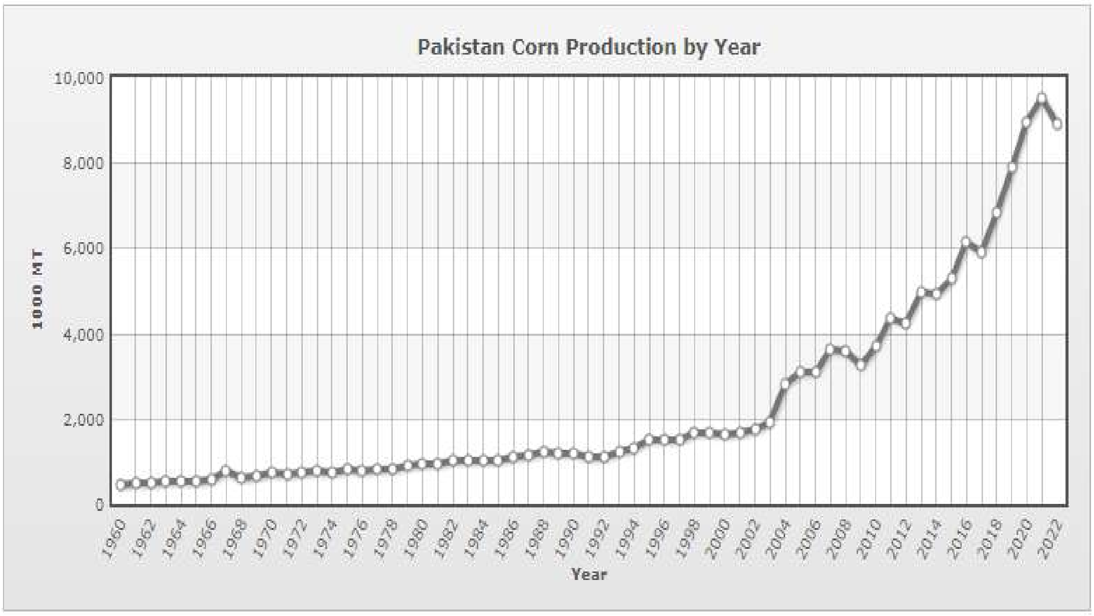


Figure 2.3: Pakistan Corn Production per Anum *[7]*

2.3.2 Worldwide Corn Production Data:

Worldwide, 1,060,247,727 tons of corn are produced each year. The largest maize producer in the world, the United States of America produces 384,777,890 tons of corn annually. China comes in second with a yearly production of 231,837,497 tons. China and the United States of America jointly create 58% of global output. [8].

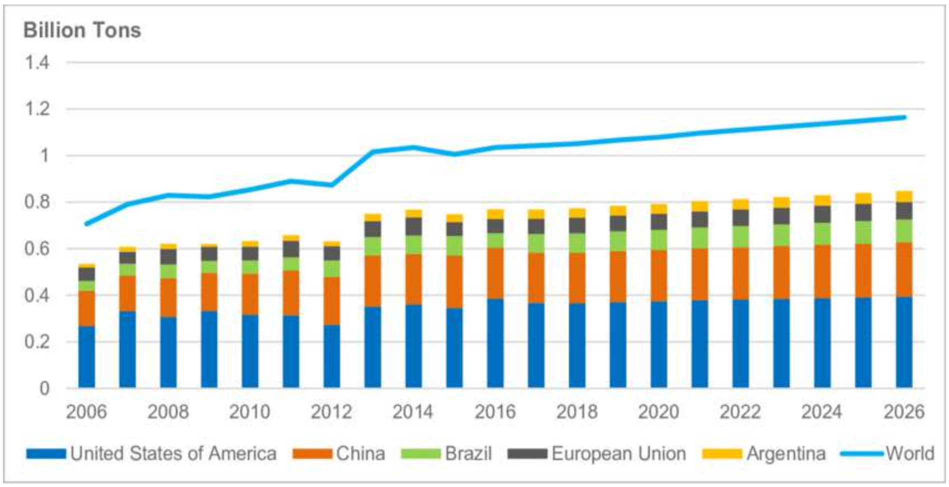
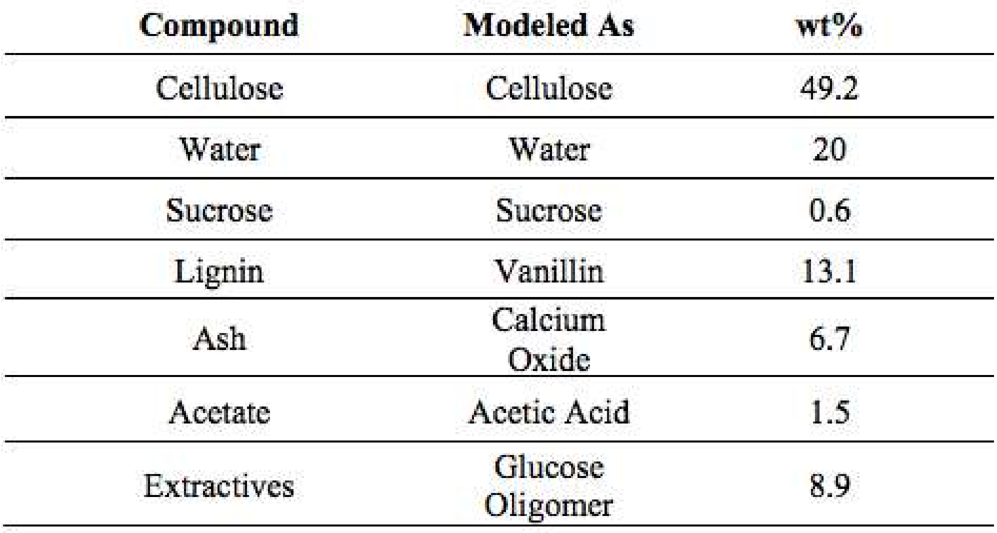


Figure 2.4: Worldwide Corn Production *[12]*

2.3.3 Raw Material:

As a raw material, zea mays, also known as corn stover, is utilized. The stalks, leaves, and husks that remain after harvesting corn are called stover. It is mostly made of cellulose, lignin, and water.



Compound	Modeled As	wt%
Cellulose	Cellulose	49.2
Water	Water	20
Sucrose	Sucrose	0.6
Lignin	Vanillin	13.1
Ash	Calcium Oxide	6.7
Acetate	Acetic Acid	1.5
Extractives	Glucose Oligomer	8.9

Figure 2.5: Raw Material Composition

2.3.4 Corn Silage Suppliers in Pakistan:

There are some companies that supply corn stover in all over Pakistan.

- Agri-complex Pakistan private limited
- Four Brothers Group Pakistan
- King Silage (Pvt) Ltd
- AIMS Agro Feeds
- A2ZEE Corporation

2.4 Capacity selection:

Corn production in Pakistan (2022 - 2023) = 8.9 million Tones

Punjab takes 76% part in corn production per year = 6.76 million Tones

2kg of corn stover contain 1kg of corn = 13.4 million Tons of Corn Stover

It is premised that we will safely and conveniently collect 40% of that waste

= 5.3 million Tones

Production of Bio-Ethylene = 3000 Tones / day

2.5 Manufacturing Processes:

The division occurring about because of the refining of gaseous petrol and oil are broken to produce ethene.
The processes are:

- ✓ The steam cracking of ethane and propane.
- ✓ The steam cracking of naphtha from crude oil
- ✓ The catalytic cracking of gas oil from crude oil.

Availability, pricing, and other products from cracking that are required will all affect the feedstock choice. Steam cracking is the primary method of ethene production.

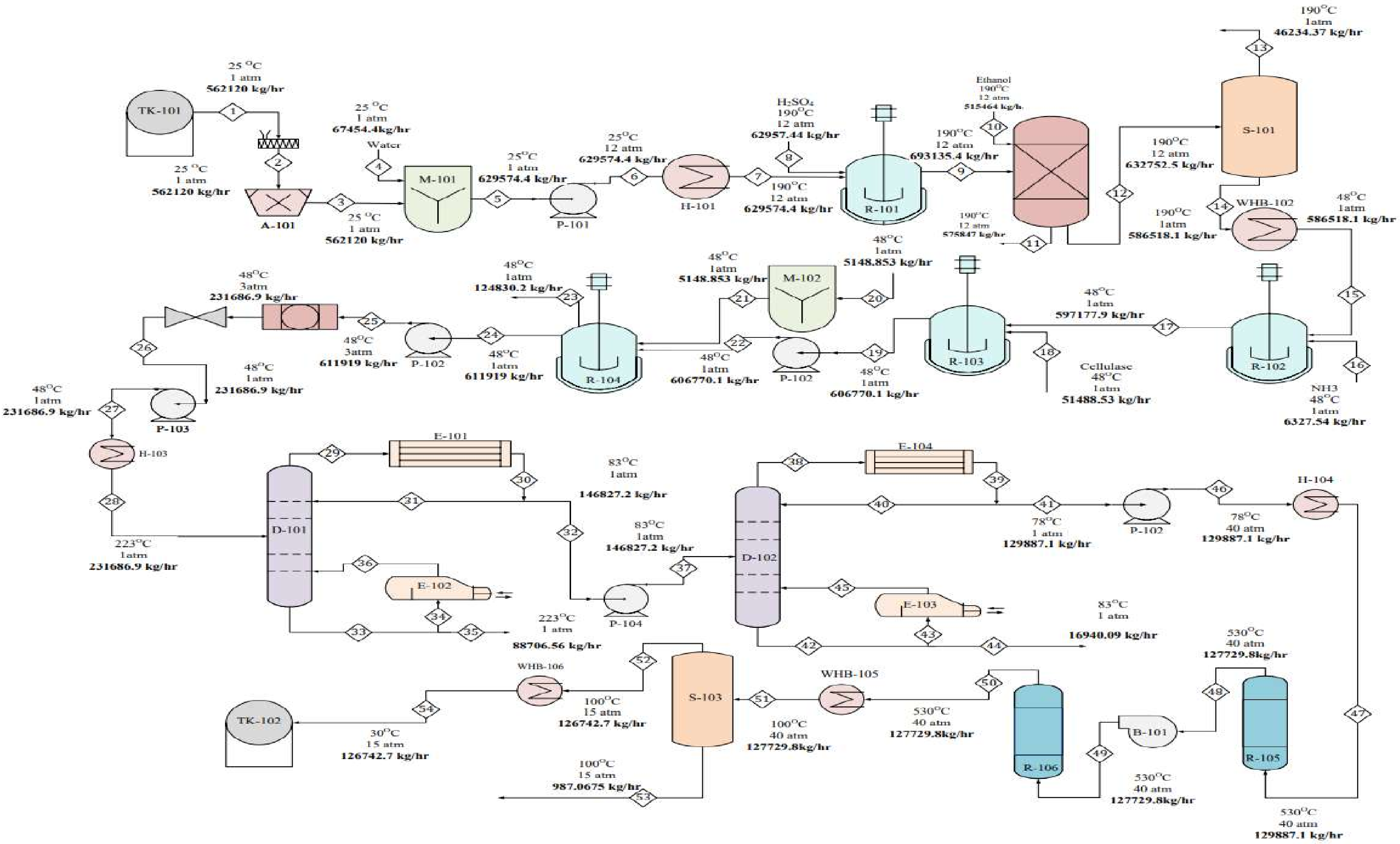
- ✓ Ethylene can also be produces by biomass such as sugarcane, wood waste, corn stover etc.

2.6 Process Description:

This procedure involves two main steps: firstly, production of bioethanol from maize stover and secondly production of ethylene from bioethanol. The procedure involves four basic stages:

- Pre-treatment of Raw Material
- Saccharification and Fermentation
- Purification of Bioethanol
- Conversion of Bioethanol to Bio-Ethylene

The physical and chemical processing of raw materials is the first phase. Corn stover is then ground to a thickness of 0.2 mm. To release hemicellulose sugars and break down biomass, maize stover is first processed with diluted sulfuric acid in the pre-treatment and conditioning unit [2]. Ammonia is then added to the pre-treated slurry to alter its acidity to be acceptable for enzymatic hydrolysis. The subsequent step involves sending the hydrolysate to an enzymatic hydrolysis and fermentation unit, where a cellulase enzyme is utilized for the enzymatic hydrolysis process. The cellulose in the hydrolyzed slurry is then fermented to create bioethanol. Using glucose as the main carbon source, an enzyme manufacturing unit on-site produces the necessary cellulase enzyme. The produced beer is then divided into bioethanol, water, and residual solids in a production recovery section using distillation and solid-liquid separation [9]. A wastewater treatment unit collects and treats wastewater streams produced during the synthesis of bioethanol using anaerobic and aerobic digestion. To form HP steam, which is utilized to produce electricity and meet the demand for process heat, solids and biogas from the product recovery unit and wastewater treatment unit are burned. Bioethanol made from maize stover is first dehydrated to produce ethylene, water, and other by-products in the ethylene manufacturing process. The dehydration reactor effluent is then pressurized and quenched. In an ethylene purification unit, the effluent is finally divided into ethylene, water, and other components [4].



Process Flow Diagram for Production of Bio-Ethylene from Zea-Mays (Corn Stover)

CHAPTER # 03

MATERIAL BALANCE

3.1 Introduction:

The quantities of all materials entering and leaving any system or process are calculated deploy on the "law of conversation of mass". This law states that the creation or destruction of matter has no effect on the overall mass. .

The basic idea behind material balancing calculations is to formulate and resolve several independent equations with several compositional and mass flow rate unknowns that are frequently seen in engineering and environmental investigations.

To develop chemical reactors, investigate substitute chemical production methods, model pollutant dispersion, and other physical system processes, for example, the mass balance theory is applied. Three complementary analysis tools are the material balance, energy balance, and the slightly more complex entropy balance. Several methodologies are required for comprehensive design and research of systems like the refrigeration cycle.

General Equation of Material Balance:

(Rate of Mass Input) - (Rate of Mass Output) + (Rate of Mass Generation) – (Rate of Mass Consumption) - (Rate of Mass Accumulation) = 0

Basis:

1 hour of operation

Assumption:

Steady state conditions.

Capacity of plant:

3000 tons /day

Reactant supplied:

14871.3 tons/ day

Yield:

21%

3.2 Feed Composition:

Table 3.1: Composition of Feed

Components	Composition %	Mass Flow Rates (Kg/hr)
Cellulose	49.2	276563.04
Water	20	112424
Sucrose	0.6	3372.72
Lignin	13.1	73637.72
Ash	6.7	37662.04
Acetic acid	1.5	8431.8
Glucose Oligomer	8.9	50028.68
Total	**100**	**562120**

3.3 Material Balance on Mixer (M-101):

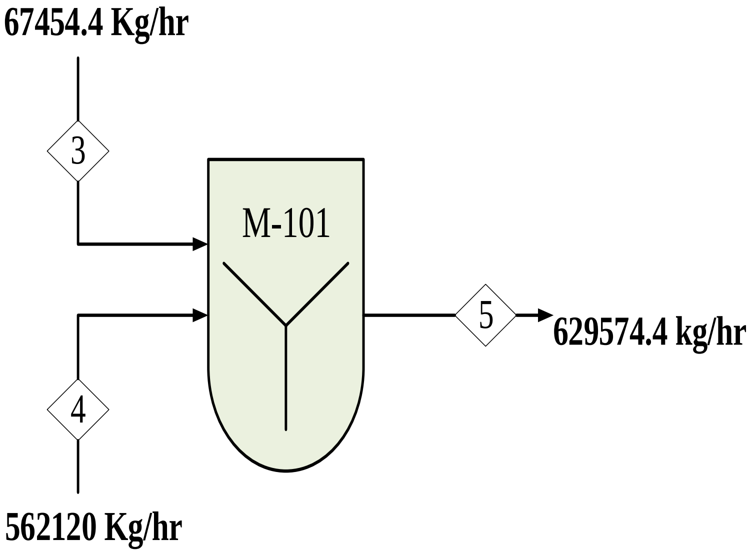


Figure 3.1: Mixer (M-101)

60% of original amount of water is added in mixture
Composition of water = 112424
= (112424×0.60)
= 67454.4 kg/hr

Table 3.2: Material Balance on Mixer (M-101)

Components	Material Input (Kg/hr)		Material Output (Kg/hr)
	Stream-03	**Stream-04**	**Stream-05**
Cellulose	-	276563.04	276563.04
Water	67454.4	112424	179878.4
Sucrose	-	3372.72	3372.72
Lignin	-	73637.72	73637.72
Ash	-	37662.04	37662.04
Acetic Acid	-	8431.8	8431.8
Glucose Oligomer	-	50028.68	50028.68
TOTAL	**629574.4 kg/hr**		**629574.4 kg/hr**

3.4 Material Balance on Flash Separator (S-101):

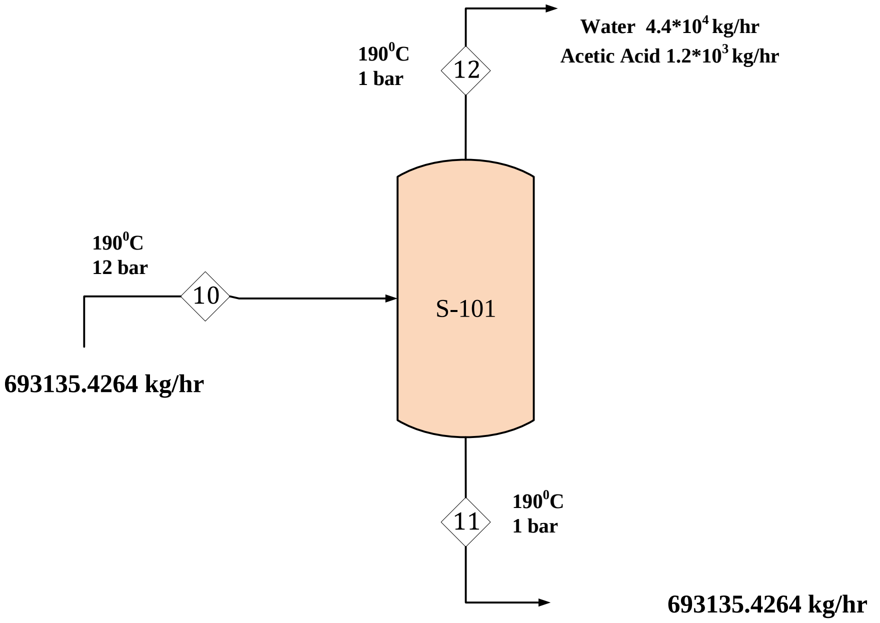


Figure 3.2: Flash Separator (S-101)

25% of water is removed in flash separator.

Composition of water = 179878.4 kg/hr

= (179878.4×0.25)

= 44969.6 kg/hr

15% of acetic acid is removed in flash separator.

Composition of acetic acid = 8431.8 kg/hr

= (8431.8× 0.15)

= 1264.77 kg/hr

Table 3.3: Material Balance on Flash Separator

	Material Input (Kg/hr)	**Material Output (Kg/hr)**	
Components	**Stream-9**	**Stream-10**	**Stream-11**
Cellulose	257442.63	-	257442.63
Water	179878.4	44969.6	134908.8
Sucrose	3372.72	-	3372.72
Lignin	73637.72	-	73637.72
Acetic Acid	8431.8	1264.77	7167.03
Glucose oligomer	50028.68	-	50028.68
H_2SO_4	1935.9413	-	1935.9413
Glucose	19359.413	-	19359.413
TOTAL	**693135.4264 kg/hr**	**693135.4264 kg/hr**	

3.5 Material Balance on Pre-Hydrolysis Reactor (R-101):

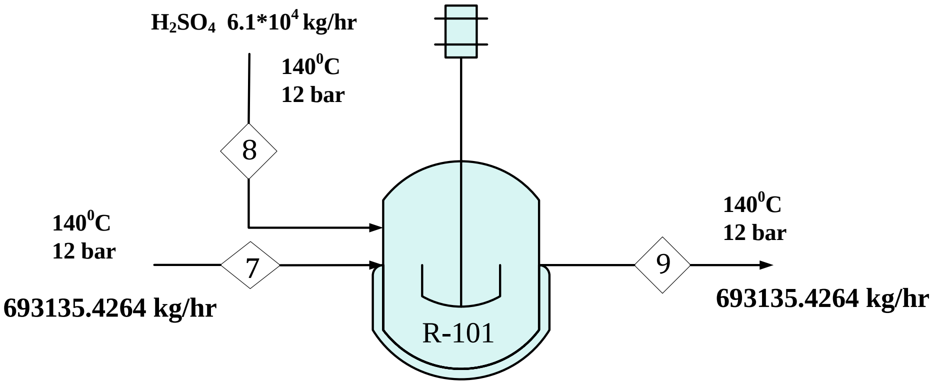


Figure 3.3: Pre-Hydrolysis Reactor (R-101)

10% of H_2SO_4 is added in reactor.

Flowrate of H_2SO_4 = 62957.44 kg/hr

Reactions:

$(C_6H_{10}O_5)n \longrightarrow C_6H_{10}O_5$ X = 0.07 *(3.1)*

$C_6H_{10}O_5 \longrightarrow C_6H_{12}O_6$ X = 0.9 *(3.2)*

Calculations:

Compositions of cellulose = 276563.04

Molar mass = 162.1406

$$\text{Moles} = \frac{Moles}{Molar\ mass}$$

$$= \frac{276563.04}{162.1406}$$

= 1705.6986 moles

= (1705.6986 ×0.07)

Cellulose reacted = 119.5025 kmoles/ hr

Cellulose reacted = 19120.41 kg/hr

Cellulose unreacted = (1705.6986 × 162.14)

= 276561.971 kg/hr

= 276561.971 – 19120.41

Cellulose unreacted = 257442.6 kg/hr

Cellulose reacted = Glucolig produced

Glucolig produced = 19376.14 kg/hr

Reacted glucolig = moles × conversion

= 119.5025× 0.9

=107.5525 kmoles/ hr

Glucolig reacted = 107.5523 kmol/hr

Glucose produced = (107.5523 × 180)

= 19359.414 kg/hr

Glucose produced = 19359.414 kg/hr

Glucolig unreacted = 119.5025-107.5523

= 11.95025 k mol/hr

Glucolig unreacted = 1935.941 kg/hr

Table 3.4: Material Balance on Pre-Hydrolysis Reactor (R-101)

Components	Material Input (Kg/hr)		Material Output (Kg/hr)
	Stream-07	**Stream-08**	**Stream-09**
Cellulose	276563.04	-	257442.63
Water	179878.4	-	179878.4
Sucrose	3372.72	-	3372.72
Lignin	73637.72	-	73637.72
Ash	37662.04	-	37662.04
Acetic Acid	8431.8	-	8431.8
Glucose Oligomer	50028.68	-	50028.68
H_2SO_4	0	62957.44	1935.9413
Glucolig	0	-	61386.08
Glucose	0	-	19359.413
TOTAL	**693135.4264 kg/hr**		**693135.4264 kg/hr**

3.6 Material Balance on Neutralization Reactor (R-102):

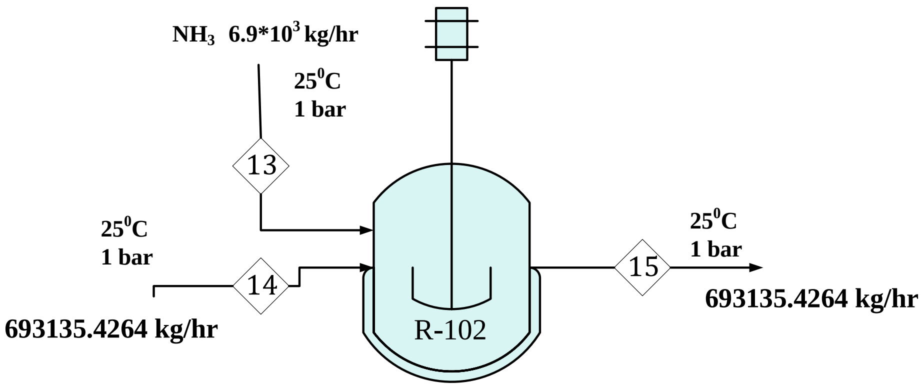


Figure 3.4: Neutralization Reactor (R-102)

1% NH_3 is added in neutralization reactor

NH_3 = 6931.354 kg/hr

Reactions:

$$CH_3COOH + NH_3 \longrightarrow NH_4COOCH_3 \quad \text{Conversion=100\%} \tag{3.3}$$

$$H_2SO_4 + 2NH_3 \longrightarrow (NH_4)_2SO_4 \quad \text{Conversion=100\%} \tag{3.4}$$

Reaction # 01:

$CH_3COOH + NH_3 \longrightarrow NH_4COOCH_3$ Conversion=100%

Limiting reactant = CH_3COOH

Flowrate of CH_3COOH = 7164.568 kg/hr

Molar mass of CH_3COOH = 60

$$\text{Moles} = \frac{\text{Mass}}{\text{Molar mass}}$$

$$= \frac{7164.568}{60}$$

CH_3COOH reacted = 119.4505 kmol/ hr

CH_3COOH reacted = 7167.03 kg/hr

NH_4COOCH_3 produced = 7167.03 × 77

NH_4COOCH_3 produced = 9197.6885 kg/hr

Reaction # 02:

$H_2SO_4 + 2NH_3 \longrightarrow (NH_4)_2SO_4$ Conversion=100%

Limiting reactant = NH_3

NH_3 unreacted = 2× 144.45

= 288.2762 kmol/hr

NH_3 unreacted = 4900.6958 kg/hr

$(NH_4)_2SO_4$ produced = 288.2762 × 132.14

$(NH_4)_2SO_4$ produced = 38092.8170 kg/hr

H_2SO_4 unreacted = 338.1124 kmol/hr

H_2SO_4 unreacted = 33135.01 kg/hr

Table 3.5: Material Balance on Neutralization Reactor

Components	Material Input (Kg/hr)	Material Output (Kg/hr)	
	Stream-9	**Stream-10**	**Stream-11**
Cellulose	257442.63	-	257442.63
Water	179878.4	44969.6	134908.8
Sucrose	3372.72	-	3372.72
Lignin	73637.72	-	73637.72
Ash	37662.04	-	37662.04
Acetic Acid	8431.8	1264.77	7167.03
Glucose oligomer	50028.68	-	50028.68
H_2SO_4	1935.9413	-	1935.9413
Glucolig	61386.08	-	61386.08
Glucose	19359.413	-	19359.413
TOTAL	**693135.4264 kg/hr**	**693135.4264 kg/hr**	

3.7 Material balance on Saccharification Reactor (R-103):

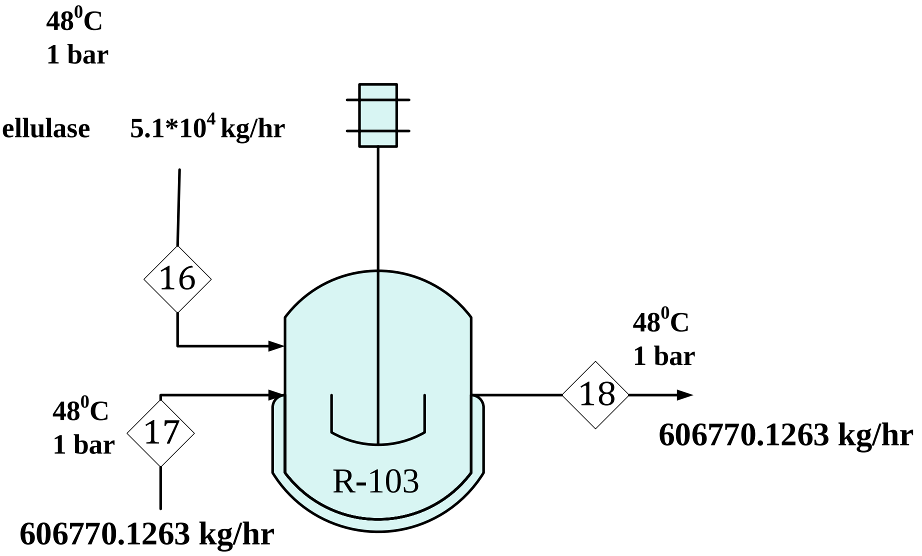


Figure 3.5: Saccharification Reactor (R-103)

Reactions:

$(C_6H_{10}O_5)n \longrightarrow C_6H_{10}O_5$ X = 0.07 *(3.5)*

$C_6H_{10}O_5 \longrightarrow C_6H_{12}O_6$ X = 0.9 *(3.6)*

Calculations:

Composition of cellulose = 257442.63

Molar mass of cellulose = 162.14

Moles = $\frac{\text{Mass}}{\text{Molar mass}} \times$ conversion

Cellulose reacted = 1430.237 kmol/ hr

Glucolig produced = (1430.237 ×162.14)

Glucolig produced = 231898.6272 kg/hr

Cellulose unreacted = 257442.6 – 231898.62

Cellulose unreacted = 25744.26 kg/hr

Glucolig reacted = moles × conversion
Glucolig reacted = 1370.078 kmol/hr
Glucose produced = 1370.078 × 162.14

Glucose produced = 246614 kg/hr
Water unreacted = 6745.44 kg/hr

Table 3.6: Material Balance on Saccharification Reactor

Components	Input (Kg/hr)		Output (Kg/hr)
	Stream-16	Stream-15	Stream-17
Cellulose	-	257442.63	25744.263
H_2O	-	134908.8	6745.44
Sucrose	-	3372.72	3372.72
lignin	-	73637.72	73637.72
Ash	-	37662.04	37662.04
Acetic Acid	-	0	0
Glucose Oligomer	-	50028.68	50028.68
H_2SO_4	-	1935.9413	11691.811
Glucolig	-	33135.01	33135.01
Glucose	-	19359.413	265973.41
$(NH_4)_2SO_4$	-	9197.6885	9197.6885
NH_4COOCH_3	-	38092.82	38092.82
Cellulase	51488.52646	-	51488.526
TOTAL	**606770.1263 kg/hr**		**606770.1263 kg/hr**

3.8 Material Balance on Fermentation Reactor (R-104):

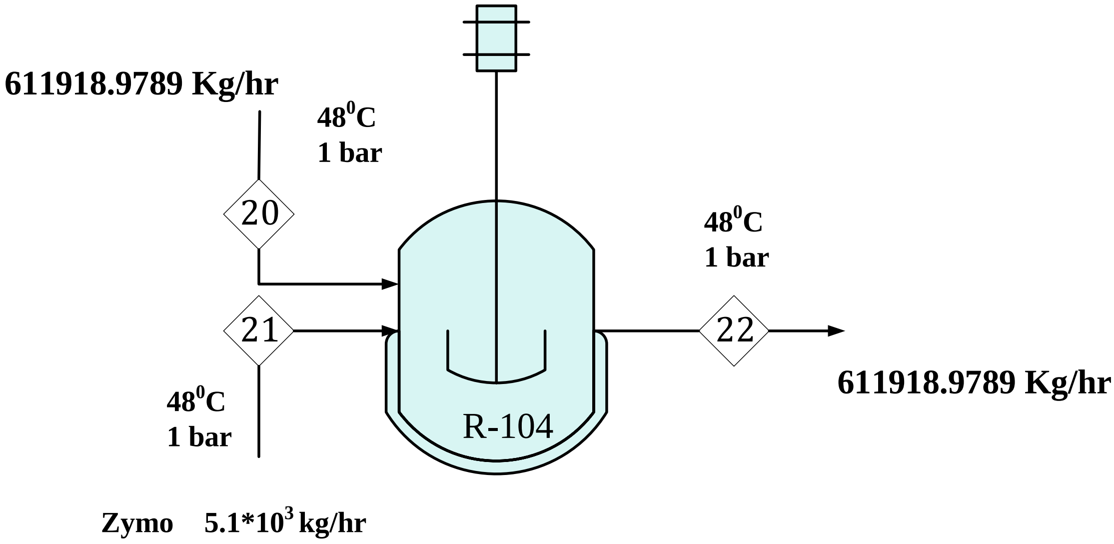


Figure 3.6: Fermentation Reactor (R-104)

Reaction:

$C_6H_{12}O_6 \longrightarrow 2C_2H_5OH + 2CO_2$ Conversion = 0.95 *(3.7)*

Calculations:

Moles = $\frac{\text{Mass}}{\text{Molar mass}}$

Glucose = 1477.63 kmol/hr

Moles = 1477.63 × 0.95

Glucose reacted = 1418.525 kmol/hr

Glucose unreacted = 1477.63 – 1418.525

= 59.105 k mol/hr

Glucose unreacted = 10638.94 kg/ hr

Ethanol produced = 2× 1419.36

Ethanol produced = 2837.05 kmol/hr

Ethanol produced = 130504.3 kg/hr
Carbon- dioxide produced = 124830.2 kg/hr

Table 3.7: Material Balance on Fermentation Reactor

Components	**Input (Kg/hr)**		**Output (Kg/hr)**
	Stream-20	**Stream-18**	**Stream-21**
Cellulose	25744.263	0	25744.263
H_2O	6745.44	-	6745.44
Sucrose	3372.72	-	3372.72
Lignin	73637.72	-	73637.72
Ash	37662.04	-	37662.04
Acetic Acid	0	-	0
Glucose Oligomer	50028.68	-	50028.68
Glucolig	11691.811	-	11691.811
H_2SO_4	33135.01	-	33135.01
Glucose	265973.41	-	10638.936
NH_4COOCH_3	9197.6885	-	9197.6885
$(NH_4)_2SO_4$	38092.82	-	38092.82
Ethanol	0	-	130504.29
CO_2	0	-	124830.19
Cellulase	51488.526	-	51488.526
Zymo	0	5148.8526	5148.8526
TOTAL	**611918.9789 Kg/hr**		**611918.9789 Kg/hr**

3.9 Material Balance on Distillation Column (D-101):

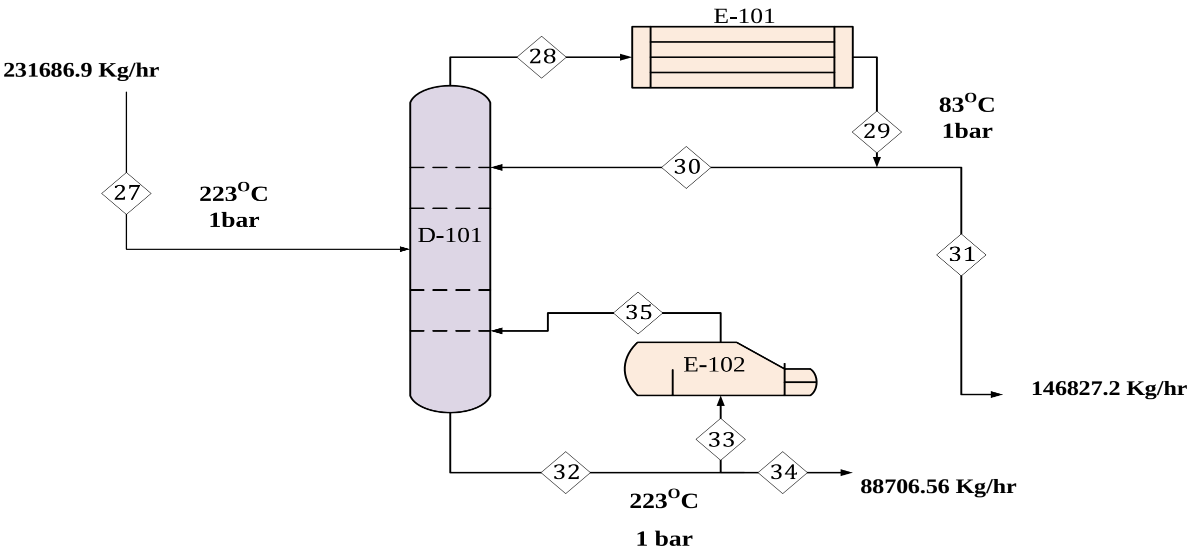


Figure 3.7: Distillation Column (D-101)

Overall material balance on distillation column

F = D+W *(3.8)*

Component balance (ethanol):

$$= \frac{0.572215-0.23}{0.77-0.23}$$

D = 0.63373F

D = 0.6337 ×(231686.9)

D = 146827.2 kg/ hr

F = D+W

231686.9 = 146827.2 +W

W = 84859.7 kg/hr

F = D+W

231686.9 = 146827.2 +84859.7

231686.9 = 231686.9

Table 3.8: Material balance on Distillation column

Components	**Input(kg/hr)**	**%**	**Distillate(kg/hr)**	**%**	**Bottoms(kg/hr)**	**%**
H_2O	6745.44	2.9576351	30833.711	21	56772.198	64
Sucrose	3372.72	1.4788176	0		887.0656	1
H2SO4	33135.01	14.528522	0		1774.1312	2
Glucose	10638.936	4.6647945	0		1774.1312	2
Ammonium Suphate	38092.82	16.702344	0		3548.2624	4
Ethanol	130504.29	57.22148	113056.94	77	20402.509	23
NH_4COOCH_3	9197.6885	4.0328587	2936.5439	2	3548.2624	4
Total	**231686.9 Kg/hr**	101.58645	**146827.2 Kg/hr**	100	**88706.56 Kg/hr**	100

3.10 Material Balance on Distillation Column (D- 102):

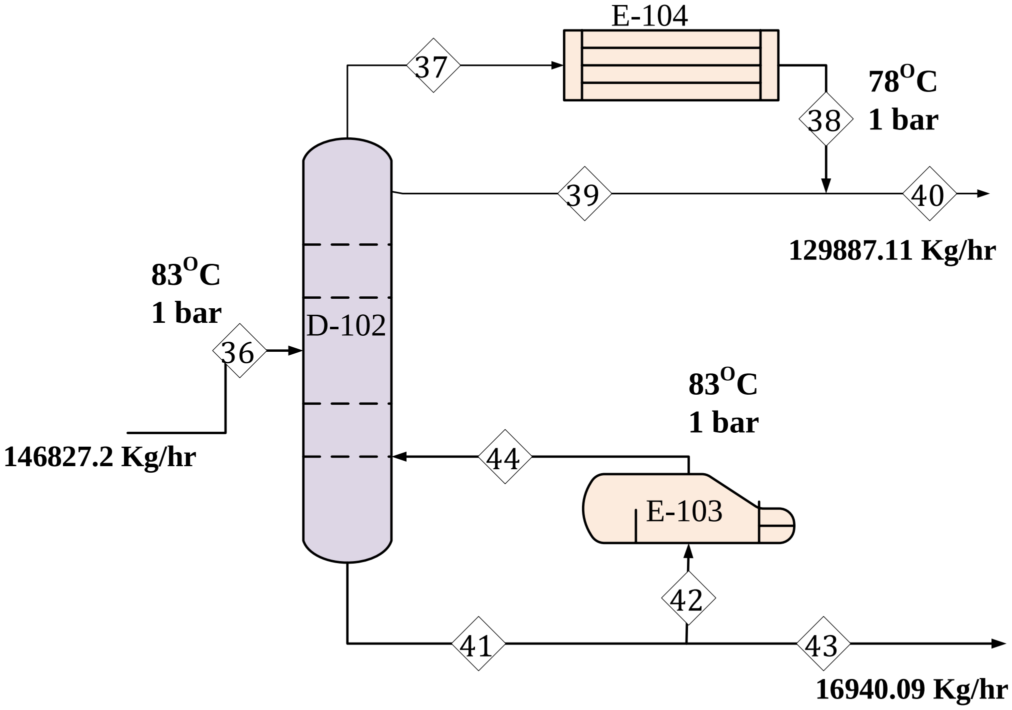


Figure 3.8: Distillation Column (D-102)

Overall material balance on distillation column

F = D+W *(3.9)*

Component balance (ethanol):

$$= \frac{0.87-0.01}{0.99-0.01}$$

D = 0.87755F

= 0.87755×(146827.2)

129887.12 kg/ hr

F = D+W

146827.2 = 12988.1 +W

W = 16940.09 kg/hr

F = D+W

146827.2 = 129887.1 +16940.09

146827.19 = 146827.19

Table 3.9: Material Balance on Distillation column

Components	**Input(kg/hr)**	**%**	**Distillate(kg/hr)**	**%**	**Bottoms(kg/hr)**	**%**
Ethanol	113056.94	87.693315	128588.23	99	419.8116	1
Water	30833.711	23.916359	1298.8711	1	37573.138	89.5
NH4COOCH3	2936.5439	2.2777485	0	0	3988.2102	9.5
Total	**146827.2 Kg/hr**		**129887.11 Kg/hr**		**16940.09 Kg/hr**	100

3.11 Material Balance on Reactor (R-105):

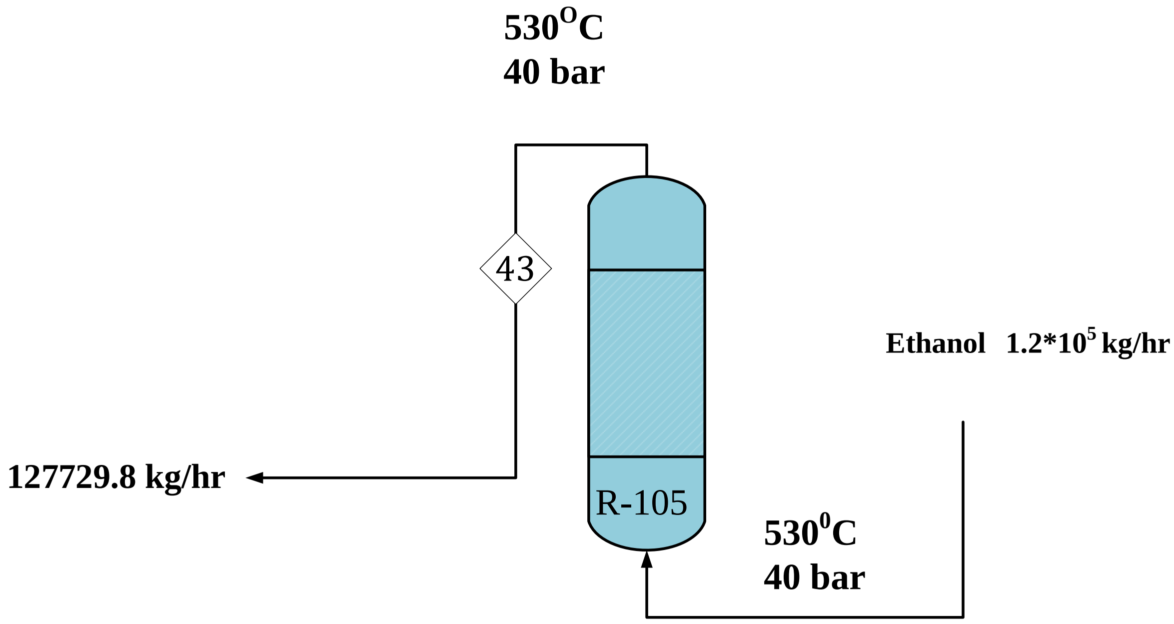


Figure 3.9: Reactor (R-105)

Reactions:

$$C_2H_5OH \longrightarrow H_2O + C_2H_4 \qquad (3.10)$$

$$2C_2H_5OH \longrightarrow H_2 + (C_2H_5)_2O \qquad (3.11)$$

$$C2H5OH \longrightarrow H_2 + C_2H_4O \qquad (3.12)$$

$$2H_2 + C_2H_5OH \longrightarrow H_2O + 2CH_4 \qquad (3.14)$$

$$H_2O + C_2H_5OH \longrightarrow CH_3COOH + H_2 \qquad (3.15)$$

$$H_2 + C_2H_5OH \longrightarrow C_2H_6 + H_2 \qquad (3.16)$$

Calculations:

The selectivity of the reactions was 98%.

Flowrate of Ethanol = 1.1×10^5 kg/hr

Ethylene Produced = 126016.5 Kg/hr

Unreacted Ethanol = 126016.5/46

Unreacted Ethanol = 2571.765 Kg/hr

Unreacted Ethanol = 55.90793 Kmoles/hr

Ethanol Reacted = 55.90793×0.7

Ethanol Reacted = 39.1355 Kmoles/hr

H_2O Produced = 39.1355 Kmoles/hr

Unreacted Ethanol = 55.90793 – 39.1355

Unreacted Ethanol = 16.77238 Kmoles/hr

Ethanol Reacted = 16.77238×0.5

Ethanol Reacted = 8.386189 Kmoles/hr

Diethyl Ether Produced = 8.3861×74.2

Diethyl Ether Produced = 622.2486 Kg/hr

Unreacted Ethanol = 8.386189 Kmol/hr

H_2 Produced = 8.386189 Kmoles/hr

Ethanol Reacted = 8.3861×0.08

Ethanol Reacted = 0.67088 Kmoles/hr

H_2 Produced = 0.67088 Kmoles/hr

Unreacted Ethanol = 8.3861 – 0.67088

Unreacted Ethanol = 7.638141 Kmoles/hr

Ethanol Reacted = 7.638141×0.01

Ethanol Reacted = 0.670895 Kmoles/hr

Acetaldehyde Produced = 29.55 Kg/hr

H_2 Reacted = 0.09051 Kmoles/hr

H_2O Produced = 0.09051×18×0.5

H_2O Produced = 0.04528 Kmoles/hr

Unreacted H_{2O} = 39.18084 Kmoles/hr

Ethanol Reacted = 10.84×0.025

Ethanol Reacted = 0.1905 Kmoles/hr

H_2 Produced = 0.190954 Kmoles/hr

Unreacted Ethanol = 342.2281 Kg/hr

H_2 Reacted = 0.009157 Kmoles/hr

Ethane Produced = 0.274729 Kg/hr

H_2O Produced = 0.009157 Kmoles/hr

Total H_2O = 687.7885 Kg/hr

Total H_2 = 9.1574 Kg/hr

Table 3.10: Material Balance on Reactor

Component	Input (kg/hr)	Output(kg/hr)
Ethanol	128588.235	342.22806
Ethylene	0	126016.47
Diethyl-Ether	0	622.255241
Acetaldehyde	0	29.5529308
Methane	0	1.4491335
Ethane	0	0.27472401
Acetic Acid	0	11.4572117
H2	0	18.2966192
H_2O	0	687.788491
TOTAL	**128588.2 kg/hr**	**127729.8 kg/hr**

3.12 Material Balance on Flash Separator (S-102):

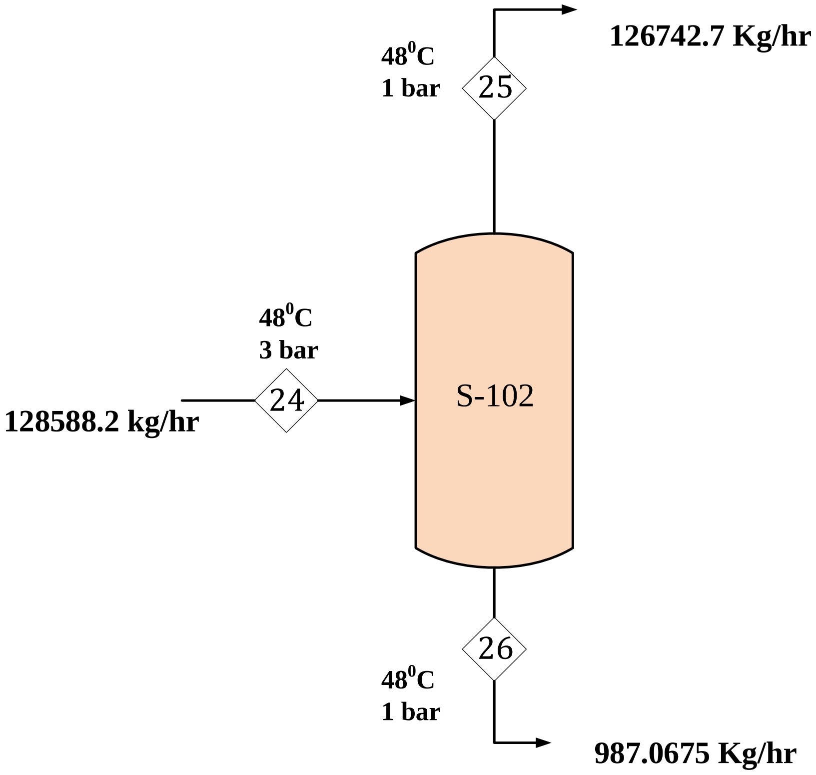


Figure 3.10: Flash Separator (S-102)

82.3% of Ethanol is removed

Composition of ethanol = 342.22806

= (342.22806 ×0.82)

Ethanol Removed = 281.6537 Kg/hr

100% of Acetaldehyde is removed

Composition of acetaldehyde = 29.5529308 = (29.5529308×1)

Acetaldehyde Removed = 29.55293 Kg/hr

96.6% of water is removed

Composition of water = 687.788491 = (687.788491×0.96)

Water Removed = 664.4037 Kg/hr

100% of Acetic Acid is removed

Composition of acetic acid = 11.4572117 = (11.4572117×1)

Acetic Acid Removed = 11.4572117 Kg/hr

Table 3.11: Material Balance on Flash Separator

Components	**Input(kg/hr)**	**Output(kg/hr)**	
	Stream-47	**Stream-48**	**stream-49**
Ethanol	342.22806	60.5743666	281.653693
Ethylene	126016.47	126016.47	0
Diethyl-Ether	622.255241	622.255241	0
Acetaldehyde	29.5529308	0	29.5529308
Methane	1.4491335	1.4491335	0
Ethane	0.27472401	0.27472401	0
Acetic Acid	11.4572117	0	11.4572117
H2	18.2966192	18.2966192	0
H_2O	687.791192	23.3848087	664.406383
TOTAL		**126742.7 Kg/hr**	**987.0675 Kg/hr**
TOTAL	**128588.2 kg/hr**	**127729.7725kg/hr**	

CHAPTER # 04
ENERGY BALANCE

4.1 General Equation of Energy Balance:

(Rate of Heat In) - (Rate of Heat Out) $\underline{+}$ (Generation / Consumption) = 0

4.2 Energy Balance on Pre-Heater (H- 101):

$T_{in}=25^0C$

$T_{out}=190^0C$

Heat Duty

$Q = mC_p\Delta T$

$Q = 2.3\times10^5$ MJ/hr

Steam Requirement

Saturated steam

P = 5 bar

T= 152°C

Steam flow rate

$Q = m\lambda$

$m = 1.141\times10^5$ kg/hr

Table 4.1: Thermodynamic Data for Pre-Heater (H-101)

Components	**Flow rates (kg/hr)**	**Cp (kJ/kg.°C)**
Corn Stover	2.7×10^5	1.37
Water	1.7×10^5	1.84

4.3 Energy Balance on Reactor -101 (Pre-hydrolysis Reactor):

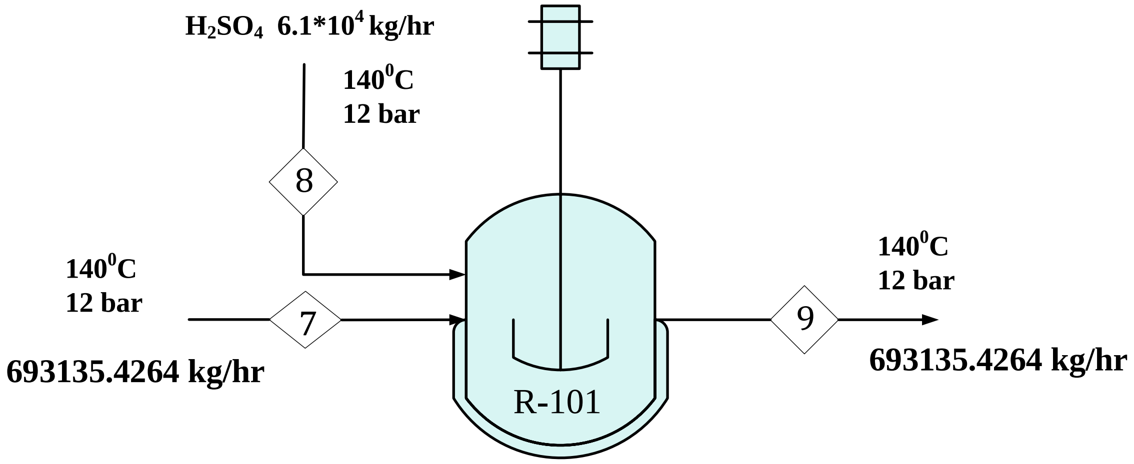


Figure 4.1: Pre-Hydrolysis Reactor (R-101)

Reaction:

$(C_6H_{10}O_5)_n \xrightarrow{\hspace{5em}} C_6H_{10}O_6$ *(4.1)*

$\Delta Hr_{25} = 3.9 \times 10^5$ kJ/hr $\Delta\hat{H}_{r,\,190}$

$$= \Delta\hat{H}_{r,25} + \int_{25}^{190} [(n\mathrm{Cp})\mathrm{P} - (\mathrm{nCp})\mathrm{R}]\mathrm{dT} \quad (4.2)$$

$\Delta Hr_{190} = 3.9\times10^5$ kJ/hr

Rate of heat in – Rate of heat out – Consumption of heat = Q

Q = 57.6 MJ/hr (Endothermic Reaction)

Steam Requirement:

Saturated steam

P = 5 bar

T= 152°C

Steam flow rate

$Q = m\lambda$ *(4.3)*

$m = 2.7\times10^4$ kg/hr

Table 4.2: Energy Balance on Reactor -101

Components	Heat In (kJ/hr)	Heat Out (kJ/hr)	Cp (kJ/kg.°C)
Corn stover	1.648×10^8	2.17×10^8	1.37
H_2SO_4	1.246×10^7	1.246×10^7	9.54

H_2O	9.9×10^7	9.9×10^7	3.36
Glucose	-	5.77×10^6	1.5
Total	**2.77×10^8**	**3.35×10^8**	**16.67**

4.4 Energy Balance on Flash Separator (S-101):

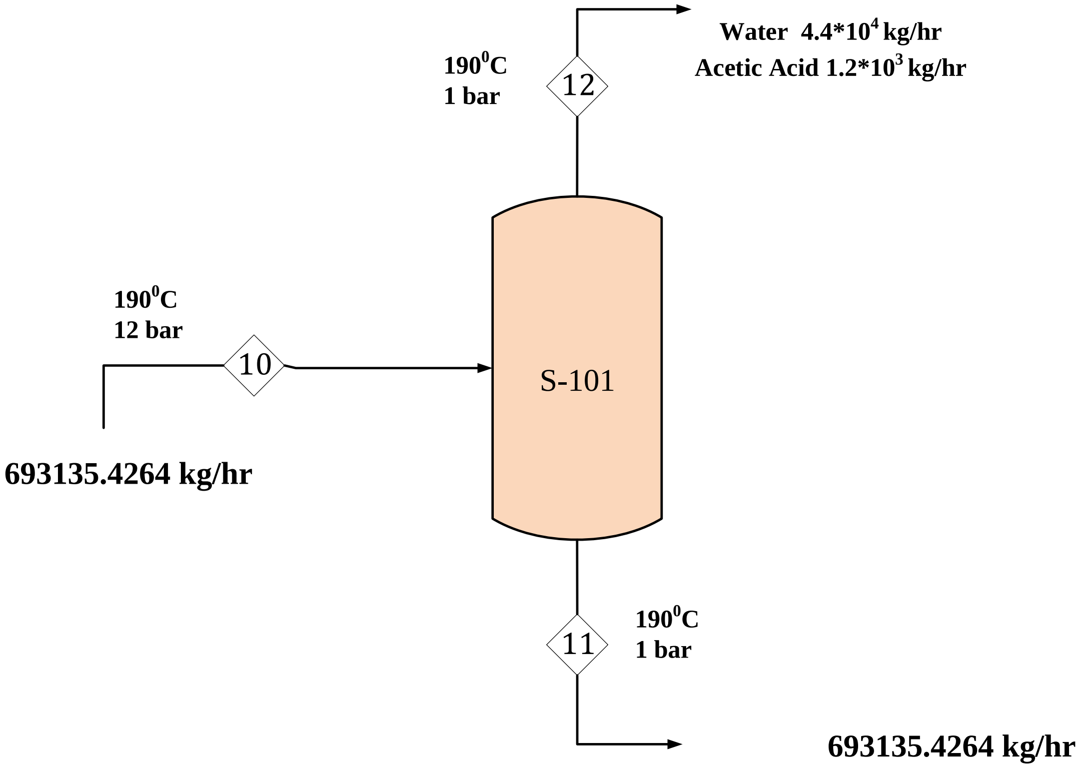


Figure 4.2: Flash Separator (S-101)

Rate of heat in = 2.78×10^5 MJ/hr

Vapor outlet:

Latent heat of vaporization for water = 2250 kJ/kg

Latent heat of vaporization for acetic acid = 870.94 kJ/Kg

Weighted λ = 693 KJ/kg

$Q = m\lambda$

$Q = 3.2\times10^4$ MJ/hr

Liquid outlet

Rate of heat outlet = 1.775×10^8 kJ/hr

Rate of heat in – Rate of heat out = Q *(4.4)*

$Q = 1\times10^5$ M J/hr

Steam Requirement

Saturated steam

P = 5 bar

T= 152°C

$Q = m\lambda$

$m = 4.8 \times 10^4$ kg/ hr

Table 4.3: Thermodynamic Data for Flash Separator -101

Components	Flow Rates (kg/hr)	Cp (kJ/kg.°C)
Cellulose	2.5×10^5	1.8
Water	1.7×10^5	3.36
Sucrose	3.3×10^3	0.42
Ash	3.7×10^4	1.5
Acetic Acid	8.4×10^3	9.54
H_2SO_4	6.1×10^3	0.11
Glucolig	1.9×10^4	11.2
Glucose	1.9×10^4	0.219

4.5 Energy Balance on Waste Heat Boiler (H-102):

Temperature input $=T_{in} = 190^oC$

Temperature output $=T_{out} = 25^oC$

Weighted Cp = 9.672 KJ/kg. K

Heat duty = $Q = mCp\Delta T$ *(4.5)*

$Q = 8.96\times10^5$ MJ/hr

Production of Saturated steam

P = 5atm

T= 152°C

$$m = \frac{Q}{Cp\Delta T+\lambda} \quad (4.6)$$

$m = 3.3\times 10^5$ kg/hr

Table 4.4: Thermodynamic Data for WHB-102

Components	Flow Rates (kg/hr)	Cp (kJ/kg.°C)
Cellulose	2.5×10^5	1.5
Water	1.3×10^5	3.36
Sucrose	3.3×10^3	0.42

Ash	3.7×10^4	
Acetic Acid	7.1×10^3	9.54
Glucose oligomer	5.0×10^4	0.2
H_2SO_4	1.9×10^3	0.13
Glucolig	1.6×10^4	185.33
Glucose	1.9×10^4	0.2

4.6 Energy Balance on Reactor- 102 (Neutralizer):

Temperature input $=T_{in} = 25^oC$
Temperature output $=T_{out} = 25^oC$
Temperature reference $=T_{ref} = 25^oC$

Reactions

$$CH_3COOH + NH_3 \longrightarrow NH_4COOCH_3 \quad (4.7)$$

$$H_2SO_4 + 2NH_3 \longrightarrow (NH_4)_2SO_4 \quad (4.8)$$

$\Delta Hr_{25} = -5.783\times10^7$ kJ/hr

Rate of heat in – Rate of Heat out + Heat generation - Q = 0

Q = -5.7×10^4 MJ/hr (Exothermic Reaction)

Cooling requirement:

$$Q = mC_p\Delta T \quad (4.9)$$

$$m = \frac{Q}{Cp\Delta T} \quad (4.10)$$

$$m = \frac{5.783\times10^7}{(4.2)(20)}$$

$m = 6.98\times10^5$ kg/hr

Table 4.5: Thermodynamic Data for R-102

Components	**Flow rates (kg/hr)**	**Cp (kJ/kg.^{o}C)**
Cellulose	2.5×10^5	1.3
Water	1.3×10^5	1.86
Sucrose	3.3×10^3	1.24
Acetic Acid	7.1×10^3	18.72
Glucose oligomer	5.0×10^4	0.4
Glucolig	1.9×10^3	185.3
H_2SO_4	6.1×10^4	2.96
Glucose	1.9×10^4	0.4

4.7 Energy Balance on Pre-Heater (H-103):

Process Stream

Temperature input =T_{in} = 25°C

Temperature output =T_{out} = 48°C

Weighted Cp = 2.0675 kJ/Kg. k

Heat duty

$$Q = mCp\Delta T \quad (4.11)$$

$Q = 2.8\times10^4$ MJ/hr

Saturated steam at

P = 5 bar

T= 152°C

$Q = m\lambda$

$m = 1.36\times 10^4$ kg/hr

Table 4.6: Thermodynamic Data for H-103

Components	Flow Rates (Kg/hr)	Cp (kJ/kg.°C)
Cellulose	2.5×10^5	1.3
Water	1.3×10^5	4.18
Sucrose	3.3×10^3	1.23
Ash	3.7×10^4	1.24
Glucose oligomer	5.0×10^4	0.4
Glucolig	1.9×10^3	185.3
H_2SO_4	3.3×10^4	0.13
Glucose	1.9×10^4	0.4
NH_4COOCH_3	9.1×10^3	7.6
$(NH_4)_2SO_4$	3.8×10^4	1.423

4.8 Energy Balance on Reactor- 103 (Saccharification Reactor):

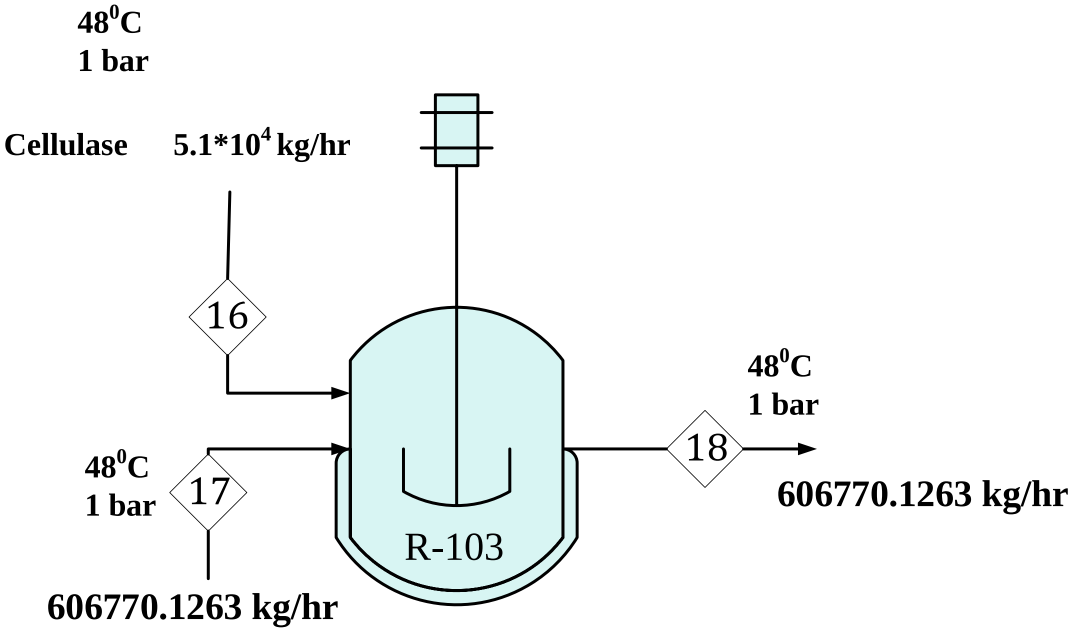


Figure 4.3: Saccharification Reactor (R-103)

Temperature input =T_{in} = 48°C

Temperature output =T_{out} = 48°C

Temperature reference =T_{ref} = 25°C

Reaction

$(C_6H_{10}O_5)_n + H_2O \longrightarrow C_6H_{12}O_6$ *(4.12)*

ΔHr_{25} =399026 kJ/hr

$\Delta Hr_{48} = 3.9\times10^5$ kJ/hr

Rate of heat in

Weighted Cp = 2.0675 kJ/kg.k

$Q = mCp\Delta T$ *(4.13)*

$Q = 2.8 \times10^4$ MJ/hr

Rate of heat out

$Q = mCp\Delta T$

$Q = (606770) \times(1.2719) \times(23)$

Q = 17.7 MJ/hr

Rate of heat in – Rate of Heat out – Consumption + Q = 0 *(4.14)*

$Q =1.1\times10^4$ MJ/hr (Endothermic Reaction)

Saturated steam at

P = 5bar

T =152^oC

Q = mλ

m = 5.4 ×10^3 kg/hr

Table 4.7: Thermodynamic Data for R-103

Components	Flow Rates (Kg/hr)	Cp (kJ/kg.^oC)
Cellulose	2.5×10^5	2.0
Water	1.3×10^5	1.84
Sucrose	3.3×10^3	0.42
Ash	3.7×10^4	1.26
Glucose oligomer	5.0×10^4	0.21
Glucolig	1.9×10^3	185.3
H_2SO_4	3.3×10^4	0.13
Glucose	1.9×10^4	0.21
NH_4COOCH_3	9.1×10^3	7.2
$(NH_4)_2SO_4$	3.8×10^4	1.42

4.9 Energy Balance on Fermentation Reactor (R- 104):

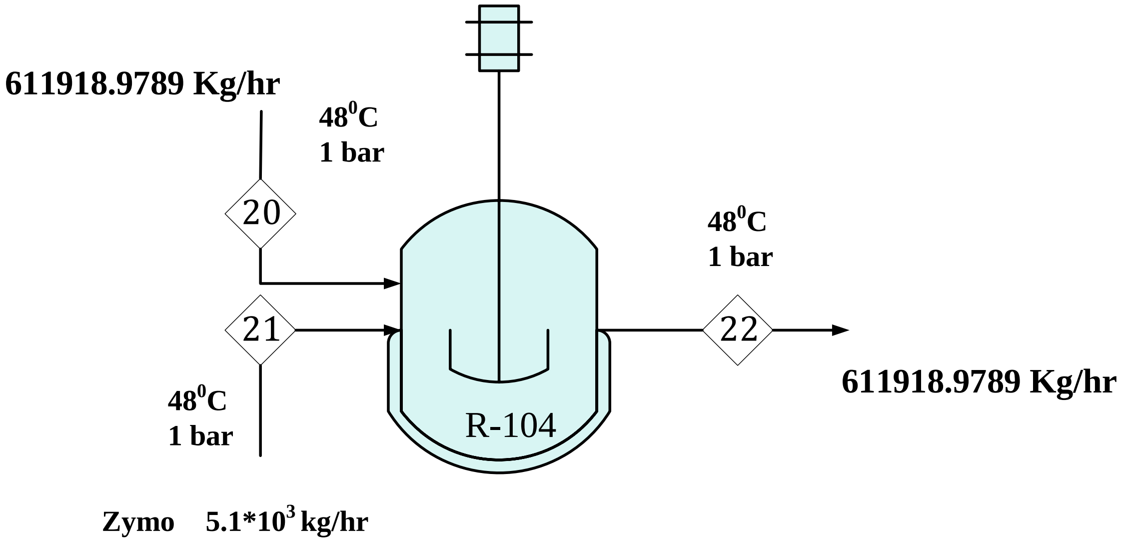


Figure 4.4: Fermentation Reactor (R-104)

Reaction

$$C_6H_{12}O_6 \longrightarrow 2C_2H_5OH+2CO_2 \qquad (4.15)$$

$\Delta Hr_{25} = -2.634\times10^8$ kJ/hr

$\Delta Hr_{48} = -2.61\times10^8$ kJ/hr

Rate of heat in

Weighted Cp = 1.2719 kJ/kg.k

$$Q = mCp\Delta T \qquad (4.16)$$

$Q = 1.7 \times10^4$ MJ/hr

Rate of heat out

$Q = mCp\Delta T$

$Q = 1.6 \times10^4$ MJ/hr

Rate of heat in – Rate of Heat out + generation - Q = 0 *(4.17)*

$Q = -2.2 \times10^5$ MJ/hr (Exothermic Reaction)

Cooling Water Flowrate

$$m = \frac{Q}{mCp\Delta T} \qquad (4.18)$$

$m = 2.7\times 10^6$ kg/hr

Table 4.8: Thermodynamic Data for R-104

Components	Flow Rates (Kg/hr)	Cp (kJ/kg.°C)
Cellulose	2.5×10^5	2.0
Water	6.7×10^5	4.2
Sucrose	3.3×10^3	0.42
Ash	3.7×10^4	1.25
Glucose oligomer	5.0×10^4	0.21
Glucolig	1.1×10^3	185.3
H_2SO_4	3.3×10^4	0.13
Glucose	2.6×10^4	0.21
NH_4COOCH_3	9.1×10^3	7.26
$(NH_4)_2SO_4$	3.8×10^4	1.42

4.10 Energy Balance on Heat Exchanger (H- 104):

Temperature input = $T_{in} = 40^oC$

Temperature output = T_{out} =223°C

Heat duty

$Q = mCp\Delta T$ *(4.19)*

$Q = 1 \times 10^5$ MJ/hr

Saturated Stream

P = 5bar

T = 152°C

Steam flow rate

$Q = m\lambda$

$m = 5 \times 10^4$ kg/hr

Table 4.9: Thermodynamic Data for H-104

Components	Flow rates (kg/hr)	Cp (kJ/kg.°C)
Ethylene	$1.2*10^5$	2.98
Di-Ethyl-Ether	$6.8*10^2$	0.24
H_2	$1.8*10^1$	14.71

4.11 Energy Balance on Distillation Column (D- 101):

Steam is saturated liquid at its bubble point i.e 223°C

Condenser duty

Dew point = 83°C

Weighted Cp = 3.4791 kJ/Kg. k

Weighted λ = -1225.8 kJ/kg

$Q = mCp\Delta T + m\lambda$ *(4.20)*

$Q = -1 \times 10^5$ MJ/hr

Cooling water requirement

$Q = mCp\Delta T$

$m = 1.2 \times 10^6$ kg/hr

Re-boiler duty:

$Q = m\lambda$

Weighted λ = 1806.17kJ/kg

$Q = 1.6 \times 10^5$ MJ/hr

Saturated Steam

P = 5bar

T = 152°C

$Q = m\lambda$

$m = 7.6\times10^4$ kg/hr

Table 4.10: Thermodynamic Data for D-101

Components	Flow Rates (Kg/hr)	Cp (kJ/kg.°C)
Water	6.7×10^5	4.2
Sucrose	3.3×10^3	0.42
H_2SO_4	3.3×10^4	0.12
Glucose	2.6×10^4	0.219
NH_4COOCH_3	9.1×10^3	6.5
$(NH_4)_2SO_4$	3.8×10^4	1.42
Ethanol	1.3×10^5	5.0

4.12 Energy Balance on Distillation Column (D- 102):

Steam is saturated liquid at its bubble point i.e 83°C

Condenser duty

Dew point = 78°C

Weighted Cp = 0.025 kJ/Kg. k

Weighted $\lambda = -926.182$ kJ/kg

$$Q = mCp\Delta T + m\lambda \qquad (4.21)$$

$Q = -1.2 \times10^5$ MJ/hr

Cooling water requirement

$Q = mCp\Delta T$

$m = 1.42 \times10^6$ kg/hr

Re-boiler duty

$Q = m\lambda$

Weighted $\lambda = 2517.5$ kJ/kg

$Q = 4.2 \times10^4$ MJ/hr

Steam at

P = 5bar

T = 152°C

$Q = m\lambda$

$m = 5.20\times10^4$ kg/hr

Table 4.11: Thermodynamic Data for D-102

Components	**Flow Rates (Kg/hr)**	**Cp (kJ/kg.°C)**
Water	3.0×10^4	15
NH_4COOCH_3	2.9×10^3	6.5
Ethanol	1.1×10^5	2.4

4.13 Energy Balance on Reactor- 105:

$$C_2H_5OH \longrightarrow H_2O + C_2H_4 \quad (4.22)$$

$$2C_2H_5OH \longrightarrow H_2 + (C_2H_5)_2O \quad (4.23)$$

$$C_2H_5OH \longrightarrow H_2 + C_2H_4O \quad (4.24)$$

$$2H_2 + C_2H_5OH \longrightarrow H_2O + 2CH_4 \quad (4.25)$$

$$H_2O + C_2H_5OH \longrightarrow CH_3COOH + H_2 \quad (4.26)$$

$$H_2 + C_2H_5OH \longrightarrow C_2H_6 + H_2 \quad (4.27)$$

Reactions

$\Delta Hr_{25} = 2.05\times10^8$ kJ/hr

$\Delta Hr_{530} = 2.924\times10^8$ kJ/hr

Rate of heat in

$Q = mCp\Delta T$

$Q = 4\times10^4$ MJ/hr

Rate of heat out

$Q = mCp\Delta T$

$Q = 1.922\times10^5$ MJ/hr

Rate of heat in – Rate of Heat out – Consumption + Q = 0

$Q = 21.4\times10^4$MJ/hr

Steam Flow-rate:

Saturated steam at

P = 5bar

T = 152°C

$Q = m\lambda$

$m = 1\times10^5$ kg/hr

4.14 Energy Balance on Waste Heat Boiler (WHB-105):

Temperature input =T_{in} = 530°C

Temperature output =T_{out} =100°C

Heat duty

$Q = mCp\Delta T + m\lambda$

$Q = 4.6 \times 10^3$ MJ/hr

Saturated Stream

P = 5bar T = 152°C

Steam flow rate

$Q = mCp\Delta T + m\lambda$

$$m = \frac{Q}{Cp\Delta T + \lambda}$$

m = 1745.2 kg/hr

Table 4.12: Thermodynamic Data for WHB-106

Components	Flow rates (kg/hr)	Cp (kJ/kg.°C)
Ethanol	$6.0*10^1$	4
Ethylene	$1.2*10^5$	2.98
Di-ethyl-ether	$6.2*10^2$	0.24
Methane	1.4	4.7
Ethane	$2.7*10^{-1}$	3.2
H_2	$1.8*10^1$	14.71
H_2O	$2.3*10^1$	10

4.15 Energy Balance on Flash Separator (S-102):

Feed

Weighted Cp = 1.754 kJ/Kg. k

$Q = mCp\Delta T$

$Q = 16.8 \times 10^3$ MJ/hr

Vapor outlet

$Q = m\lambda$

$Q = 6.1 \times 10^4$ MJ/hr

Liquid outlet

Weighted Cp = 3.63 kJ/Kg. k

$Q = mCp\Delta T$

Q = 243 MJ/hr

Rate of heat out = 6.17×10^{7} kJ/hr

Rate of heat in – Rate of heat out = Q

Q = -44670 MJ/hr

Cooling requirement

$Q = mCp\Delta T$

$m=\frac{Q}{Cp\Delta T}$

m = 5.2×10^{5} kg/ hr

Table 4.13: Thermodynamic Data for S-102

Components	Flow rates (kg/hr)	Cp (kJ/kg.^{o}C)
Ethanol	$6.0*10^{1}$	6
Ethylene	$1.2*10^{5}$	2.98
Di-ethyl-ether	$6.2*10^{2}$	0.24
Methane	1.4	4.7
Ethane	$2.7*10^{-1}$	3.2
H_2	$1.8*10^{1}$	14.71
H_2O	$2.3*10^{1}$	10

CHAPTER # 05

EQUIPMENT DESIGN

5.1 Reactor Design:

5.1.1 Introduction:

The heart of a chemical reaction is the reactor. Reactor design is an essential part of the process' overall design because it is the only area where raw materials are transformed into finished goods. The industrial chemical reactor's design must adhere to the some of the requirements:

- The chemical factors: The design must permit the desired reaction to progress to the necessary degree of conversion by allowing enough residence time.
- Transfer factors of mass: In heterogeneous reactions, the diffusion rate of the species that are reacting must in control of the reaction rate rather than chemical kinetics.
- Transfer factors of heat: the reduction or addition of reaction heat.
- The safety factors: the containment of potentially dangerous reactants and products, as well as the regulation of reaction and process conditions.

5.1.2 Principal Types of Reactors:

Reactor designs are typically categorized using the following features:

- Batch or continuous **operating modes**.
- Homogeneous or heterogeneous **phases** are observe.
- **Geometry of the reactor:** The flow pattern and the method of contacting the phases

Reactors are classified in the following broad category:

- Stirred tank reactor
- Tubular reactor
- Packed bed, stationary and mobile

There are further varieties, such as fluidized beds and micro channel reactors, in addition to these. The interaction of the reactor design with the other process processes must not be disregarded while choosing the reactor conditions, especially the conversion, and optimizing the design.

The size and expense of any equipment required to separate and recycle unreacted materials will depend on the degree of conversion of the reactor's input materials. In these situations, it is necessary to optimize the reactor as a whole with the supporting machinery.

5.1.3 Selection of Reactor:

There are many factors to take into account while choosing the reactor type for a particular procedure.

- Temperature and pressure is mandatory for chemical reaction.
- The requirement for the reactants and products to be removed or added.
- Reaction Phase.
- The required product delivery strategy.
- The use of catalysts should take into consideration aspects like the need for solid catalyst particles and contact with fluid reactants and products.

- Reactor Relative cost.

The design of the reactor must Adhere the following requirements:

i. **Chemical factors:** Enough residence time must be allowed in the design for the desired reaction to proceed to the necessary level of conversion.
ii. **Mass transfer factors:** For instance, in heterogeneous reactions, the rates of diffusion of the species that are reacting may be in control of the reaction rate rather than chemical kinetics.
iii. **Heat transfer factors:** Taking away or adding heat from the reaction.
iv. **Safety factors:** The containment of potentially harmful reactants and products, as well as the regulation of reaction and process conditions.

5.1.4 Design of Multi-Tubular Reactor (R-105):

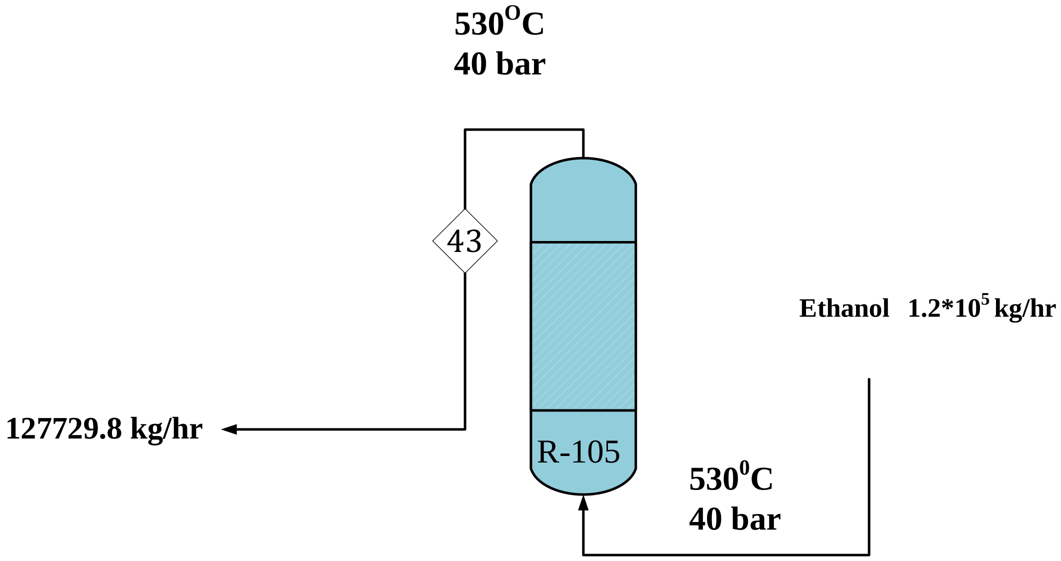


Figure 5.1: Multi-Tubular Reactor (R-105)

For fixed-bed catalytic reactor, multi-tubular reactors are frequently used in the chemical and refining sectors. In the reactor, a gas phase reaction is taking place. Since the dehydrogenation of ethanol is a strongly endothermic reaction, steam must be added to the reactor to keep it at a constant temperature, necessitating the use of a shell and tube configuration.

$$C_2H_5OH \longrightarrow H_2O + C_2H_4 \qquad (5.1)$$

$$2C_2H_5OH \longrightarrow H_2 + (C_2H_5)_2O \qquad (5.2)$$

$$C_2H_5OH \longrightarrow H_2 + C_2H_4O \qquad (5.3)$$

$$2H_2 + C_2H_5OH \longrightarrow H_2O + 2CH_4 \qquad (5.4)$$

$$H_2O + C_2H_5OH \longrightarrow CH_3COOH + H_2 \qquad (5.5)$$

$$H_2 + C_2H_5OH \longrightarrow C_2H_6 + H_2 \qquad (5.6)$$

Rate Equation:

$-r_A = k *((P_e)/(P_e + (K_w/K_a)*P_w + (K_{aw}/K_a)*P_e*P_w))$ *(5.7)*

$n = 1$, $k = 1500\ hr^{-1}$

$P_e = 4048$ kPa , $P_w = 4.053$ kPa

$(K_w/K_a) = 0.63$, $(K_{aw}/K_a) = 0.2$

$-r_A = 828.16$ kPa/hr

$-r_A = 125.69\ kmol/m^3hr$

Design Equation:

$$W/F_{AO} = \int_0^{0.5} \mathrm{d}XA/{-\mathrm{rA}} \tag{5.8}$$

Weight of Catalyst:

W = 1760 kg

Volume of Catalyst:

$V_c = W/\rho_{bulk}$ *(5.9)*

$V_c = 2.2\ m^3$

Volume of Reactor:

$V_r = V_c/1- \varepsilon$ *(5.10)*

$V_r = 3\ m^3$

Space Time:

$\tau = V_r/V_o$ *(5.11)*

$\tau = 3.3$ s

Tube Dia:

L = 16 ft or 4.86 m

Tube dia = $D_t = 0.12$ m

Particle size = $D_p = 0.0014$ m

$D_t/D_p = 85.7$

Volume of Tube:

$V_t = \pi D_t^2L/4$ *(5.12)*

$V_t = 0.054m^3$

No. of Tubes:

$N_t = V_r/V_t$

$N_t = 56$ tubes

Shell Dia:

$N_t = ((D_s – k_1)^2 * \pi/4 + k_2) – P_t*(D_s – k_1)*(nk_3 + k_4))/1.223\ P_t$ *(5.13)*

$k_1 = 1.080$, $k_2 = -0.9000$, $k_3 = 0.690$, $k_4 = -0.8000$

$D_s = 0.26$ m

Shell Height:

$L_t = 4.8$ m

$L_s = 40\%$ of $L_t + L_t$

$L_s = 6.8$ m

Pressure Drop:

$\Delta P/L = [(150*\mu*(1-\varepsilon)^2*G)/(\varepsilon^3*D_p^2*\rho)] + [(1.75*(1-\varepsilon)*G^2)/(\varepsilon^3*D_p*\rho)]$ *(5.14)*

$\mu = 9.135* 10^{-6}$ kg/ms

$\rho = 800$ kg/m^3

$D_p = 0.0014$ m

$G = 146$ kg/sm^2

ΔP = 5.29 psi

SPECIFICATION SHEET	
Identification	
Item	Reactor
Item no.	R-105
No. required	2
Operation	Continuous
Type	Multi Tubular Packed Bed Reactor
Catalyst	Al_2O_3
Function	
Dehydration of Ethanol to Ethylene	
Chemical Reactions	
$C_2H_5OH \rightarrow H2O + C_2H_4$ $2C_2H_5OH \rightarrow H_2 + (C_2H_5)_2O$ $C_2H_5OH \rightarrow H_2 + C_2H_4O$ $2H_2 + C_2H_5OH \rightarrow H_2O + 2CH_4$ $H_2O + C_2H_5OH \rightarrow CH_3COOH + H_2$ $H_2 + C_2H_5OH \rightarrow C_2H_6 + H_2$	
Weight of bed	1760 Kg
Volume of Catalyst	2.2 m^3
Volume of Reactor	3 m^3
Space Time	3.3 sec
Diameter of tube	0.12 m
Number of Tube	56
Shell Diameter	0.26 m
Pressure Drop	5.29 psi

5.1.5 Design of Pre-Hydrolysis Reactor (R-101):

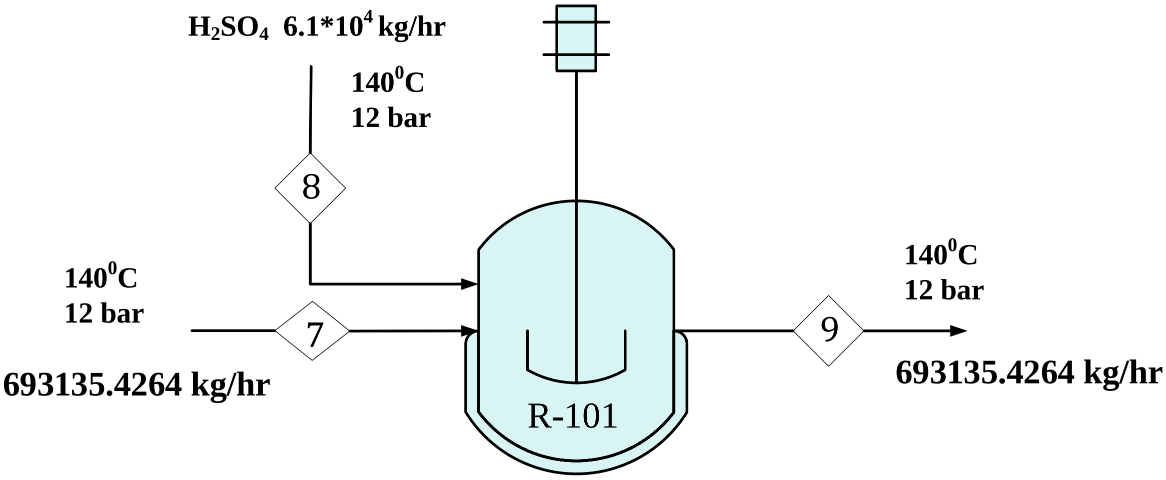


Figure 5.2: Pre-Hydrolysis Reactor (R-101)

In the pre-hydrolysis reactor, the slurry of maize stover is treated with diluted sulfuric acid to dissolve the lignin shield and expose the cellulose for additional enzymatic hydrolysis. The following factors led to the selection of the continuous stirred tank reactor.

- ✓ Reaction That Is Moderately Endothermic
- ✓ Accurate Temperature Regulation
- ✓ Superior heat and mass transmission rates.
- ✓ Reduced labor requirements
- ✓ Effective for delayed reactions that require a lot of hold time
- ✓ Spread of the Catalyst
- ✓ To Control Liquid-Gas Systems

Reaction:

$(C_6H_{10}O_5)n \longrightarrow C_6H_{12}O_6$ Conversion = 0.07 *(5.15)*

Rate Equation:

$-r_A = k\,C_c$ *(5.16)*

$n = 1$, $k = 9.8\ hr^{-1}$

$-r_A = k\,C_{co}\,(1-X_c)$

$X_c = 0.07$, $C_{co} = 2.6\ kmol/m^3$

$-r_A = 24.5\ kmol/m^3hr$

Design Equation:

$V/F_{co} = \Delta X_c / -r_A$ *(5.17)*

Volume Calculations:

$F_{co} = 1705.7\ kmol/hr$

$V = 5\ m^3$

5% Safety allowance

$V = 5.1\ m^3$

$\tau = 28$ s

$L/D = 1.2$

$L = 2$ m , $D = 1.74$ m

Impeller Design:

Pitched blade turbine is selected because:

1. For mixing of slurries
2. Viscosity is moderately high
3. Maximum Radial Flow
4. No back mixing
5. Promote Heat Transfer

- **Impeller Specifications:**

Dia of impeller = $D_a = 0.58$ m

Height of impeller = $W = 0.11$ m

Length of impeller = $L_a = 0.145$ m

Distance of impeller from bottom = $E = 0.58$ m

Thickness of Baffles = $J = 0.145$ m

- **Power Requirements:**

Agitator Speed= $n = 100$ rpm

$Re = 9768$

$Np = 1.5$

$r = 1088.8\ kg/m^3$

$$P = (Np)*(D_a)^5*(r)*(n)^3 \tag{5.18}$$

$P = 5.7$ hp

Jacket Selection:

Dimple Jacket is selected because

1. High pressure steam
2. Induce turbulence
3. Low pressure drop
4. Cost effective

$La = 0.145$ m , $Da = 0.58$ m

$n = 100$ rpm

$Tr = 140°C$ $Ts = 152°C$

$\rho = 1088.8$ kg/m3

$\mu = 0.06$ kg/m s $k = 0.0316$ W/m K

$Cp = 1.37$ KJ/kg K

$Re = (La2*n*r)/\mu$ *(5.19)*

$Re = 610$

$j_h = 15$

$h_i = j_h *(k/D_a)*((c_p* \mu)/k)^{1/3}$ *(5.20)*

$h_i = 3.327$ W/m^{2}K

$h_o = 8520$ W/m^{2}K

$U_c = (h_i*h_o)/(h_i+h_o)$

$U_c = 3.325$ W/m^{2}K

$R_d = 0.003$

$1/U_d = (1/U_c) + R_d$

$U_d = 3.27$ W/m^{2}K

Jacket Covers 95% of reactor area

$A= 0.95*(\pi D_L + \pi D^2/4) = 13$ m^2

SPECIFICATION SHEET			
Identification			
Item		Reactor	
Item no.		R-104	
No. required		3	
Operation		Continuous	
Type		Continuous stirred type reactor	
Function			
Fermentation of Glucose to Ethanol			
Chemical Reaction $C_6H_{12}O_6 \rightarrow 2C_2H_5OH + 2CO_2$			
Reactor Specifications		**Jacket Specifications**	
Length of Reactor	2 m	Process Fluid Temp	48℃
Diameter of Reactor	1.36 m	t_1	25℃
Volume of Reactor	3.05 m^3	t_2	45℃
Speed of Impeller	100 rev/min	hi	4.26 W/m^2K
Length of Impeller	0.1125 m	ho	1.61 W/m^2K
Diameter of Impeller	0.45 m	Ud	1.15 W/m^2K
Power	6.46 hp	Area	9.5 m^2

<table>
<tr><th colspan="4">SPECIFICATION SHEET</th></tr>
<tr><th colspan="4">Identification</th></tr>
<tr><td colspan="2">Item</td><td colspan="2">Reactor</td></tr>
<tr><td colspan="2">Item no.</td><td colspan="2">R-103</td></tr>
<tr><td colspan="2">No. required</td><td colspan="2">1</td></tr>
<tr><td colspan="2">Operation</td><td colspan="2">Continuous</td></tr>
<tr><td colspan="2">Type</td><td colspan="2">Continuous stirred type reactor</td></tr>
<tr><th colspan="4">Function</th></tr>
<tr><td colspan="4">Hydrolysis of Cellulose to Glucose</td></tr>
<tr><td colspan="4">Chemical Reaction
$(C_6H_{10}O_5)\ n + H_2O \rightarrow C_6H_{12}O_6$</td></tr>
<tr><th colspan="2">Reactor Specifications</th><th colspan="2">Jacket Specifications</th></tr>
<tr><td>Volume of Reactor</td><td>1.2 m^3</td><td>Jacket Type</td><td>Dimple Jacket</td></tr>
<tr><td>Diameter of Reactor</td><td>1 m</td><td>Steam Temperature</td><td>152℃</td></tr>
<tr><td>Length of Reactor</td><td>1.5 m</td><td>Steam Pressure</td><td>5 bar</td></tr>
<tr><td>Type of Impeller</td><td>Rushton Impeller</td><td>hi</td><td>3.927 W/m^2K</td></tr>
<tr><td>Speed of Impeller</td><td>100 rev/min</td><td>ho</td><td>8520 W/m^2K</td></tr>
<tr><td>Diameter of Impeller</td><td>0.33 m</td><td>Ud</td><td>3.72 W/m^2K</td></tr>
<tr><td>Power</td><td>1.3 hp</td><td>Area</td><td>5 m^2</td></tr>
</table>

SPECIFICATION SHEET

Identification			
Item		Reactor	
Item no.		R-101	
No. required		1	
Operation		Continuous	
Type		Continuous stirred type reactor	
Function			
Pre-Hydrolysis of Cellulose to Glucose			
Chemical Reaction $(C_6H_{10}O_5)\ n \rightarrow C_6H_{12}O_6$			
Reactor Specifications		**Jacket Specifications**	
Volume of Reactor	5.1 m^3	Jacket Type	Dimple Jacket
Diameter of Reactor	1.7 m	Steam Temperature	152°C
Length of Reactor	2 m	Steam Pressure	5 bar
Type of Impeller	Pitched Blade Impeller	hi	3.327 W/m^2K
Speed of Impeller	100 rev/min	ho	8520 W/m^2K
Diameter of Impeller	0.58 m	Ud	3.27 W/m^2K
Power	5.7 hp	Area	13 m^2

5.2 Distillation Column Design:

5.2.1 Introduction:

In industry it is common practice to separate a liquid mixture by distilling the components, which have lower boiling points when they are in pure condition from those having higher boiling points. This process is accomplished by partial vaporization and subsequent condensation.

"***It is a process in which a liquid or vapor mixture of two or more substances is separated into its component fractions of desired purity, by the application and removal of heat".***

The creation or addition of another phase in distillation is obtained by the repeated vaporization and condensation of the fluid. The separation process exploits the differences in vapor pressure of key components in the mixture initiate the separation.

The advantages of distillation are its simple flow sheet, low capital investment and low risk. The separation process can handle wide ranges of feed concentrations and throughputs while producing a high purity product.

5.2.2 Types of Distillation Column:

Distillation columns come in a variety of forms, most of which is created to carry out a particular kind of separation and varies in complexity.

- Batch columns
- Continuous columns

Batch Columns:

In a batch operation, the feed is introduced to the column in batches. For more information, the distillation process is carried out once the column is charged with a "batch" of material. A new batch of feed is supplied once the product has undergone the necessary level of purification.

Continuous Columns:

In contrast, a continuous feed stream is processed by continuous columns. Unless there is a problem with the column or nearby process units, there are no interruptions. They are the more prevalent of the two varieties and are able to manage high throughputs. We will solely focus on this category of columns.

5.2.3 Choice Between Packed and Plate Column:

The mass transfer of vapors and liquids can be done in a packed column or a plate column. These two separate sorts of operations are very dissimilar. a selection process that takes into account the elements under four areas,

- ✓ Scale, foaming, pressure loss, and liquid holdup are factors that rely on the system.
- ✓ Elements that are affected by the fluid flow moment.
- ✓ Factors that depend on the column's internal structure and physical properties, such as cost, size, weight, and side stream.

✓ Factors that are affected by the mode of operation, such as batch and continuous distillation turndown, and intermittent distillation.

Following are the advantages of plate over packed column:

1. Plate columns are used to manage dispersion issues when liquid flow rates are lower than gas flow rates.
2. For large columns, the packed column weighs more than the plate column.
3. Manholes shall be fed for cleaning if recurring maintenance is necessary. Before cleaning packed columns, packaging must be taken out.
4. The plate column is ideal for non-foaming systems.
5. Compared to packed columns, design information for plates columns is more accessible and trustworthy.
6. To reduce the temperature of a reaction or solution in a plate column, inter-stage cooling can be offered.
7. Packing may be harmed when there is a temperature fluctuation.

5.2.4 Choice of Plates in Column:

In distillation column, Four main tray types, the bubble cap, sieve tray, ballast or valve trays and the counter flow trays are used.

I have selected sieve tray because:

✓ They are lighter in weight and less expensive. Easily install.

✓ Cleaning is so much easy as compared to other types of trays so maintenance cost is reduced as compared to others.

5.2.5 Selection of Trays:

Cost:

Cost of plate depends upon material of construction used.

Valve plate : Sieve plate : Bubble-cap plate

1.5 : 3 : 1.0

Operating Range:

Comparison of operating range flexibility is,

Bubble cape tray > Valve tray > Sieve tray

Sieve plates provide a reasonable operating range with good design.

Pressure Drop:

Bubble-cap tray > valve tray > sieve tray

5.2.6 Main Components of Distillation Column:

1. Column internals such as trays/plates and/or packaging which are utilized to enhance component separations.
2. A reboiler to supply the required vaporization for the distillation process. The liquid taken out of the reboiler is referred to as bottoms, or just bottoms.

3. A condenser that will cool and condense the vapor that is leaving the column's top. The distillate, referred as the top product, is the condensed liquid that is taken out of the system.

5.2.7 Factors Affecting the Distillation Column Operation:

1. **Foaming:** The term foaming describes how a liquid expands when a vapor or gas passes through it. Excessive foaming frequently causes liquid to accumulate on trays even if it offers high interfacial liquid-vapor contact. Foaming can occasionally get so terrible that it mixes with the liquid in the tray above. The physical characteristics of the liquid mixtures have the most role in determining whether foaming will happen, but tray designs and conditions can also play a role.
2. **Entrainment:** Once more, entrainment is brought on by high vapor flow rates and refers to the liquid that the vapor carries up to the tray above. It is harmful because tray efficiency is decreased and less volatile material is transferred to a plate holding a more volatility. High purity distillate might also become contaminated. Over-entrainment might result in floods.
3. **Weeping/Dumping:** Low vapor flow is the cause of this condition. The liquid on the tray cannot be supported by the vapor's pressure alone. Liquid begins to leak as a result through perforations. Weeping too much will result in dumping. In other words, a domino effect will cause the liquid on all trays to crash (dump) through to the base of the column, necessitating a restart of the column.
4. **Flooding:** Excessive vapor flow causes liquid to be entrained in the vapor up the column, which results in flooding. An increase in liquid holdup on the plate above results from the pressure buildup brought on by too much vapor, which also backs up the liquid in the down comer. The column's maximum capacity may be significantly lowered depending on the level of flooding. Sharp rises in column differential pressure and a sizable decline in separation efficiency are indicators of flooding.
5. **Feed Conditions:** The operating lines and, thus, the number of stages needed for separation are influenced by the state of the feed mixture and feed composition. The position of the feed tray is also impacted. Trays and packaging condition: Keep in mind that the effectiveness of the plate determines the precise number of trays needed for a given separation duty. Therefore, any elements that lower tray efficiency will also affect how well the column performs. The rates at which fouling, wear and tear, and corrosion affect tray efficiency vary on the characteristics of the liquids being processed.
6. **Column Diameter:** Column diameter affects vapor flow velocity. Column capacity is calculated by weeping, which establishes the minimal vapor flow necessary, and flooding, which establishes the highest vapor flow permitted. As a result, the column will not function effectively if the column diameter is not sized properly.

5.2.8 Design calculations:

1. Designing steps of distillation column:
2. Bubble point and dew point calculations.
3. Key components selection.
4. Determining the Minimum number of stages.(N min).
5. Minimum Reflux Ratio (Rm) calculations.

6. Determining the Actual Reflux Ratio R.
7. Calculate the theoretical number of stages and actual number of stages.
8. Establishing the physical characteristics of the top and bottom products.
9. Calculating the column's diameter.
10. Calculation of entrainment, weeping point, etc.
11. Calculating pressure drop.
12. Determining the column's height.

5.2.9 Design Calculations:

Temperature of feed = 83°C

Temperature of top product = 78°C

Temperature of bottom product = 83°C

Heavy Key Component = Water

Light Key Component = Ethanol

Table 5.1: Feed Composition

Components	Feed (X_f) %	Distillate (X_d) %	Bottom (X_b) %	Relative Volatility
Ethanol	77	99.5	0.005	1
Water	21	0.5	90.5	0.969
Ammonium Acetate	2	-	9.8	5.4226

Calculation of Minimum Reflux Ratio R_m

We are using Underwood equation,

$$\frac{x_{fA}\propto_A}{\propto_A-\theta}+\frac{x_{fB}\propto_B}{\propto_B-\theta}+\frac{x_{fC}\propto_C}{\propto_C-\theta}=1-q \tag{5.21}$$

As feed is entering as its boiling point so, q = 1

By trial, θ = 0.9757

We are using eq. of min. reflux ratio,

$$\frac{x_{fA}\propto_A}{\propto_A-\theta}+\frac{x_{fB}\propto_B}{\propto_B-\theta}+\frac{x_{fC}\propto_C}{\propto_C-\theta}=R_m-1 \tag{5.22}$$

Putting all values $R_m = 2.6907$

Actual Reflux Ratio

We follow the rule of thumb is:

$R = (1.2 - 1.5)\ R_{min}$

$R = 3.2284$

Minimum No. of Plates

For minimum no. of stages N_{min} is obtained.

Using Fenske relation which is,

$$N_{min} = \frac{\log\left[\left(\frac{x_B}{x_C}\right)_D\left(\frac{x_C}{x_B}\right)_B\right]}{\log(\propto_{BC})_{avg}} \tag{5.23}$$

Theoretical No. of Plates

For theoretical number of plates,

$$\frac{N-N_{min}}{N-1} = 0.75\left[1-\left(\frac{R-R_{min}}{R-1}\right)^{0.566}\right] \tag{5.24}$$

Theoretical no. of stages to be,

N = 24 trays

We removed One plates for Re-boiler, so

N = 24-1 = 23 trays

N_{min} = 10

Location of Feed Plate

$$\log\frac{N_D}{N_B} = 0.206\log\left[\left(\frac{B}{D}\right)\left(\frac{x_{HK}}{x_{LK}}\right)\left(\frac{(x_{LK})_B}{(x_{HK})_D}\right)^2\right] \tag{5.25}$$

$N_D = 0.27N_B$

$N_A = N_D + N_R$

Number of Plates above the feed tray N_R = 26

Number of Plates below the feed tray N_D = 8

So, the feed enters at 9th plate

Actual Number of Stages

N_A = 34 trays

$$\text{Eo} = \frac{\text{Number of thoretical stages}}{\text{Actual number of stages}} \tag{5.26}$$

Eo = 70%

Determination of the Column Diameter

Table 5.2: Product Conditions

Top Conditions	Bottom Conditions
$L_n = 34.9\times10^4$ Kg/hr $V_n = 47.9\times10^4$ Kg/hr Average mol.wt = 46.07 g/mol $T = 78^0$ C $\rho V = 3.134$ Kg/ m^3 $\rho L = 701.4$ Kg/ m^3	$L_m = 49.6\times10^4$ Kg/hr $V_m = 47.9\times10^4$ Kg/hr Average mol.wt = 77.01 g/mol $T = 83^0$ C $\rho V = 1.57$ Kg/ m^3 $\rho L = 950.1$ Kg/ m^3

Flow Parameters

Liquid and vapor flow rates are larger at bottom so based upon bottom flow rates.

$$F_{LV} = \left(\frac{L_m}{V_m}\right)\left(\frac{\rho_v}{\rho_L}\right)^{0.5} \tag{5.27}$$

F_{LV} = Liquid Vapor Factor = 0.0420

Capacity Parameter

Assumed tray spacing = 14 in. = 0.3m

$C_{sb(20)}$ = 0.061 m/Sec (Capacity parameter for liquids (σ = 20 dynes/cm))

Surface tension of system = α = 27.3 dynes/Cm

$$C_{sb} = C_{sb(20)}\left(\frac{\propto}{20}\right)^{0.2} \tag{5.28}$$

$$= 0.103 \text{ m/s}$$

$$U_{nf} = C_{sb}\left(\frac{\rho_L-\rho_v}{\rho_v}\right)^{0.5}$$

$$= 1.41 \text{ m/s}$$

Tray Selection

We have selected single cross flow sieve tray with segmental down comer.

Down comer area = $A_d = 0.12\ A_T$

Weir length = $L_w = 0.77\ D_T$

weir height = h_w = 50 mm

Hole size (range 1/8“ to ½”) = 1/8” = 3.015 mm

Spacing between trays = 14 in. = 0.3m

Tower Diameter

Let flooding = 80% (by trial)

$F^* = 0.8$

$U_n^* = U_{nf} \times F^* = 1.128$ m/Sec

Net area.

Net Area = [column area (cross sectional) - (Down-comer area)]

$A_n = A_T - A_d = 0.88\ A_T$ *(5.29)*

$A_T = \frac{A_n}{0.88} = \frac{Q_v}{0.88 U_n}$

$A_T = 2.45\ m^2$

$A_T = (\pi/4)*D^2$

D = 1.76 m

Tower area = $A_T = 2.45\ m^2$

Net area = $A_n = 0.88\ A_T$

= $2.156\ m^2$

Down comer area = $A_d = 0.12\ A_T$

= $0.294\ m^2$

Hole area = $A_h = 0.1\ A_T$

= $0.245\ m^2$

Flooding Check

$V_{max} = K_1 \sqrt{\rho L - \rho v / \rho v}$ *(5.30)*

$V_{max} = 1.09$

$F = \frac{Un}{Uf} \times 100$

F = 78%

Calculation of Entrainment

As FLV = 0.042

F = 80%

$\psi = 0.06$

Since $\psi < 0.1$, process is satisfactory

Tray Pressure Drop

Ht =Hd + (Hw + How) + Hr *(5.31)*

$H_w = 50$ mm

Dry Tray Pressure drop

$H_d = 51\ (U_h/Co)^{2}\ (\rho_V/\rho_L)$ *(5.32)*

U_h = Hole velocity = Q_v/A_h

$U_h = Q_v/A_h = 11.40$ m/sec

Using Fig, we find "C_o"

$C_o = 0.80$

$H_d = 51*(Uh/Co)^2*(\rho_v/\rho_L)$

$H_d = 17.11$ mm

Weir Crest

$H_{ow} = 750\ (L_w/\rho_L * l_w)^{2/3}$ *(5.33)*

$L_w = 0.77D_T$

$L_w = 1.35$ m

$H_{ow} = 750\ (L_w/\rho_L * l_w)^{2/3}$

$H_{ow} = 18.61$ mm

Residual Head (Hr)

$H_r = \left(\frac{12.5\times10^3}{\rho_L}\right)$ *(5.34)*

$H_r = 17.83$ mm

$H_t = H_d + (H_w + H_{ow}) + H_r$

$= 17.11 + 50 + 18.61 + 17.83$

$= 103$ mm

Total Pressure Drop

$P_t = (9.81 \times 10^{-3}\ 10)\ H_t \times \rho_L$

$= 9.81 \times 10^4 \times 78.9 \times 701$

$= 708$ Pa

$= 0.1$ psi

Estimation of Weep point

$$\bar{U}_{h(min)} = \frac{K_2 - [0.90 - (25.4 - d_h]}{\left(\rho_V\right)^{0.5}} \quad (5.35)$$

$H_W = 25.4$ mm

$H_{OW} = 18.61$ mm

$H_W + H_{OW} = 25.4+18.61= 44.18$ mm

Using graph,

$K_2 = 30.2$

$Uh_{(min)} =$ 9.2 m/sec

Actual Min. Vapour Velocity $\geq Uh_{(min)}$

No Weeping.

Total no. of holes

Total no. of holes = A_h/a_h

Diameter of 1 hole = 5mm

= 0.005m

Area of one hole = $(3.14 \times 2.5 \times 10^{-5})/ 4$

$= 1.96 \times 10^{-5}$ m^2

Total no. of holes = A_h/a_h

Total number of holes = 12500

Height of Distillation Column

No. of plates = 34

Tray spacing = 0.30m

Distance between 12 plates = 0.30 × 34 = 10.2 m

Tray thickness = 3 mm/plate

Total Height of column = $[(34-1) \times 0.3] +0.5$

= 10 m

SPECIFICATION SHEET			
Identification			
Item		Distillation column	
Item no		D-102	
No of required		2	
Type		Multi- components	
Calculations			
No of plates	34	Tray spacing	0.3m
Height of column	10m	Efficiency	70%
Flooding	80%	Hole area	1.96×10^{-5} m^2
Weeping	No weeping	Hole size	3.175 mm
Fractional entrainment	0.06	Pressure drop	0.1 psi
Reflux ratio	3.22	Diameter of column	1.76 m

<table>
<tr><th colspan="4">SPECIFICATION SHEET</th></tr>
<tr><th colspan="4">Identification</th></tr>
<tr><td colspan="2">Item</td><td colspan="2">Distillation column</td></tr>
<tr><td colspan="2">Item no</td><td colspan="2">D-101</td></tr>
<tr><td colspan="2">No of required</td><td colspan="2">2</td></tr>
<tr><td colspan="2">Type</td><td colspan="2">Multi- components</td></tr>
<tr><th colspan="4">Calculations</th></tr>
<tr><td>No of plates</td><td>16</td><td>Tray spacing</td><td>0.3m</td></tr>
<tr><td>Height of column</td><td>5.43m</td><td>Efficiency</td><td>68%</td></tr>
<tr><td>Flooding</td><td>75%</td><td>Hole area</td><td>1.96×10^{-5} m^2</td></tr>
<tr><td>Weeping</td><td>No weeping</td><td>Hole size</td><td>3.175 mm</td></tr>
<tr><td>Fractional entrainment</td><td>0.04</td><td>Pressure drop</td><td>0.09 psi</td></tr>
<tr><td>Reflux ratio</td><td>0.712</td><td>Diameter of column</td><td>1.01 m</td></tr>
</table>

5.3 Design of Waste Heat Boiler (WHB-102):

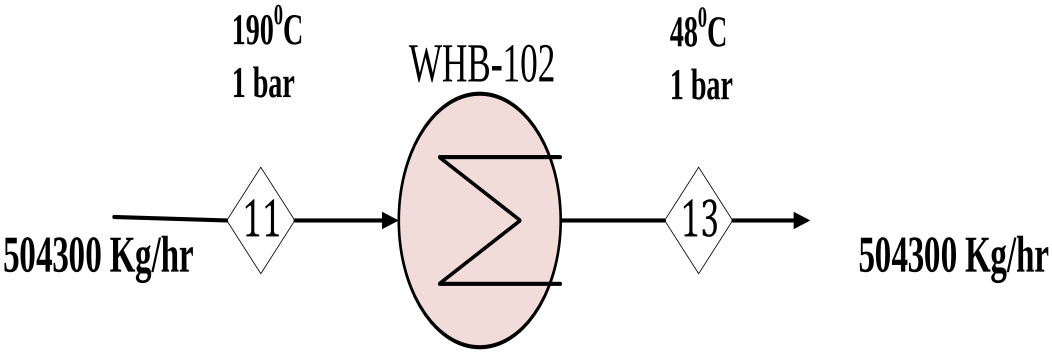


Figure 5.3: Waste Heat Boiler (WHB-102)

5.3.1 Introduction:

A waste heat boiler converts heat generated as a byproduct of another operation, heat that would otherwise be squandered, into steam. Steam may be used to power turbines that generate energy. The boiler can also be used to merely heat water or other types of fluid. A waste heat boiler, also known as a waste heat recovery boiler, can lower a system's fossil fuel consumption and operating costs by recycling part of the energy utilized. This also implies that less greenhouse gases enter the atmosphere.

A waste heat boiler with a water-tube design can handle significantly greater steam pressures than a boiler with a fire-tube design, but it is more complicated to build and install. The tubes in this type of boiler are thinner than those in a fire-tube boiler, and they hold water rather than hot gases. Waste heat, in the form of hot gases or furnace flames, surrounds the water-filled tubes in a reversal of the system within a fire-tube boiler. To prevent the boiler tubes from flame damage, insulating materials are utilized. A water-tube waste heat boiler can withstand high pressures while also responding swiftly to variations in heat input.

We have selected water tube waste heat boiler due to following reasons

5.3.2 Advantages:

- ✓ Working pressures is high
- ✓ Superheated steam generation
- ✓ Heat recovery is faster
- ✓ Better turn down

5.3.3 Design Calculations:

Heat Duty

Temperature Input= 190^{o}C

Temperature Output = 48^{o}C

As the relation for latent heat is given as

Q = 8.96×10^5 MJ/hr

Steam Requirement

Pressure = 5 bar

m = 3.3×10^5 kg/hr

Energy required and heat production

- ✓ 2% heat loses

Using the relation for energy required and heat loss

$Q_T = Q_u + 0.02Q_T$ *(5.36)*

Rearranging the above the equation and putting values

$Q_T = 9.14\times10^7$ kg/hr

Water Requirement

$\text{Mass of Feed water }(M_F) = \text{Mass of Stream}(M_F) + \text{Blow} - \text{down}(M_B)$ *(5.37)*

Blow-down up to 10% (Proposefull discharge)

$M_F - 0.1M_F = M_S$

$M_S = 36.6\times10^4$ kg/hr

$M_S = 101.6$ kg/sec

Estimation of Surface Area

For system having heavy organics assume $U_D = 700 - 1140$

So assuming,

$U_D = 1140$ W/ m^{2}k

$U_D = 4104$ kJ/hr.m^{2}k

LMTD Calculations

$$\Delta T_{ln} = \frac{(T_1 - t_1) - (T_2 - t_2)}{l_n\left(\frac{T_1 - t_1}{T_2 - t_2}\right)} \quad (5.38)$$

$\Delta T_1 = 38^oC$

$\Delta T_2 = 23^oC$

LMTD = 30^oC

Correction Factor

$R = \frac{T_a - T_b}{t_b - t_a} = 1.11$

$S = \frac{t_b - t_a}{T_a - T_a} = 0.139$

$F_{LMTD} = 0.91$

$\Delta T_c = \text{LMTD} \times F_{LMTD}$ *(5.39)*

ΔT_c = 301k

Applying Correction factor

A = $\frac{Q}{U\Delta T_{ln} f}$

A =74 m^2 = 796 ft^2

Tube Dimensions:

Length of tube = 16 ft = 4.87m

Tube outer diameter = $\frac{3}{4}$in = 0.019m

Tube inner diameter = 0.620in = 0.016m

BWG = 16

Tube side (Process Slurry)

No of tubes

$N_t = \frac{A}{\pi d_o L}$ *(5.40)*

N_t = 254

Tube outer dia = $\frac{3}{4} in$ = 0.019m

1 shell and 2 passes

Triangular pitch

Inner diameter of shell = 19.25in = 0.488m

Bundle diameter

K_1 = 0.249

n_1 = 2.207

$D_b = d_o \left(\frac{N_t}{k_1}\right)^{\frac{1}{n1}}$

D_b = 0.44m

Tube cross-sectional area

Cross – sectional area = $\frac{\pi d_o^2}{4}$ *(5.41)*

Cross sectional area =$2.8 \times 10^{-4} m^2$

Tube per passes

For 2 passes

N_t = 127

Area per passes = tube per area × cross-sectional area

$= 0.03\ \mathrm{m}^2$

Volumetric flow rate:

$$\text{Volumetric flow rate} = \frac{\text{Feed flowrate}}{\text{Density of water}} \quad (5.42)$$

$= 0.11\ \mathrm{m}^3/\mathrm{s}$

Tube velocity:

$$\text{Tube velocity} = \frac{\text{volumetric flow rate}}{\text{Area per passes}} \quad (5.43)$$

$G_t = 3.3\ \mathrm{m/s}$

Reynold Number:

$$\text{Reynold Number} = \mathrm{Re} = \frac{\rho u d_i}{\mu} \quad (5.44)$$

$\mathrm{Re} = 14671.8$

Prandtl number:

$$\text{Prandtl number} = \mathrm{Pr} = \frac{\mu C_p}{k} \quad (5.45)$$

$\mathrm{Pr} = 224$

$$\frac{L}{D_i} = \frac{4.87}{0.019} = 256$$

From graph,

$j_H = 1.9 \times 10^{-3}$

Tube side coefficient:

$$h_i = \frac{k}{d_i} j_H \mathrm{Re}(\mathrm{Pr})^{0.33}\left(\frac{\mu}{\mu_w}\right)^{0.14} \quad (5.46)$$

$h_i = 1837\ \mathrm{W/m^2k}$

Shell side (Water)

Shell diameter

For fixed tube, diametrical clearance

$= 13 \times 10^{-3}\mathrm{m}$

$\text{Shell diametr} = \mathrm{Db} + \text{diametrical clearance}$

$\mathrm{Ds} = 0.45\mathrm{m}$

Ideal cross flow coefficient (H_{oc})

Tube pitch = $P_t = 1.25do$

$P_t = 0.024\mathrm{m}$

Area of shell:

$$As = \left(\frac{\rho_t - d_o}{\rho_t}\right)(D)\ (L_B) \qquad (5.47)$$

$L_B = \frac{d_s}{5} = 0.1m$

$As = 0.01\ m^2$

Shell side velocity

$$\text{Shell velocity} = \frac{\text{flow rate}}{\text{Area}} \qquad (5.48)$$

$G_t = 706\ kg/m^2s$

Reynold Number

$\text{Reynold Number} = Re = \frac{Gsd_o}{\mu}$

$Re = 31008$

Prandtl number

$\text{Prandtl number} = Pr = \frac{\mu C_p}{k}$

$Pr = 3.5$

Shell side coefficient:

$$h_{oc} = \frac{k}{d_o} j_H Re(Pr)^{0.33}\left(\frac{\mu}{\mu_w}\right)^{0.14} \qquad (5.49)$$

$h_{oc} = 4460\ w/m^2/s$

Tube row correction factor (Fn):

Tube vertical pitch = $Pt' = 0.87pt$

$Pt' = 0.02m$

Baffle height cut = baffle cut × D_s

$hc = 0.12m$

Height b/w baffle cut = shell ID $- 2 \times hc$

= 0.25m

$N_{cv} = \frac{\text{height between baffle cut}}{\text{tube vertical pitc}}$

$N_{cv} = 12.5$

From fig, 12.32 we get value of Fn

Fn = 1.04

$j_H = 3.0 \times 10^{-2}$

Window correction factor:

$H_b = \frac{Db}{2} - D_S\ (0.5 - B_c)$ *(5.50)*

$H_b = 0.11m$

Bundle cut:

$B_b = \frac{H_B}{Db} = 25\%$

From fig, 12.41 Ra = 0.19

Tube in one window area:

Nw = Nt× Ra

= 48

Tube in cross flow area

Nc = Nt-2Nw

= 158

$Rw = \frac{2\times Nw}{Nt} = 0.37$

From Figure, 12.33 Fw = 1.03

Bypass correction:

$A_B = l_b \times (D_S - D_b)$

$A_B = 2.2 \times 10^{-3} m^2$

$F_B = \exp[-\alpha \times \frac{Ab}{As}\ (1 - (\frac{2Ns}{Ncv}\ )0.33]$ *(5.51)*

$F_B = 0.99$

Leakage correction:

$Ct = 7 \times 10^{-4} m$

$Cs = 4.7 \times 10^{-3} m$

$A_{tb} = \frac{Ct\ \pi Do}{2} \times (Nt-Nw)$ *(5.52)*

$A_{tb} = 4.2 \times 10^{-3} m^2$

$A_{sb} = \frac{CsDs}{2}\ (2\pi - Qb)$

$A_{sb} = 3.8 \times 10^{-3}$

$A_L = A_{tb} + A_{sb}$

$A_L = 7.8 \times 10^{-3}$

From fig, $\beta_l = 0.43$

$F_L = 1 - \beta_l \left(\frac{A_{tb} + 2A_{sb}}{A_L}\right)$

$F_L = 0.5$

Shell side coefficient:

$$H_S = H_{oc} + F_L + Fw + Fn + Fb \qquad (5.53)$$

$H_S = 2364\ \mathrm{W/m^2 s}$

Overall heat transfer coefficient:

$$\left(\frac{1}{U_o}\right) = \left(\frac{1}{h_o}\right) + \left(\frac{1}{h_{od}}\right) + \left(\frac{d_o}{2k_m}\right) + ln\left(\frac{d_o}{d_i}\right) + \left(\frac{d_o}{d_i}\right)\left(\frac{1}{h_{id}}\right) + \left(\frac{d_o}{d_i}\right)\left(\frac{1}{h_i}\right) \qquad (5.54)$$

$h_{od} = 5000\ \mathrm{w/m^2}$

$Kw = 16\ \mathrm{W/m^2 s}$

Putting values

$U_D = 940\ \mathrm{W/\ m^2 k}$

Pressure Drop:

Tube Side (Heavy Organic)

No of tubes =254

1 shell and 2 passes

ID of tube = 0.016m

$U_t = 3.3\ \mathrm{m/s}$

$J_f = 1.9 \times 10^{-3}$

$$\Delta P_t = N_P\ [8\ J_f\ \frac{l}{Di}\ (\frac{\mu}{\mu w})^{-0.14} + 2.5]\frac{\rho\ U_t}{2} \qquad (5.55)$$

$\Delta P_t = 8$ psi

Shell Side (Water)

$$\Delta P_t = 2\ \Delta P_e + \Delta P_c\ (\ N_b - 1) + \ N_b\ \Delta P_w \qquad (5.56)$$

Cross flow zone

$J_h = 8.3 \times 10^{-3}$

$\mu s = 0.7\ \mathrm{m/s}$

$\Delta P_i = 8\ \ J_f\ \ N_{cv}\ \square\ \frac{(us)2}{2}$

$\Delta P_i = 158$

$F'_B = \exp[-\alpha \times \frac{Ab}{As}\ (1 - (\frac{2Ns}{Ncv}\ )0.33]$

$F_B = 1.00$

From figure, $\beta_l = 0.66$

$F_L = 1 - \beta_l \left(\frac{A_{tb} + 2A_{sb}}{A_L}\right)$ *(5.57)*

$F'_L = 0.06$

$\Delta P_c = F'_B\, \Delta P_i F'_L$

$\Delta P_c = 10.45$

Window zone

$A_w = \left(\frac{\pi (Ds)^2}{4} \times Ra\right) - \left(Nw \times \frac{\pi (Do)^2}{4}\right)$

$A_w = 9.1 \times 10^{-3}$

$Nw = 10$

$\Delta P_w = 2.35$

$\Delta P_e = \Delta P_i \frac{Nw + Ncv}{Nw} \times F'_b$ *(5.58)*

$\Delta P_e = 316$

$\Delta P_s = 2\, \Delta P_e + \Delta P_c (Nb - 1) + Nb \Delta P_w$ *(5.59)*

$\Delta P_s = 0.10$psi

<table>
<tr><th colspan="4">SPECIFICATION SHEET</th></tr>
<tr><th colspan="4">Identification</th></tr>
<tr><td colspan="2">Item</td><td colspan="2">Waste Heat Boiler</td></tr>
<tr><td colspan="2">Item no.</td><td colspan="2">WHB- 102</td></tr>
<tr><td colspan="2">No. of required</td><td colspan="2">3</td></tr>
<tr><td colspan="2">Type</td><td colspan="2">1-2 horizontal heat exchanger</td></tr>
<tr><td colspan="2">Operation</td><td colspan="2">Continuous</td></tr>
<tr><td colspan="4">Utilization of extra heat in output gases by generating steam</td></tr>
<tr><td colspan="4">Heat Duty = $8.9 \times 10^5 \frac{\mathrm{MJ}}{\mathrm{hr}}$</td></tr>
<tr><td colspan="4">Heat Transfer area = $74\ m^2$</td></tr>
<tr><td>Operating Pressure</td><td>1 bar</td><td>Operating Pressure</td><td>1 bar</td></tr>
<tr><td>Temperature In/ out</td><td>21- 152 ℃</td><td>Temperature In/ out</td><td>190- 48 ℃</td></tr>
<tr><td>Diameter</td><td>0.45m</td><td>Tube inner Diameter</td><td>0.016</td></tr>
<tr><td>Passes</td><td>1</td><td>No of tubes</td><td>254</td></tr>
<tr><td>Shell Diameter</td><td>0.488 m</td><td>Tube outer Diameter</td><td>0.019 m</td></tr>
<tr><td>Pressure Drop</td><td>0.10 Pa</td><td>Pressure Drop</td><td>8 Pa</td></tr>
</table>

5.4 Design of Heat Exchanger:

5.4.1 Introduction:

A Heat Exchanger is a heat transfer device that is used for transfer of internal thermal energy between two or more fluids available at different temperatures. In most of the exchangers the fluids are separated by a heat transfer surface and ideally don't mix with each other.

5.4.2 Basic Principle of Heat Exchangers:

The main principle of a heat exchanger is the exchange of thermal energy on the basis of thermal gradients between two bodies (fluids). The mechanism of heat transfer in heat exchangers is the combination of the basic heat transfer mechanisms. The basic heat transfer mechanisms are:

- ✓ Conduction
- ✓ Convection
- ✓ Radiation

5.4.3 Types of Flow Arrangements in Heat Exchangers:

One of the following flow types may be used in a heat exchanger, which is the most typical in practice.

- ✓ Parallel flow
- ✓ Counter-flow
- ✓ Crossflow

5.4.4 Selection Criteria of Heat Exchanger:

Proper selection of heat exchanger depends upon following factors:

- ✓ Heat transfer rate
- ✓ Operating temperature
- ✓ Cost
- ✓ Pumping power
- ✓ Material of construction
- ✓ Flow rates
- ✓ Flow arrangements
- ✓ Phases of fluids

5.4.5 Why Shell and Tube Heat Exchanger Is Selected:

The reasons of selection of this heat exchanger are as follows:

- ✓ Easy maintenance
- ✓ Having great heat transfer
- ✓ Well-proven design procedures.
- ✓ Can be built from a variety of materials.
- ✓ Applicable for large heat transfer coefficients

5.4.6 Design Steps of Shell & Tube Heat Exchanger:

In designing the shell and tube heat exchanger the following steps are involved.

1. Q is available from energy balance so with heat balance the flow rate of utility and process stream determined by heat balance.
2. Calculate log mean temperature difference i.e.

$$\text{LMTD} = \frac{\Delta T_2 - \Delta T_1}{\log\frac{\Delta T_2}{\Delta T_1}} \tag{5.60}$$

$\Delta T_2 = T_2 - t_1$; $\Delta T_1 = T_1 - t_2$

3. Then calculate R & S for FT i.e.

$$R = \frac{T_1 - T_2}{t_2 - t_1}, \qquad S = \frac{t_2 - t_1}{T_1 - t_1}$$

4. Get F_T from Fig.
5. True temperature difference by multiplying F_T & LMTD.
6. When there is no available exchanger and only the process conditions are known Q and true temperature difference are fixed by the process conditions
7. Only A and U_D are unknown
8. If U_D is considered to have a trial value and A can be determined to have a trial value
   i. $A = Q/U_D.\ \Delta Tt$
9. For trial value of U_D see Appendix Table 8
10. U_D is related to U_c by a reasonable dirt factor R_d
11. The criterion of performance R_d was then obtained from U_D and U_c
12. Except where both coefficient are approximately equal, the lower film coefficient determines the range of U_c and U_D
13. From fluid flow conditions h_o, h_{io}, U_c, and pressure drops were calculated
14. The tube counts in Appendix Table 9 become a list of all conceivable exchanger shells when the value of A is paired with tube length and pitch.
15. Having decided which fluid will flow in the tubes and which one in shell the trial number of tube passes and number shell passes can be approximated.
16. Then get internal diameter of shell and the internal diameter of tubes from table 9 according number tubes and equivalent diameter from figure 28 according taking square pitch. The outer diameter of tube is selected from table 10.
17. The length of tubes range up to 16 in.ft. The BWG is up to 16".
18. After getting all calculate the area of shell and tube side.
19. The mass velocity is also calculated for both sides.
20. Then calculate the equivalent diameter of both sides.
21. Reynolds number across shell side and tube side
22. Calculating Prandtl's number
23. Calculating factors for heat transfer coefficients
24. Individual heat transfer coefficients calculation for both sides
25. Then overall clean coefficients i.e.

$$U_c = \frac{h_{io} h_o}{h_{io} + h_o} \tag{5.61}$$

26. Dirt factor calculations i.e.

$$R_d = \frac{U_c - U_D}{U_c U_D}$$

27. Pressure drop calculations are also done for shell side and tube side.

5.4.7 Design of Heat Exchanger (H-103):

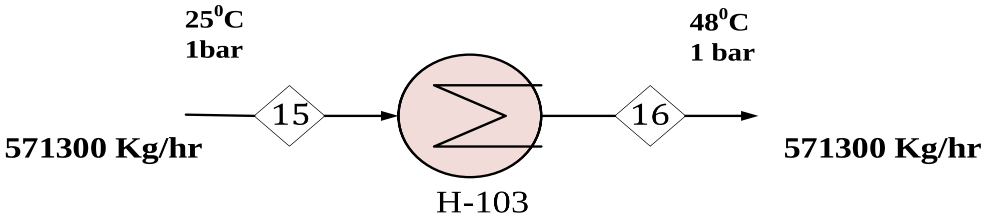


Figure 5.4: Heat Exchanger (H-103)

Steam:
Temperature = 152^{o}C
Pressure = 5 bar
Shell side

Aqueous solution is used on shell side because it is less corrosive.

Tube Side

Steam is used on tube side as it is corrosive in nature.

Calculation of Heat Duty:

Aqueous solution:

Q = mass flow rate × specific heat × temperature difference

Q = $m\ C_p\ \Delta T$

= 1.0×10^8 Btu/hr

Steam:

Q = $m\lambda$

Q = $1.0*10^8$ BTU/hr

Q = Enthalpy of Hot stream

C_p = mass heat capacity

Calculation of LMTD:

LMTD = $\{\Delta T_2 - \Delta T_1\}/ \ln(\Delta T_2/\Delta T_1)$ *(5.62)*

LMTD = 186^oF

Heat Transfer Area

Q = UA ΔT

Where,

U_D = Heat transfer coefficient

Assumed U_D = 680 BTU/hrft^{2}F [From table 8]

(Range = 200-700 BTU/hrft^{2}F)

Heat transfer area,

A = 790 ft^2

Tube specification:

16 BWG

Outside Diameter = OD = 1.5 in

Inside Diameter = ID = 1.37 in

Length of Tube = L = 16 ft

Pitch = 15/ 8 (square)

Passes = 1

(Triangular pitch)

Correction of Heat transfer Area and U_D:

Corrected Area:

$A = N_t \times L \times a$

$A_c = 823 \text{ ft}^2$

Corrected Coefficient U_D:

U_D = 653 BTU/hrft^{2}F

Number of Tubes N_t:

Number of Tubes = $N_t = A/L*a$ *(5.63)*

Nt = 131

Table 5.3: Dimensions of Shell and Tube Side

Shell Side	Tube Side
ID =29 in	N_t =131
Baffle =5.8 in	Length =16 ft
Passes =1	OD =1.5 in
Clearance =1.8 in	16 BWG
	Pitch =15/8 (square)
	Passes =1

Shell and tube side's calculations:

Shell side (Aqueous solution)	**Tube side (steam)**
Flow area:	**Flow area:**
a_s = ID*C*B/144P_t	a_t = N_t*A_t/144n
= 1.2 ft^2	= 1.3 ft^2
Mass velocity:	**Mass velocity:**
G_s = w/a_s	G_t = w/a_t
= 425000 lb/hr ft^2	= 84615 lb/hrft^2
Re =D_e*G_s / μ	**Re =D_i*G_t / μ**
D_e = 0.018m	D_i = 0.015m
Re_s=106250	Re_t=38782
J_H Factor [From Fig 8]	**J_H Factor**
J_H = 220	**Tube Side Coefficient (h_i)**
From Figure	
k= 0.06 BTU/hr ft^2 F	h_{io} = 1500 Btu/hrft^{2o}F
Cp=0.57 BTU/lb F	
$(Cp\mu/k)^{1/3}$=1.6	
Shell Side Coefficient (h_o)	
h_0 = J_h.k/D_e $(C_p\ \mu/k)^{1/3}$.	
= 176 BTU/hr ft^2F	

Clean Overall Coefficient:

$U_c = h_o h_{io}/(h_o+h_{io})$

$= 158$ BTU/hr ft^2 F

$R_d = U_c - U_D/U_C * U_D$ *(5.64)*

$R_d = 0.003$

$U_D = 107$ BTU/hr ft^{2}F

This is very close to the assumed value. Thus, the Design is Satisfactory.

Pressure Drop Calculations:

Shell Side (Aqueous Solution)	Tube Side (Steam)
$\Delta Ps= fG_S^2(N+1)Ds/5.22x10^{10}De\Phi s$ $Re_t=106250$ $f= 0.0013$ $n_b+1= 12*L/B=33$ $\Delta P_s=fG^2d_s(n_b+1)/7.50*10^{12}D_e\text{Ȣ}\Phi s$ $\Delta P_s=0.028$ psi	$Re_t= 38782$ $\text{Ƒ}=0.4137Re^{-0.2585}$ $\text{Ƒ}= 0.02$ $SG=0.0931$ $\Delta P_t=fn_pLG^2/7.50*10^{12}D_i\text{Ȣ}\Phi_t$ $\Delta P_t= 0.03$ psi

5.5 Design of Flash Separator (S-102):

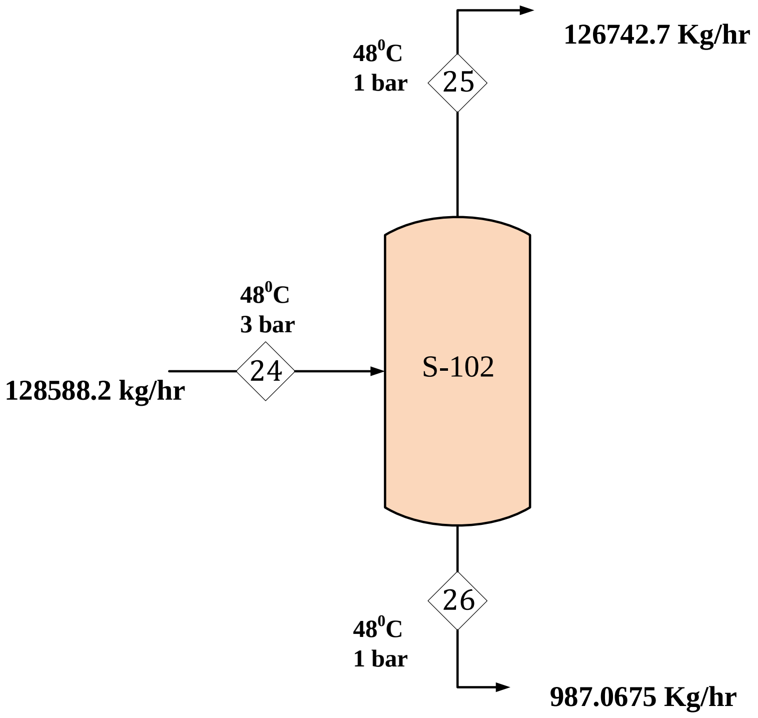


Figure 5.5: Flash Separator (S-102)

A flash separator is a tool used in chemical engineering to divide a vapor-liquid mixture into its component phases. It can function as a two-phase or three-phase separator and can be a vertical or horizontal tank.

The terms flash drum, break pot, knock-out pot, compressor suction drum, suction scrubber or compressor inlet drum, or vent scrubber are also used to describe vapor-liquid separators. It is frequently referred to as a demister when used to remove suspended water droplets from streams of air.

Gravity is used in vapor-liquid separators to induce the less dense fluid (vapor) to be removed from the top of the vessel while the denser fluid (liquid) settles to the bottom of the vessel.

A typical liquid separator will not work in low gravity settings like a space station since gravity is ineffective as a separation mechanism. In this scenario, liquid must be forced toward the outer edge of the chamber for removal using centrifugal force in a rotating centrifugal separator. Gaseous parts go inward, toward the center.

Vertical Flash Separator is selected because:

L/d ratio is less than 5

F = 128119.8 kg/hr

D = 127072.4 kg/hr

W = 1047.408 kg/hr

- **Vapor Velocity**

$V_v = K_v ((\rho_L - \rho_v)/ \rho_v)^{0.5}$ *(5.65)*

$S = (D/W)*(\rho_V/ \rho_L)^{0.5}$

$S = 1.1$

$K_v = 0.12$

$V_v = 0.5$ m/s

- **Vapor Flow Rate**

$V = D/ \rho_V$ *(5.66)*

$V = 0.07$ m^3/s

- **Vapor Cross Sectional Area**

$A_v = V/V_v$ *(5.67)*

$A_v = 0.14$ m^2

- **Diameter**

$d= (4*A_v/\pi)$ *(5.68)*

d = 0.42 m

- **Liquid Flow Rate**

$L = W/ \rho_L$

$L = 3.12*10^{-4}$ m^3/s

Allow a hold up time of 10 min

- **Volume Holdup**

  $V_L = L*10*60$ *(5.69)*

  $V_L = 0.187\ m^3$

- **Vessel Cross Sectional Area**

  $A = (\pi*d^2/4)$ *(5.70)*

  $A = 0.11\ m^2$

- **Vessel Height**

  $L_v = V_L/A$ *(5.71)*

  $L_v = 1.7$ m

  Increase 0.3 m to allow space for positioning

  $L_t = L_v + 0.3$

  $L_t = 2$ m

- **L/d Ratio**

  L/d = 4.7 m

SPECIFICATION SHEET	
Identification	
Item	Flash Tank
Item no.	S-102
No. required	1
Operation	Continuous
Type	Vertical Flash Separator
Function	
Separation of Ethylene	
Vapor Velocity	0.5 m/s
Vessel Cross Sectional Area	0.11 m^2
Diameter	0.42 m
Liquid Holdup	0.187 m^3
Vessel Height	2 m
L/D Ratio	4.7

5.6 Design of Spiral Tube Heat Exchanger:

The design of spiral tube heat exchangers consists of many tubes arranged in multiple layers of helical coils, around a center pipe. This tube bundle is fitted in a cylindrical pressure vessel. The fluid on the shell as well as tube side flows in opposite directions, making the equipment a true countercurrent heat exchanger.

Shell side

The high turbulence flow is created by the patented design of the tube coils. The variation of the fluid velocity between the tubes creates a pulse-surge collision flow regime increasing subsequently the heat exchange coefficient outside the tubes.

The possibility of fouling is greatly limited by the non-baffle design, the turbulence of the fluid and the very low surface roughness of the tubes.

Tube side

The helix-pattern flow in the tubes creates, thanks to the centrifugal forces, a secondary flow consisting of a pair of vortices enhancing the coefficient of heat transfer at the peripheral of the tubes.

Spiral tubes are coiled layer by layer in opposite direction to have a homogeneous heat transfer all along the exchanger.

- **Selection of Appropriate Type**

Spiral tube heat exchanger is selected so here is the reason behind selection:

- ✓ It is effective for slurries, sludges and viscous liquids
- ✓ Highly resistant to thermal and hydraulic shock
- ✓ Suitable for fluids that tend to cause fouling because of
  - Continuous Curving
  - High Turbulence
  - High shear stress
- ✓ Self-cleaning ability

Hot Fluid (Steam)

$T_1 = 152$ °C $T_2 = 152$ °C

- **Cold Fluid (Process Slurry)**

$t_1 = 25$ °C $t_2 = 140$ °C

OD of tube = d_o = 19 mm

Thickness of tube = 2.7 mm

ID of tube = d_i = 13.5 mm

Number of Spiral Coils = n = 3

Number of Turns = N = 4

Spiral Pitch = P = 25 mm

ID of Spiral = D_i = 114.02 mm

OD of Straight Tube = d_{ho} = 27 mm

ID of Straight Tube = d_{hi} = 25 mm

Shell Inside Dia

$D_{is} = 2*(R_o + Rh_o)$ *(5.72)*

D_{is} = 261 mm

Length of Shell

$L_s = (R_o^2 – R_i^2)/a$ *(5.73)*

$a = P\pi/4$

L_s = 332 mm

Curvature Ratio

ȣ = d_i/D_i

ȣ = 0.087

Developed length of spiral

$L_o = 3.14*n\,(R_o + R_i)$ *(5.74)*

L_o = 1639 mm

Total Length

$L_t = N_t*L_o$

L_t = 6556 mm

Area

$A_s = n*\pi*d_o*L_o$ *(5.75)*

$A_s = 0.185\ m^2$

LMTD

$LMTD = (\Delta t_1 – \Delta t_2)/\ln(\Delta t_1 / \Delta t_2)$

LMTD = 49 °C

LMTD = 322 K

Steam Heat Transfer Coefficient

$h_o = 8520\ W/m^2K$

Slurry Heat Transfer Coefficient

Mass Velocity

$G = m/A_s$

$G = 940\ kg/sm^2$

Reynolds Number

$Re = G*d_i / \mu$ *(5.75)*

Re = 3666

Equivalent Diameter:

$D_e = R_e\ (r_i/R_i)^{1/2}$

$D_e = 1261$ mm

Nusselt Number:

$N_u = 0.836 * D_e^{0.5} * Pr^{0.1}$

$N_u = 61.2$

$h_i = 278\ W/m^2K$

Clean Overall Heat Transfer Coefficient:

$U_c = (h_i * h_o)/(h_i + h_o)$ *(5.76)*

$U_c = 269\ W/m^2K$

Design Overall Heat Transfer Coefficient :

$1/U_d = (1/U_c) + R_d$

$U_d = 148\ W/m^2K$

Area:

$A = A_{so} + A_{si}$

$A = [n*\pi*d_o*L_o + n_1*\pi*dh_o*L_t] + [n*\pi*d_i*L_o + n_1*\pi*dh_i*L_t]$ *(5.77)*

$A = 2.5\ m^2$

For Steam:

$\Delta P = [(0.0789*(L/\rho_h))*(m_h/H*dh_o)*((1.3*(\mu h)^{0.33})/(dh_o+0.032))*(H/m_h)^{0.33}+1.5+16/L]$

$\Delta P = 0.0154$ bar

$\Delta P = 0.22$ psi

For Slurry:

$\Delta P = [(0.0789*(L/\rho_c))*(m_c/H*dh_i)*((1.3*(\mu c)^{0.33})/(dh_i+0.032))*(H/m_c)^{0.33}+1.5+16/L]$

$\Delta P = 0.382$ bar

$\Delta P = 5.5$ psi

SPECIFICATION SHEET			
Identification			
Item		Heat Exchanger	
Item no.		H-101	
No. required		1	
Operation		Continuous	
Type		Spiral Tube	
Function			
Pre-heating of corn stover slurry			
Heat Duty		$9.9*10^4$ MJ/hr	
Heat Transfer Area		2.5 m^2	
Shell Side (Steam)		**Tube Side (Slurry)**	
Steam Temperature	2 m	t_1	25℃
Steam Pressure	1.36 m	t_2	140℃
h_o	8520 W/m^2K	hi	278 W/m^2K
Pressure Drop	0.22 psi	Pressure Drop	5.5 psi

CHAPTER # 06

MECHNICAL DESIGN

6.1 Mechanical Design of Pre-Hydrolysis Reactor (R-101):

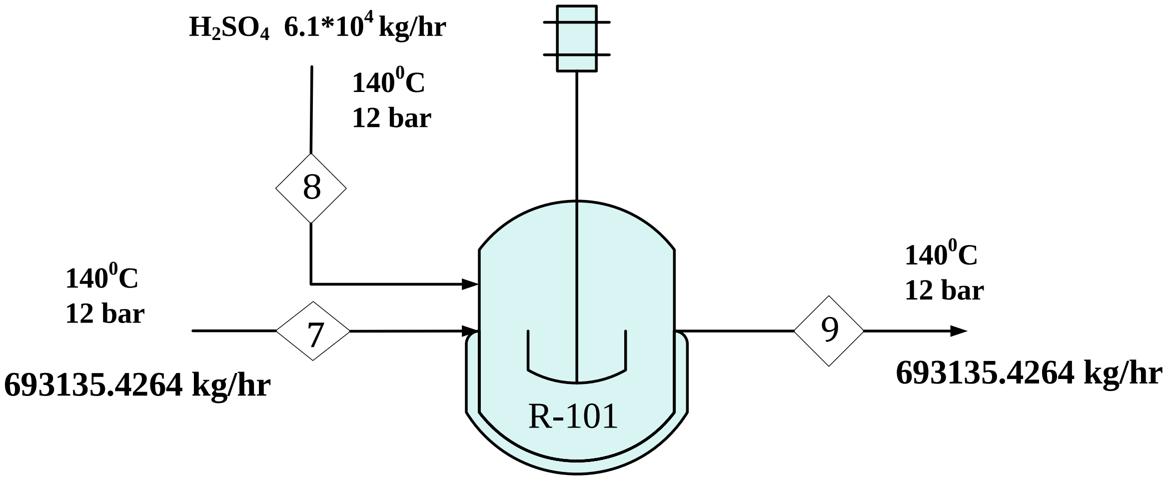


Figure 6.1: Pre-Hydrolysis Reactor (R-101)

Material Selection:

Austenitic Stainless-Steel Grade 304 is selected because of :

- ✓ High strength
- ✓ Resistant to scaling at high temperatures
- ✓ Resistant to corrosion
- ✓ Used for high pressure reactors

Chemical Composition:

Cr = 24-26% Ni = 19-23% C = 0.25%

Baffle Spacing:

4 radial baffles are used

$B = \pi D_t/4$ *(6.1)*

B = 1.36 m

$J = D_t/12$ *(6.2)*

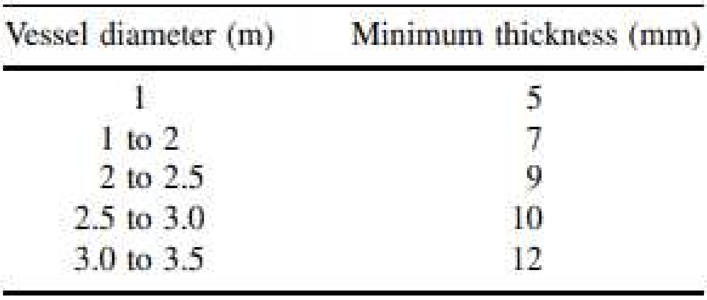
J = 0.145 m

Distance from bottom = $D_t/2$

E = 0.87 m

Minimum Practical Wall Thickness

As vessel diameter is between 1-2 mm so minimum practical thickness is 7mm



Vessel diameter (m)	Minimum thickness (mm)
1	5
1 to 2	7
2 to 2.5	9
2.5 to 3.0	10
3.0 to 3.5	12

Wall Thickness

D_i = 1.744 m

L = 2 m

P_i = 12 bar or 1.32 N/mm^2

Corrosion allowance = 2 mm

Stress Factor = S = 135 N/mm^2

Joint Efficiency = E = 1

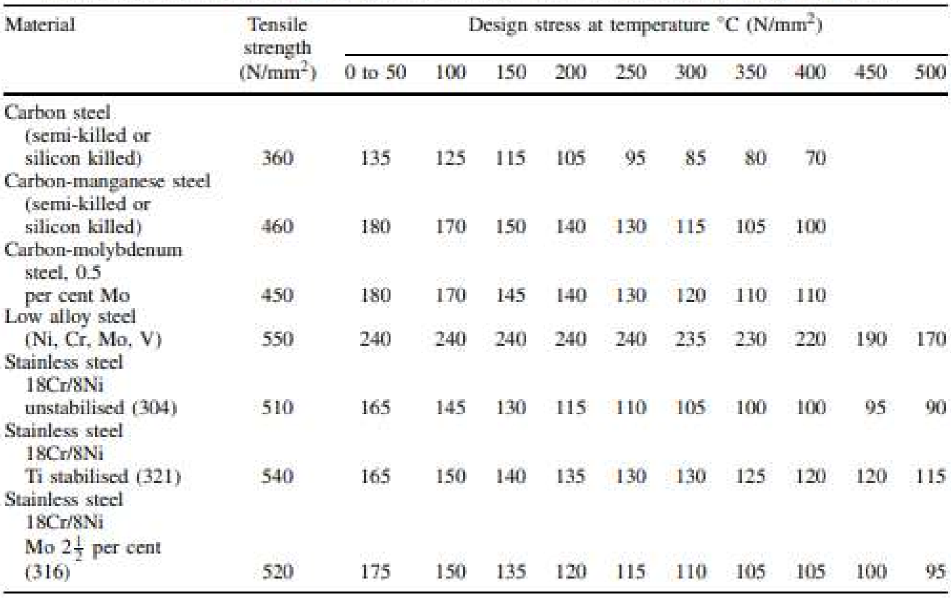


Table 13.2. Typical design stresses for plate
(The appropriate material standards should be consulted for particular grades and plate thicknesses)

Material	Tensile strength (N/mm^2)	Design stress at temperature °C (N/mm^2)									
		0 to 50	100	150	200	250	300	350	400	450	500
Carbon steel (semi-killed or silicon killed)	360	135	125	115	105	95	85	80	70		
Carbon-manganese steel (semi-killed or silicon killed)	460	180	170	150	140	130	115	105	100		
Carbon-molybdenum steel, 0.5 per cent Mo	450	180	170	145	140	130	120	110	110		
Low alloy steel (Ni, Cr, Mo, V)	550	240	240	240	240	240	235	230	220	190	170
Stainless steel 18Cr/8Ni unstabilised (304)	510	165	145	130	115	110	105	100	100	95	90
Stainless steel 18Cr/8Ni Ti stabilised (321)	540	165	150	140	135	130	130	125	120	120	115
Stainless steel 18Cr/8Ni Mo $2\frac{1}{2}$ per cent (316)	520	175	150	135	120	115	110	105	105	100	95

$t = (D_i*P_i)/(2SE-P_i)$ *(6.3)*

t = 8.24 mm

t = 10.24 mm

Outer Diameter of Shell

$D_o = D_i + 2t$ *(6.4)*

$D_o = 1.764$ m

Ellipsoidal Heads

$t = (D_i*P_i)/(2SE-0.2P_i)$ *(6.5)*

$t = 8.21$ mm

$t = 10.2$ mm

Vessel Support

For reactors we use bracket supports

Weight Loads

$W_v = 240*C_v*D_i*(L+0.8(D_i)*t$ *(6.6)*

$W_v = 15.65$ N

Wind Loads

$F = P_w * D_o$ *(6.7)*

$F = 1816.92$ N/m

Longitudinal Stress

$\sigma_h = (P_i*D_i)/2t$ *(6.8)*

$\sigma_h = 112$ N/mm^2

Circumferential Stress

$\sigma_L = (P_i*D_i)/4t$

$\sigma_L = 56.29$ N/mm^2

Dead Weight Stress

$\sigma_L = W/p*(D_i+t)*t$ *(6.9)*

$\sigma_L = 0.27$ N/mm^2

Radial Stress

$\sigma_d = P_i/2$

$\sigma_d = 0.66$ N/mm^2

Bending Moment

$M_x = F*H$

$M_x = 1816.92$ N/m

Bending Stress

$\sigma_b = (M_x/Iv)*((D_i/2)+t)$ *(6.10)*

$\sigma_b = 0.15$ N/mm^2

6.2 Mechanical Design of Saccharification Reactor (R-103):

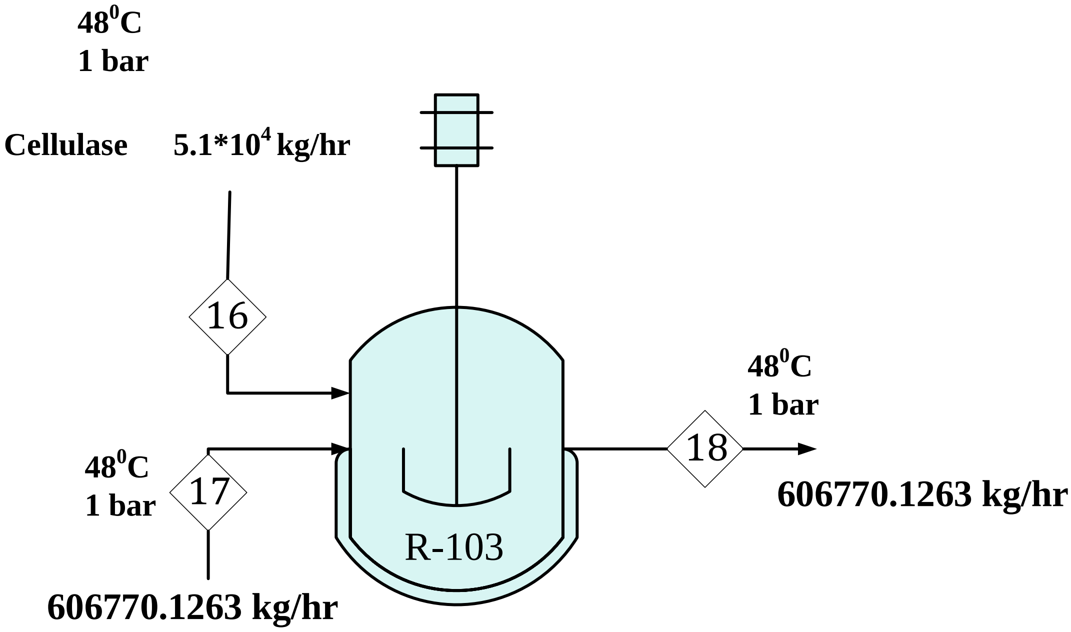


Figure 6.2: Saccharification Reactor (R-103)

Material Selection

- ✓ Austenitic Stainless-Steel Grade 304 is selected because of:
- ✓ High strength
- ✓ Resistant to scaling at high temperatures
- ✓ Resistant to corrosion
- ✓ Used for high pressure reactors

Chemical Composition

Cr = 24-26% Ni = 19-23% C = 0.25%

Baffle Spacing

4 radial baffles are used

B = pDt/4

B = 0.785 m

J = Dt/12

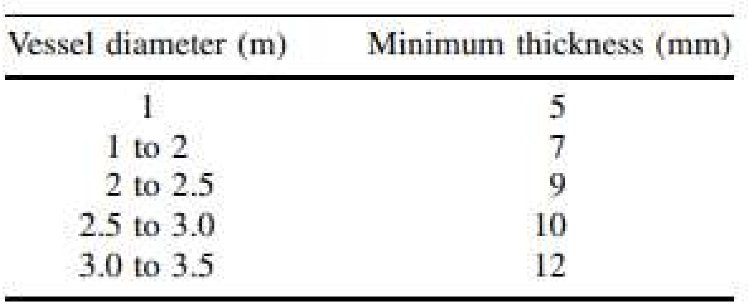
J = 0.08 m

Distance from bottom = Dt/2

E = 0.5 m

Minimum Practical Wall Thickness

As vessel diameter is 1 mm so minimum practical thickness is 5 mm



Vessel diameter (m)	Minimum thickness (mm)
1	5
1 to 2	7
2 to 2.5	9
2.5 to 3.0	10
3.0 to 3.5	12

Wall Thickness

$D_i = 1$ m

$L = 1.5$ m

$P_i = 1$ bar or $0.11\ N/mm^2$

Corrosion allowance = 2 mm

Stress Factor = $S = 135\ N/mm^2$

Joint Efficiency = $E = 1$

$t = (D_i*P_i)/(2SE-P_i)$ *(6.11)*

$t = 0.40$ mm

$t = 2.4$ mm

Outer Diameter of Shell

$D_o = D_i + 2t$

$D_o = 1$ m

Ellipsoidal Heads

$t = (D_i*P_i)/(2SE-0.2P_i)$ *(6.12)*

$t = 0.40$ mm

$t = 2.4$ mm

Vessel Support

For reactors we use bracket supports

Weight Loads

$W_v = 240*C_v*D_i*(L+0.8(D_i)*t$ *(6.13)*

$W_v = 1.43$ N

Wind Loads

$F = P_w * D_o$

$F = 1030$ N/m

Longitudinal Stress

$\sigma_h = (P_i*D_i)/2t$ *(6.14)*

$\sigma_h = 22.9\ N/mm^2$

Circumferential Stress

$\sigma_L = (P_i*D_i)/4t$ *(6.15)*

$\sigma_L = 11.4\ N/mm^2$

Dead Weight Stress

$\sigma_L = W/p*(D_i+t)*t$ *(6.16)*

$\sigma_L = 0.189\ N/mm^2$

Radial Stress

$\sigma_d = P_i/2$

$\sigma_d = 0.055\ N/mm^2$

Bending Moment

$M_x = F*H$

$M_x = 515\ N/m$

Bending Stress

$\sigma_b = (M_x/I_v)*((D_i/2)+t)$ *(6.17)*

$\sigma_b = 0.059\ N/mm^2$

6.3 Mechanical Design of Fermentation Reactor (R-104):

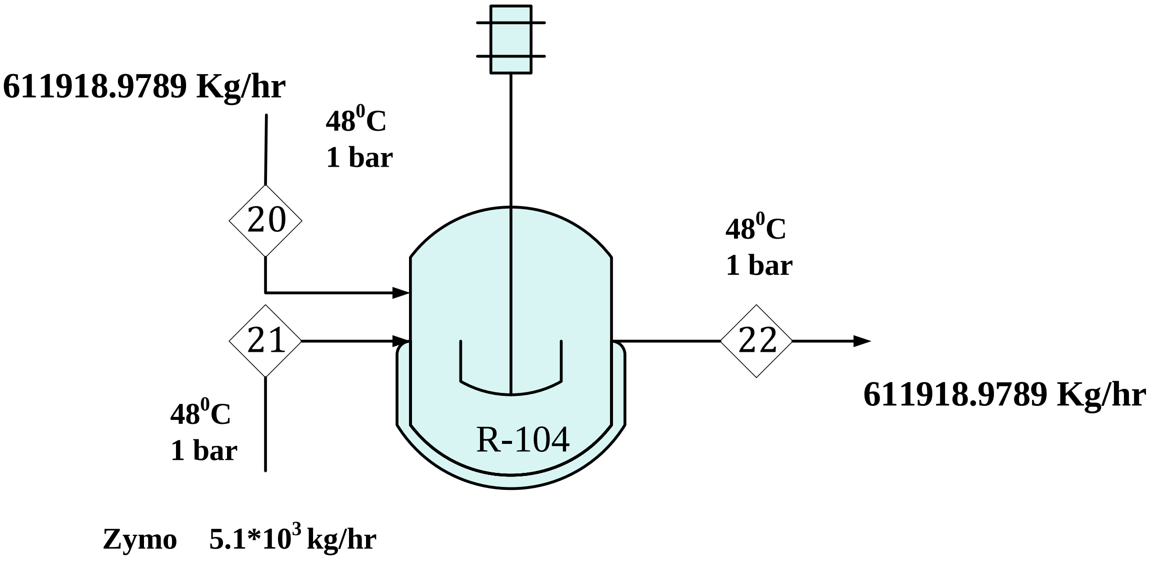


Figure 6.3: Fermentation Reactor (R-104)

Material Selection

- ✓ Austenitic Stainless-Steel Grade 304 is selected because of :
- ✓ High strength
- ✓ Resistant to scaling at high temperatures
- ✓ Resistant to corrosion
- ✓ Used for high pressure reactors

Chemical Composition

Cr = 24-26% Ni = 19-23% C = 0.25%

Baffle Spacing

4 radial baffles are used

B = pDt/4

B = 1.06 m

J = Dt/12

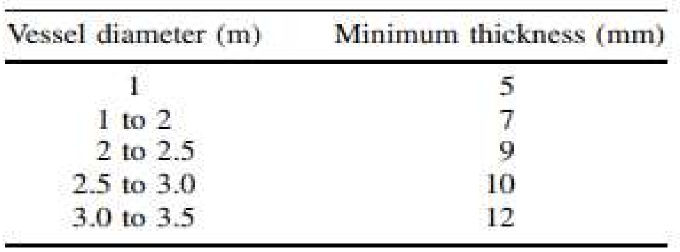
J = 0.11m

Distance from bottom = Dt/2

E = 0.68 m

Minimum Practical Wall Thickness

As vessel diameter is between 1-2 mm so minimum practical thickness is 5 mm



Vessel diameter (m)	Minimum thickness (mm)
1	5
1 to 2	7
2 to 2.5	9
2.5 to 3.0	10
3.0 to 3.5	12

Wall Thickness

$D_i = 1.36$ m

L = 2 m

$P_i = 1$ bar or 0.11 N/mm^2

Corrosion allowance = 2 mm

Stress Factor = S = 135 N/mm^2

Joint Efficiency = E = 1

$t = (D_i * P_i)/(2SE - P_i)$ *(6.18)*

t = 0.55 mm

t = 2.55 mm

Outer Diameter of Shell

$D_o = D_i + 2t$

$D_o = 1.36$ m

Ellipsoidal Heads

$t = (D_i*P_i)/(2SE-0.2P_i)$ *(6.19)*

$t = 0.55$ mm

$t = 2.55$ mm

Vessel Support

Weight Loads

$W_v = 240*C_v*D_i*(L+0.8(D_i)*t$ *(6.20)*

$W_v = 2.77$ N

Wind Loads

$F = P_w * D_o$

$F = 1400.8$ N/m

Longitudinal Stress

$\sigma_h = (P_i*D_i)/2t$ *(6.21)*

$\sigma_h = 29.3$ N/mm^2

Circumferential Stress

$\sigma_L = 14.6$ N/mm^2

Dead Weight Stress

$\sigma_L = W/p*(D_i+t)*t$ *(6.22)*

$\sigma_L = 0.25$ N/mm^2

Radial Stress

$\sigma_d = P_i/2$

$\sigma_d = 0.055$ N/mm^2

Bending Moment

$M_x = F*H$

$M_x = 1400.8$ N/m

Bending Stress

$\sigma_b = (M_x/I_v)*((D_i/2)+t)$ *(6.23)*

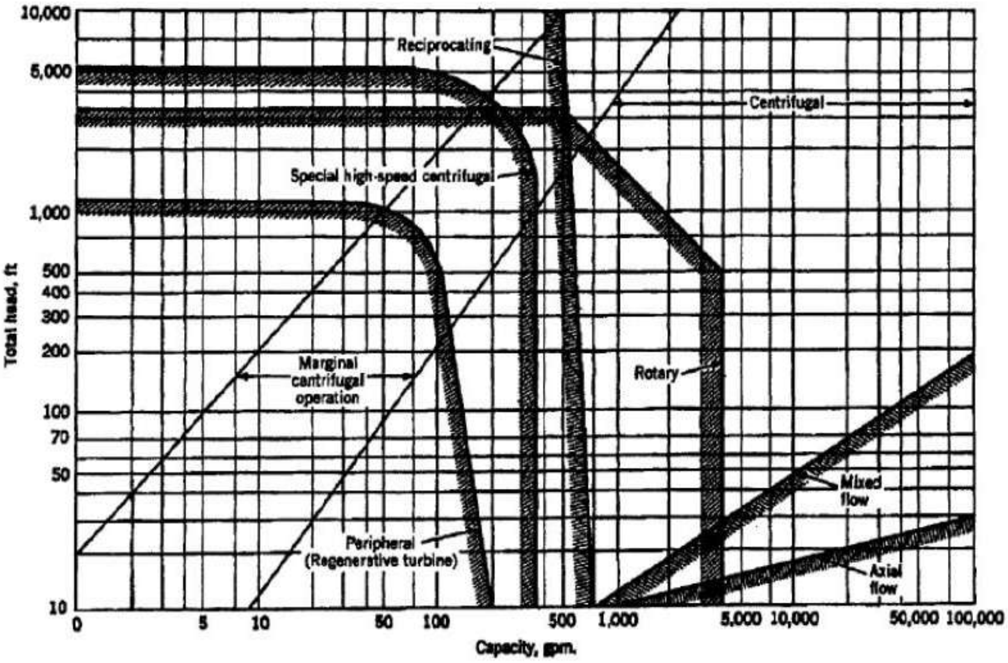
$\sigma_b = 0.059$ N/mm^2

CHAPTER # 07
POWER CALCULATIONS

7.1 Introduction:

The liquid's mechanical energy is increased by centrifugal action in centrifugal pumps. In a correctly working pump, the liquid enters by a suction connection that is centered on the excess of the impeller, a high-speed rotating element; the distance between the vanes is totally filled with a liquid that is flowing without cavitation. The fluid that exits the impeller's outer perimeter is gathered in a spiral casing known as the volute and is discharged from the pumps through a tangential discharge connection. The liquid's velocity head from the impeller is changed into pressure head in the volute. Single-stage or multistage, propeller, mixed-flow, and peripheral centrifugal pumps are all possible.

We have selected centrifugal pumps for a process because of the following advantages:



- They are easy in operation and cheep
- Fluid is transferred at uniform pressure without shocks or pulsation
- No valves involved in pump operation
- They operate at high speed (up to 4000 rpm) so they can be coupled linerly to an electric motor.
- Without altering the pump, the discharge pipe can be totally or partially closed off.
- Compared to other kinds of pumps, this one requires less maintenance.

7.2 Design Calculations:

Design steps for pump sizing:

- Define the flow system, i-e locate points 1 and 2. The pressures P_1 and P_2 will be known at these points.

- ✓ Position the process equipment follows to the rules-of-thumbs.
- ✓ Estimate Z_1 and Z_2.
- ✓ Estimate Frictional pressure losses E_D and E_S.
- ✓ Determine Pump Work.
- ✓ Find out Pump shaft horsepower & estimate its Efficiency.

Step 1: Pressure at suction and Discharge of the pump:

Inlet pressure= $P_1 = 1.3$ bar
Outlet pressure= $P_2 = 8.2$ bar (required)

Step 2: Rule of thumbs for locating the process equipment:

Table 7.1: Rule of thumb for locating the process equipment

Process equipment	Location above ground level, ft
Pumps	0
Condensers	20
Reflux drums	10
Phase Separators	3 to 5
Skirt height for Columns	3 to 6
Heat Exchangers	1 to 4

As the discharge of our pump is at the top of absorber so we take skirt height within the range given above.
Height of skirt = 4ft = 1.21m

Step 3: Estimate Z_1 and Z_2 that is height at suction and discharge:

Suction height = Z_1 = 0ft = 0m (As by rule of thumb pump will always consider at ground level) **Discharge height** = Z_2 = height of equipment + skirt height = 10+1.21 = 11.21m

Step 4: Estimate Frictional pressure losses E_D and E_S.

Flow system components	Pressure drop (bar)
Pipeline	0.35
Control valve	0.70
Interchanger	0.35
Air cooler	0.60
Surge vessel	Small

So $E_D = 0.35 + 0.70 + 0.35 = 1.40$ bar

Step 5: Calculate Pump Work:
Density = 1024 kg/m^3

$$W = \frac{g}{g_c}(z_1 - z_2) + \frac{p_1 - p_2}{\rho} - (E_s + E_D)$$

$$W = 9.8(0 - 11.21) + \frac{(1.3 - 8.82) \times 10^5}{1024} - \frac{(0.35 + 0.35 + 0.70) \times 10^5}{1024}$$

$$W = (109.85)\text{-}(734.37)\text{-}(136.71) = -980.93\ \frac{N-m}{kg}$$

The negative sign shows that the work is done on the system.

Step 6: Pump Power Calculations:

Flow rate = m = 4050Kg/hr = 1.125kg/s

Pump efficiency = η =0.45

$$P_P = \frac{m \times W}{\eta}$$

$$P_p = \frac{1.125 \times 980.93}{0.45}$$

P_p = 2452.32 Watt

P_p = 3.26 hp

Step 7: Calculate electric-motor horsepower & estimate its Efficiency:

An induction motor with an efficiency of motor is selected to be 86%.

$$P_E = \frac{P_P}{\eta}$$

$$P_E = \frac{3.26}{0.86}$$

$$\mathbf{P_E = 3.79hp}$$

Safety factor: A safety factor of 10% is taken

Safety factor = 1.1

So P_E = 3.79 × 1.1 = 4.16 hp

Select a standard electric-motor horsepower

The standard motor selected is of 5 hp.

Net Positive Suction Head:

We know that

$$NPSH = \frac{1}{g}\left(\frac{P_a - P_V}{\rho} - h_{fs}\right) - Z_a$$

Where,

H_{fs} = friction in suction line = 0bar

Z_a = 0

P_a= 1.3atm = 130000 Pa **Vapor pressure:**

Process stream at 116 ℃ = 94392.2 Pa

So, by putting the values, we get

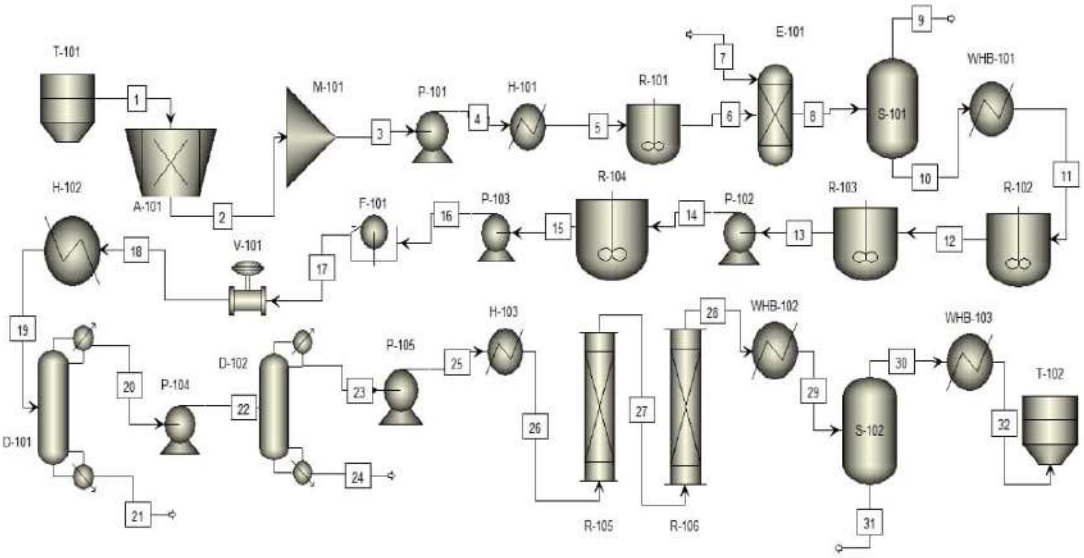
$$NPSH = \frac{1}{9.8}\left(\frac{130000 - 94392.2}{1024} - 0\right) - 0 = 3.24\ m$$

Specification Sheet P-101	
Total Mass flow rate	405056 kg/hr
Discharge height = Z_2	12 m
Work required by pump	--980.93 N-m/kg
Pump Power	3.26 hp
Electric-motor Power	5 hp
NPSH	3.24m

CHAPTER 8

PROCESS SIMULATION

8.1 Process Flow Diagram:



8.2 Introduction to ASPEN HYSYS:

Aspen HYSYS (or simply HYSYS) is a chemical process simulator currently developed by Aspen Tech used to mathematically model chemical processes, from unit operations to full chemical plants and refineries. HYSYS is able to perform many of the core calculations of chemical engineering, including those concerned with mass balance, energy balance, vapor-liquid equilibrium, heat transfer, mass transfer, chemical kinetics, fractionation, and pressure drop. HYSYS is used extensively in industry and academia for steady-state and dynamic simulation, process design, performance modeling, and optimization

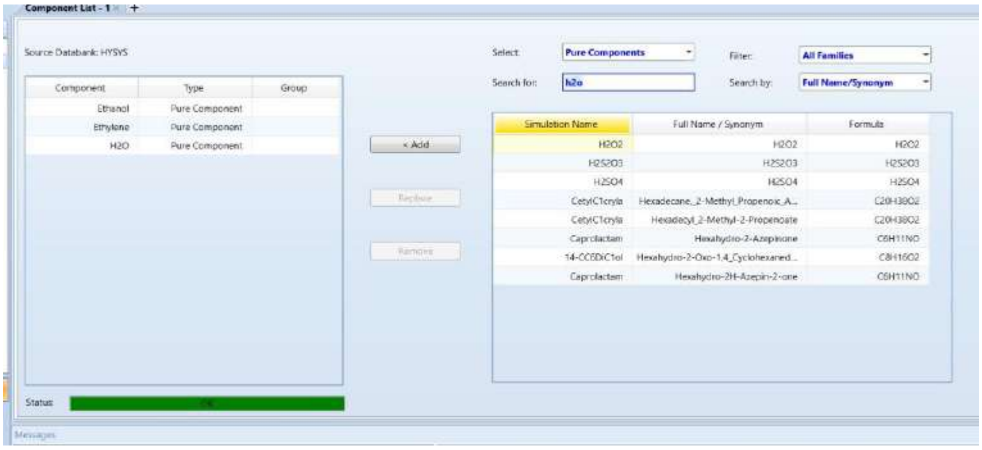
8.3 Introduction to ASPEN Plus:

ASPEN PLUS allows you to create your own process model, starting with the flow sheet, then specifying the chemical components and operating conditions. ASPEN PLUS will take all of your specifications and, with a click of the mouse button, simulate the model. The process simulation is the action that executes all necessary calculations needed to solve the outcome of the system, hence predicting its behavior. When the calculations are complete, ASPEN PLUS lists the results, stream by stream and unit by unit, so you can observe what happened to the chemical species of your process model.

8.4 Simulation of Dehydration Reactor:

Selection and addition of the components of the process.

- ✓ Ethanol
- ✓ Ethylene
- ✓ Water



Source Databank: HYSYS

Select: Pure Components

Filter: All Families

Search for: h2o

Search by: Full Name/Synonym

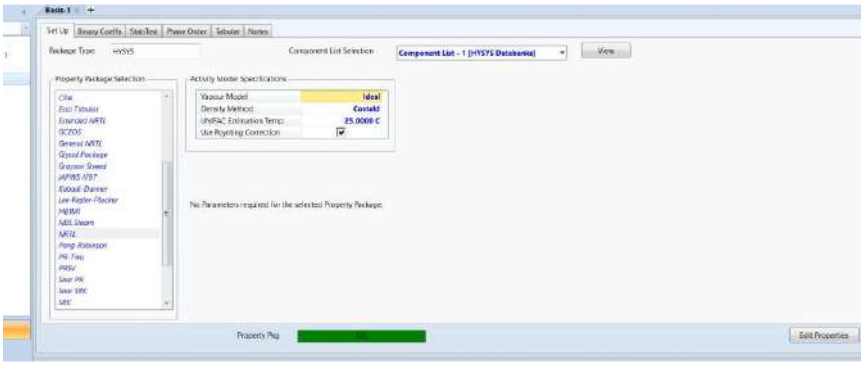
Component	Type	Group
Ethanol	Pure Component	
Ethylene	Pure Component	
H2O	Pure Component	

Simulation Name	Full Name / Synonym	Formula
H2O2	H2O2	H2O2
H2S2O3	H2S2O3	H2S2O3
H2SO4	H2SO4	H2SO4
CetylC1cryla	Hexadecane_2-Methyl_Propenoic_A...	C20H38O2
CetylC1cryla	Hexadecyl_2-Methyl-2-Propenoate	C20H38O2
Caprolactam	Hexahydro-2-Azepinone	C6H11NO
14-CC6DiC1ol	Hexahydro-2-Oxo-1,4_Cyclohexaned...	C8H16O2
Caprolactam	Hexahydro-2H-Azepin-2-one	C6H11NO

Status: OK

Selection of the proper fluid package. In our case we have selected NTRL as fluid package.

Establish reactions. Click Add after going to the Reactions folder. This will produce Set-1



Package Type: HYSYS

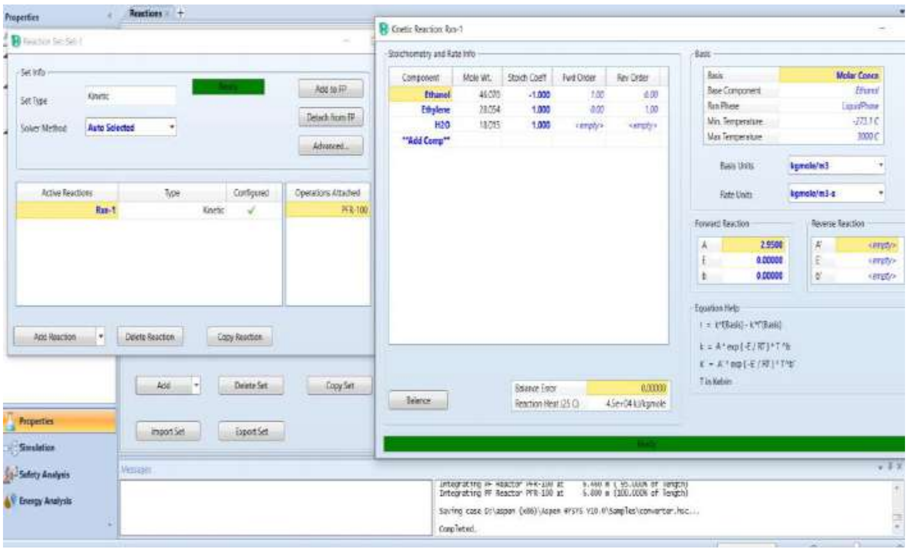
Component List Selection: Component List - 1 [HYSYS Databanks]

Vapour Model	Ideal
Density Method	Costald
UNIFAC Estimation Temp	25.0000 C
Use Poynting Correction	☑

No Parameters required for the selected Property Package.

Property Pkg: OK

a new reaction set. Select Add Reaction and Hysys, Conversion in Set 1. Rxn-1 will be a novel reaction that results from this. To open the Rxn-1 window, double-click on the Rxn-1 icon. Enter the information below. When finished, close this window. We now need to connect the reaction set to the fluid package in Set-1. To add Basis-1 to your FP, click the Add to FP button. Now, the response system ought to be prepared.



Set Type: Kinetic

Solver Method: Auto Selected

Active Reactions	Type	Configured	Operations Attached
Rxn-1	Kinetic	✓	PFR-100

Component	Mole Wt.	Stoich Coeff	Fwd Order	Rev Order
Ethanol	46.070	-1.000	1.00	0.00
Ethylene	28.054	1.000	0.00	1.00
H2O	18.015	1.000	<empty>	<empty>
Add Comp				

Basis	Molar Concn
Base Component	Ethanol
Rxn Phase	LiquidPhase
Min. Temperature	-273.1 C
Max Temperature	3000 C

Basis Units: kgmole/m3

Rate Units: kgmole/m3-s

Forward Reaction	
A	2.9500
E	0.00000
b	0.00000

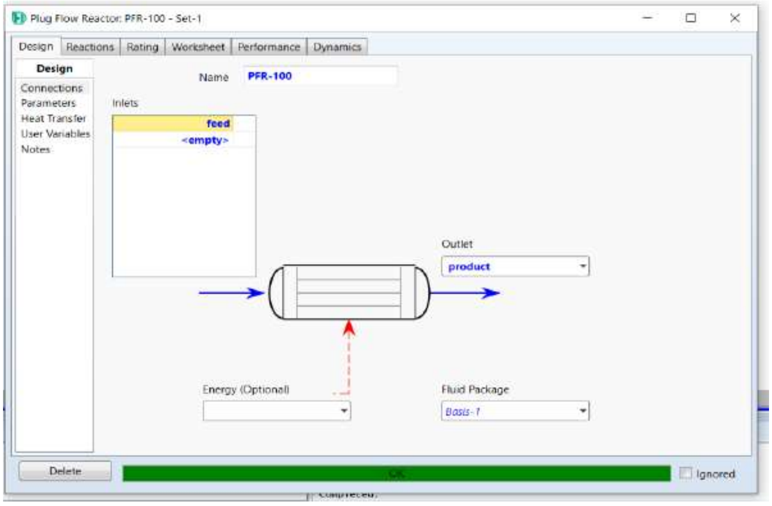
Reverse Reaction	
A'	<empty>
E'	<empty>
b'	<empty>

Balance Error: 0.00000

Reaction Heat (25 C): 4.5e+04 kJ/kgmole

```
Integrating PF Reactor PFR-100 at     6.000 m (100.000% of length)
Saving case D:\aspen (x86)\Aspen HYSYS V10.0\Samples\convertor.hsc...
Completed.
```

Go to the simulation environment and select a PFR-100 and add to worksheet. By double clicking on the PFR a window will open.



Name: PFR-100

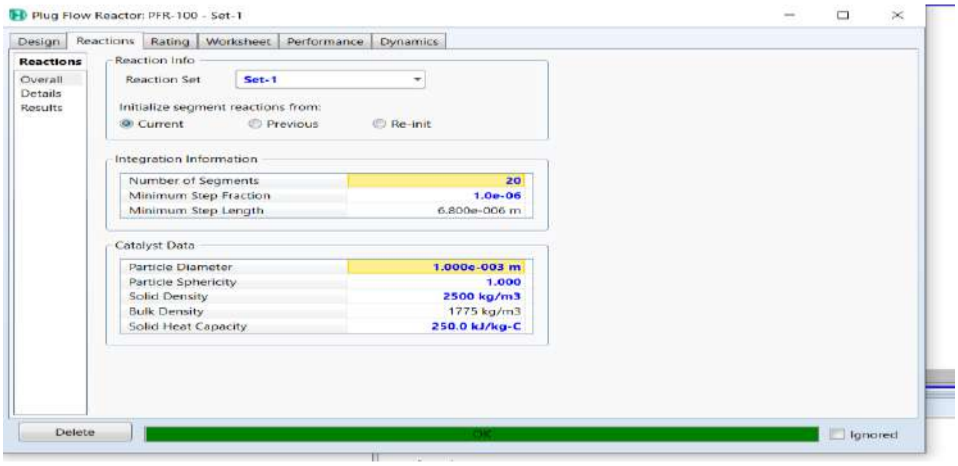
Inlets: feed, <empty>

Outlet: product

Energy (Optional):

Fluid Package: Basis-1

Go to the reactions tab in the already opened window and insert the required data.



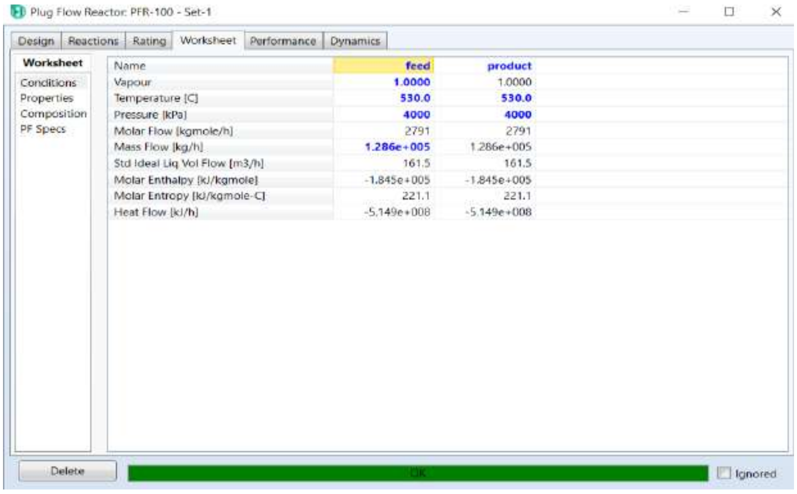
Reaction Set: Set-1

Initialize segment reactions from: ● Current ○ Previous ○ Re-init

Integration Information	
Number of Segments	20
Minimum Step Fraction	1.0e-06
Minimum Step Length	6.800e-006 m

Catalyst Data	
Particle Diameter	1.000e-003 m
Particle Sphericity	1.000
Solid Density	2500 kg/m3
Bulk Density	1775 kg/m3
Solid Heat Capacity	250.0 kJ/kg-C

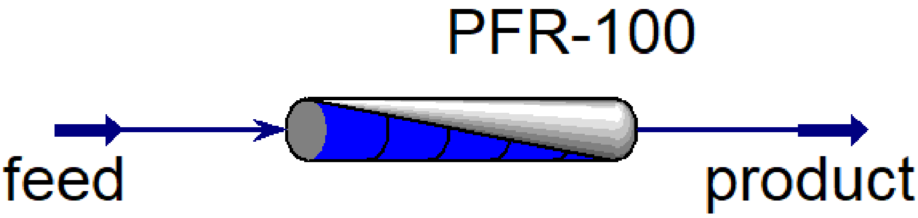
Click on the worksheet tab and enter the data of the temperature and pressure of both input and output streams.



Name	feed	product
Vapour	1.0000	1.0000
Temperature [C]	530.0	530.0
Pressure [kPa]	4000	4000
Molar Flow [kgmole/h]	2791	2791
Mass Flow [kg/h]	1.286e+005	1.286e+005
Std Ideal Liq Vol Flow [m3/h]	161.5	161.5
Molar Enthalpy [kJ/kgmole]	-1.845e+005	-1.845e+005
Molar Entropy [kJ/kgmole-C]	221.1	221.1
Heat Flow [kJ/h]	-5.149e+008	-5.149e+008

Go to the ratings tab in the already opened window.

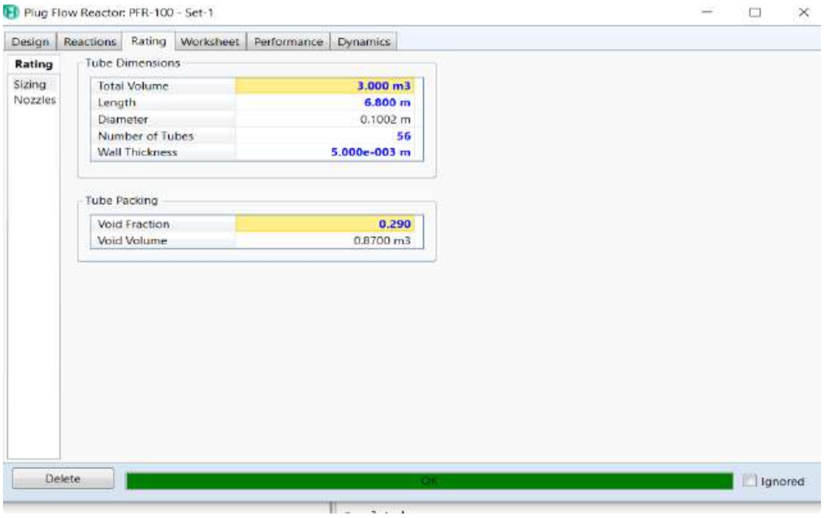
The PFR is simulated and should look like this now.



8.5 Simulation of Fermenter:

Selection and addition of the components of the process

- ✓ Glucose



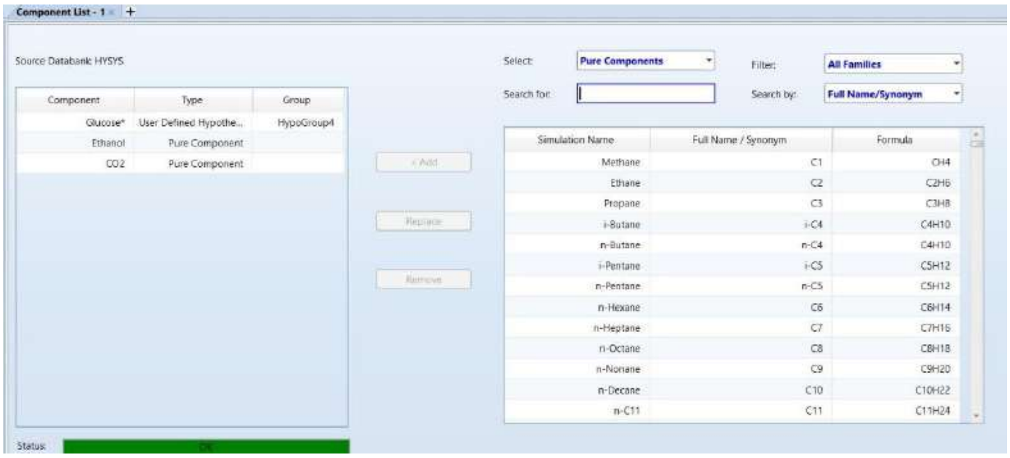
Tube Dimensions

Total Volume	3.000 m3
Length	6.800 m
Diameter	0.1002 m
Number of Tubes	56
Wall Thickness	5.000e-003 m

Tube Packing

Void Fraction	0.290
Void Volume	0.8700 m3

- ✓ Ethanol
- ✓ CO_2



Source Databank: HYSYS

Component	Type	Group
Glucose*	User Defined Hypothe...	HypoGroup4
Ethanol	Pure Component	
CO2	Pure Component	

Select: Pure Components

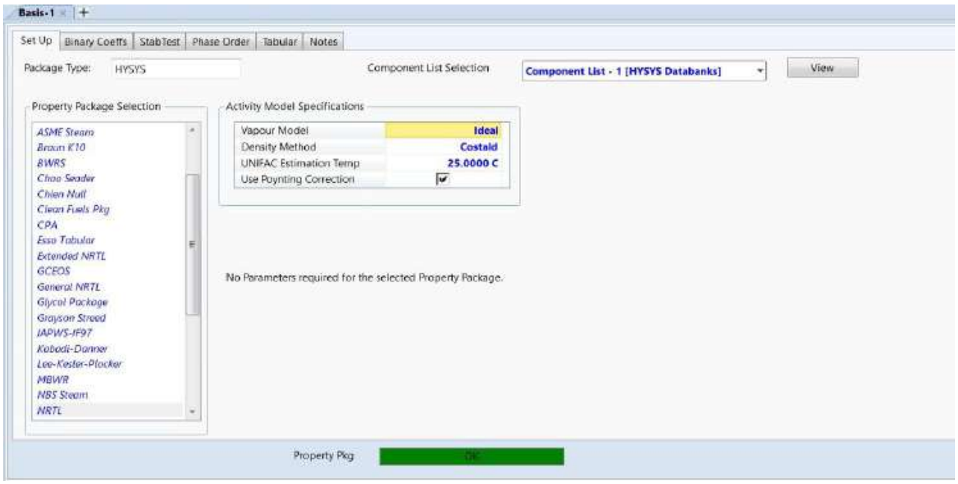
Filter: All Families

Search for:

Search by: Full Name/Synonym

Simulation Name	Full Name / Synonym	Formula
Methane	C1	CH4
Ethane	C2	C2H6
Propane	C3	C3H8
i-Butane	i-C4	C4H10
n-Butane	n-C4	C4H10
i-Pentane	i-C5	C5H12
n-Pentane	n-C5	C5H12
n-Hexane	C6	C6H14
n-Heptane	C7	C7H16
n-Octane	C8	C8H18
n-Nonane	C9	C9H20
n-Decane	C10	C10H22
n-C11	C11	C11H24

Selection of the proper fluid package. In our case we have selected NTRL as fluid package.



Package Type: HYSYS

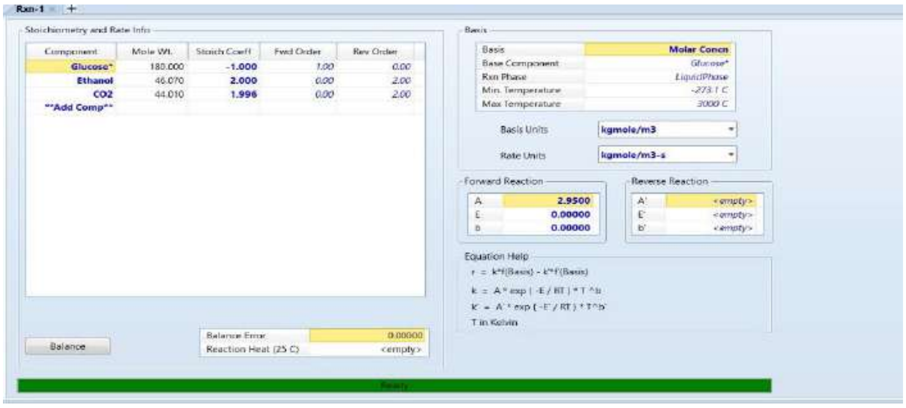
Component List Selection: Component List - 1 [HYSYS Databanks]

Vapour Model	Ideal
Density Method	Costald
UNIFAC Estimation Temp	25.0000 C
Use Poynting Correction	☑

No Parameters required for the selected Property Package.

Property Pkg: OK

Establish reactions. Click Add after going to the Reactions folder. This will produce Set-1, a new reaction set. Select Add Reaction and Hysys, Conversion in Set 1. Rxn-1 will be a novel reaction that results from this. To open the Rxn-1 window, double-click on the Rxn-1 icon. Enter the information below. When finished, close this window. We now need to connect the reaction set to the fluid package in Set-1. To add Basis-1 to your FP, click the Add to FP button. The reaction set should now be ready.



Component	Mole Wt.	Stoich Coeff	Fwd Order	Rev Order
Glucose*	180.000	-1.000	1.00	0.00
Ethanol	46.070	2.000	0.00	2.00
CO2	44.010	1.996	0.00	2.00
Add Comp				

Basis	Molar Concn
Base Component	Glucose*
Rxn Phase	LiquidPhase
Min. Temperature	-273.1 C
Max Temperature	3000 C

Basis Units: kgmole/m3

Rate Units: kgmole/m3-s

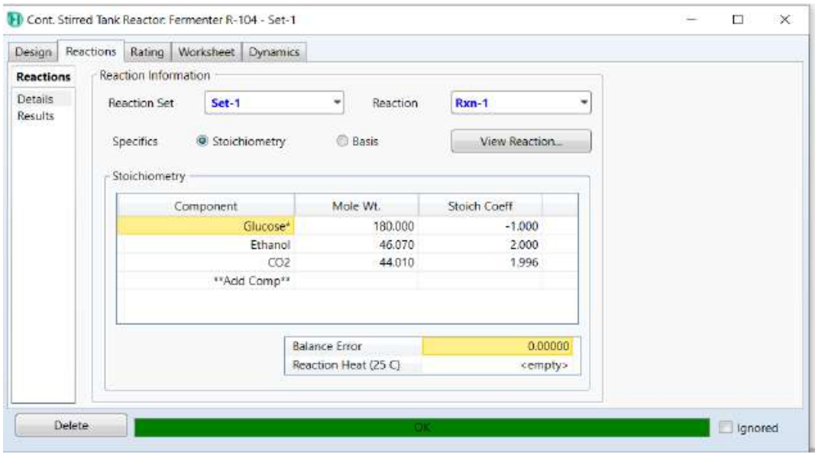
Forward Reaction	
A	2.9500
E	0.00000
b	0.00000

Reverse Reaction	
A'	<empty>
E'	<empty>
b'	<empty>

r = k*f(Basis) - k'*f'(Basis)
k = A * exp (-E / RT) * T ^b
k' = A' * exp (-E' / RT) * T^b'
T in Kelvin

Balance Error	0.00000
Reaction Heat (25 C)	<empty>

Ready



Reaction Set: Set-1

Reaction: Rxn-1

Specifics: ● Stoichiometry ○ Basis

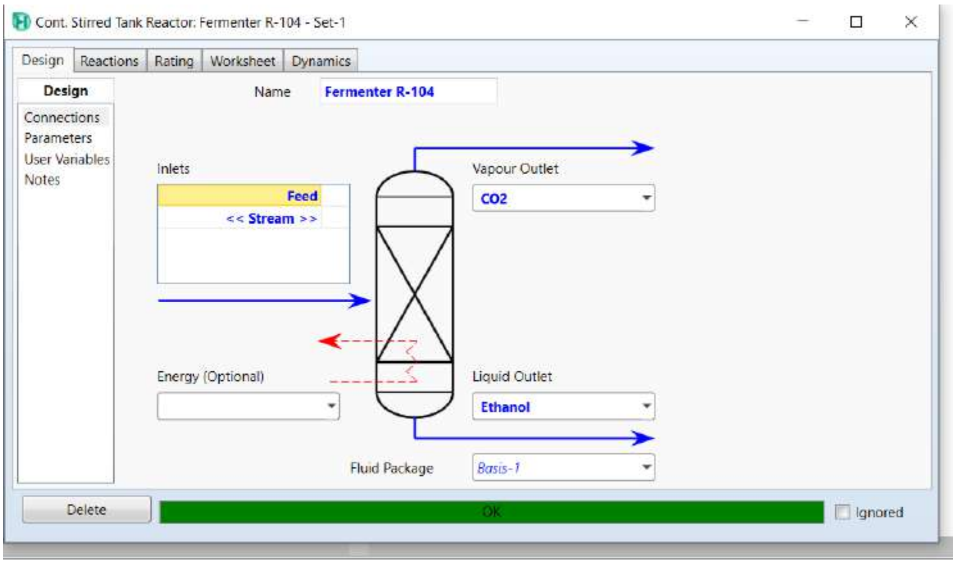
Component	Mole Wt.	Stoich Coeff
Glucose*	180.000	-1.000
Ethanol	46.070	2.000
CO2	44.010	1.996
Add Comp		

Balance Error	0.00000
Reaction Heat (25 C)	<empty>

OK

☐ Ignored

Go to the simulation environment and select Fermenter R-101 by double clicking a window will open label the input and output streams.



Name: **Fermenter R-104**

Inlets: **Feed**

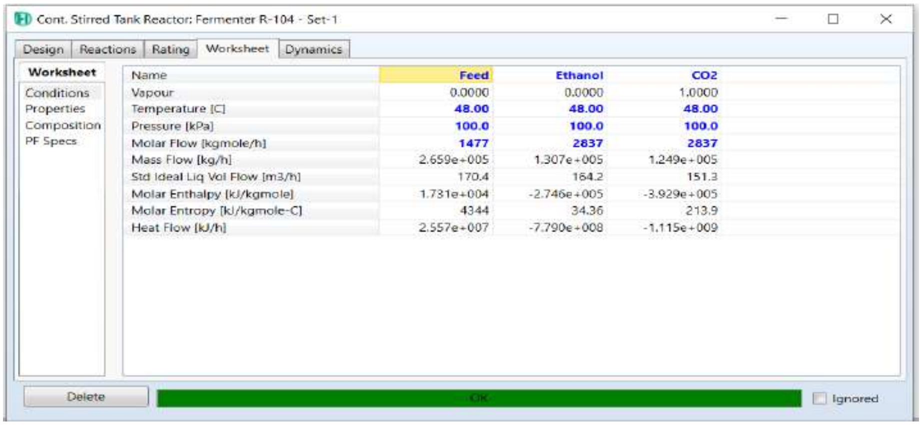
Vapour Outlet: **CO2**

Energy (Optional):

Liquid Outlet: **Ethanol**

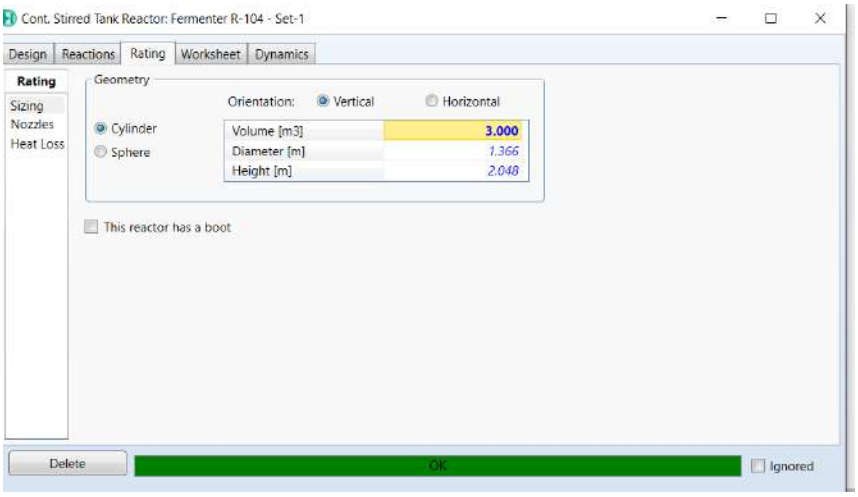
Fluid Package: *Basis-1*

Click on the worksheet tab and enter the data of the temperature and pressure of both input and output streams.



Name	Feed	Ethanol	CO2
Vapour	0.0000	0.0000	1.0000
Temperature [C]	**48.00**	**48.00**	**48.00**
Pressure [kPa]	**100.0**	**100.0**	**100.0**
Molar Flow [kgmole/h]	**1477**	**2837**	**2837**
Mass Flow [kg/h]	2.659e+005	1.307e+005	1.249e+005
Std Ideal Liq Vol Flow [m3/h]	170.4	164.2	151.3
Molar Enthalpy [kJ/kgmole]	1.731e+004	-2.746e+005	-3.929e+005
Molar Entropy [kJ/kgmole-C]	4344	34.36	213.9
Heat Flow [kJ/h]	2.557e+007	-7.790e+008	-1.115e+009

Click on the ratings tab in the already opened window.



Orientation: ● Vertical ○ Horizontal

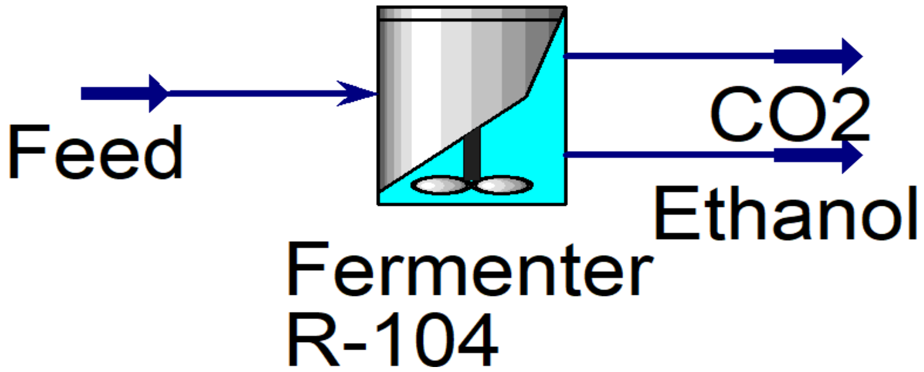
● Cylinder

○ Sphere

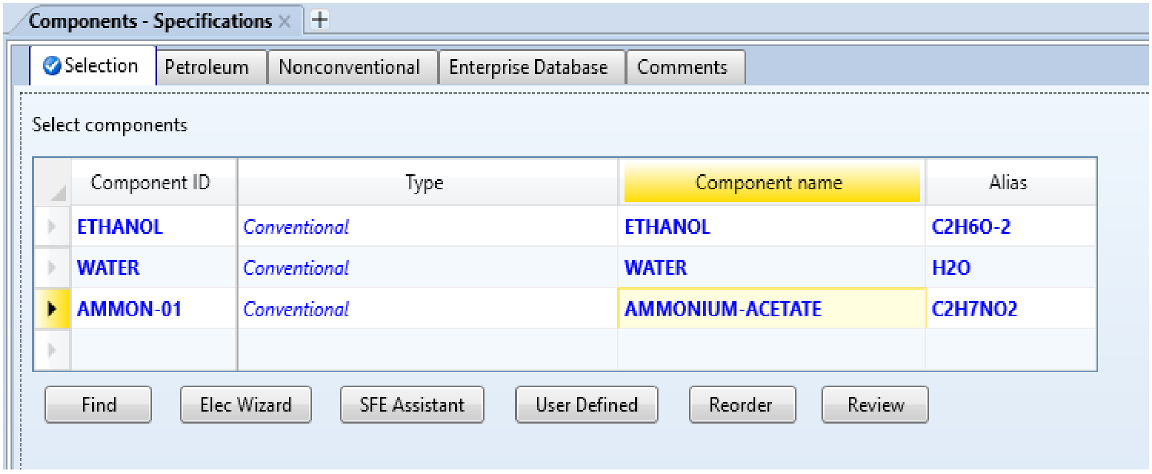
Volume [m3]	3.000
Diameter [m]	*1.366*
Height [m]	*2.048*

☐ This reactor has a boot

The Fermenter Reactor should be simulated now and look like this.



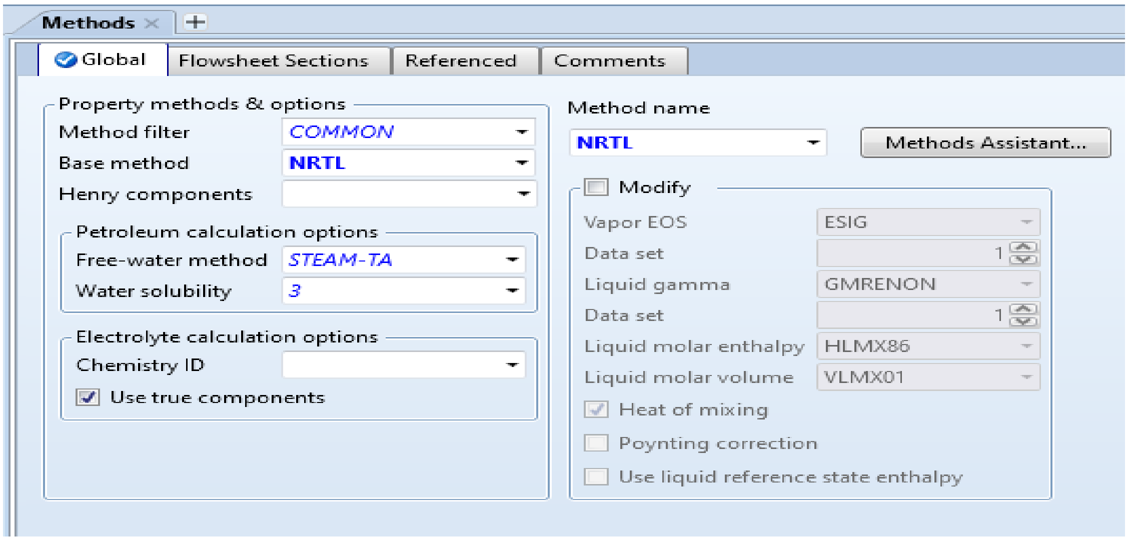
8.6 Simulation of Distillation Column (D-101):



Select components

Component ID	Type	Component name	Alias
ETHANOL	Conventional	ETHANOL	C2H6O-2
WATER	Conventional	WATER	H2O
AMMON-01	Conventional	AMMONIUM-ACETATE	C2H7NO2

Click on the Methods in the navigation pane. Select NTRL as the Base Method.



Property methods & options
Method filter: COMMON
Base method: NRTL
Henry components:

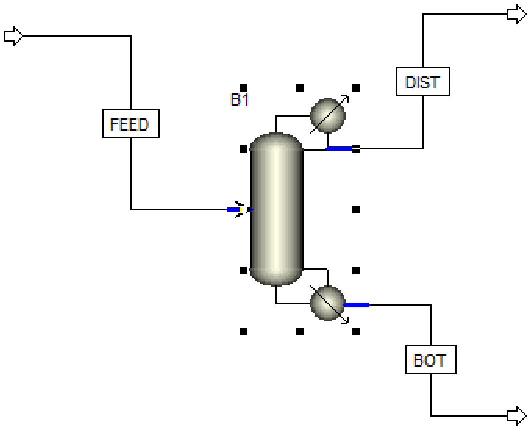
Petroleum calculation options
Free-water method: STEAM-TA
Water solubility: 3

Electrolyte calculation options
Chemistry ID:
☑ Use true components

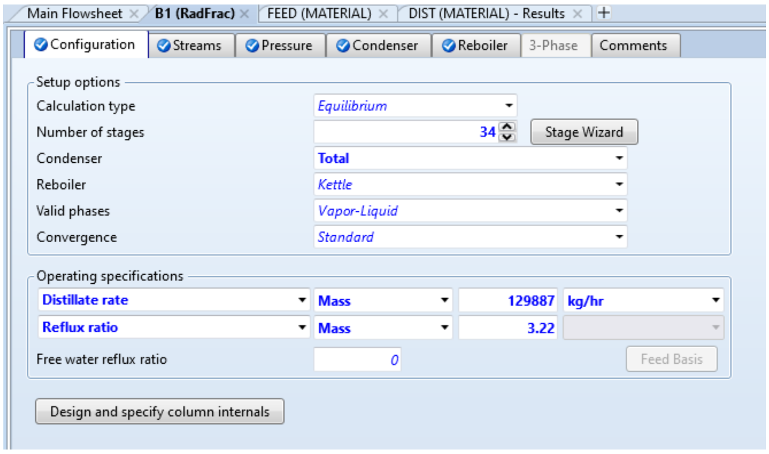
Method name: NRTL

☐ Modify
Vapor EOS: ESIG
Data set: 1
Liquid gamma: GMRENON
Data set: 1
Liquid molar enthalpy: HLMX86
Liquid molar volume: VLMX01
☑ Heat of mixing
☐ Poynting correction
☐ Use liquid reference state enthalpy

Click a DSTWU block onto the flowchart in the Columns area of the Model Palette, then link the FEED stream to the input of column. Create a bottoms stream and a distillate stream. Your flow-sheet should resemble the illustration below.



Now in the configuration tab add your desired data.



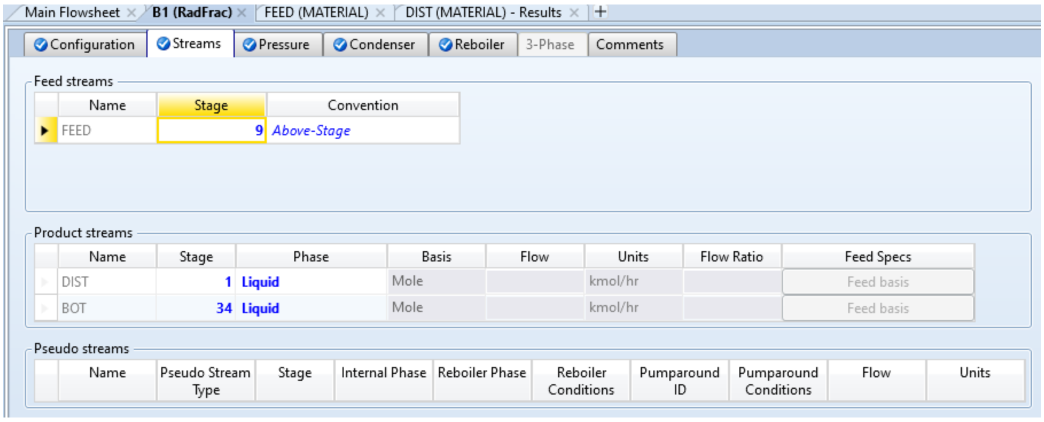
Setup options

Calculation type	Equilibrium
Number of stages	34
Condenser	Total
Reboiler	Kettle
Valid phases	Vapor-Liquid
Convergence	Standard

Operating specifications

Distillate rate	Mass	129887	kg/hr
Reflux ratio	Mass	3.22	
Free water reflux ratio		0	

In the Streams Tab give the Feed at stage 9 and also give the remaining data.



Feed streams

Name	Stage	Convention
FEED	9	Above-Stage

Product streams

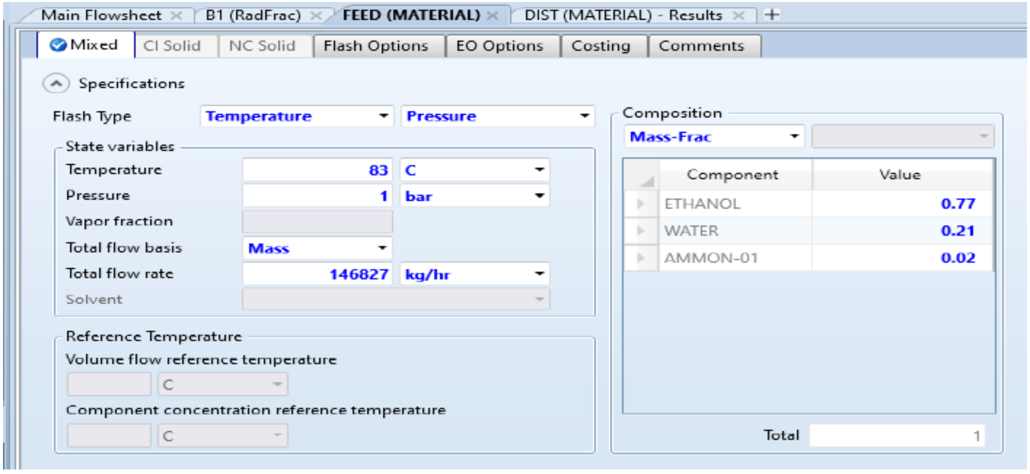
Name	Stage	Phase	Basis	Flow	Units	Flow Ratio	Feed Specs
DIST	1	Liquid	Mole		kmol/hr		Feed basis
BOT	34	Liquid	Mole		kmol/hr		Feed basis

Pseudo streams

Name	Pseudo Stream Type	Stage	Internal Phase	Reboiler Phase	Reboiler Conditions	Pumparound ID	Pumparound Conditions	Flow	Units

In the Pressures Tab give the pressure for condenser.

Choose the Flash Type variables to be Pressure and Temperature.



Flash Type: Temperature, Pressure

State variables

Temperature: 83 C

Pressure: 1 bar

Vapor fraction:

Total flow basis: Mass

Total flow rate: 146827 kg/hr

Solvent:

Reference Temperature

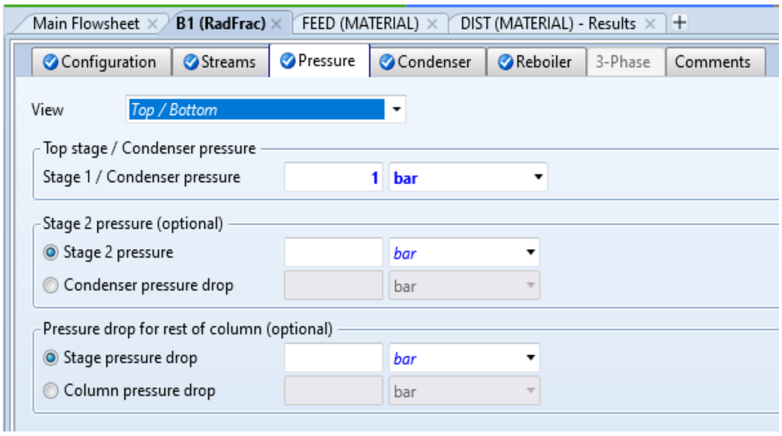
Volume flow reference temperature: C

Component concentration reference temperature: C

Composition: Mass-Frac

Component	Value
ETHANOL	0.77
WATER	0.21
AMMON-01	0.02

Total: 1



View: Top / Bottom

Top stage / Condenser pressure

Stage 1 / Condenser pressure: 1 bar

Stage 2 pressure (optional)

● Stage 2 pressure: bar

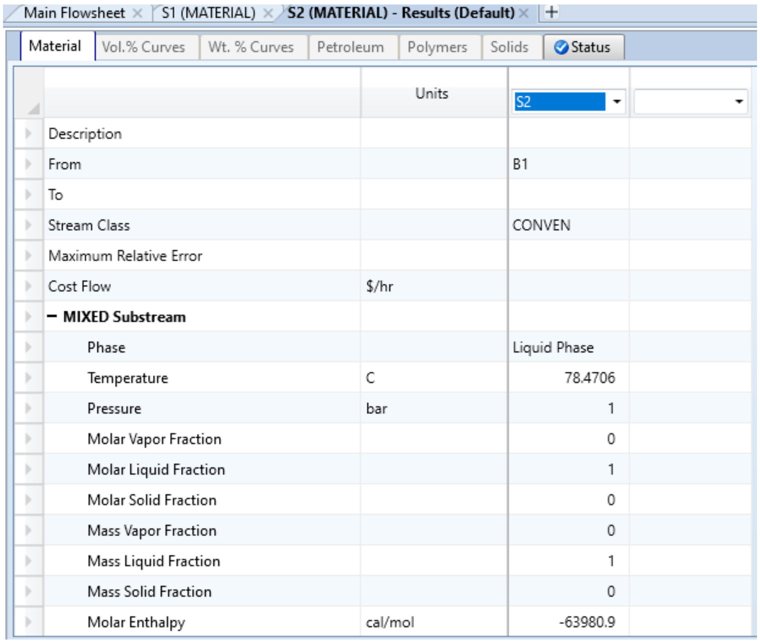
○ Condenser pressure drop: bar

Pressure drop for rest of column (optional)

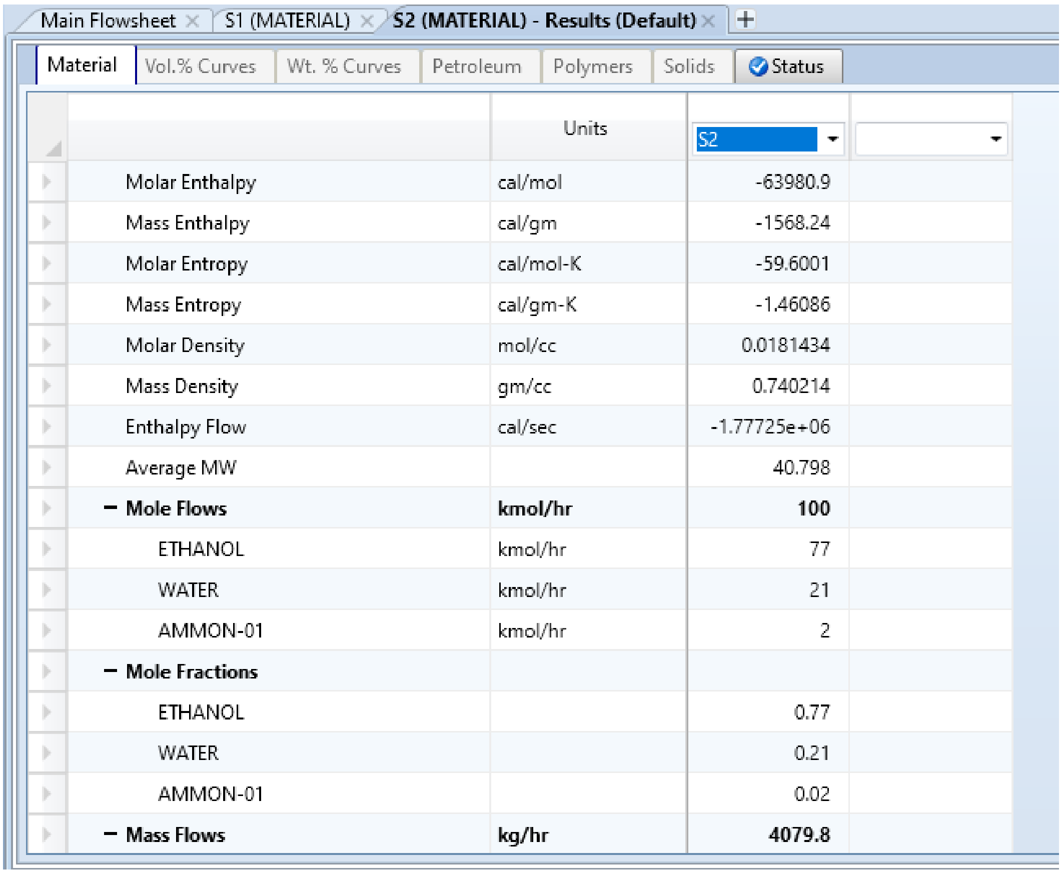
● Stage pressure drop: bar

○ Column pressure drop: bar

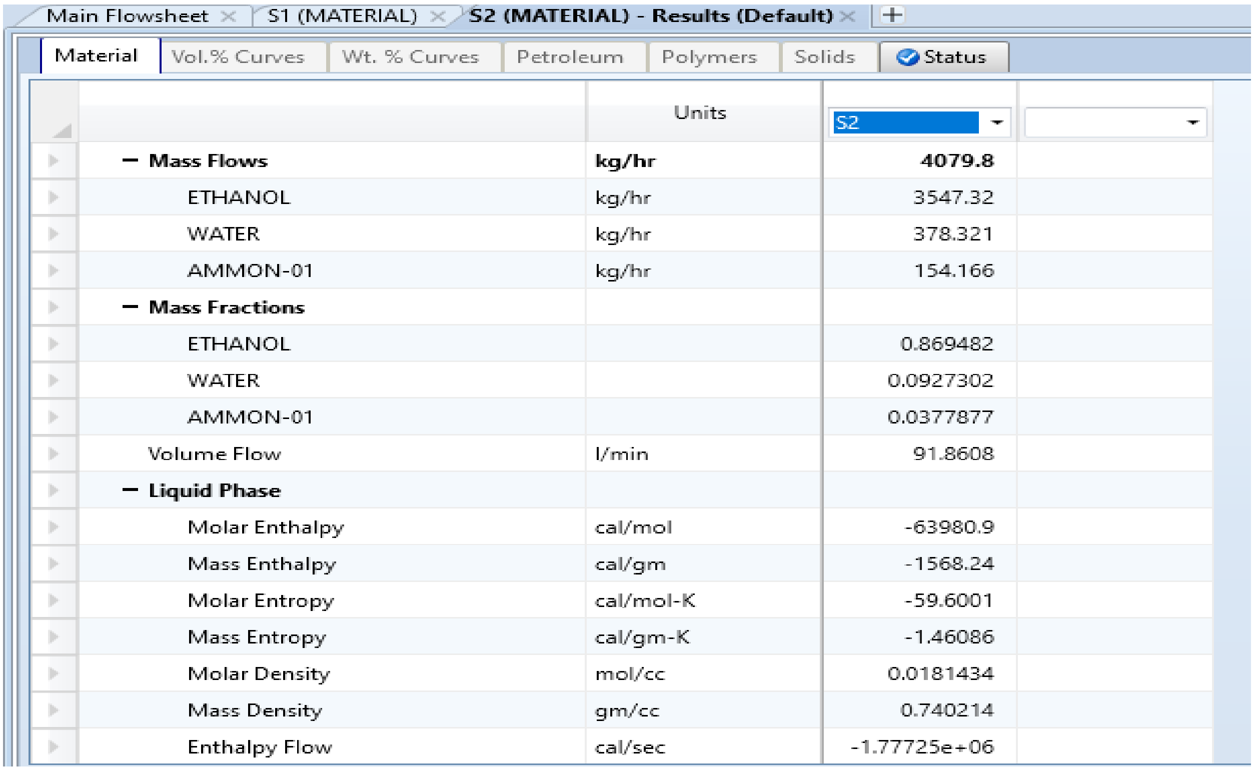
Run the simulation.



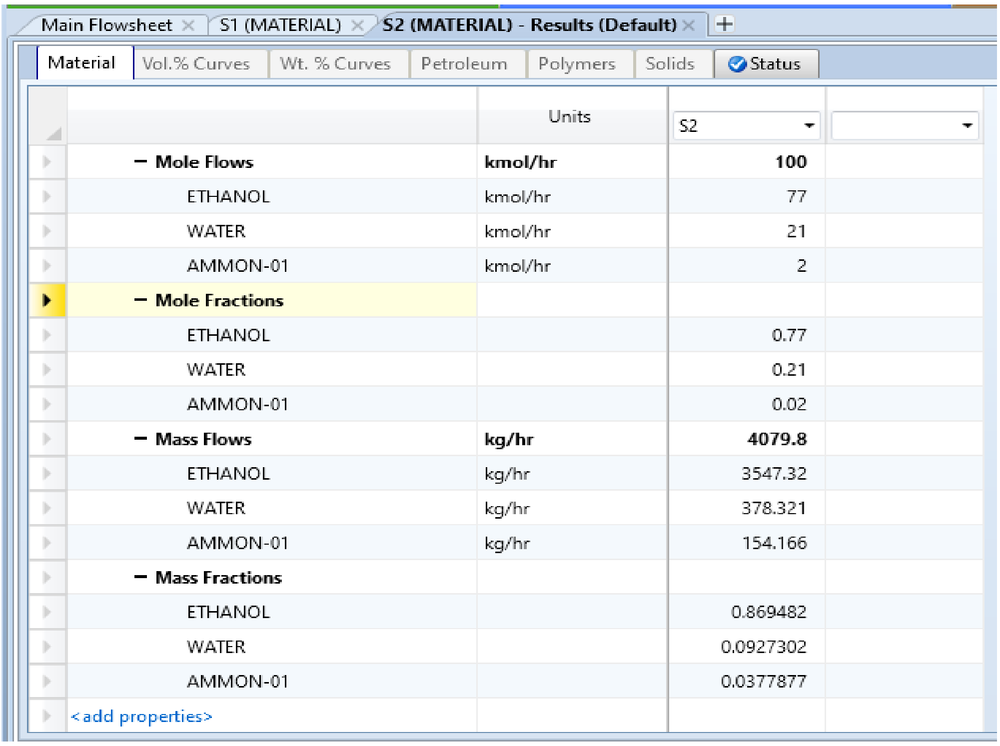
	Units	S2
Description		
From		B1
To		
Stream Class		CONVEN
Maximum Relative Error		
Cost Flow	$/hr	
− MIXED Substream		
Phase		Liquid Phase
Temperature	C	78.4706
Pressure	bar	1
Molar Vapor Fraction		0
Molar Liquid Fraction		1
Molar Solid Fraction		0
Mass Vapor Fraction		0
Mass Liquid Fraction		1
Mass Solid Fraction		0
Molar Enthalpy	cal/mol	-63980.9



	Units	S2
Molar Enthalpy	cal/mol	-63980.9
Mass Enthalpy	cal/gm	-1568.24
Molar Entropy	cal/mol-K	-59.6001
Mass Entropy	cal/gm-K	-1.46086
Molar Density	mol/cc	0.0181434
Mass Density	gm/cc	0.740214
Enthalpy Flow	cal/sec	-1.77725e+06
Average MW		40.798
− Mole Flows	**kmol/hr**	**100**
ETHANOL	kmol/hr	77
WATER	kmol/hr	21
AMMON-01	kmol/hr	2
− Mole Fractions		
ETHANOL		0.77
WATER		0.21
AMMON-01		0.02
− Mass Flows	**kg/hr**	**4079.8**



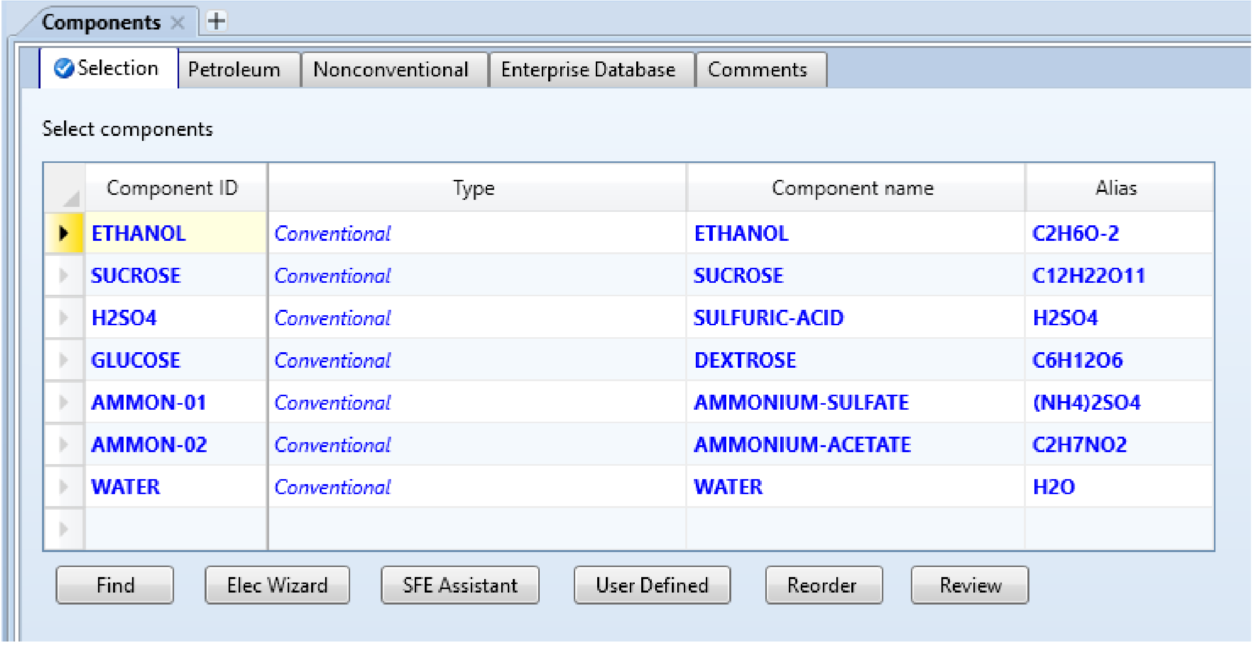
	Units	S2
− Mass Flows	**kg/hr**	**4079.8**
ETHANOL	kg/hr	3547.32
WATER	kg/hr	378.321
AMMON-01	kg/hr	154.166
− Mass Fractions		
ETHANOL		0.869482
WATER		0.0927302
AMMON-01		0.0377877
Volume Flow	l/min	91.8608
− Liquid Phase		
Molar Enthalpy	cal/mol	-63980.9
Mass Enthalpy	cal/gm	-1568.24
Molar Entropy	cal/mol-K	-59.6001
Mass Entropy	cal/gm-K	-1.46086
Molar Density	mol/cc	0.0181434
Mass Density	gm/cc	0.740214
Enthalpy Flow	cal/sec	-1.77725e+06



	Units	S2
− Mole Flows	**kmol/hr**	**100**
ETHANOL	kmol/hr	77
WATER	kmol/hr	21
AMMON-01	kmol/hr	2
− Mole Fractions		
ETHANOL		0.77
WATER		0.21
AMMON-01		0.02
− Mass Flows	**kg/hr**	**4079.8**
ETHANOL	kg/hr	3547.32
WATER	kg/hr	378.321
AMMON-01	kg/hr	154.166
− Mass Fractions		
ETHANOL		0.869482
WATER		0.0927302
AMMON-01		0.0377877
<add properties>		

8.7 Simulation of Distillation Column (D-102):

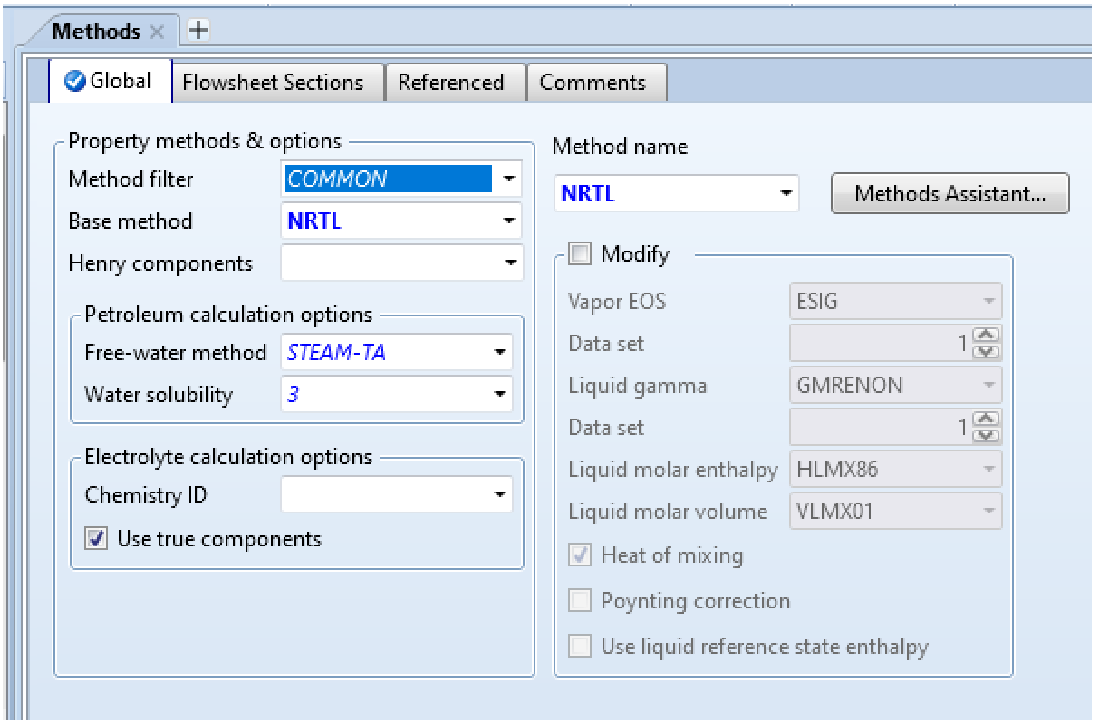
In the Components Specifications selection tab, enter the components needed for this simulation.



Select components

Component ID	Type	Component name	Alias
ETHANOL	Conventional	ETHANOL	C2H6O-2
SUCROSE	Conventional	SUCROSE	C12H22O11
H2SO4	Conventional	SULFURIC-ACID	H2SO4
GLUCOSE	Conventional	DEXTROSE	C6H12O6
AMMON-01	Conventional	AMMONIUM-SULFATE	(NH4)2SO4
AMMON-02	Conventional	AMMONIUM-ACETATE	C2H7NO2
WATER	Conventional	WATER	H2O

Click on Methods in the navigation pane. Select NTRL as the Base Method.



Property methods & options
Method filter: COMMON
Base method: NRTL
Henry components:

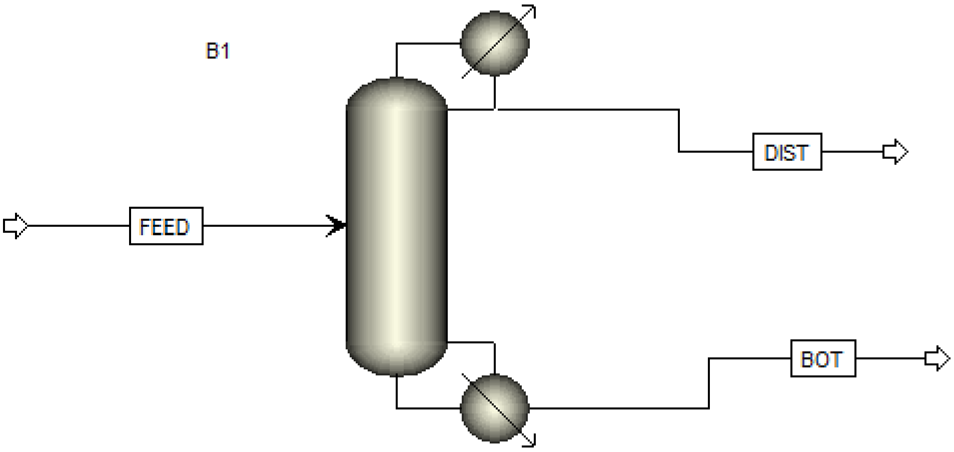
Petroleum calculation options
Free-water method: STEAM-TA
Water solubility: 3

Electrolyte calculation options
Chemistry ID:
☑ Use true components

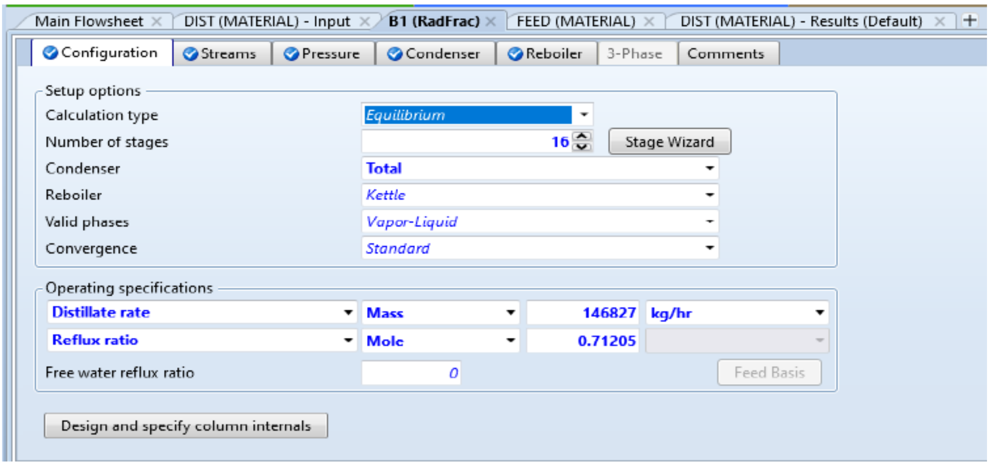
Method name: NRTL

☐ Modify
Vapor EOS: ESIG
Data set: 1
Liquid gamma: GMRENON
Data set: 1
Liquid molar enthalpy: HLMX86
Liquid molar volume: VLMX01
☑ Heat of mixing
☐ Poynting correction
☐ Use liquid reference state enthalpy

In the Columns section of the Model Palette, insert a DSTWU block onto the flowsheet and connect the FEED stream to the column input. Make a distillate stream and a bottoms stream. Your flowsheet should look like the example below.



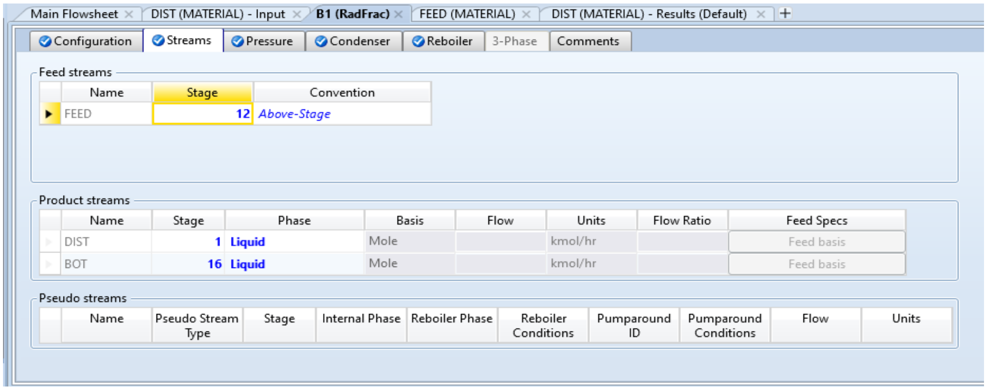
Now in the configuration tab add your desired data.



Setup options
Calculation type: Equilibrium
Number of stages: 16
Condenser: Total
Reboiler: Kettle
Valid phases: Vapor-Liquid
Convergence: Standard

Operating specifications
Distillate rate: Mass 146827 kg/hr
Reflux ratio: Mole 0.71205
Free water reflux ratio: 0

In the Streams Tab give the Feed at stage 9 and give the remaining data.



Feed streams

Name	Stage	Convention
FEED	12	Above-Stage

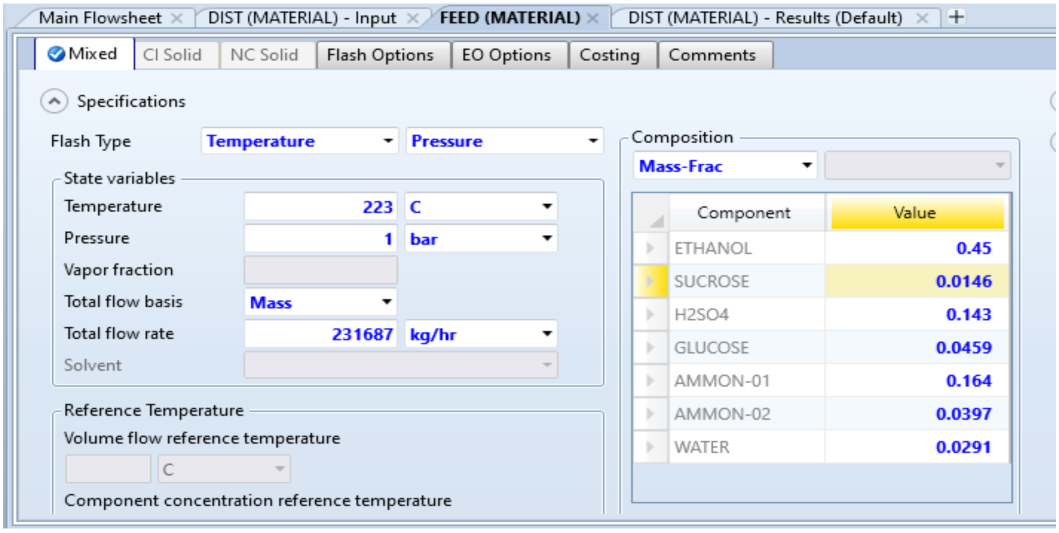
Product streams

Name	Stage	Phase	Basis	Flow	Units	Flow Ratio	Feed Specs
DIST	1	Liquid	Mole		kmol/hr		Feed basis
BOT	16	Liquid	Mole		kmol/hr		Feed basis

Pseudo streams

Name	Pseudo Stream Type	Stage	Internal Phase	Reboiler Phase	Reboiler Conditions	Pumparound ID	Pumparound Conditions	Flow	Units

Choose the Flash Type variables to be Pressure and Temperature.



Flash Type: Temperature Pressure

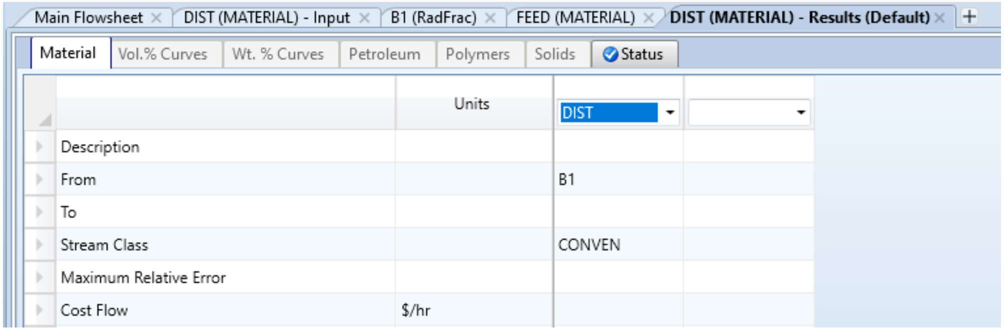
State variables
Temperature: 223 C
Pressure: 1 bar
Vapor fraction
Total flow basis: Mass
Total flow rate: 231687 kg/hr
Solvent

Reference Temperature
Volume flow reference temperature: C
Component concentration reference temperature

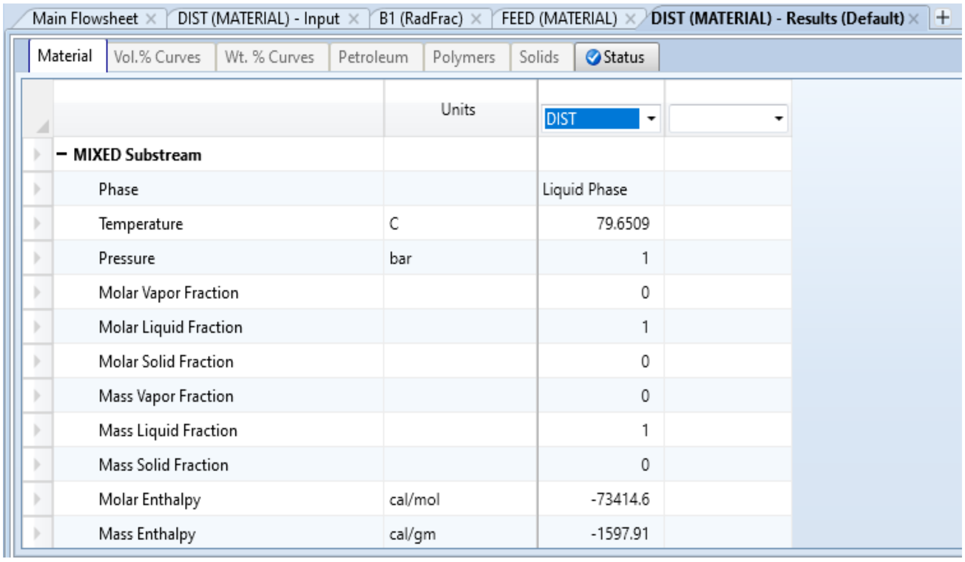
Composition: Mass-Frac

Component	Value
ETHANOL	0.45
SUCROSE	0.0146
H2SO4	0.143
GLUCOSE	0.0459
AMMON-01	0.164
AMMON-02	0.0397
WATER	0.0291

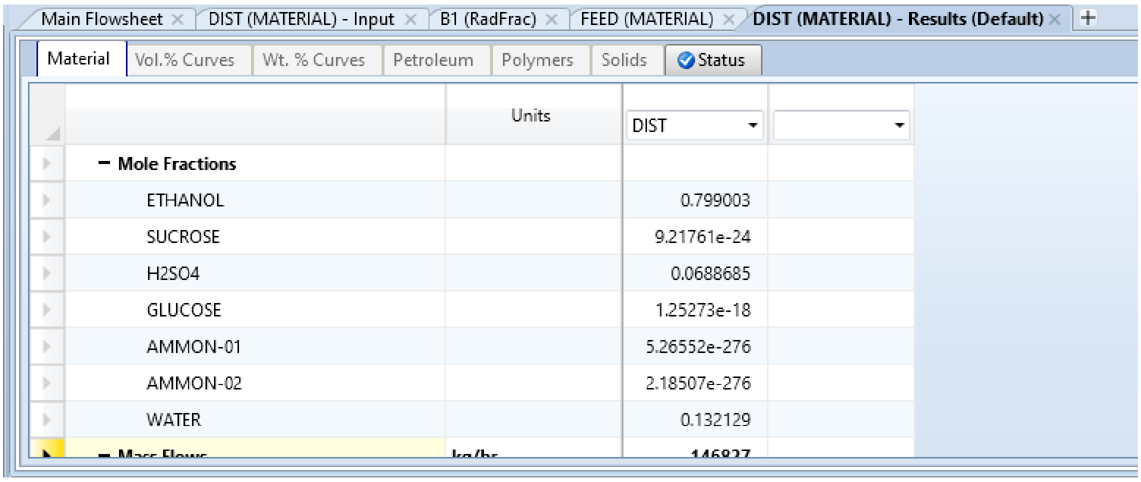
Run the simulation



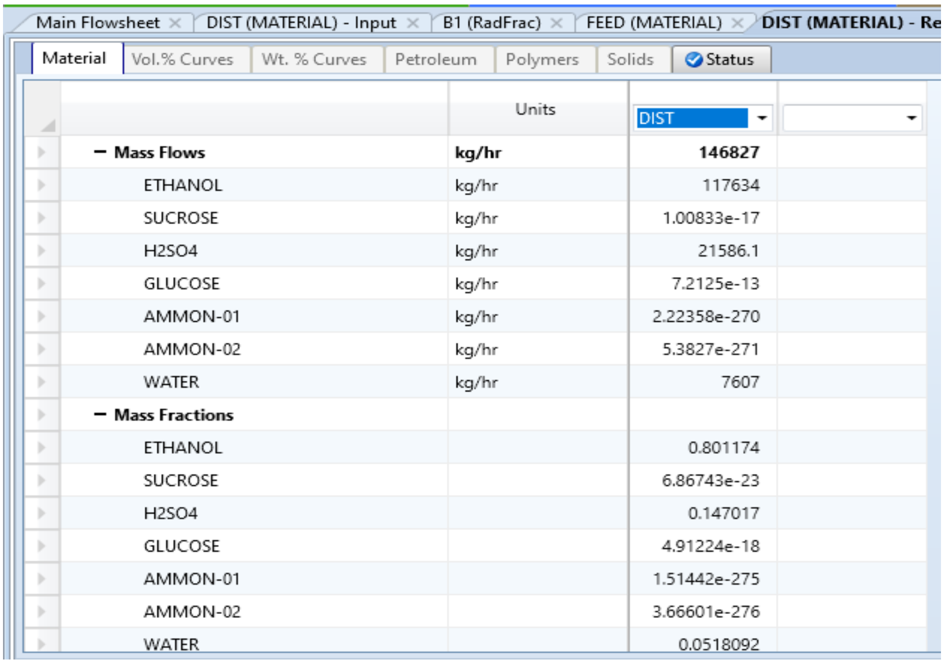
	Units	DIST	
Description			
From		B1	
To			
Stream Class		CONVEN	
Maximum Relative Error			
Cost Flow	$/hr		



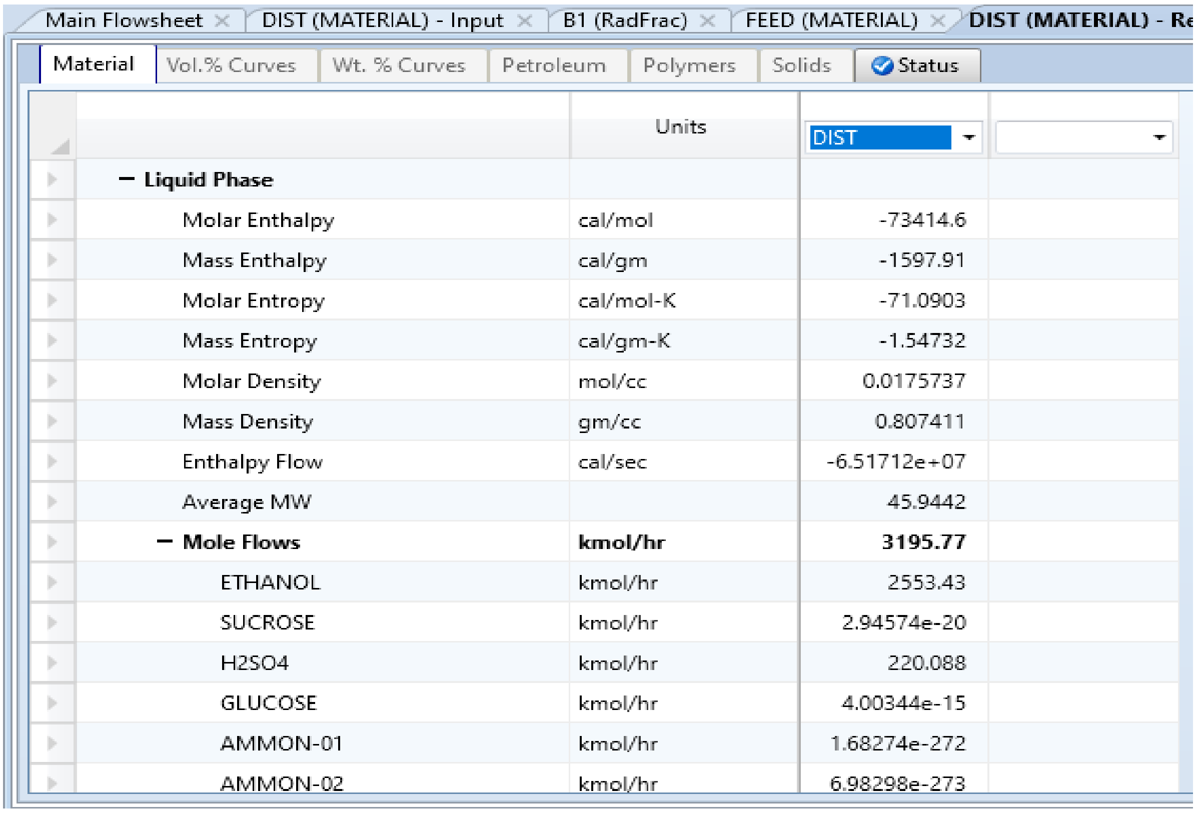
	Units	DIST	
− MIXED Substream			
Phase		Liquid Phase	
Temperature	C	79.6509	
Pressure	bar	1	
Molar Vapor Fraction		0	
Molar Liquid Fraction		1	
Molar Solid Fraction		0	
Mass Vapor Fraction		0	
Mass Liquid Fraction		1	
Mass Solid Fraction		0	
Molar Enthalpy	cal/mol	-73414.6	
Mass Enthalpy	cal/gm	-1597.91	



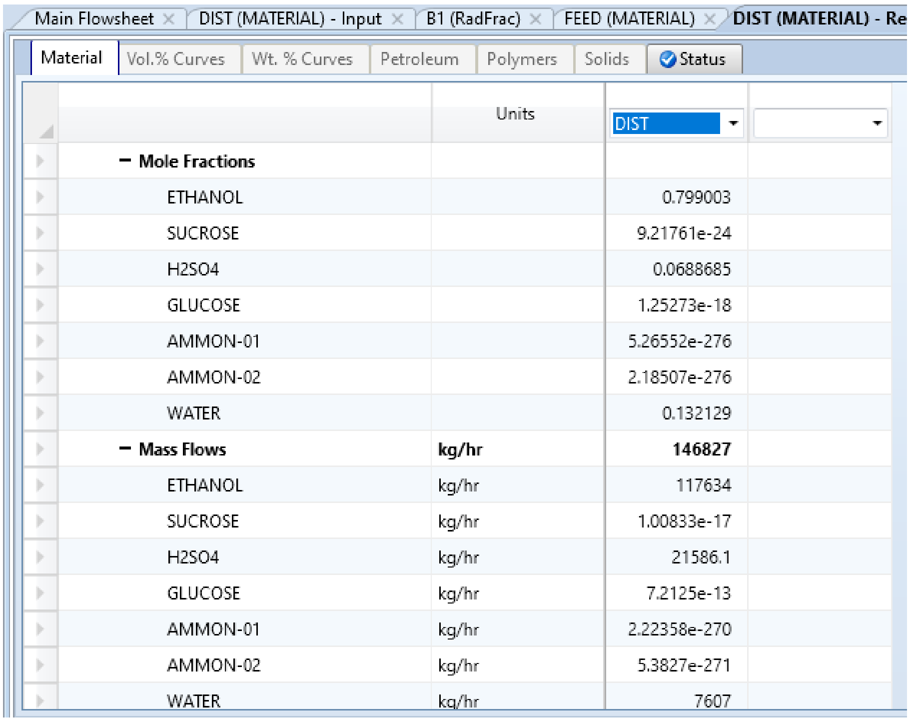
	Units	DIST	
− Mole Fractions			
ETHANOL		0.799003	
SUCROSE		9.21761e-24	
H2SO4		0.0688685	
GLUCOSE		1.25273e-18	
AMMON-01		5.26552e-276	
AMMON-02		2.18507e-276	
WATER		0.132129	
− Mass Flows	kg/hr	**146927**	



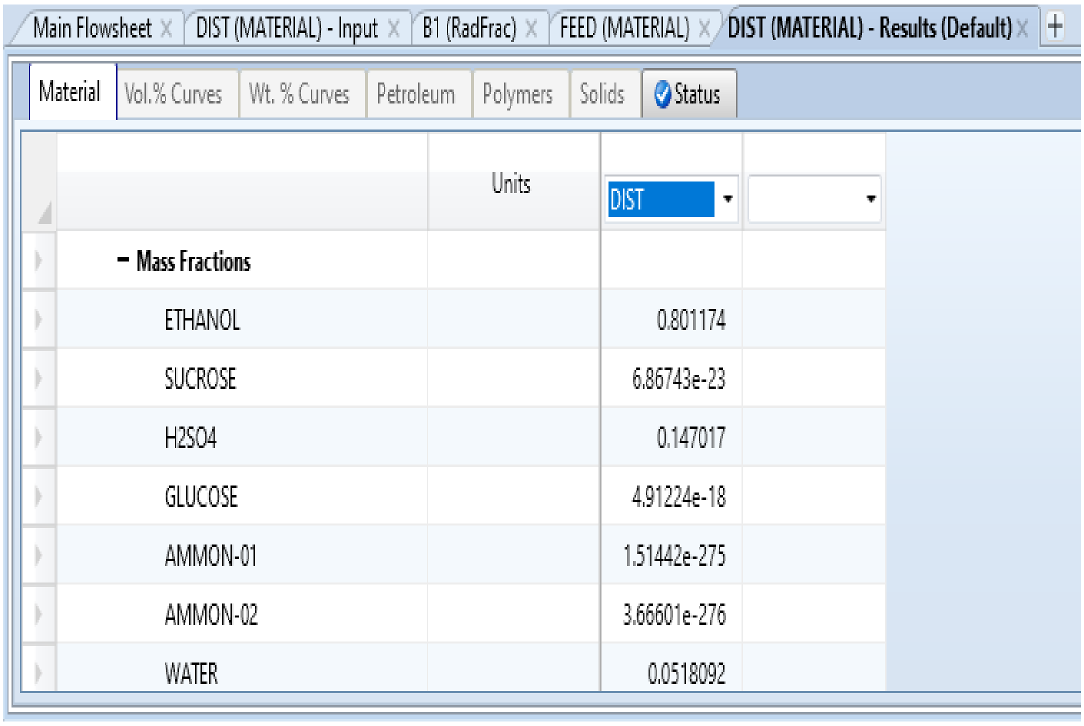
	Units	DIST
− Mass Flows	**kg/hr**	**146827**
ETHANOL	kg/hr	117634
SUCROSE	kg/hr	1.00833e-17
H2SO4	kg/hr	21586.1
GLUCOSE	kg/hr	7.2125e-13
AMMON-01	kg/hr	2.22358e-270
AMMON-02	kg/hr	5.3827e-271
WATER	kg/hr	7607
− Mass Fractions		
ETHANOL		0.801174
SUCROSE		6.86743e-23
H2SO4		0.147017
GLUCOSE		4.91224e-18
AMMON-01		1.51442e-275
AMMON-02		3.66601e-276
WATER		0.0518092



	Units	DIST
− Liquid Phase		
Molar Enthalpy	cal/mol	-73414.6
Mass Enthalpy	cal/gm	-1597.91
Molar Entropy	cal/mol-K	-71.0903
Mass Entropy	cal/gm-K	-1.54732
Molar Density	mol/cc	0.0175737
Mass Density	gm/cc	0.807411
Enthalpy Flow	cal/sec	-6.51712e+07
Average MW		45.9442
− Mole Flows	**kmol/hr**	**3195.77**
ETHANOL	kmol/hr	2553.43
SUCROSE	kmol/hr	2.94574e-20
H2SO4	kmol/hr	220.088
GLUCOSE	kmol/hr	4.00344e-15
AMMON-01	kmol/hr	1.68274e-272
AMMON-02	kmol/hr	6.98298e-273

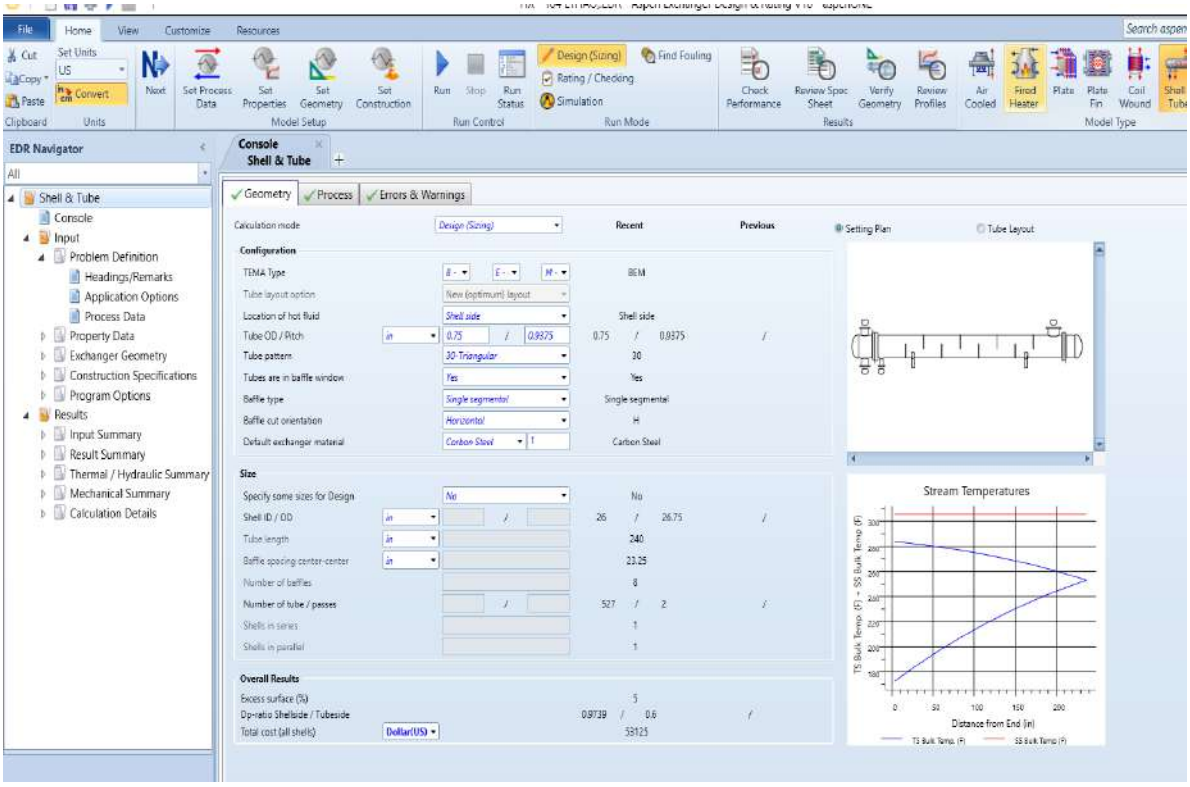


	Units	DIST
− Mole Fractions		
ETHANOL		0.799003
SUCROSE		9.21761e-24
H2SO4		0.0688685
GLUCOSE		1.25273e-18
AMMON-01		5.26552e-276
AMMON-02		2.18507e-276
WATER		0.132129
− Mass Flows	**kg/hr**	**146827**
ETHANOL	kg/hr	117634
SUCROSE	kg/hr	1.00833e-17
H2SO4	kg/hr	21586.1
GLUCOSE	kg/hr	7.2125e-13
AMMON-01	kg/hr	2.22358e-270
AMMON-02	kg/hr	5.3827e-271
WATER	kg/hr	7607



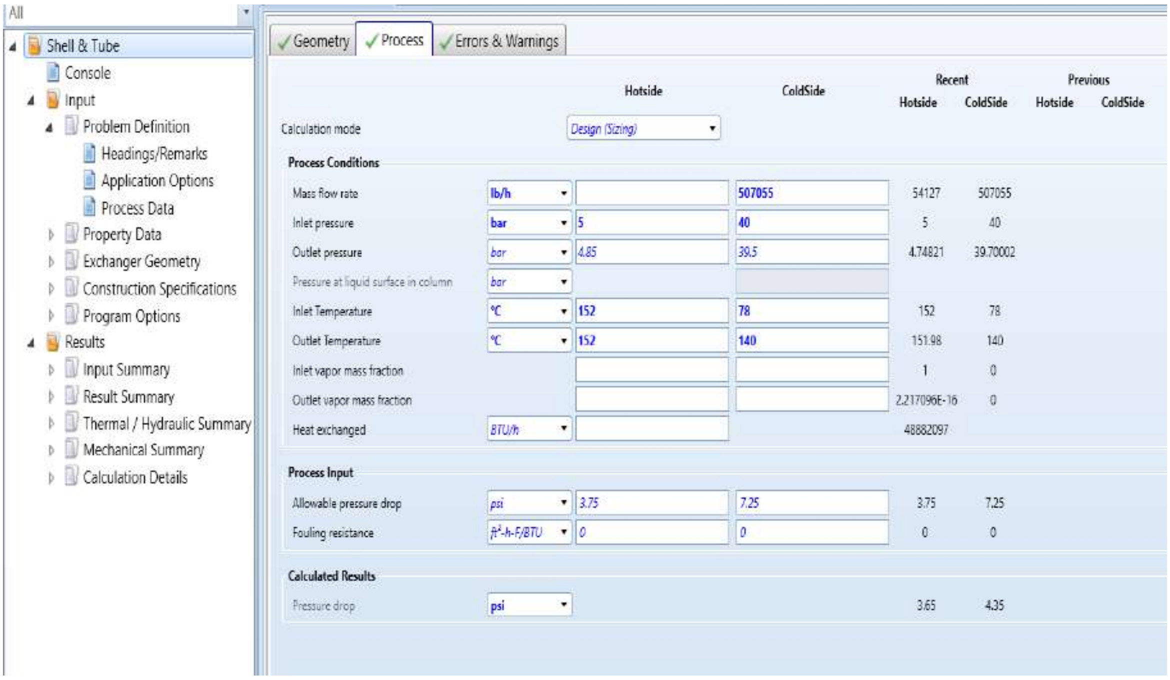
	Units	DIST
− Mass Fractions		
ETHANOL		0.801174
SUCROSE		6.86743e-23
H2SO4		0.147017
GLUCOSE		4.91224e-18
AMMON-01		1.51442e-275
AMMON-02		3.66601e-276
WATER		0.0518092

8.8 Simulation of Heat Exchanger:



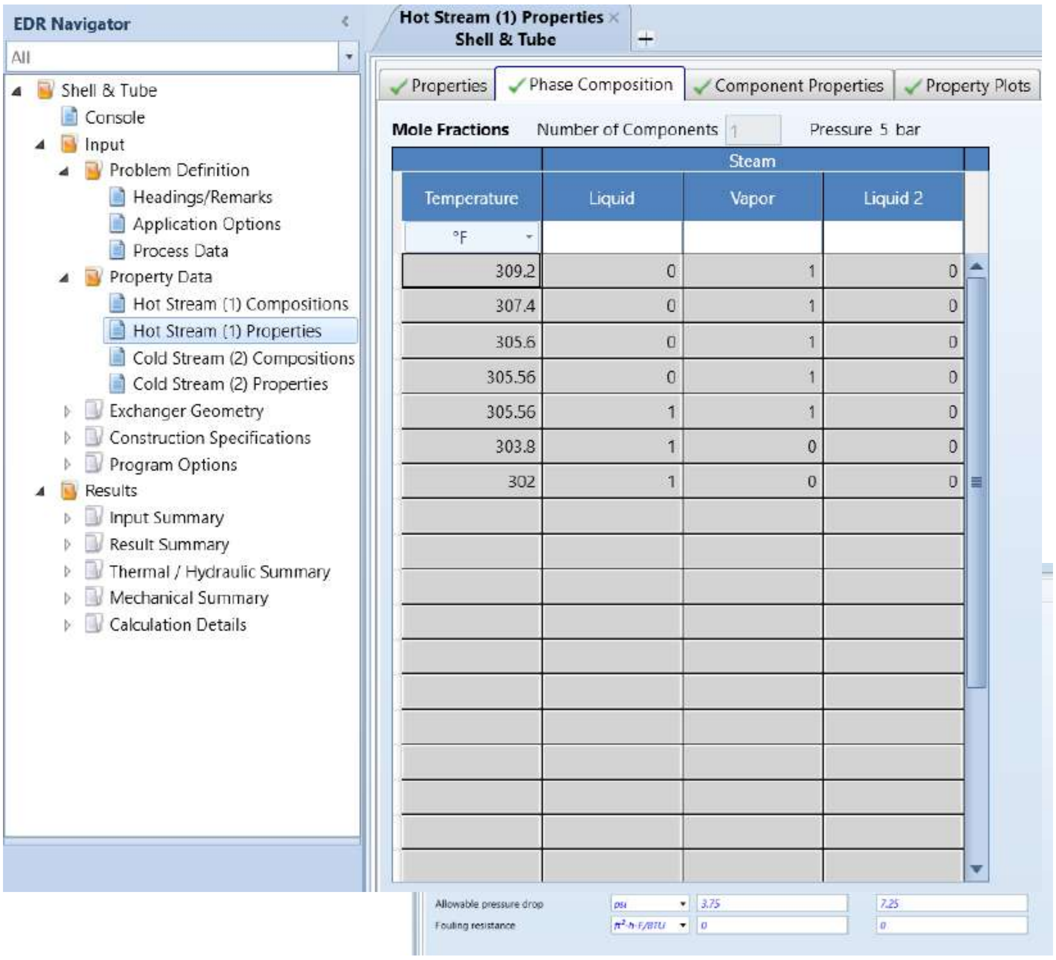
Calculation mode: Design (Sizing)

			Recent	Previous
Configuration				
TEMA Type		B - / E - / M -	BEM	
Tube layout option		New (optimum) layout		
Location of hot fluid		Shell side	Shell side	
Tube OD / Pitch	in	0.75 / 0.9375	0.75 / 0.9375	/
Tube pattern		30-Triangular	30	
Tubes are in baffle window		Yes	Yes	
Baffle type		Single segmental	Single segmental	
Baffle cut orientation		Horizontal	H	
Default exchanger material		Carbon Steel 1	Carbon Steel	
Size				
Specify some sizes for Design		No	No	
Shell ID / OD	in	/	26 / 26.75	/
Tube length	in		240	
Baffle spacing center-center	in		23.25	
Number of baffles			8	
Number of tube / passes		/	527 / 2	/
Shells in series			1	
Shells in parallel			1	
Overall Results				
Excess surface (%)			5	
Dp-ratio Shellside / Tubeside			0.9739 / 0.6	/
Total cost (all shells)	Dollar(US)		53125	



Calculation mode: Design (Sizing)

		Hotside	ColdSide	Recent Hotside	Recent ColdSide	Previous Hotside	Previous ColdSide
Process Conditions							
Mass flow rate	lb/h		507055	54127	507055		
Inlet pressure	bar	5	40	5	40		
Outlet pressure	bar	4.85	39.5	4.74821	39.70002		
Pressure at liquid surface in column	bar						
Inlet Temperature	°C	152	78	152	78		
Outlet Temperature	°C	152	140	151.98	140		
Inlet vapor mass fraction				1	0		
Outlet vapor mass fraction				2.217096E-16	0		
Heat exchanged	BTU/h			48882097			
Process Input							
Allowable pressure drop	psi	3.75	7.25	3.75	7.25		
Fouling resistance	ft²-h-F/BTU	0	0	0	0		
Calculated Results							
Pressure drop	psi			3.65	4.35		

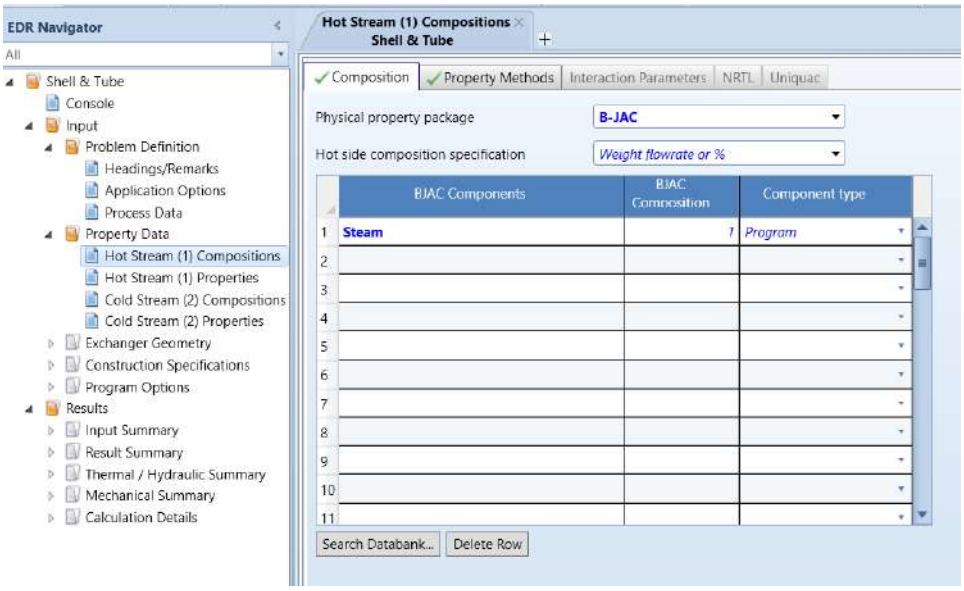


Mole Fractions Number of Components 1 Pressure 5 bar

	Steam		
Temperature	Liquid	Vapor	Liquid 2
°F			
309.2	0	1	0
307.4	0	1	0
305.6	0	1	0
305.56	0	1	0
305.56	1	1	0
303.8	1	0	0
302	1	0	0

Allowable pressure drop psi 3.75 7.25

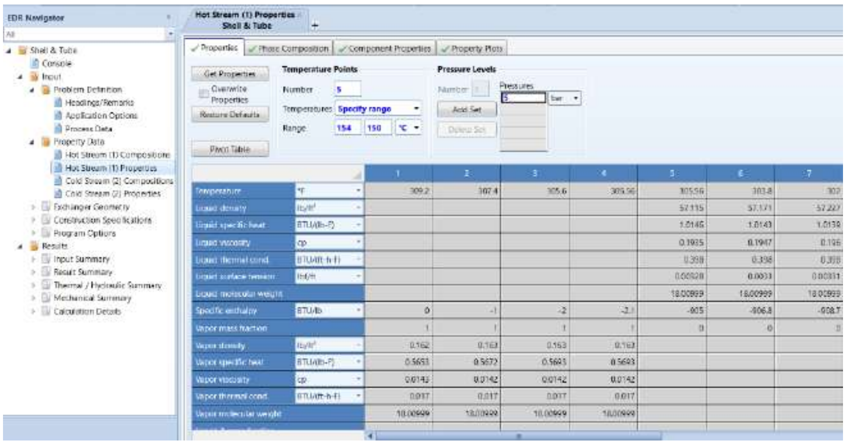
Fouling resistance ft²-h-F/BTU 0 0



Physical property package B-JAC

Hot side composition specification *Weight flowrate or %*

	BJAC Components	BJAC Composition	Component type
1	**Steam**	1	*Program*
2			
3			
4			
5			
6			
7			
8			
9			
10			
11			



Temperature Points

Number: 5

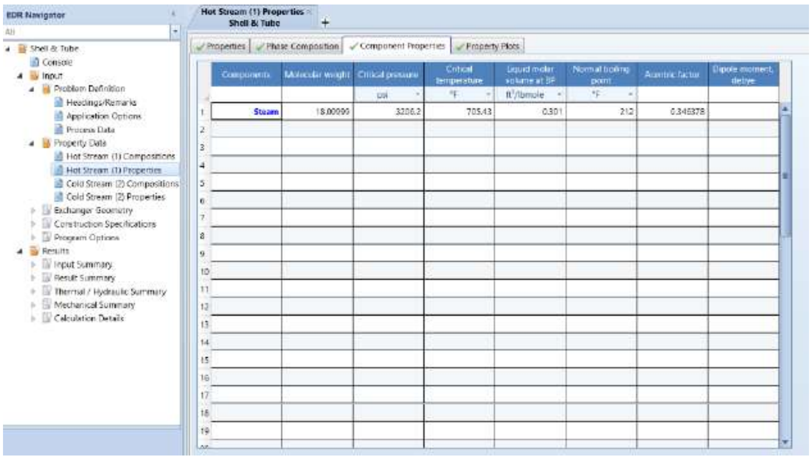
Temperatures: Specify range

Range: 154 150 °C

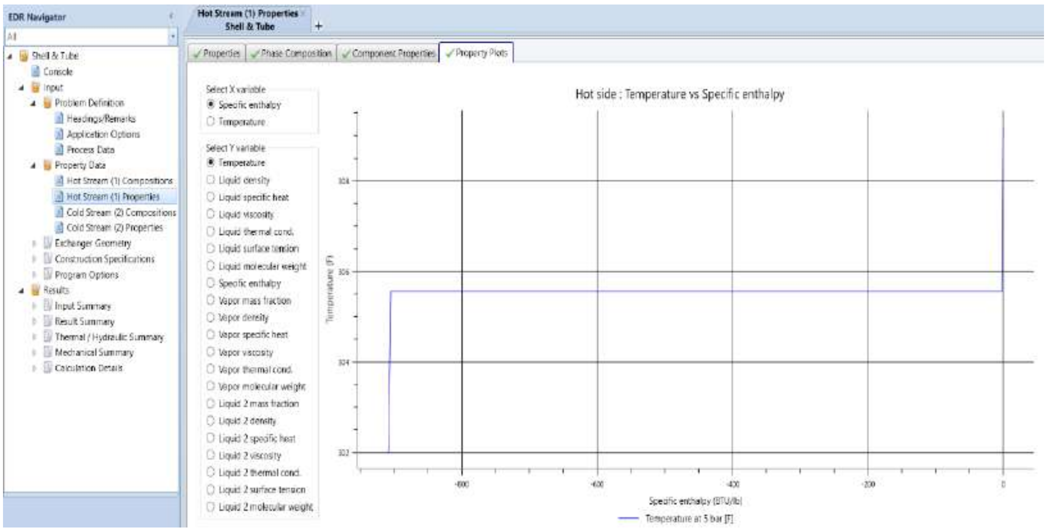
Pressure Levels

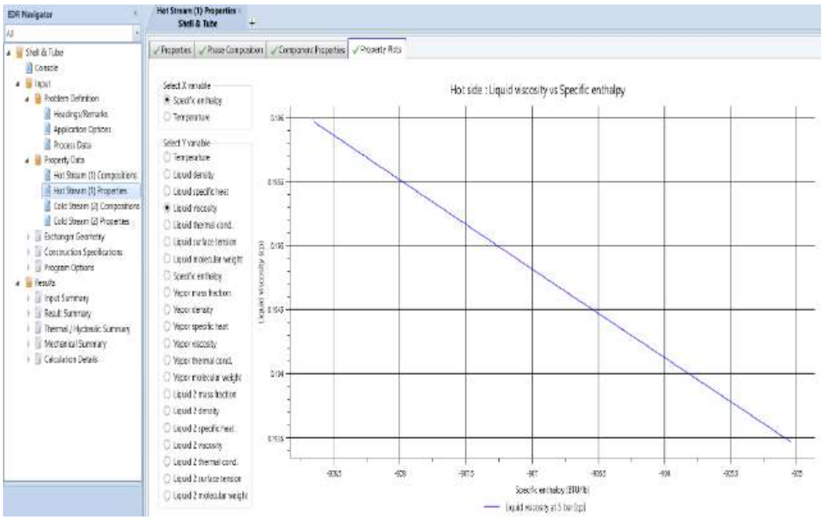
Pressures: 5 bar

		1	2	3	4	5	6	7
Temperature	°F	309.2	307.4	305.6	305.56	305.56	303.8	302
Liquid density	lb/ft³					57.115	57.171	57.227
Liquid specific heat	BTU/(lb-F)					1.0145	1.0143	1.0139
Liquid viscosity	cp					0.1935	0.1947	0.196
Liquid thermal cond.	BTU/(ft-h-F)					0.398	0.398	0.398
Liquid surface tension	lbf/ft					0.00328	0.0033	0.00331
Liquid molecular weight						18.00999	18.00999	18.00999
Specific enthalpy	BTU/lb	0	-1	-2	-2.1	-905	-906.8	-908.7
Vapor mass fraction		1	1	1	1	0	0	0
Vapor density	lb/ft³	0.162	0.163	0.163	0.163			
Vapor specific heat	BTU/(lb-F)	0.5653	0.5672	0.5693	0.5693			
Vapor viscosity	cp	0.0143	0.0142	0.0142	0.0142			
Vapor thermal cond.	BTU/(ft-h-F)	0.017	0.017	0.017	0.017			
Vapor molecular weight		18.00999	18.00999	18.00999	18.00999			



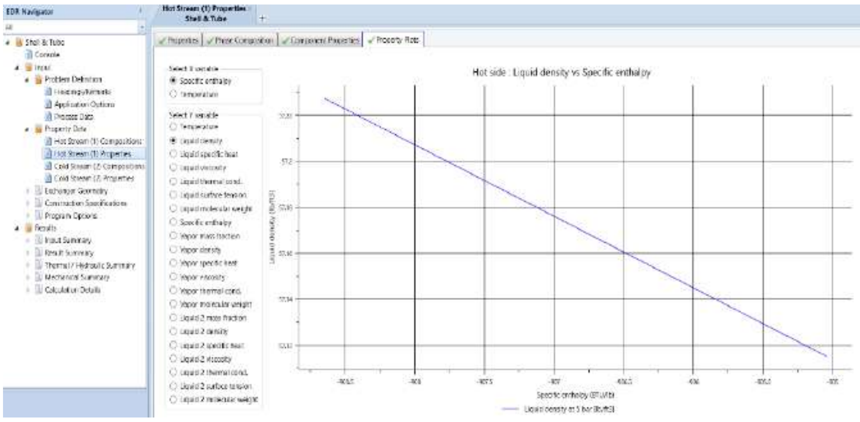
	Components	Molecular weight	Critical pressure	Critical temperature	Liquid molar volume at BP	Normal boiling point	Acentric factor	Dipole moment, debye
			psi	°F	ft³/lbmole	°F		
1	Steam	18.00999	3208.2	705.43	0.301	212	0.348378	
2								
3								
4								
5								
6								
7								
8								
9								
10								
11								
12								
13								
14								
15								
16								
17								
18								
19								





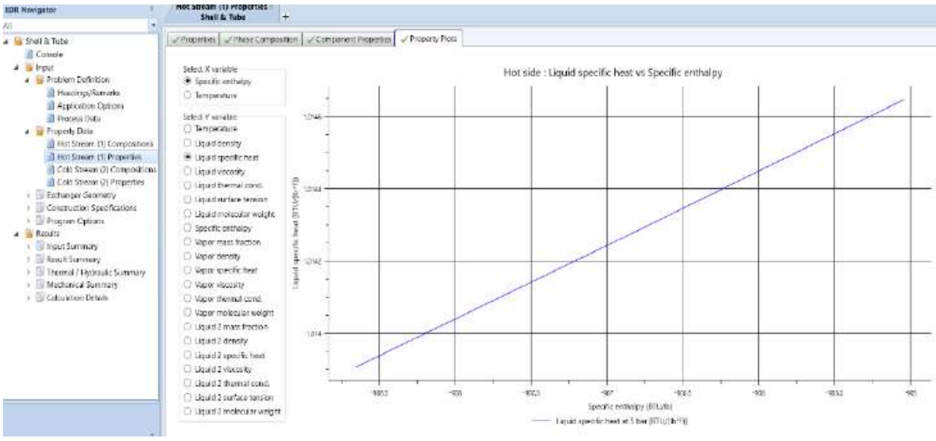
Select X variable
● Specific enthalpy
○ Temperature

Select Y variable
○ Temperature
○ Liquid density
○ Liquid specific heat
● Liquid viscosity
○ Liquid thermal cond.
○ Liquid surface tension
○ Liquid molecular weight
○ Specific enthalpy
○ Vapor mass fraction
○ Vapor density
○ Vapor specific heat
○ Vapor viscosity
○ Vapor thermal cond.
○ Vapor molecular weight
○ Liquid 2 mass fraction
○ Liquid 2 density
○ Liquid 2 specific heat
○ Liquid 2 viscosity
○ Liquid 2 thermal cond.
○ Liquid 2 surface tension
○ Liquid 2 molecular weight



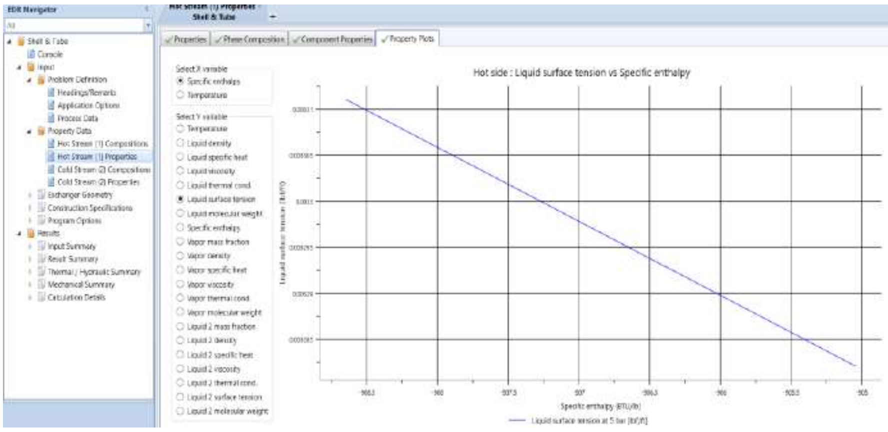
Select X variable
● Specific enthalpy
○ Temperature

Select Y variable
○ Temperature
● Liquid density
○ Liquid specific heat
○ Liquid viscosity
○ Liquid thermal cond.
○ Liquid surface tension
○ Liquid molecular weight
○ Specific enthalpy
○ Vapor mass fraction
○ Vapor density
○ Vapor specific heat
○ Vapor viscosity
○ Vapor thermal cond.
○ Vapor molecular weight
○ Liquid 2 mass fraction
○ Liquid 2 density
○ Liquid 2 specific heat
○ Liquid 2 viscosity
○ Liquid 2 thermal cond.
○ Liquid 2 surface tension
○ Liquid 2 molecular weight



Select X variable
● Specific enthalpy
○ Temperature

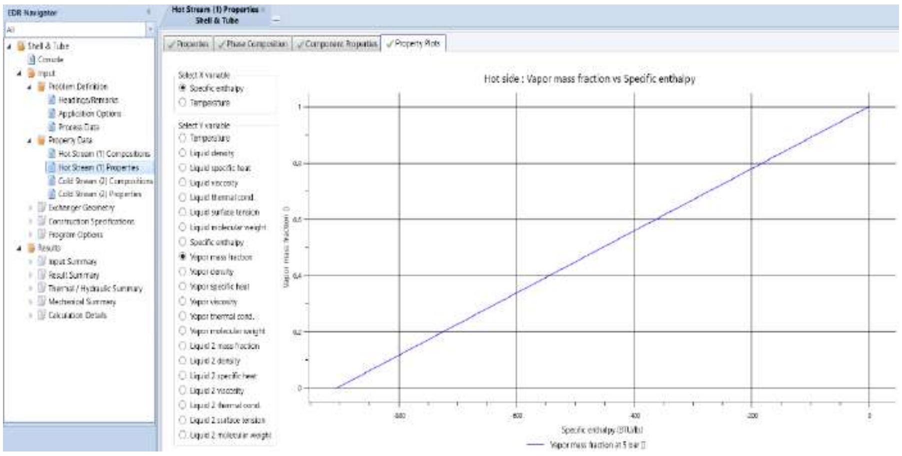
Select Y variable
○ Temperature
○ Liquid density
● Liquid specific heat
○ Liquid viscosity
○ Liquid thermal cond.
○ Liquid surface tension
○ Liquid molecular weight
○ Specific enthalpy
○ Vapor mass fraction
○ Vapor density
○ Vapor specific heat
○ Vapor viscosity
○ Vapor thermal cond.
○ Vapor molecular weight
○ Liquid 2 mass fraction
○ Liquid 2 density
○ Liquid 2 specific heat
○ Liquid 2 viscosity
○ Liquid 2 thermal cond.
○ Liquid 2 surface tension
○ Liquid 2 molecular weight



Hot Stream (1) Properties – Shell & Tube – Property Plots

Select X variable: ● Specific enthalpy

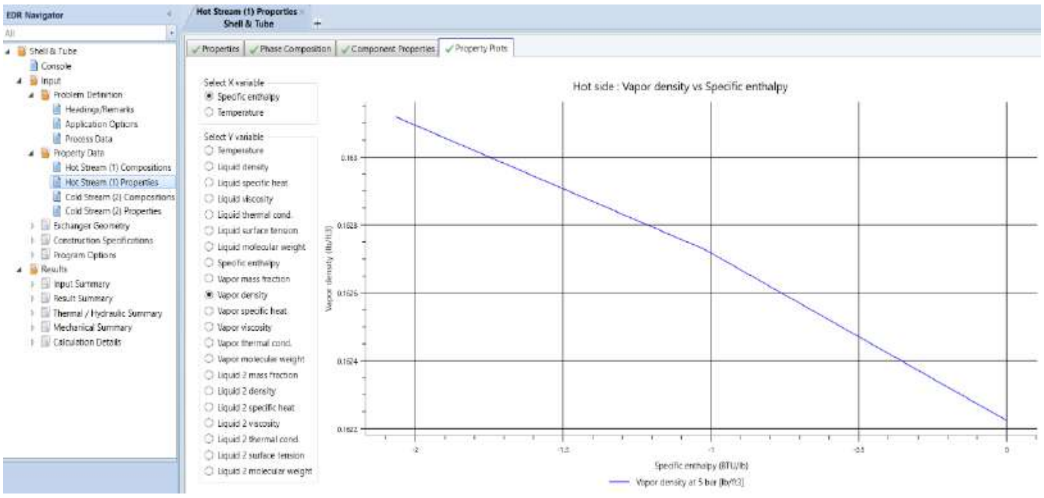
Select Y variable: ● Liquid surface tension



Hot Stream (1) Properties – Shell & Tube – Property Plots

Select X variable: ● Specific enthalpy

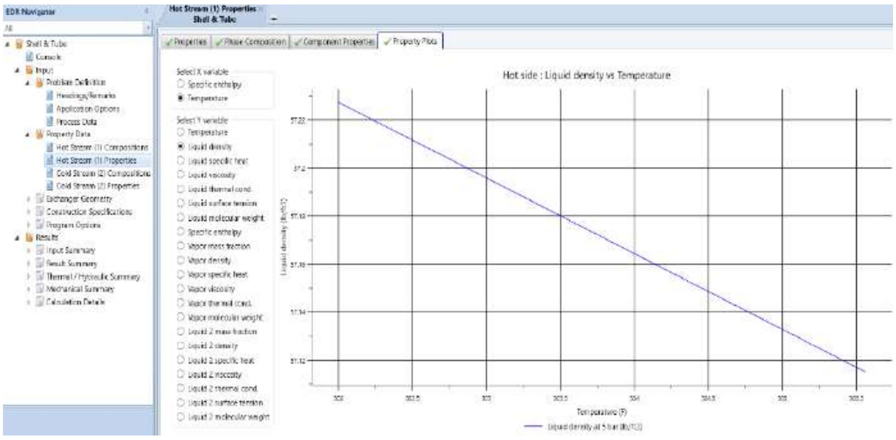
Select Y variable: ● Vapor mass fraction



Hot Stream (1) Properties – Shell & Tube – Property Plots

Select X variable: ● Specific enthalpy

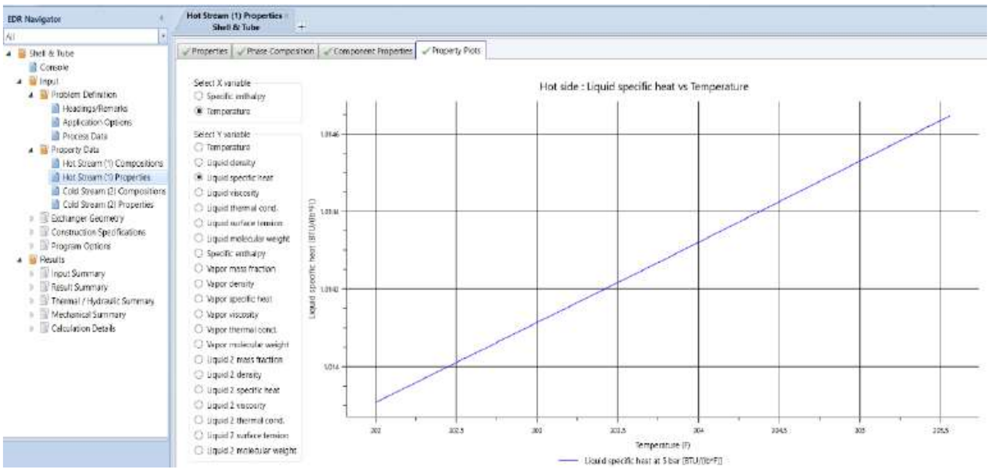
Select Y variable: ● Vapor density



Hot Stream (1) Properties
Shell & Tube

Select X variable: ● Temperature

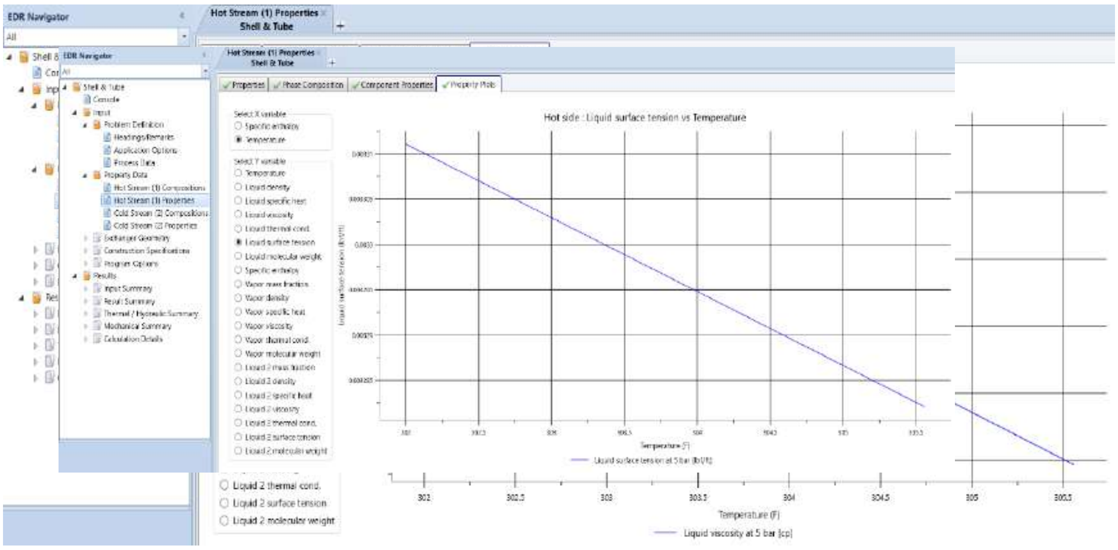
Select Y variable: ● Liquid density



Hot Stream (1) Properties
Shell & Tube

Select X variable: ● Temperature

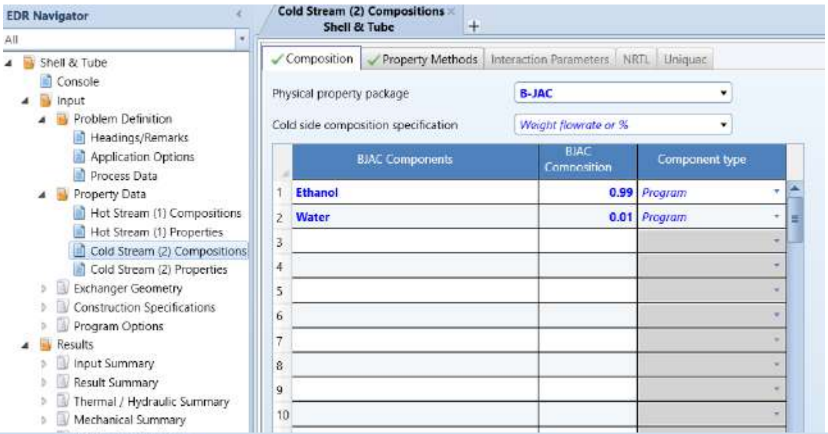
Select Y variable: ● Liquid specific heat



Hot Stream (1) Properties
Shell & Tube

Select X variable: ● Temperature

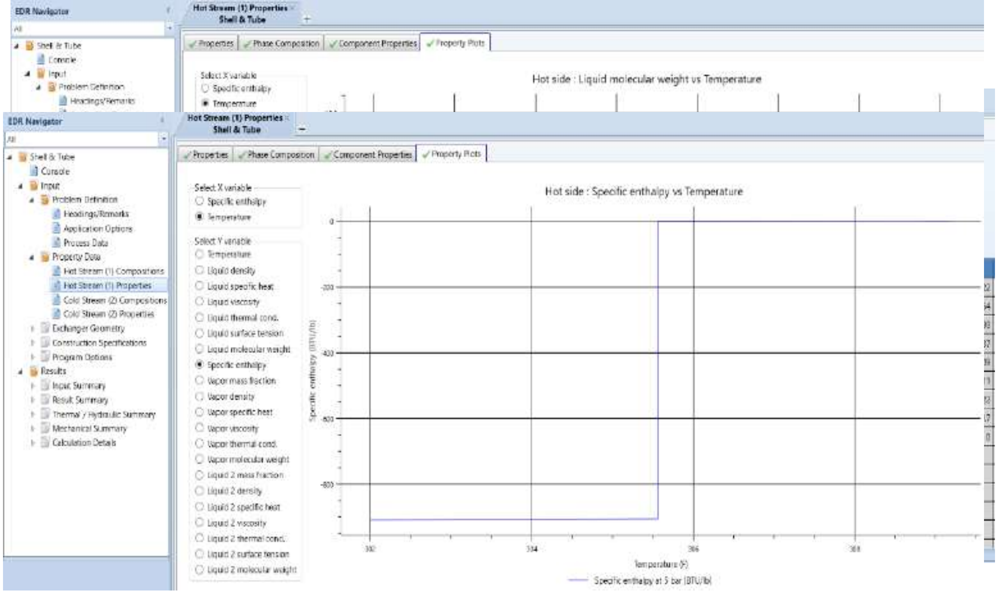
Select Y variable: ● Liquid surface tension



Physical property package: B-JAC

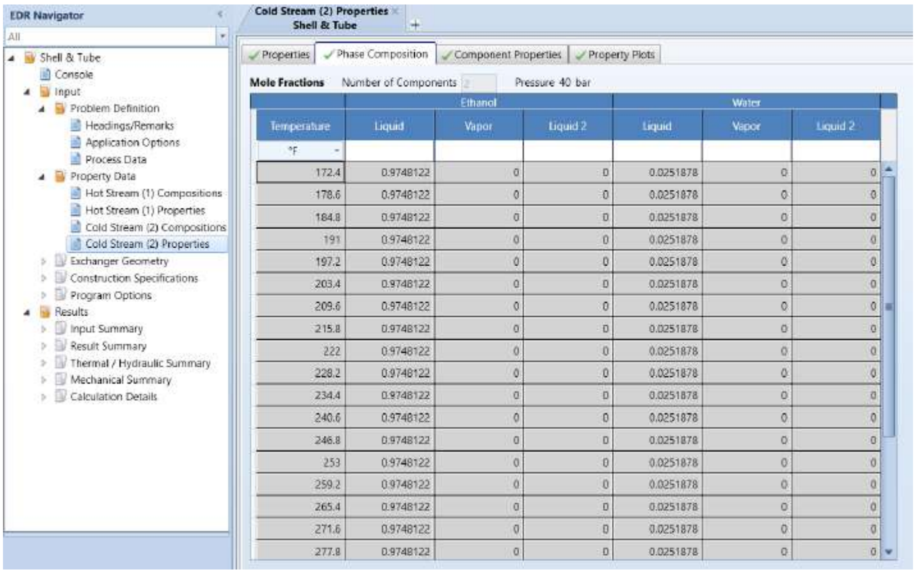
Cold side composition specification: Weight flowrate or %

	BJAC Components	BJAC Composition	Component type
1	**Ethanol**	**0.99**	*Program*
2	**Water**	**0.01**	*Program*
3			
4			
5			
6			
7			
8			
9			
10			



Select X variable
○ Specific enthalpy
● Temperature

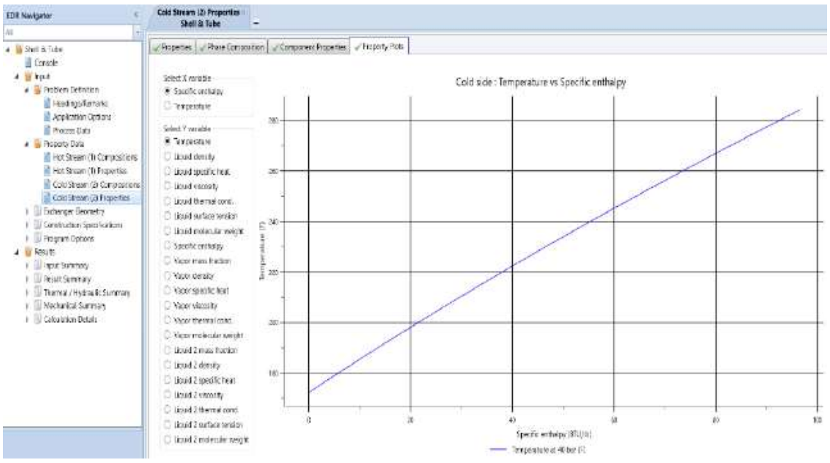
Select Y variable
○ Temperature
○ Liquid density
○ Liquid specific heat
○ Liquid viscosity
○ Liquid thermal cond.
○ Liquid surface tension
○ Liquid molecular weight
● Specific enthalpy
○ Vapor mass fraction
○ Vapor density
○ Vapor specific heat
○ Vapor viscosity
○ Vapor thermal cond.
○ Vapor molecular weight
○ Liquid 2 mass fraction
○ Liquid 2 density
○ Liquid 2 specific heat
○ Liquid 2 viscosity
○ Liquid 2 thermal cond.
○ Liquid 2 surface tension
○ Liquid 2 molecular weight



Mole Fractions Number of Components 2 Pressure 40 bar

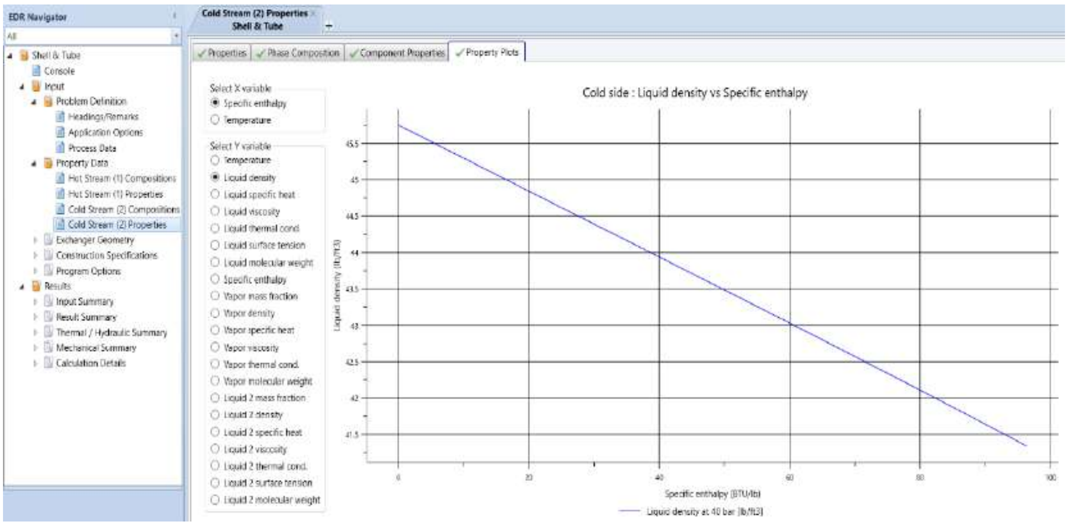
	Ethanol			Water		
Temperature	Liquid	Vapor	Liquid 2	Liquid	Vapor	Liquid 2
°F						
172.4	0.9748122	0	0	0.0251878	0	0
178.6	0.9748122	0	0	0.0251878	0	0
184.8	0.9748122	0	0	0.0251878	0	0
191	0.9748122	0	0	0.0251878	0	0
197.2	0.9748122	0	0	0.0251878	0	0
203.4	0.9748122	0	0	0.0251878	0	0
209.6	0.9748122	0	0	0.0251878	0	0
215.8	0.9748122	0	0	0.0251878	0	0
222	0.9748122	0	0	0.0251878	0	0
228.2	0.9748122	0	0	0.0251878	0	0
234.4	0.9748122	0	0	0.0251878	0	0
240.6	0.9748122	0	0	0.0251878	0	0
246.8	0.9748122	0	0	0.0251878	0	0
253	0.9748122	0	0	0.0251878	0	0
259.2	0.9748122	0	0	0.0251878	0	0
265.4	0.9748122	0	0	0.0251878	0	0
271.6	0.9748122	0	0	0.0251878	0	0
277.8	0.9748122	0	0	0.0251878	0	0

	Components	Molecular weight	Critical pressure	Critical temperature	Liquid molar volume at BP	Normal boiling point	Acentric factor	Dipole moment, debye
			psi	°F	ft³/lbmole	°F		
1	**Ethanol**	46.06999	924	469.93	1.01	173.1	0.6579835	
2	**Water**	18.00999	3206.2	705.43	0.301	212	0.346378	
3								
4								
5								
6								
7								
8								
9								
10								
11								
12								
13								



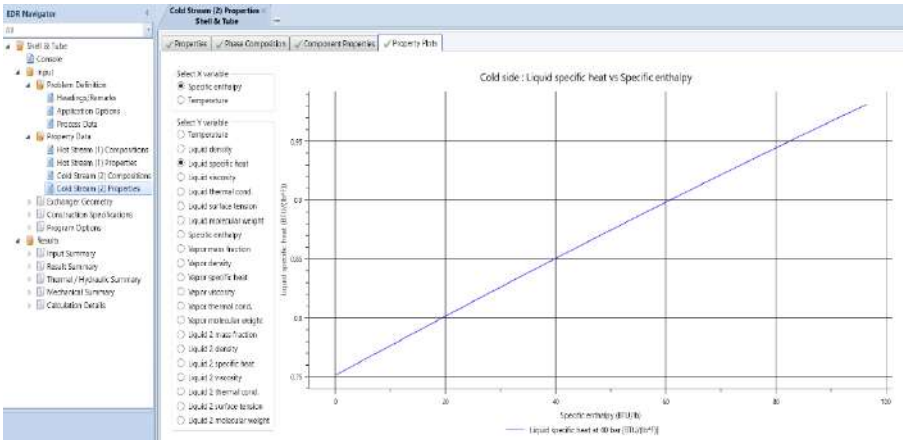
Select X variable: ● Specific enthalpy ○ Temperature

Select Y variable: ● Temperature



Select X variable: ● Specific enthalpy ○ Temperature

Select Y variable: ● Liquid density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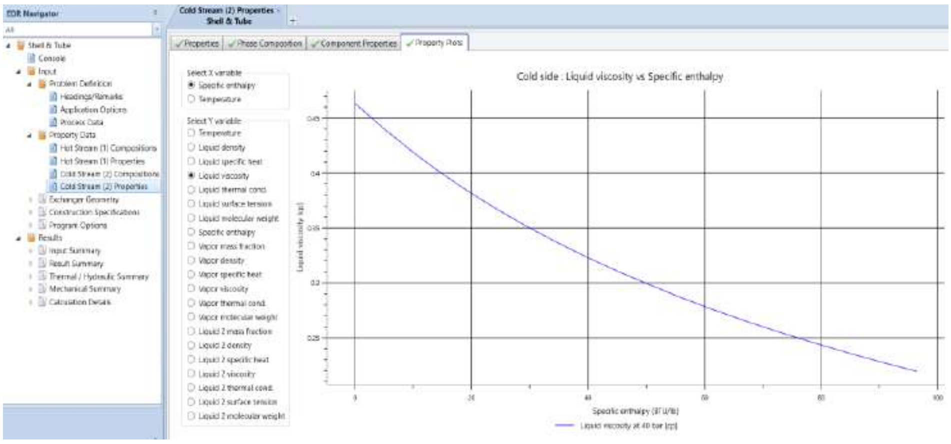
Select X variable
● Specific enthalpy
○ Temperature
Select Y variable
○ Temperature
○ Liquid density
● Liquid specific heat
○ Liquid viscosity
○ Liquid thermal cond.
○ Liquid surface tension
○ Liquid molecular weight
○ Specific enthalpy
○ Vapor mass fraction
○ Vapor density
○ Vapor specific heat
○ Vapor viscosity
○ Vapor thermal cond.
○ Vapor molecular weight
○ Liquid 2 mass fraction
○ Liquid 2 density
○ Liquid 2 specific heat
○ Liquid 2 viscosity
○ Liquid 2 thermal cond.
○ Liquid 2 surface tension
○ Liquid 2 molecular weight
Cold side : Liquid specific heat vs Specific enthalpy
0.95
0.9
0.85
0.8
0.75
Liquid specific heat (BTU/(lb*F))
0
20
40
60
80
100
Specific enthalpy (BTU/lb)
Liquid specific heat at 40 bar [BTU/(lb*F)]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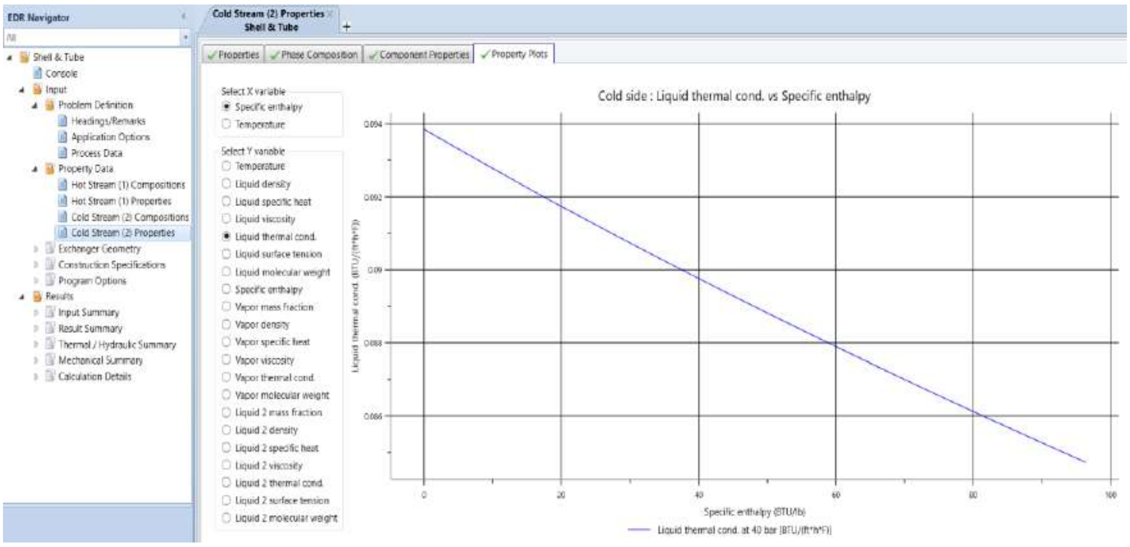
● Specific enthalpy
○ Temperature
Select Y variable
○ Temperature
○ Liquid density
○ Liquid specific heat
● Liquid viscosity
○ Liquid thermal cond.
○ Liquid surface tension
○ Liquid molecular weight
○ Specific enthalpy
○ Vapor mass fraction
○ Vapor density
○ Vapor specific heat
○ Vapor viscosity
○ Vapor thermal cond.
○ Vapor molecular weight
○ Liquid 2 mass fraction
○ Liquid 2 density
○ Liquid 2 specific heat
○ Liquid 2 viscosity
○ Liquid 2 thermal cond.
○ Liquid 2 surface tension
○ Liquid 2 molecular weight
Cold side : Liquid viscosity vs Specific enthalpy
0.45
0.4
0.35
0.3
0.25
Liquid viscosity (cp)
0
20
40
60
80
100
Specific enthalpy (BTU/lb)
Liquid viscosity at 40 bar [cp]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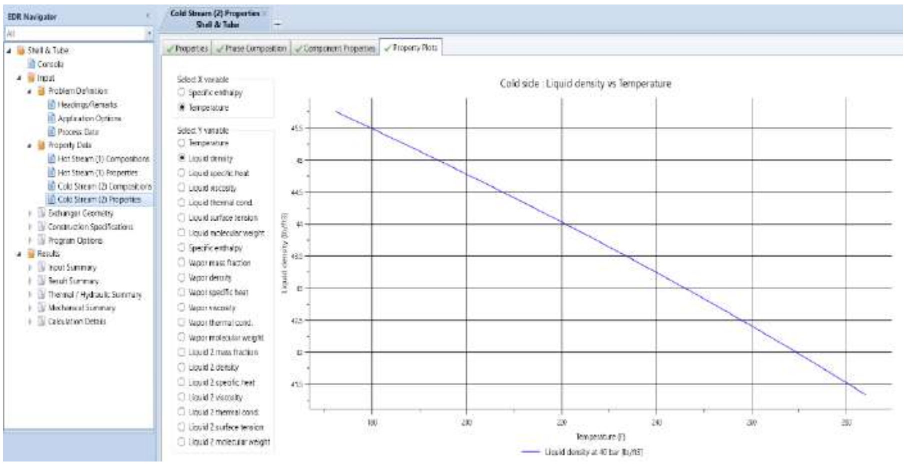
✓ Properties
✓ Phase Composition
✓ Component Properties
✓ Property Plots
Select X variable
● Specific enthalpy
○ Temperature
Select Y variable
○ Temperature
○ Liquid density
○ Liquid specific heat
○ Liquid viscosity
● Liquid thermal cond.
○ Liquid surface tension
○ Liquid molecular weight
○ Specific enthalpy
○ Vapor mass fraction
○ Vapor density
○ Vapor specific heat
○ Vapor viscosity
○ Vapor thermal cond.
○ Vapor molecular weight
○ Liquid 2 mass fraction
○ Liquid 2 density
○ Liquid 2 specific heat
○ Liquid 2 viscosity
○ Liquid 2 thermal cond.
○ Liquid 2 surface tension
○ Liquid 2 molecular weight
Cold side : Liquid thermal cond. vs Specific enthalpy
0.094
0.092
0.09
0.088
0.086
Liquid thermal cond. (BTU/(ft*h*F))
0
20
40
60
80
100
Specific enthalpy (BTU/lb)
Liquid thermal cond. at 40 bar [BTU/(ft*h*F)]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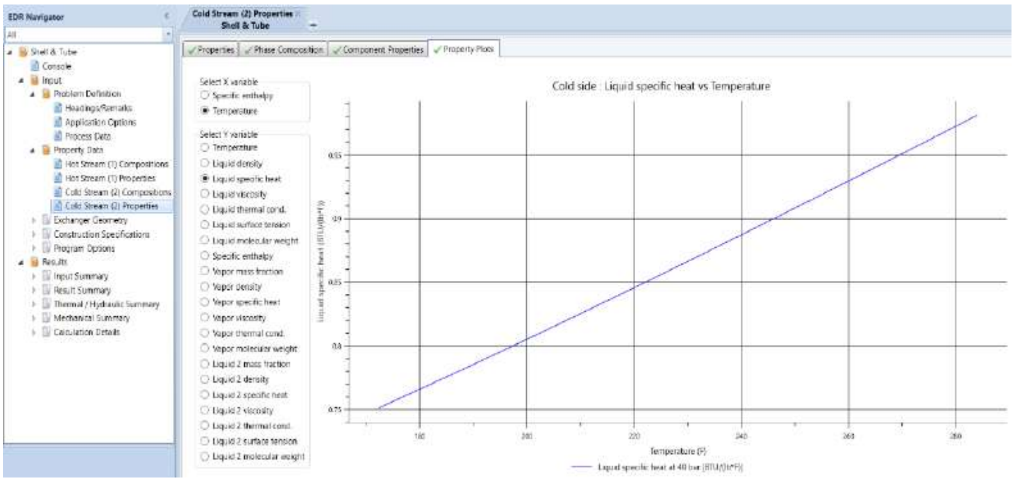
✓ Property Plots
Select X variable
○ Specific enthalpy
● Temperature
Select Y variable
○ Temperature
● Liquid density
○ Liquid specific heat
○ Liquid viscosity
○ Liquid thermal cond.
○ Liquid surface tension
○ Liquid molecular weight
○ Specific enthalpy
○ Vapor mass fraction
○ Vapor density
○ Vapor specific heat
○ Vapor viscosity
○ Vapor thermal cond.
○ Vapor molecular weight
○ Liquid 2 mass fraction
○ Liquid 2 density
○ Liquid 2 specific heat
○ Liquid 2 viscosity
○ Liquid 2 thermal cond.
○ Liquid 2 surface tension
○ Liquid 2 molecular weight
Cold side : Liquid density vs Temperature
Liquid density (lb/ft3)
45.5
45
44.5
44
43.5
43
42.5
42
41.5
180
200
220
240
260
280
Temperature (F)
Liquid density at 40 bar (lb/ft3)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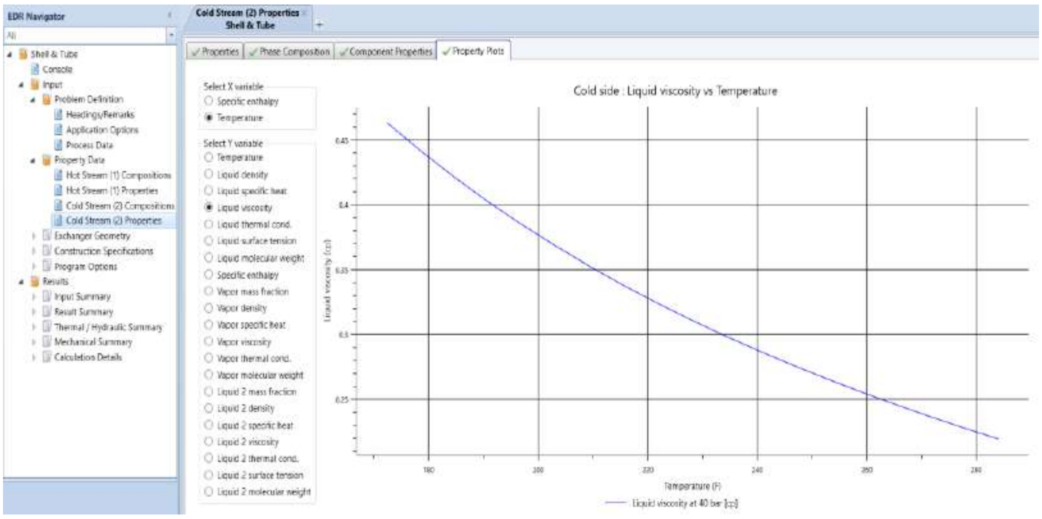
Select X variable
○ Specific enthalpy
● Temperature
Select Y variable
○ Temperature
○ Liquid density
● Liquid specific heat
○ Liquid viscosity
○ Liquid thermal cond.
○ Liquid surface tension
○ Liquid molecular weight
○ Specific enthalpy
○ Vapor mass fraction
○ Vapor density
○ Vapor specific heat
○ Vapor viscosity
○ Vapor thermal cond.
○ Vapor molecular weight
○ Liquid 2 mass fraction
○ Liquid 2 density
○ Liquid 2 specific heat
○ Liquid 2 viscosity
○ Liquid 2 thermal cond.
○ Liquid 2 surface tension
○ Liquid 2 molecular weight
Cold side : Liquid specific heat vs Temperature
Liquid specific heat (BTU/(lb*F))
0.95
0.9
0.85
0.8
0.75
180
200
220
240
260
280
Temperature (F)
Liquid specific heat at 40 bar (BTU/(lb*F))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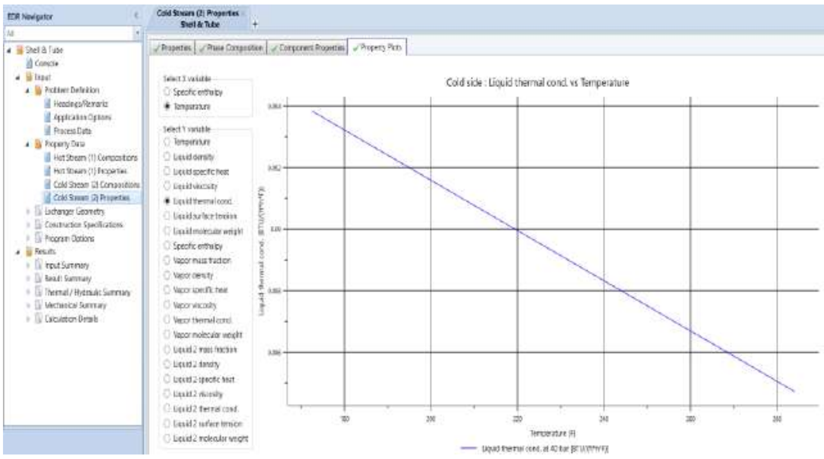
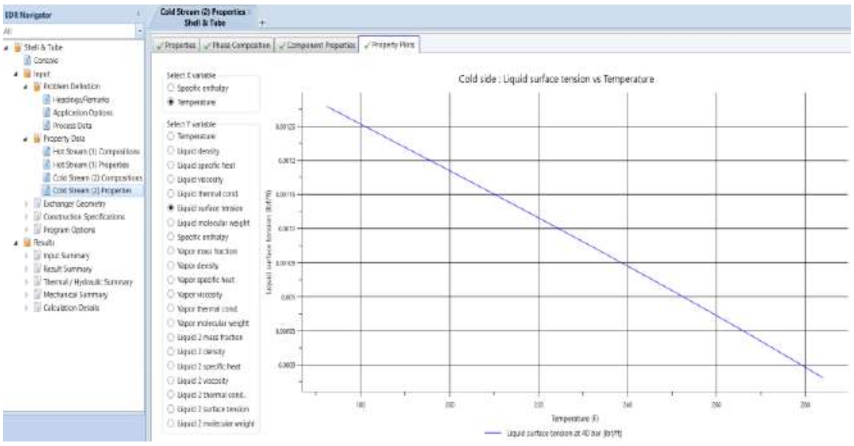
Shell & Tube
✓ Properties
✓ Phase Composition
✓ Component Properties
✓ Property Plots
Select X variable
○ Specific enthalpy
● Temperature
Select Y variable
○ Temperature
○ Liquid density
○ Liquid specific heat
● Liquid viscosity
○ Liquid thermal cond.
○ Liquid surface tension
○ Liquid molecular weight
○ Specific enthalpy
○ Vapor mass fraction
○ Vapor density
○ Vapor specific heat
○ Vapor viscosity
○ Vapor thermal cond.
○ Vapor molecular weight
○ Liquid 2 mass fraction
○ Liquid 2 density
○ Liquid 2 specific heat
○ Liquid 2 viscosity
○ Liquid 2 thermal cond.
○ Liquid 2 surface tension
○ Liquid 2 molecular weight
Cold side : Liquid viscosity vs Temperature
Liquid viscosity (cp)
0.45
0.4
0.35
0.3
0.25
180
200
220
240
260
280
Temperature (F)
Liquid viscosity at 40 bar (cp)



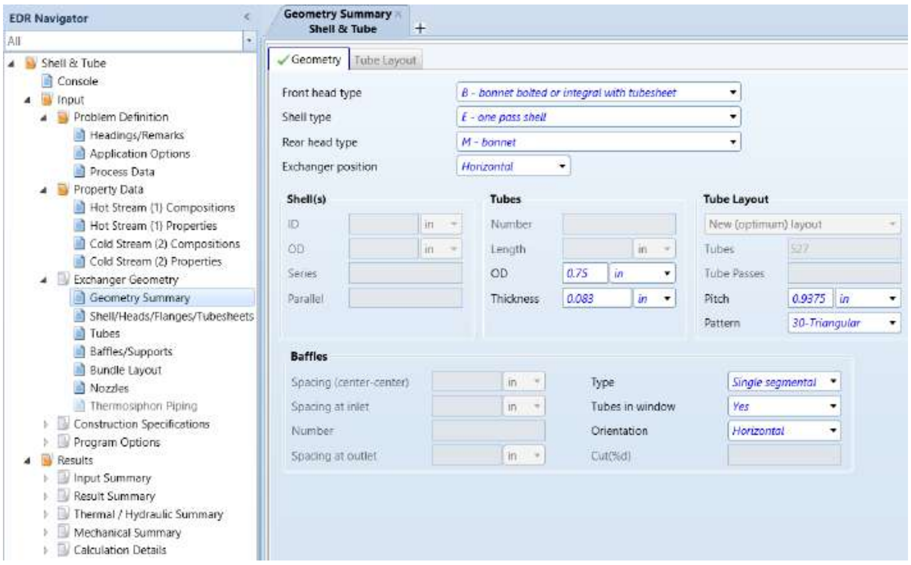
Select X variable: ● Temperature

Select Y variable: ● Liquid thermal cond.



Select X variable: ● Temperature

Select Y variable: ● Liquid surface tension



Front head type: B - bonnet bolted or integral with tubesheet

Shell type: E - one pass shell

Rear head type: M - bonnet

Exchanger position: Horizontal

Shell(s)

ID: in

OD: in

Series:

Parallel:

Tubes

Number:

Length: in

OD: 0.75 in

Thickness: 0.083 in

Tube Layout

New (optimum) layout

Tubes: 527

Tube Passes:

Pitch: 0.9375 in

Pattern: 30-Triangular

Baffles

Spacing (center-center): in

Spacing at inlet: in

Number:

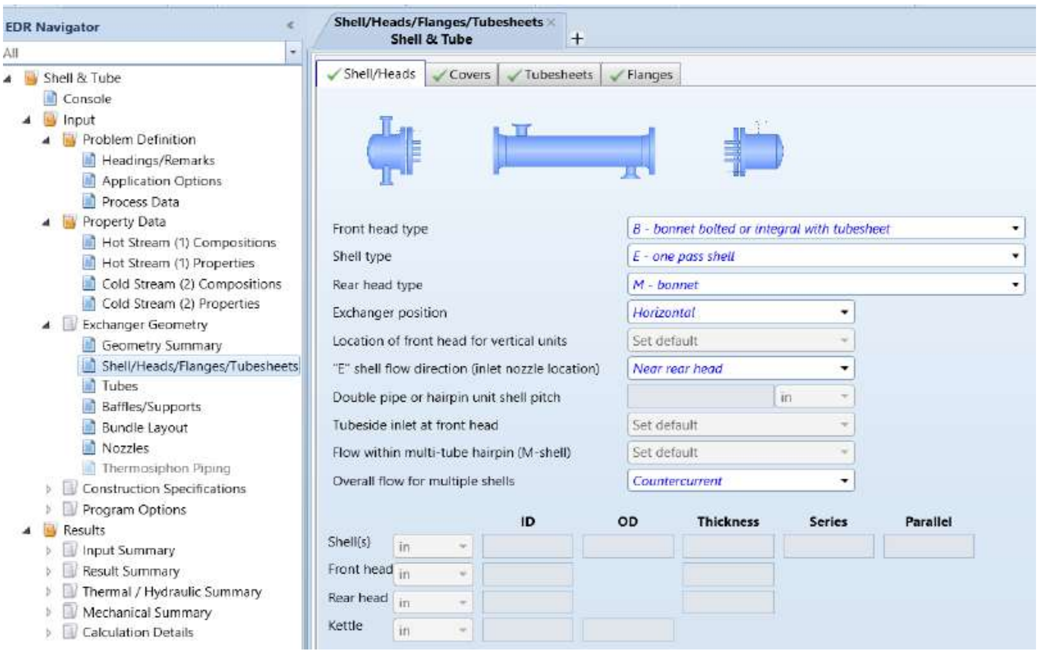
Spacing at outlet: in

Type: Single segmental

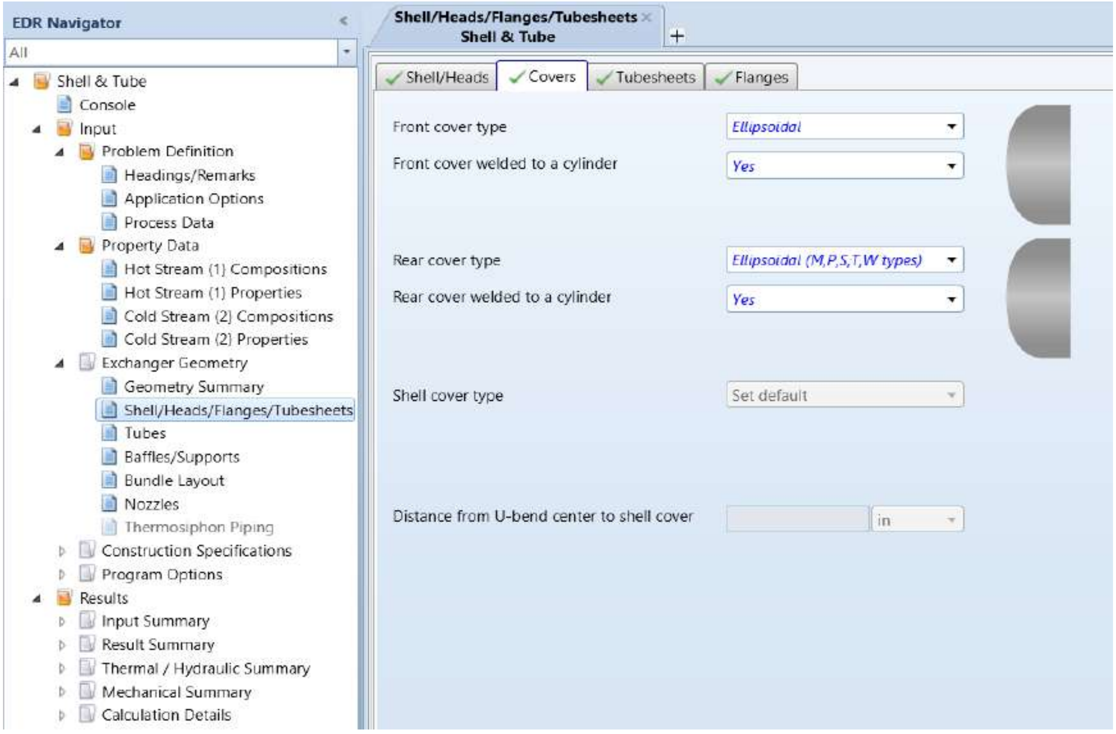
Tubes in window: Yes

Orientation: Horizontal

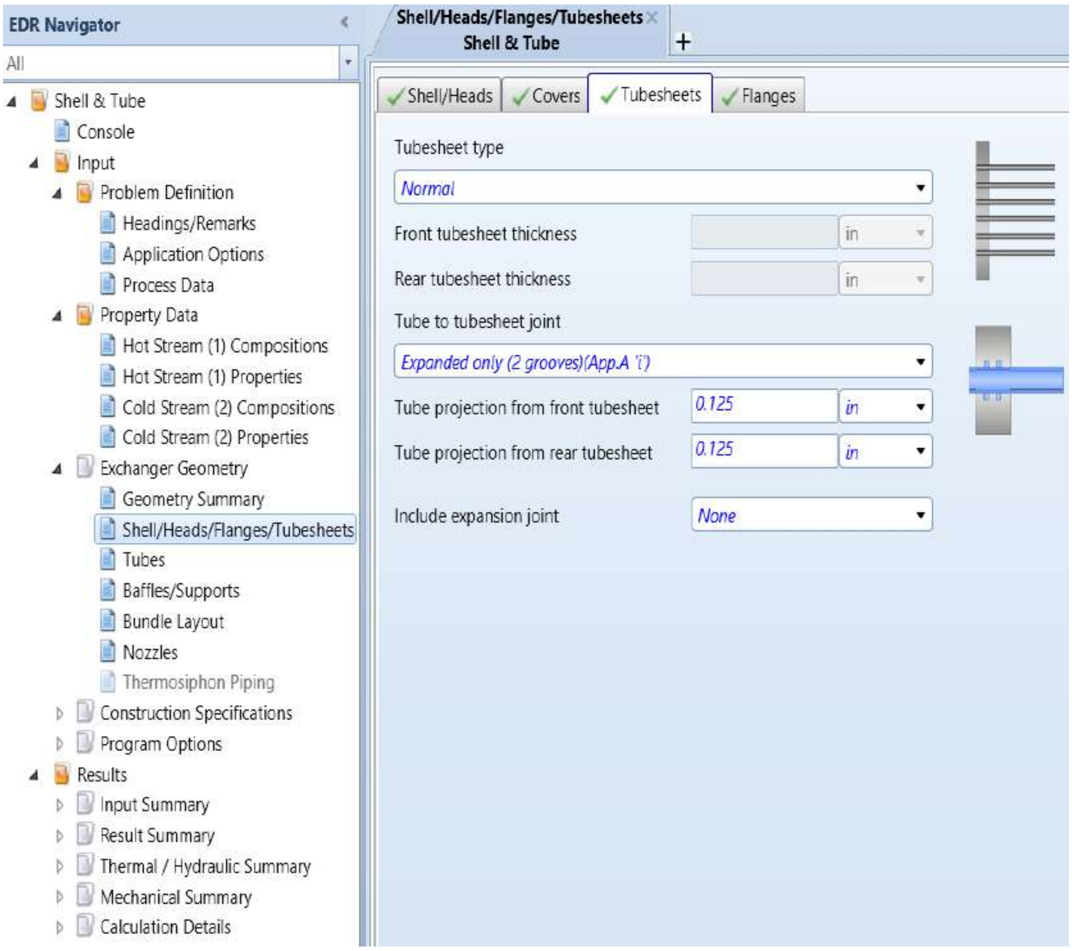
Cut(%d):



Front head type: B - bonnet bolted or integral with tubesheet
Shell type: E - one pass shell
Rear head type: M - bonnet
Exchanger position: Horizontal
Location of front head for vertical units: Set default
"E" shell flow direction (inlet nozzle location): Near rear head
Double pipe or hairpin unit shell pitch: in
Tubeside inlet at front head: Set default
Flow within multi-tube hairpin (M-shell): Set default
Overall flow for multiple shells: Countercurrent



Front cover type: Ellipsoidal
Front cover welded to a cylinder: Yes
Rear cover type: Ellipsoidal (M,P,S,T,W types)
Rear cover welded to a cylinder: Yes
Shell cover type: Set default
Distance from U-bend center to shell cover: in



Tubesheet type: Normal

Front tubesheet thickness: in

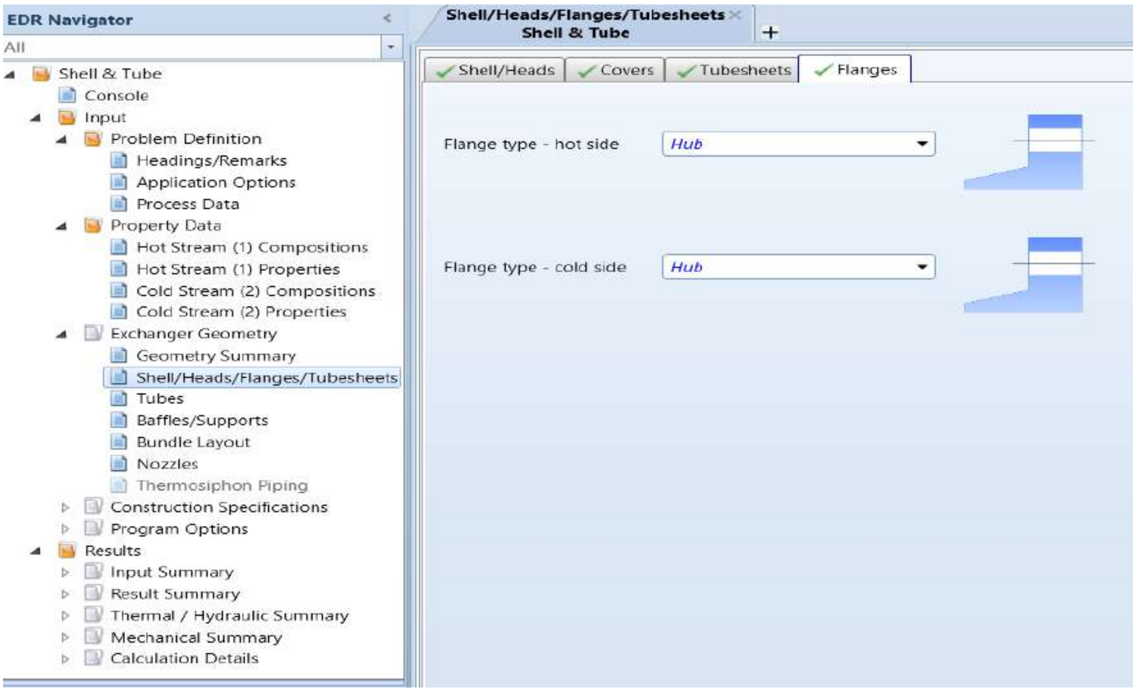
Rear tubesheet thickness: in

Tube to tubesheet joint: Expanded only (2 grooves)(App.A 'i')

Tube projection from front tubesheet: 0.125 in

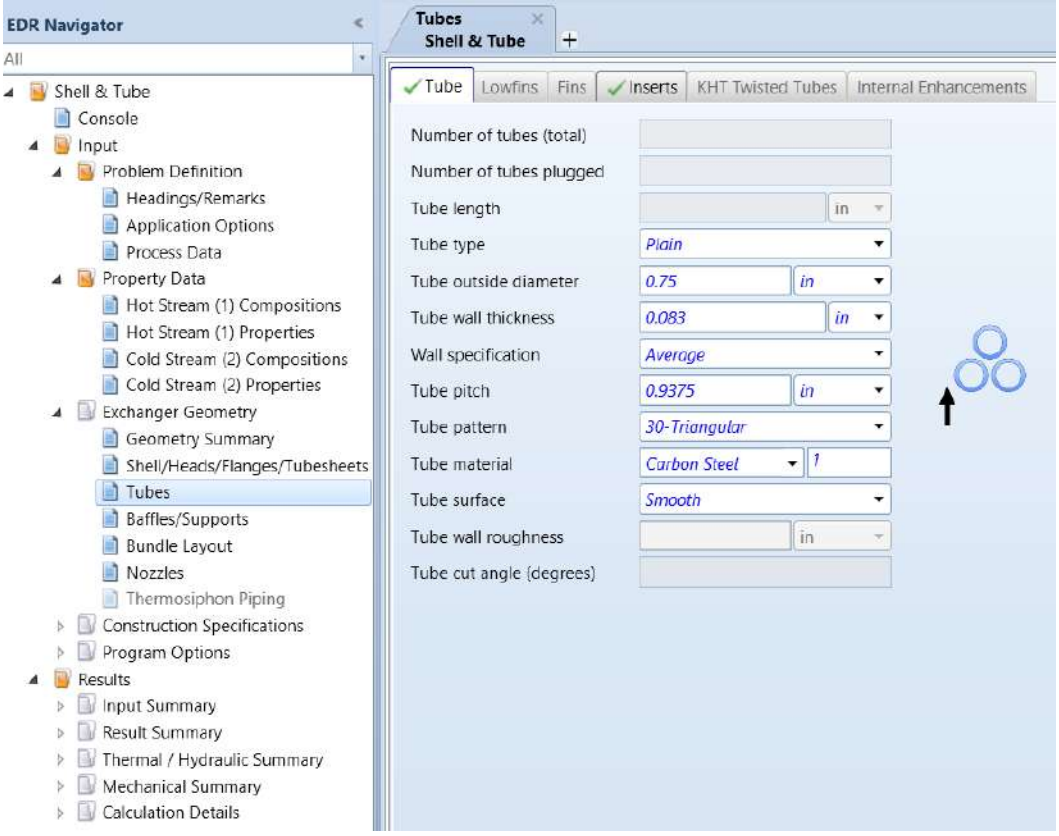
Tube projection from rear tubesheet: 0.125 in

Include expansion joint: None

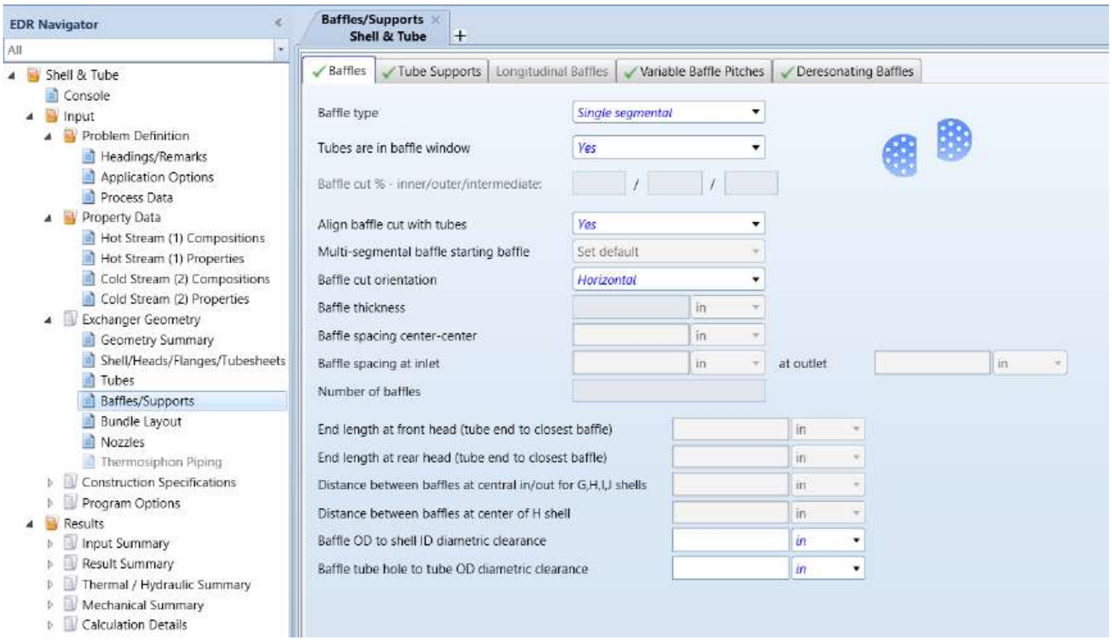


Flange type - hot side: Hub

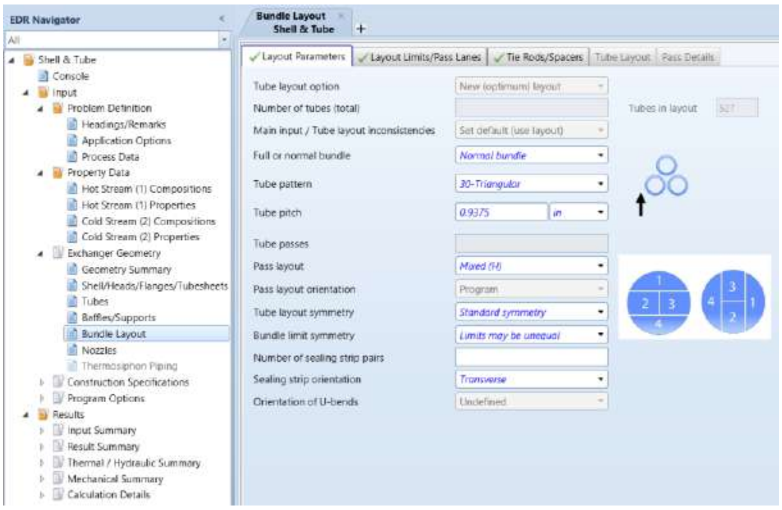
Flange type - cold side: Hub



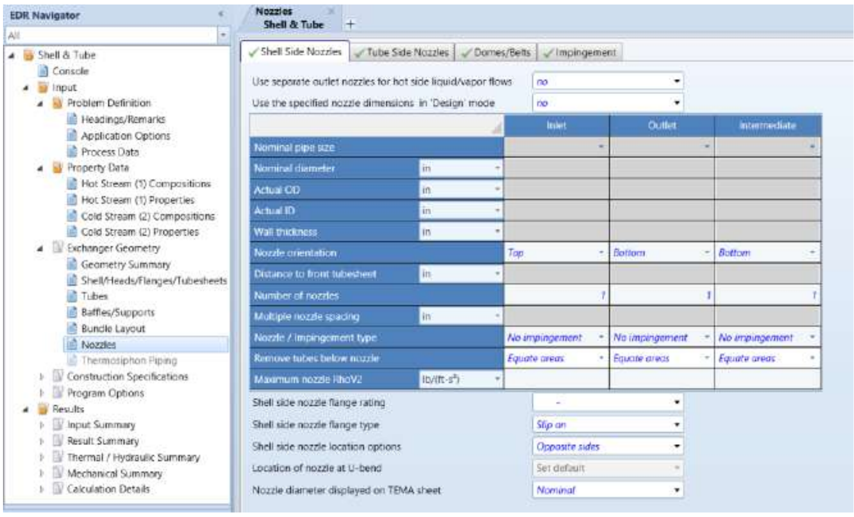
Number of tubes (total)
Number of tubes plugged
Tube length in
Tube type Plain
Tube outside diameter 0.75 in
Tube wall thickness 0.083 in
Wall specification Average
Tube pitch 0.9375 in
Tube pattern 30-Triangular
Tube material Carbon Steel 1
Tube surface Smooth
Tube wall roughness in
Tube cut angle (degrees)



Baffle type Single segmental
Tubes are in baffle window Yes
Baffle cut % - inner/outer/intermediate: / /
Align baffle cut with tubes Yes
Multi-segmental baffle starting baffle Set default
Baffle cut orientation Horizontal
Baffle thickness in
Baffle spacing center-center in
Baffle spacing at inlet in at outlet in
Number of baffles
End length at front head (tube end to closest baffle) in
End length at rear head (tube end to closest baffle) in
Distance between baffles at central in/out for G,H,I,J shells in
Distance between baffles at center of H shell in
Baffle OD to shell ID diametric clearance in
Baffle tube hole to tube OD diametric clearance in



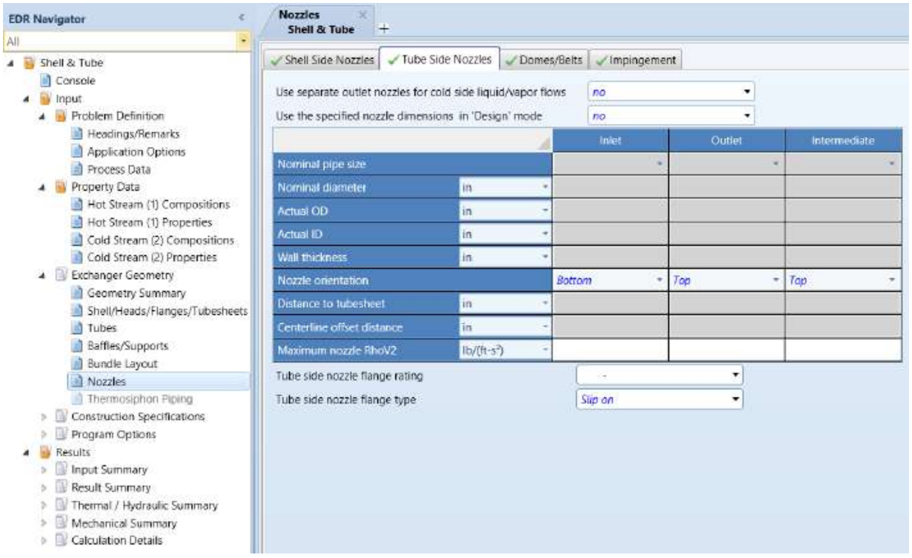
Tube layout option: New (optimum) layout
Number of tubes (total):
Tubes in layout: 527
Main input / Tube layout inconsistencies: Set default (use layout)
Full or normal bundle: Normal bundle
Tube pattern: 30-Triangular
Tube pitch: 0.9375 in
Tube passes:
Pass layout: Mixed (H)
Pass layout orientation: Program
Tube layout symmetry: Standard symmetry
Bundle limit symmetry: Limits may be unequal
Number of sealing strip pairs:
Sealing strip orientation: Transverse
Orientation of U-bends: Undefined



Use separate outlet nozzles for hot side liquid/vapor flows: no
Use the specified nozzle dimensions in 'Design' mode: no

		Inlet	Outlet	Intermediate
Nominal pipe size				
Nominal diameter	in			
Actual OD	in			
Actual ID	in			
Wall thickness	in			
Nozzle orientation		Top	Bottom	Bottom
Distance to front tubesheet	in			
Number of nozzles		1	1	1
Multiple nozzle spacing	in			
Nozzle / Impingement type		No impingement	No impingement	No impingement
Remove tubes below nozzle		Equate areas	Equate areas	Equate areas
Maximum nozzle RhoV2	lb/(ft-s²)			

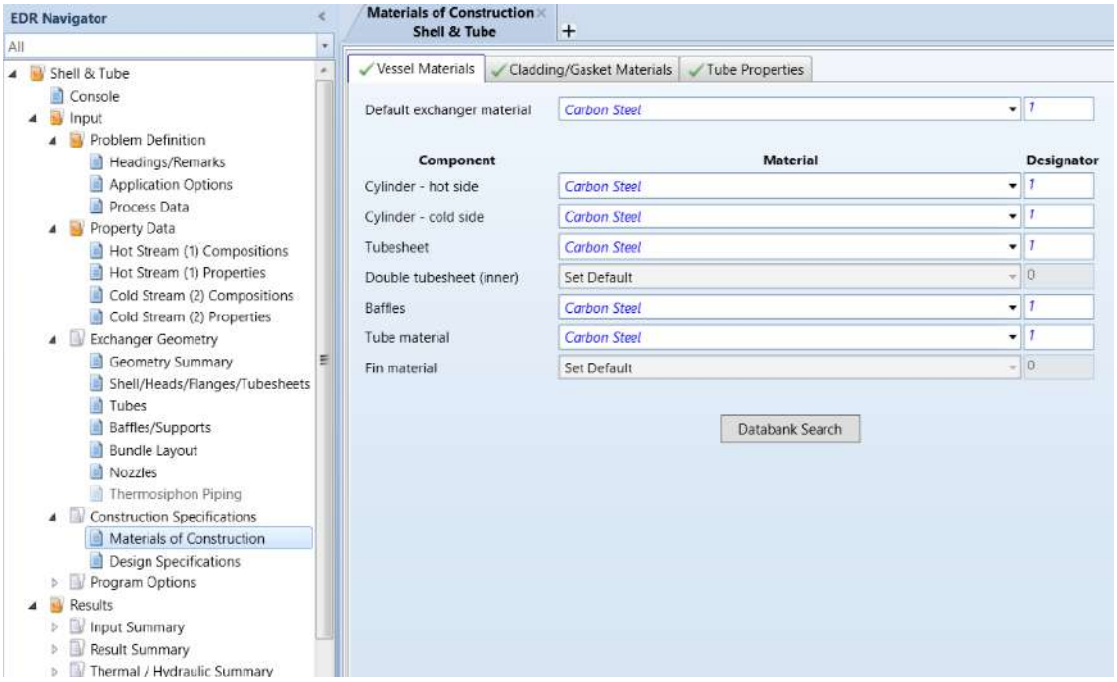
Shell side nozzle flange rating: -
Shell side nozzle flange type: Slip on
Shell side nozzle location options: Opposite sides
Location of nozzle at U-bend: Set default
Nozzle diameter displayed on TEMA sheet: Nominal



Use separate outlet nozzles for cold side liquid/vapor flows: no
Use the specified nozzle dimensions in 'Design' mode: no

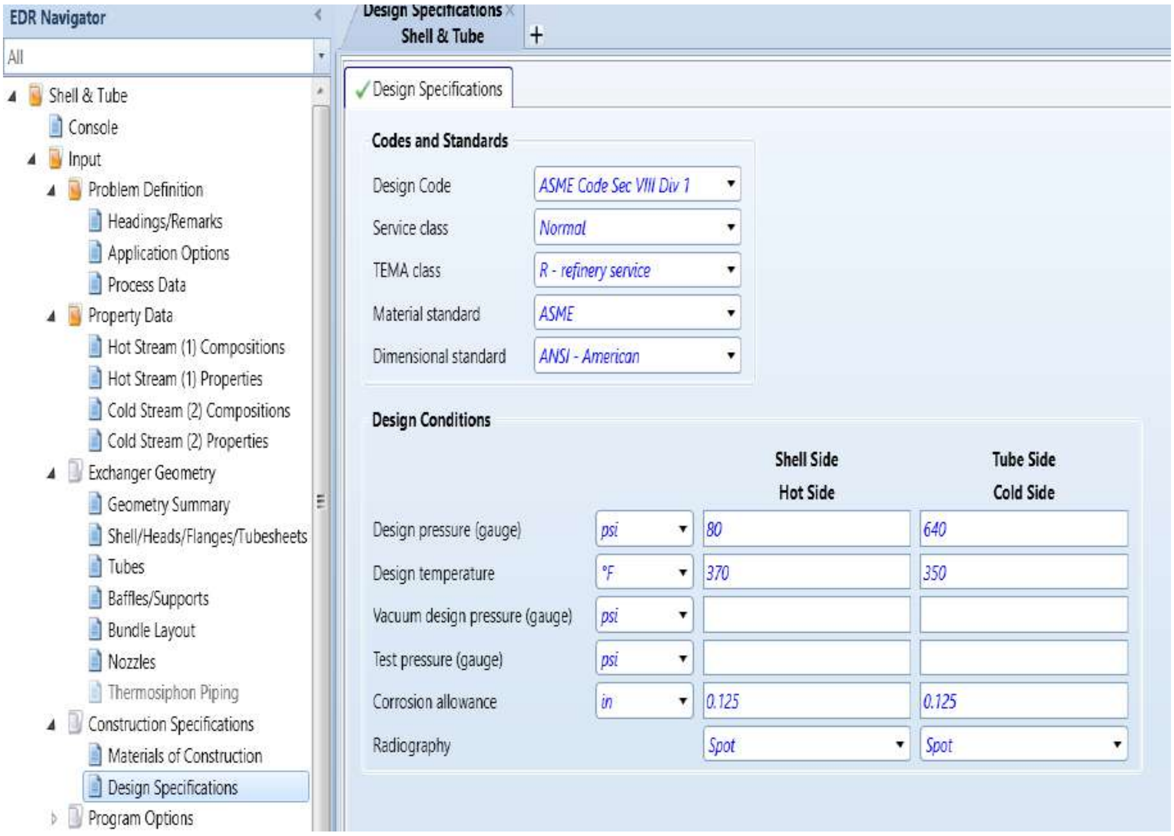
		Inlet	Outlet	Intermediate
Nominal pipe size				
Nominal diameter	in			
Actual OD	in			
Actual ID	in			
Wall thickness	in			
Nozzle orientation		Bottom	Top	Top
Distance to tubesheet	in			
Centerline offset distance	in			
Maximum nozzle RhoV2	lb/(ft-s²)			

Tube side nozzle flange rating: -
Tube side nozzle flange type: Slip on



Default exchanger material: Carbon Steel, 1

Component	Material	Designator
Cylinder - hot side	Carbon Steel	1
Cylinder - cold side	Carbon Steel	1
Tubesheet	Carbon Steel	1
Double tubesheet (inner)	Set Default	0
Baffles	Carbon Steel	1
Tube material	Carbon Steel	1
Fin material	Set Default	0



Codes and Standards

Design Code: ASME Code Sec VIII Div 1

Service class: Normal

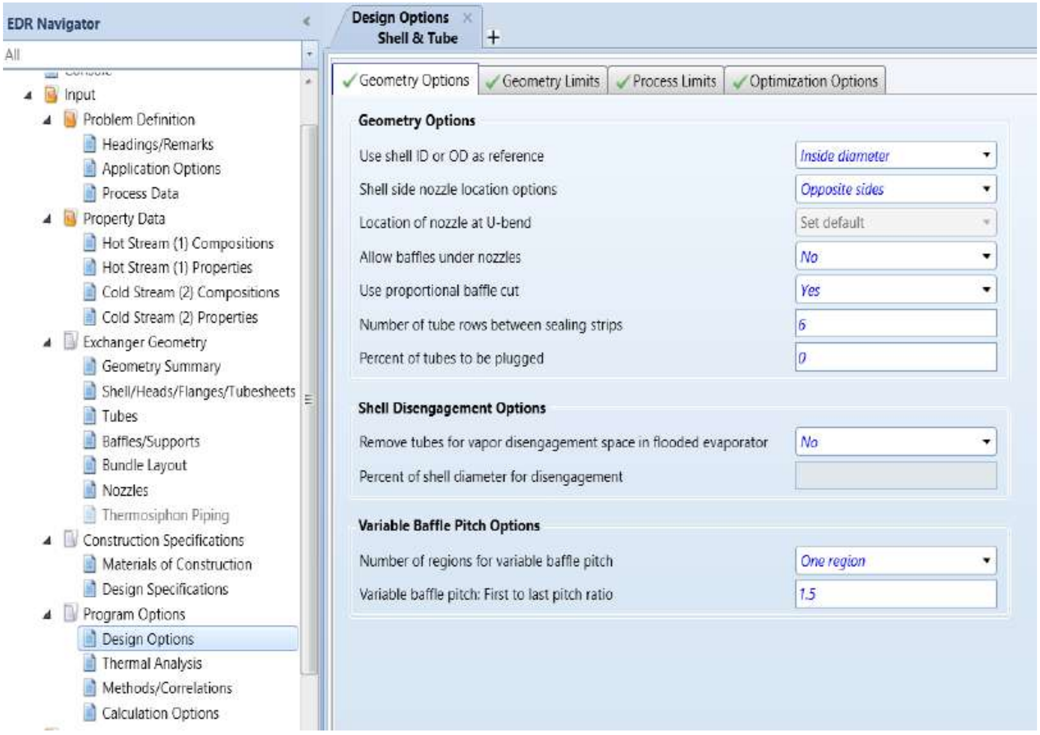
TEMA class: R - refinery service

Material standard: ASME

Dimensional standard: ANSI - American

Design Conditions

		Shell Side Hot Side	Tube Side Cold Side
Design pressure (gauge)	psi	80	640
Design temperature	°F	370	350
Vacuum design pressure (gauge)	psi		
Test pressure (gauge)	psi		
Corrosion allowance	in	0.125	0.125
Radiography		Spot	Spot



Geometry Options

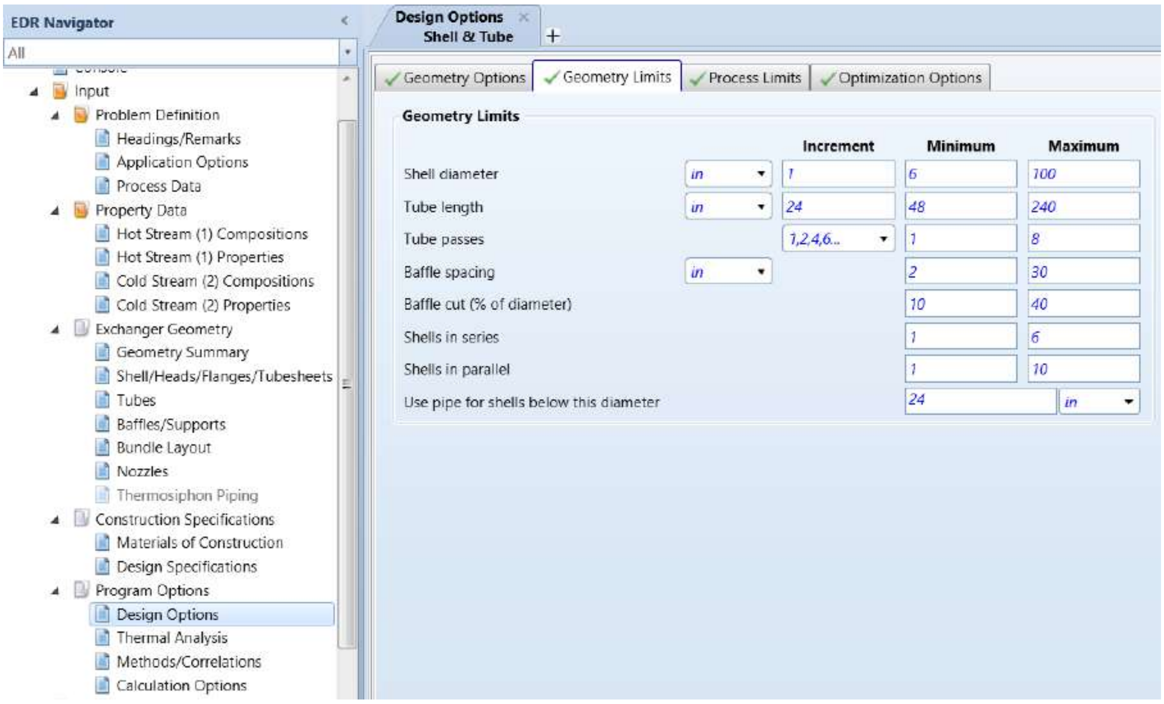
Use shell ID or OD as reference	Inside diameter
Shell side nozzle location options	Opposite sides
Location of nozzle at U-bend	Set default
Allow baffles under nozzles	No
Use proportional baffle cut	Yes
Number of tube rows between sealing strips	6
Percent of tubes to be plugged	0

Shell Disengagement Options

Remove tubes for vapor disengagement space in flooded evaporator	No
Percent of shell diameter for disengagement	

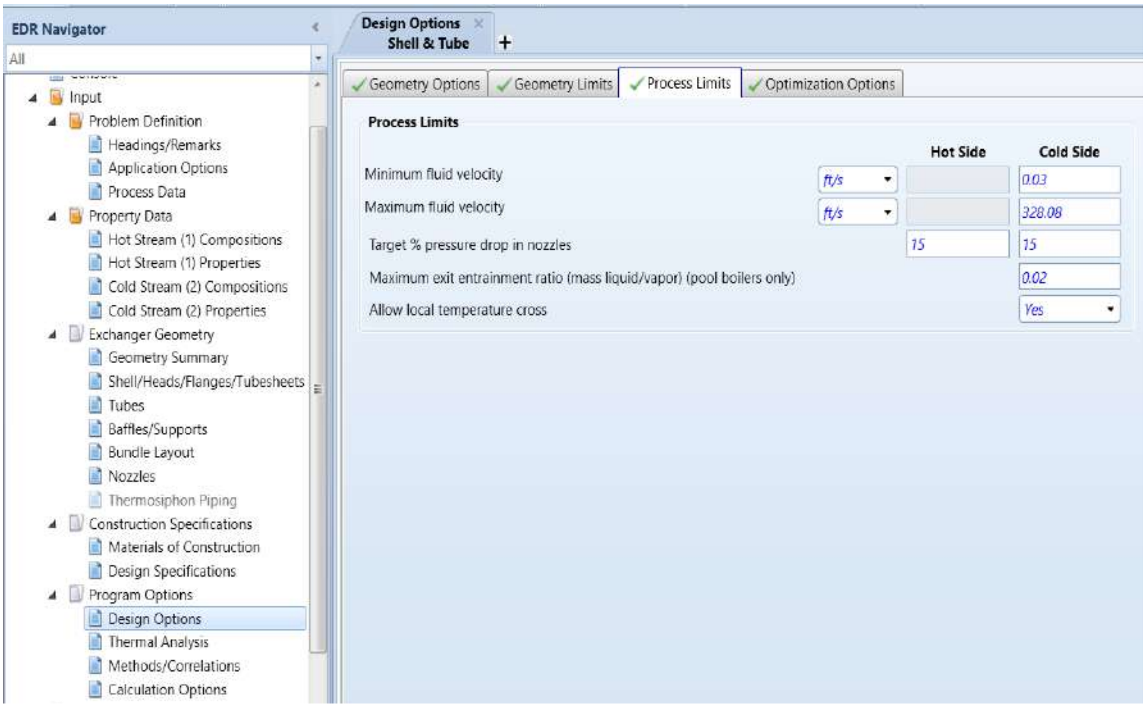
Variable Baffle Pitch Options

Number of regions for variable baffle pitch	One region
Variable baffle pitch: First to last pitch ratio	1.5



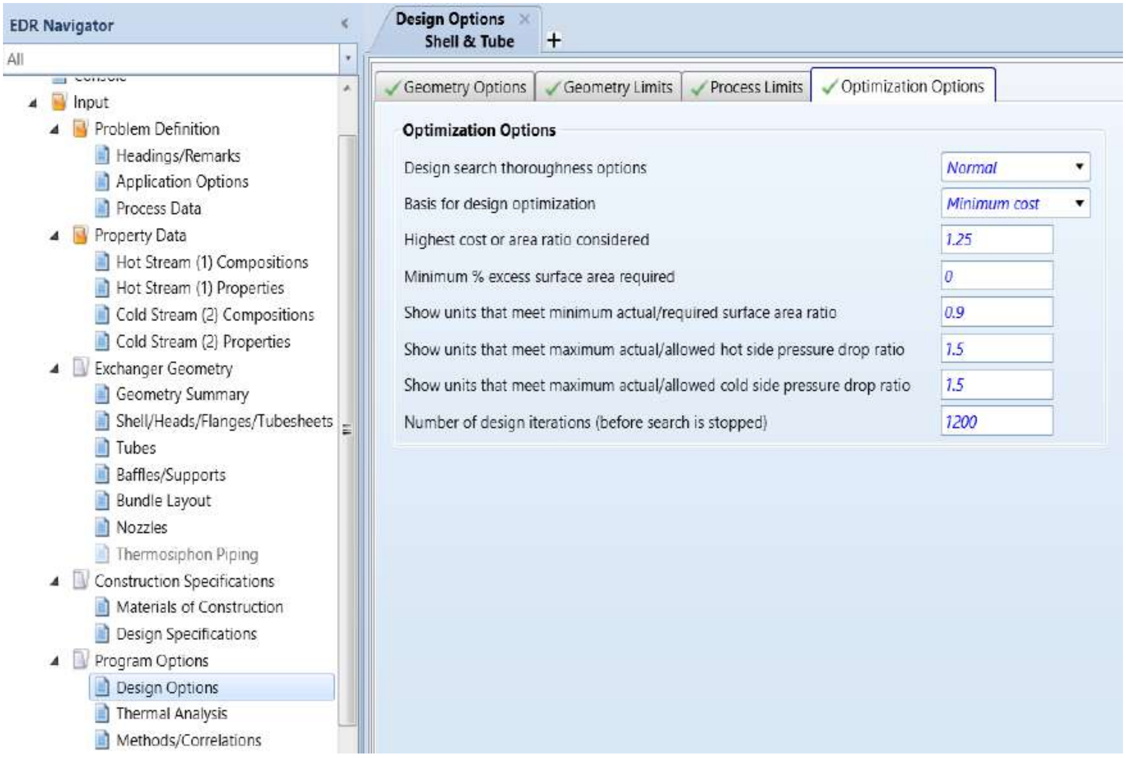
Geometry Limits

		Increment	Minimum	Maximum
Shell diameter	in	1	6	100
Tube length	in	24	48	240
Tube passes		1,2,4,6...	1	8
Baffle spacing	in		2	30
Baffle cut (% of diameter)			10	40
Shells in series			1	6
Shells in parallel			1	10
Use pipe for shells below this diameter			24	in



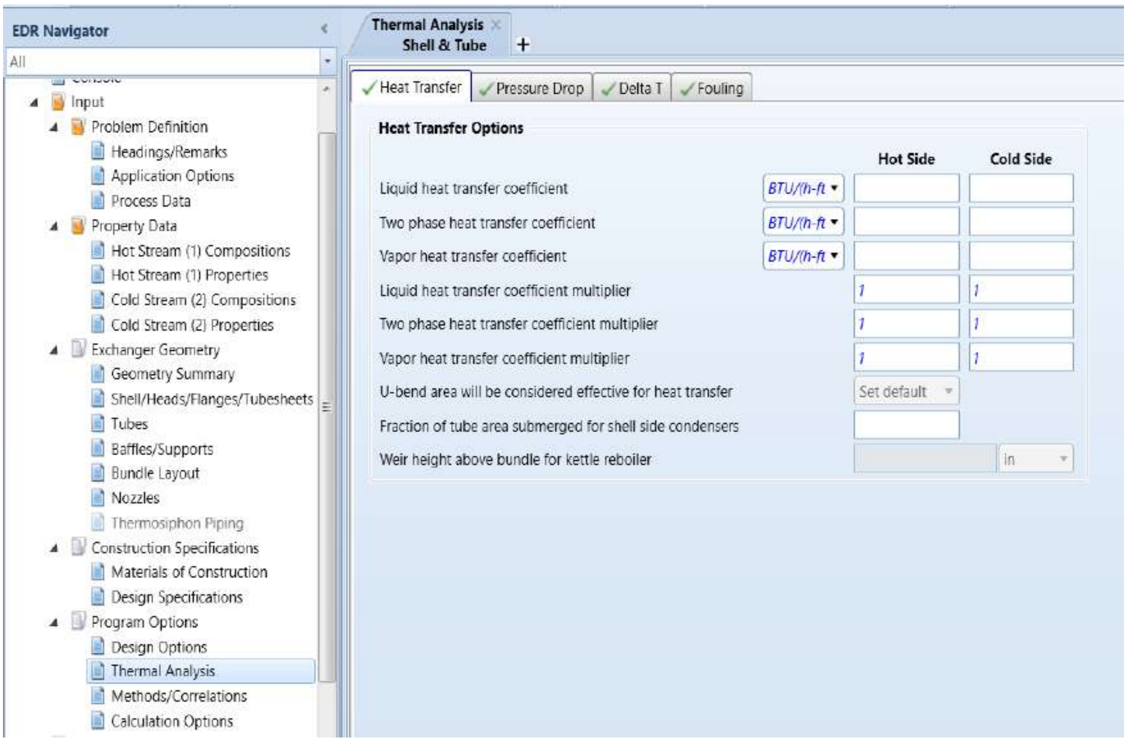
Process Limits

		Hot Side	Cold Side
Minimum fluid velocity	ft/s		0.03
Maximum fluid velocity	ft/s		328.08
Target % pressure drop in nozzles		15	15
Maximum exit entrainment ratio (mass liquid/vapor) (pool boilers only)			0.02
Allow local temperature cross			Yes



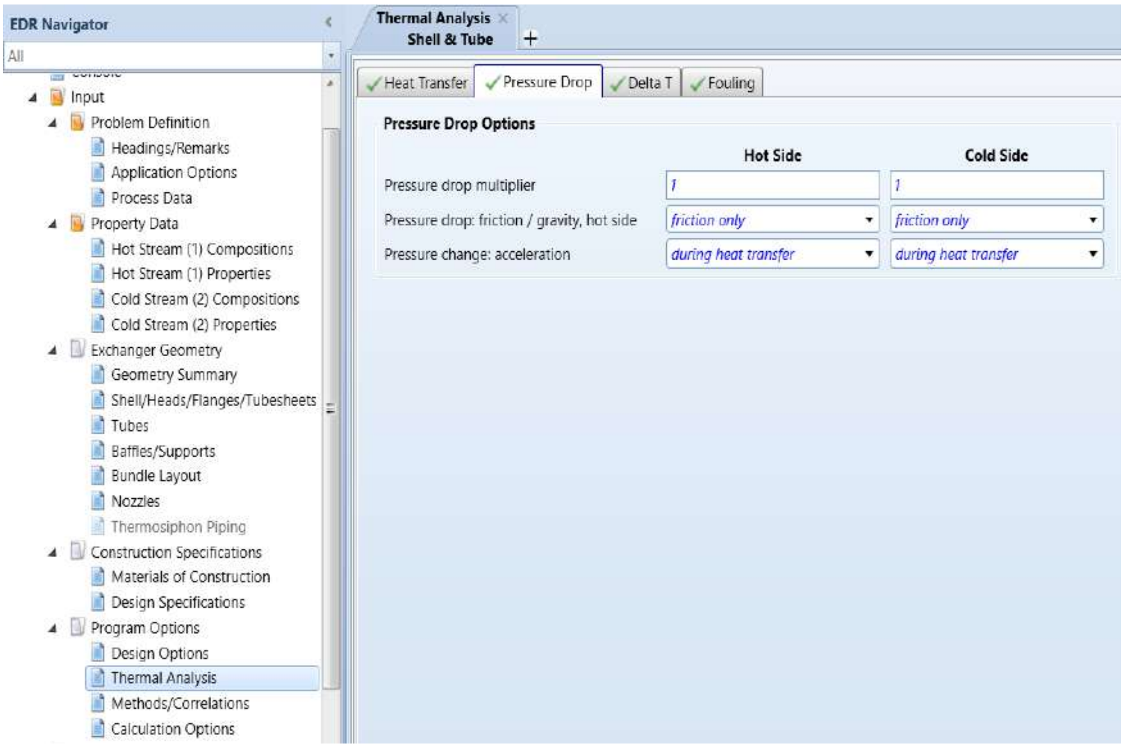
Optimization Options

Design search thoroughness options	Normal
Basis for design optimization	Minimum cost
Highest cost or area ratio considered	1.25
Minimum % excess surface area required	0
Show units that meet minimum actual/required surface area ratio	0.9
Show units that meet maximum actual/allowed hot side pressure drop ratio	1.5
Show units that meet maximum actual/allowed cold side pressure drop ratio	1.5
Number of design iterations (before search is stopped)	1200



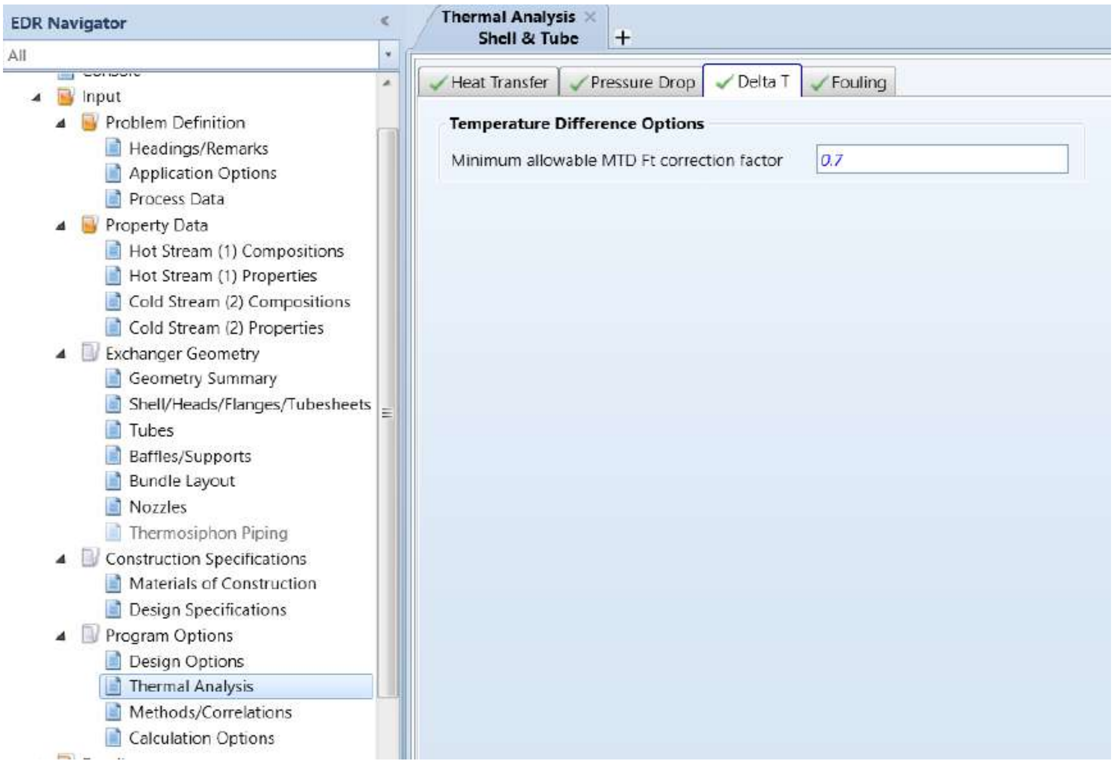
Heat Transfer Options

		Hot Side	Cold Side
Liquid heat transfer coefficient	BTU/(h-ft		
Two phase heat transfer coefficient	BTU/(h-ft		
Vapor heat transfer coefficient	BTU/(h-ft		
Liquid heat transfer coefficient multiplier		1	1
Two phase heat transfer coefficient multiplier		1	1
Vapor heat transfer coefficient multiplier		1	1
U-bend area will be considered effective for heat transfer		Set default	
Fraction of tube area submerged for shell side condensers			
Weir height above bundle for kettle reboiler			in



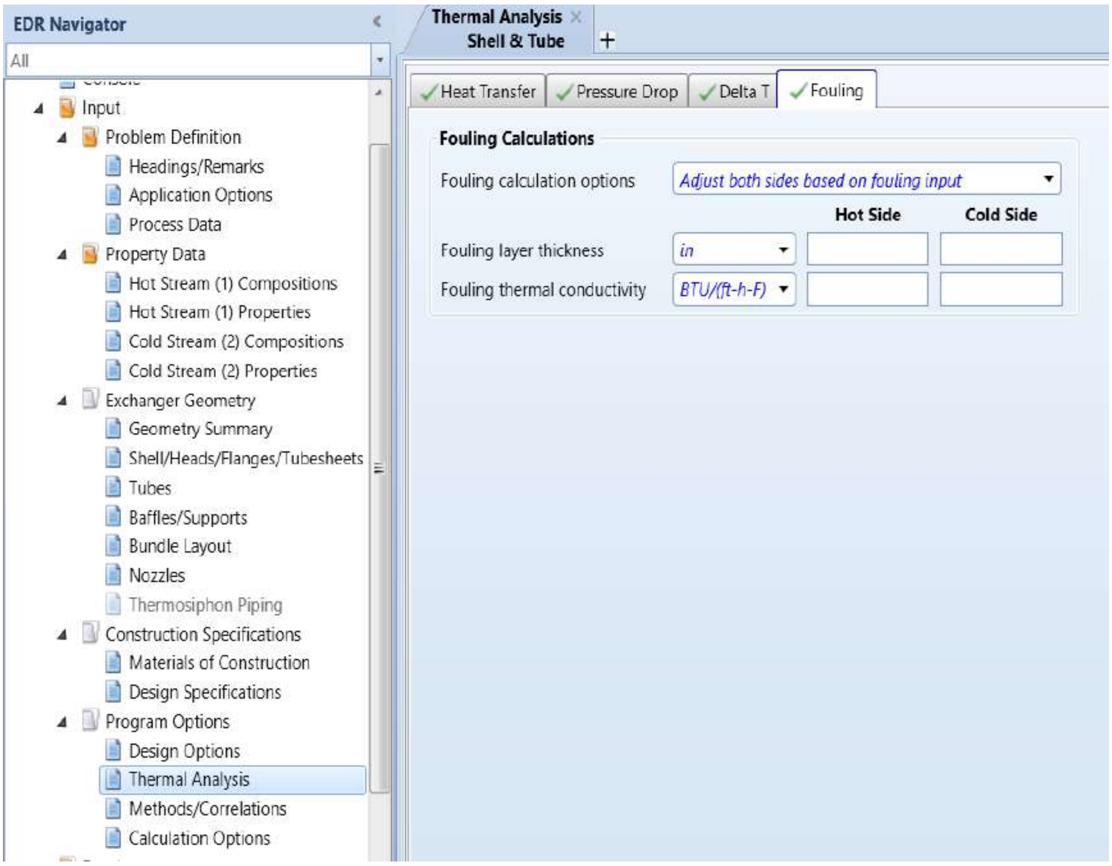
Pressure Drop Options

	Hot Side	Cold Side
Pressure drop multiplier	1	1
Pressure drop: friction / gravity, hot side	friction only	friction only
Pressure change: acceleration	during heat transfer	during heat transfer



Temperature Difference Options

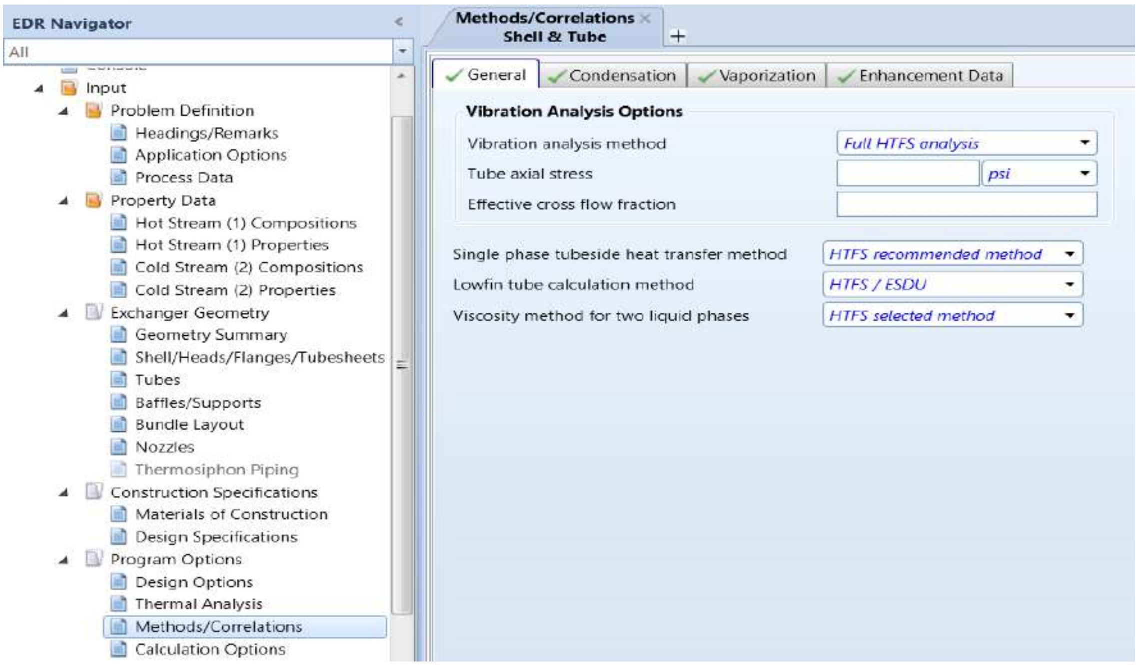
Minimum allowable MTD Ft correction factor: 0.7



Fouling Calculations

Fouling calculation options: Adjust both sides based on fouling input

	Unit	Hot Side	Cold Side
Fouling layer thickness	in		
Fouling thermal conductivity	BTU/(ft-h-F)		



Vibration Analysis Options

Vibration analysis method: *Full HTFS analysis*

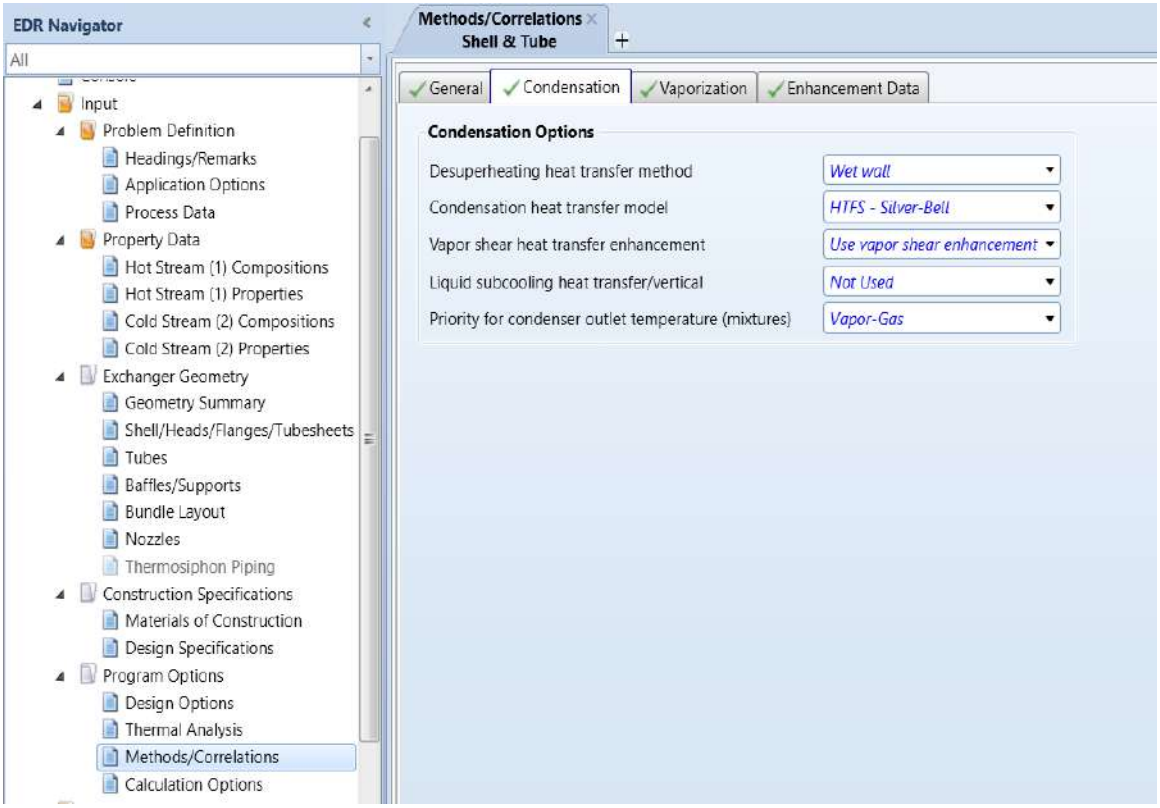
Tube axial stress: *psi*

Effective cross flow fraction:

Single phase tubeside heat transfer method: *HTFS recommended method*

Lowfin tube calculation method: *HTFS / ESDU*

Viscosity method for two liquid phases: *HTFS selected method*



Condensation Options

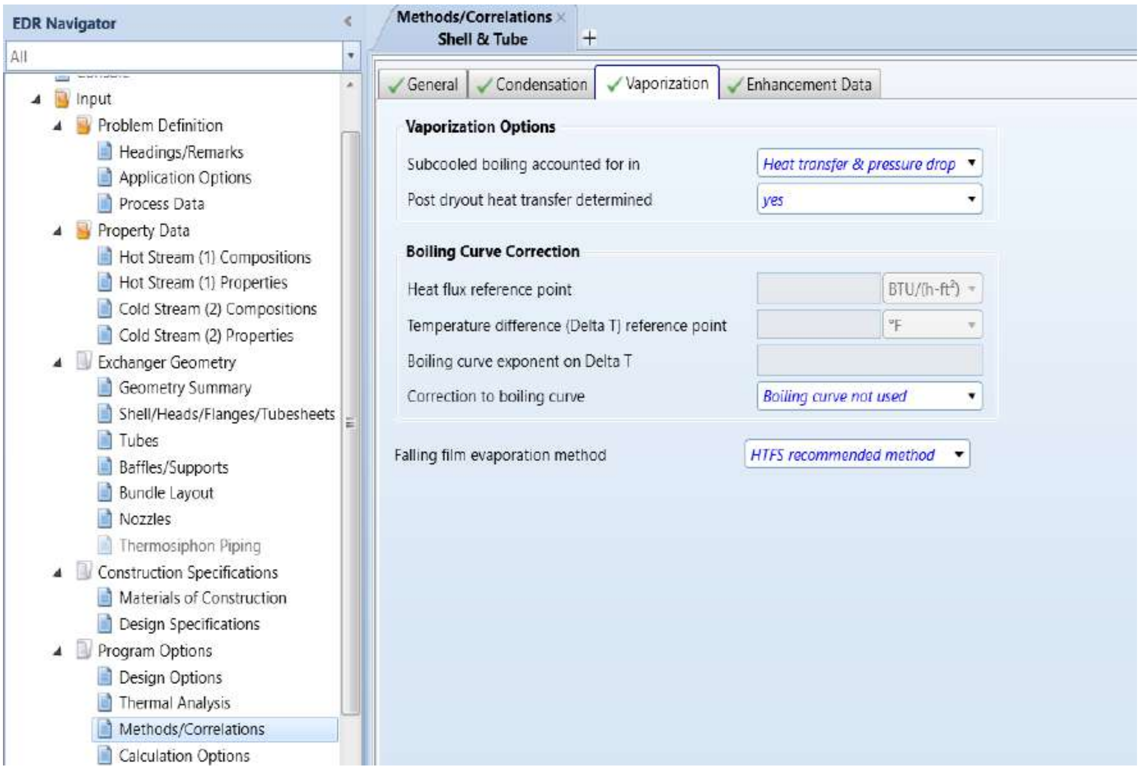
Desuperheating heat transfer method: *Wet wall*

Condensation heat transfer model: *HTFS - Silver-Bell*

Vapor shear heat transfer enhancement: *Use vapor shear enhancement*

Liquid subcooling heat transfer/vertical: *Not Used*

Priority for condenser outlet temperature (mixtures): *Vapor-Gas*



Vaporization Options

Subcooled boiling accounted for in: Heat transfer & pressure drop

Post dryout heat transfer determined: yes

Boiling Curve Correction

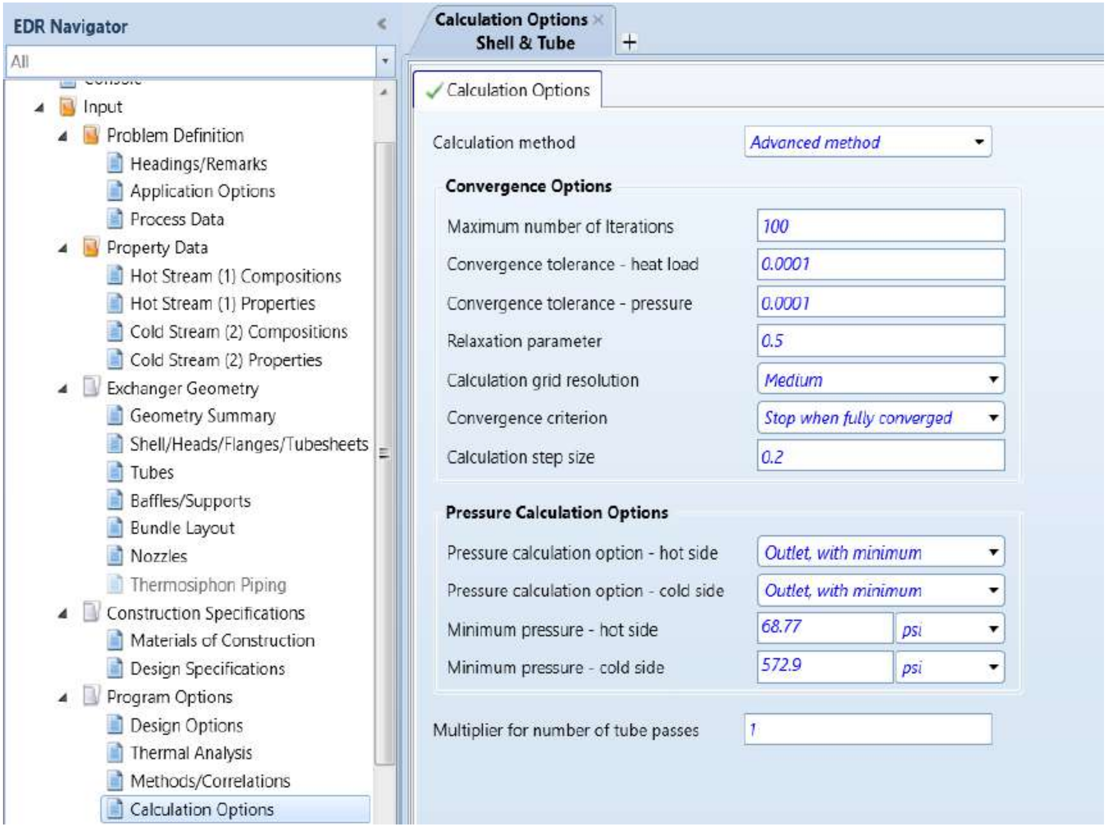
Heat flux reference point: BTU/(h-ft²)

Temperature difference (Delta T) reference point: °F

Boiling curve exponent on Delta T

Correction to boiling curve: Boiling curve not used

Falling film evaporation method: HTFS recommended method



Calculation method: Advanced method

Convergence Options

Maximum number of Iterations: 100

Convergence tolerance - heat load: 0.0001

Convergence tolerance - pressure: 0.0001

Relaxation parameter: 0.5

Calculation grid resolution: Medium

Convergence criterion: Stop when fully converged

Calculation step size: 0.2

Pressure Calculation Options

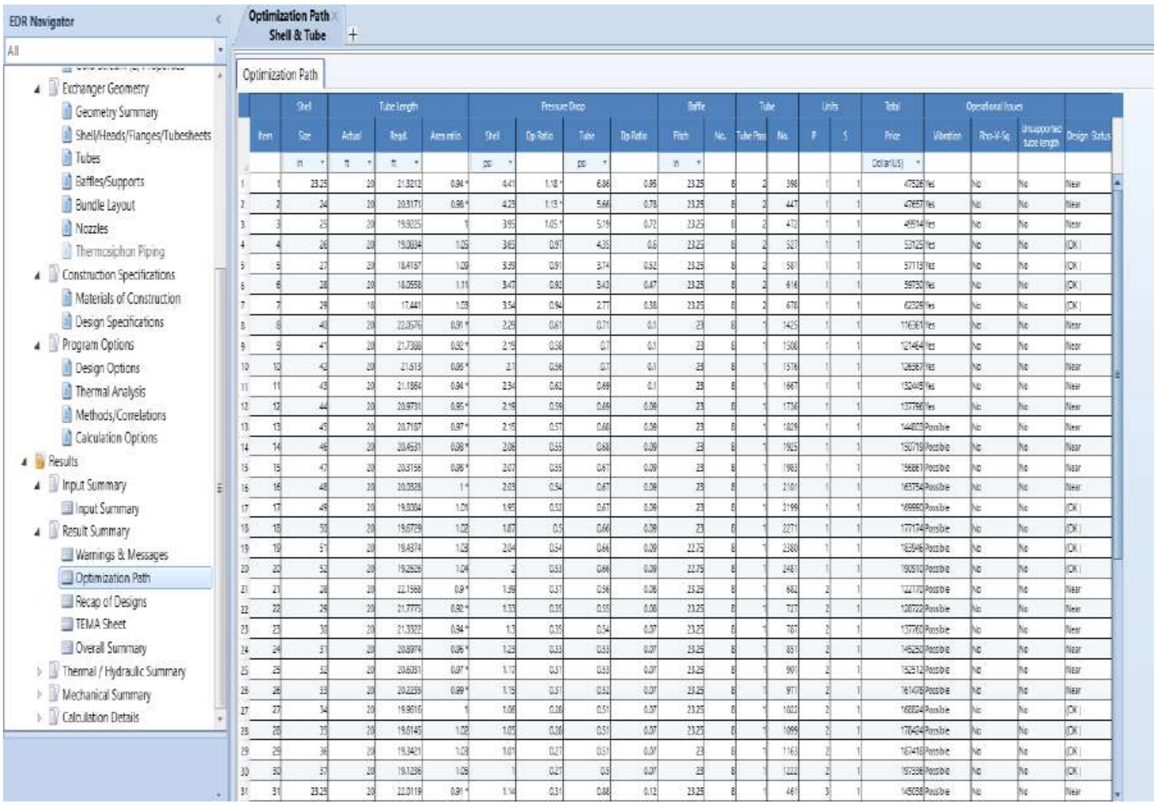
Pressure calculation option - hot side: Outlet, with minimum

Pressure calculation option - cold side: Outlet, with minimum

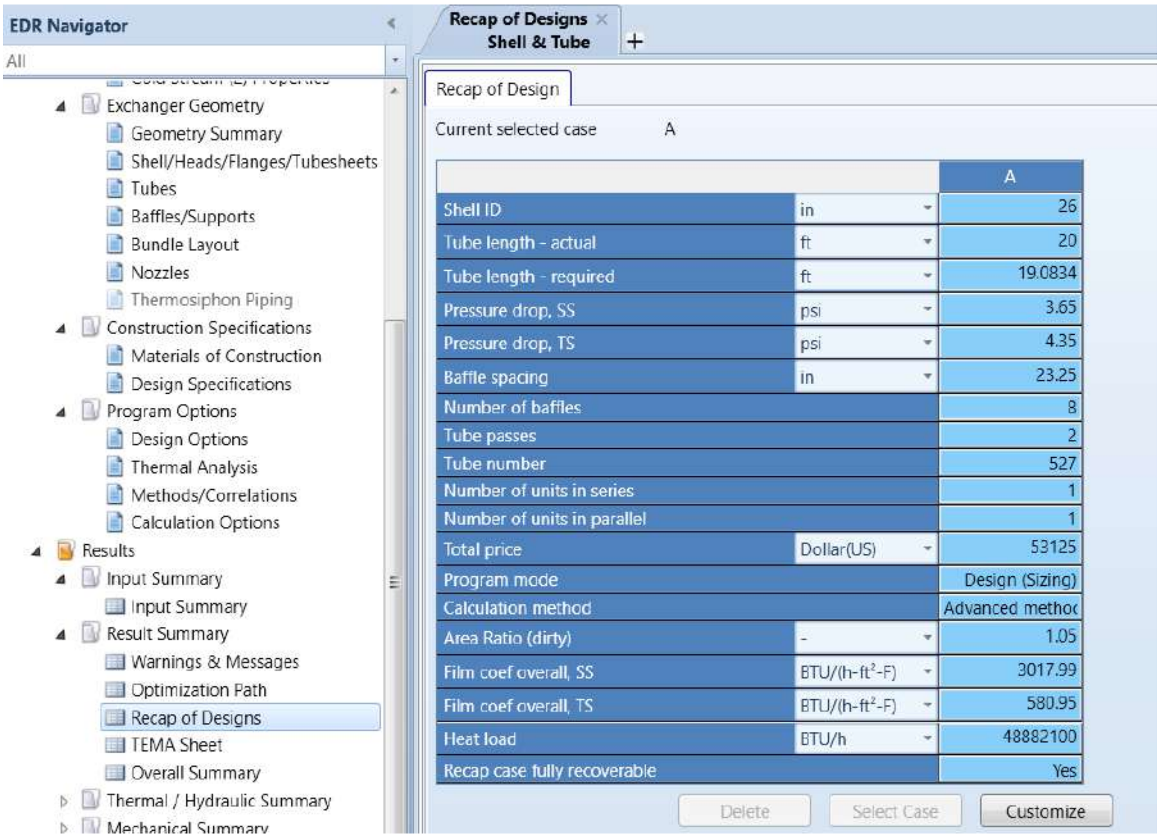
Minimum pressure - hot side: 68.77 psi

Minimum pressure - cold side: 572.9 psi

Multiplier for number of tube passes: 1

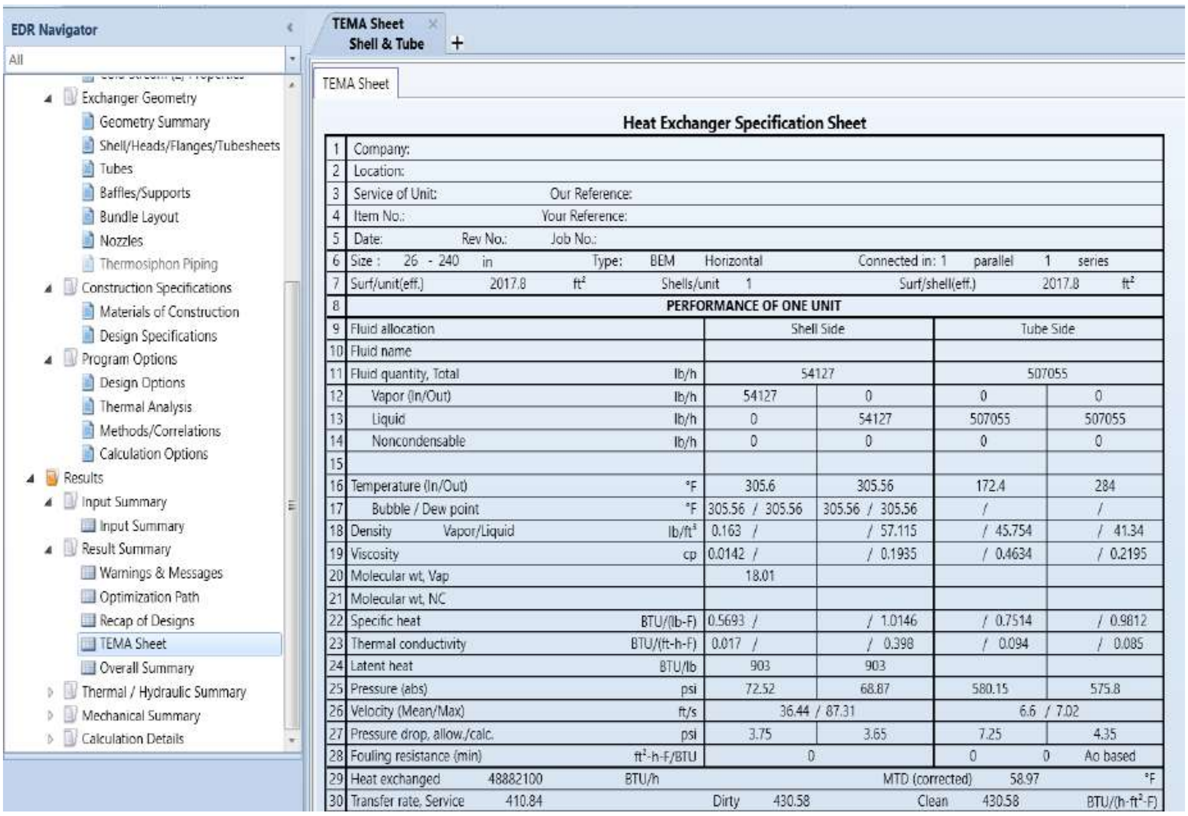


		Shell	Tube Length			Pressure Drop				Baffle		Tube		Units		Total	Operational Issues			
	Item	Size	Actual	Reqd.	Area ratio	Shell	Dp Ratio	Tube	Dp Ratio	Pitch	No.	Tube Pass	No.	P	S	Price	Vibration	Rho-V-Sq	Unsupported tube length	Design Status
		in	ft	ft		psi		psi		in						Dollar(US)				
1	1	23.25	20	21.3212	0.94 *	4.41	1.18 *	6.86	0.96	23.25	8	2	398	1	1	47526	Yes	No	No	Near
2	2	24	20	20.3171	0.98 *	4.23	1.13 *	5.66	0.78	23.25	8	2	447	1	1	47657	Yes	No	No	Near
3	3	25	20	19.9225	1	3.95	1.05 *	5.19	0.72	23.25	8	2	472	1	1	49514	Yes	No	No	Near
4	4	26	20	19.0834	1.05	3.65	0.97	4.35	0.6	23.25	8	2	527	1	1	53125	Yes	No	No	(OK)
5	5	27	20	18.4167	1.09	3.39	0.91	3.74	0.52	23.25	8	2	581	1	1	57113	Yes	No	No	(OK)
6	6	28	20	18.0558	1.11	3.47	0.93	3.43	0.47	23.25	8	2	616	1	1	59730	Yes	No	No	(OK)
7	7	29	18	17.441	1.03	3.54	0.94	2.77	0.38	23.25	8	2	678	1	1	62329	Yes	No	No	(OK)
8	8	40	20	22.0576	0.91 *	2.29	0.61	0.71	0.1	23	8	1	1425	1	1	116361	Yes	No	No	Near
9	9	41	20	21.7388	0.92 *	2.19	0.58	0.7	0.1	23	8	1	1508	1	1	121464	Yes	No	No	Near
10	10	42	20	21.513	0.96 *	2.1	0.56	0.7	0.1	23	8	1	1576	1	1	126367	Yes	No	No	Near
11	11	43	20	21.1864	0.94 *	2.34	0.62	0.69	0.1	23	8	1	1667	1	1	132445	Yes	No	No	Near
12	12	44	20	20.9731	0.95 *	2.19	0.59	0.69	0.08	23	8	1	1736	1	1	137796	Yes	No	No	Near
13	13	45	20	20.7187	0.97 *	2.15	0.57	0.68	0.08	23	8	1	1829	1	1	144803	Possible	No	No	Near
14	14	46	20	20.4631	0.98 *	2.06	0.55	0.68	0.09	23	8	1	1925	1	1	150719	Possible	No	No	Near
15	15	47	20	20.3156	0.98 *	2.07	0.55	0.67	0.09	23	8	1	1983	1	1	156861	Possible	No	No	Near
16	16	48	20	20.0328	1 *	2.03	0.54	0.67	0.08	23	8	1	2101	1	1	163754	Possible	No	No	Near
17	17	49	20	19.8084	1.01	1.95	0.52	0.67	0.08	23	8	1	2199	1	1	169990	Possible	No	No	(OK)
18	18	50	20	19.6729	1.02	1.87	0.5	0.66	0.08	23	8	1	2271	1	1	177174	Possible	No	No	(OK)
19	19	51	20	19.4874	1.03	2.04	0.54	0.66	0.09	22.75	8	1	2380	1	1	183946	Possible	No	No	(OK)
20	20	52	20	19.2626	1.04	2	0.53	0.66	0.09	22.75	8	1	2481	1	1	190910	Possible	No	No	(OK)
21	21	28	20	22.1568	0.9 *	1.39	0.37	0.56	0.08	23.25	8	1	682	2	1	122170	Possible	No	No	Near
22	22	29	20	21.7775	0.92 *	1.33	0.35	0.55	0.08	23.25	8	1	727	2	1	128722	Possible	No	No	Near
23	23	30	20	21.3322	0.94 *	1.3	0.35	0.54	0.07	23.25	8	1	787	2	1	137760	Possible	No	No	Near
24	24	31	20	20.8974	0.96 *	1.23	0.33	0.53	0.07	23.25	8	1	851	2	1	145250	Possible	No	No	Near
25	25	32	20	20.6081	0.97 *	1.17	0.31	0.53	0.07	23.25	8	1	901	2	1	152512	Possible	No	No	Near
26	26	33	20	20.2235	0.99 *	1.15	0.31	0.52	0.07	23.25	8	1	971	2	1	161478	Possible	No	No	Near
27	27	34	20	19.9616	1	1.06	0.28	0.51	0.07	23.25	8	1	1022	2	1	168824	Possible	No	No	(OK)
28	28	35	20	19.6145	1.02	1.05	0.28	0.51	0.07	23.25	8	1	1099	2	1	178424	Possible	No	No	(OK)
29	29	36	20	19.3421	1.03	1.01	0.27	0.51	0.07	23	8	1	1163	2	1	187418	Possible	No	No	(OK)
30	30	37	20	19.1236	1.05	1	0.27	0.5	0.07	23	8	1	1222	2	1	197556	Possible	No	No	(OK)
31	31	23.25	20	22.0119	0.91 *	1.14	0.31	0.88	0.12	23.25	8	1	461	3	1	145058	Possible	No	No	Near



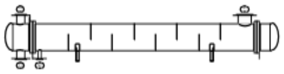
Current selected case A

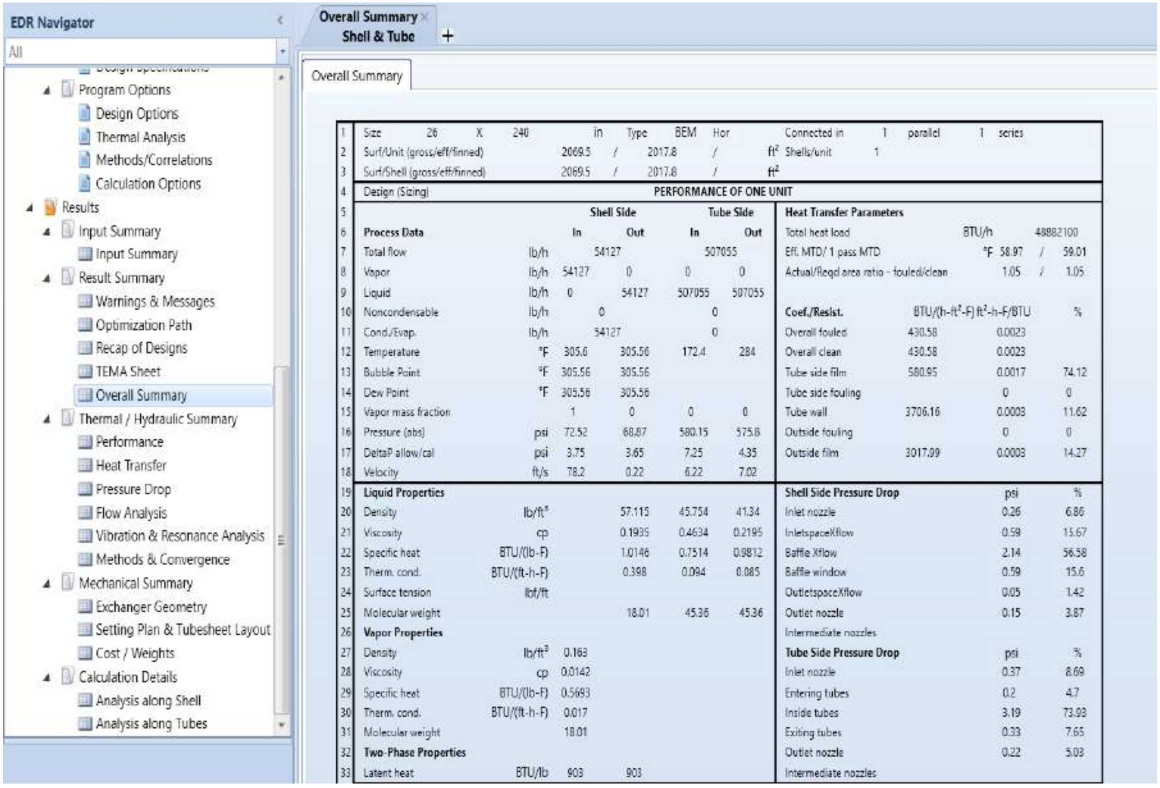
		A
Shell ID	in	26
Tube length - actual	ft	20
Tube length - required	ft	19.0834
Pressure drop, SS	psi	3.65
Pressure drop, TS	psi	4.35
Baffle spacing	in	23.25
Number of baffles		8
Tube passes		2
Tube number		527
Number of units in series		1
Number of units in parallel		1
Total price	Dollar(US)	53125
Program mode		Design (Sizing)
Calculation method		Advanced methoc
Area Ratio (dirty)	-	1.05
Film coef overall, SS	BTU/(h-ft^2-F)	3017.99
Film coef overall, TS	BTU/(h-ft^2-F)	580.95
Heat load	BTU/h	48882100
Recap case fully recoverable		Yes



Heat Exchanger Specification Sheet

1	Company:					
2	Location:					
3	Service of Unit: Our Reference:					
4	Item No.: Your Reference:					
5	Date: Rev No.: Job No.:					
6	Size : 26 - 240 in Type: BEM Horizontal Connected in: 1 parallel 1 series					
7	Surf/unit(eff.) 2017.8 ft² Shells/unit 1 Surf/shell(eff.) 2017.8 ft²					
8	**PERFORMANCE OF ONE UNIT**					
9	Fluid allocation		Shell Side		Tube Side	
10	Fluid name					
11	Fluid quantity, Total	lb/h	54127		507055	
12	Vapor (In/Out)	lb/h	54127	0	0	0
13	Liquid	lb/h	0	54127	507055	507055
14	Noncondensable	lb/h	0	0	0	0
15						
16	Temperature (In/Out)	°F	305.6	305.56	172.4	284
17	Bubble / Dew point	°F	305.56 / 305.56	305.56 / 305.56	/	/
18	Density Vapor/Liquid	lb/ft³	0.163 /	/ 57.115	/ 45.754	/ 41.34
19	Viscosity	cp	0.0142 /	/ 0.1935	/ 0.4634	/ 0.2195
20	Molecular wt, Vap		18.01			
21	Molecular wt, NC					
22	Specific heat	BTU/(lb-F)	0.5693 /	/ 1.0146	/ 0.7514	/ 0.9812
23	Thermal conductivity	BTU/(ft-h-F)	0.017 /	/ 0.398	/ 0.094	/ 0.085
24	Latent heat	BTU/lb	903	903		
25	Pressure (abs)	psi	72.52	68.87	580.15	575.8
26	Velocity (Mean/Max)	ft/s	36.44 / 87.31		6.6 / 7.02	
27	Pressure drop, allow./calc.	psi	3.75	3.65	7.25	4.35
28	Fouling resistance (min)	ft²-h-F/BTU	0		0	0 Ao based
29	Heat exchanged 48882100 BTU/h MTD (corrected) 58.97 °F					
30	Transfer rate, Service 410.84 Dirty 430.58 Clean 430.58 BTU/(h-ft²-F)					

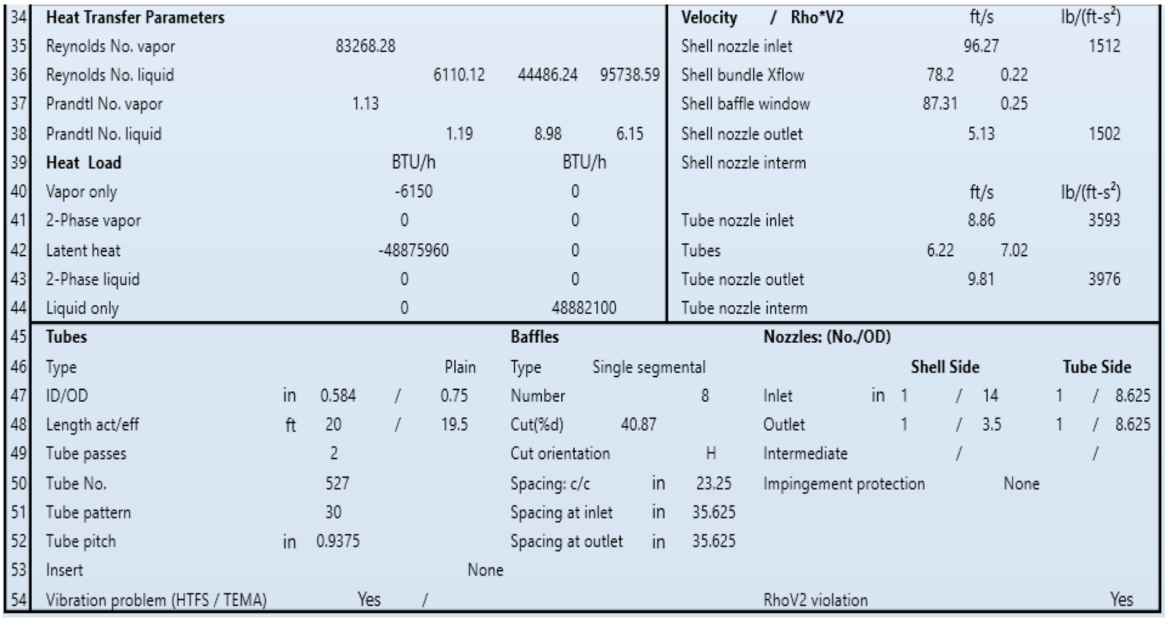
31	**CONSTRUCTION OF ONE SHELL**				**Sketch**
32			Shell Side	Tube Side	
33	Design/Vacuum/test pressure	psi	80 / /	640 / /	
34	Design temperature	°F	370	350	
35	Number passes per shell		1	2	
36	Corrosion allowance	in	0.125	0.125	
37	Connections	In in	1 14 / -	1 8 / -	
38	Size/Rating	Out	1 3 / -	1 8 / -	
39	Nominal	Intermediate	1 / -	1 / -	
40	Tube #: 527 OD: 0.75 Tks. Average 0.083 in Length: 240 in Pitch: 0.9375 in Tube pattern: 30				
41	Tube type: Plain Insert: None Fin#: #/in Material: Carbon Steel				
42	Shell Carbon Steel ID 26 OD 26.75 in			Shell cover -	
43	Channel or bonnet Carbon Steel			Channel cover -	
44	Tubesheet-stationary Carbon Steel -			Tubesheet-floating -	
45	Floating head cover -			Impingement protection None	
46	Baffle-cross Carbon Steel Type Single segmental Cut(%d) 40.87 H Spacing: c/c 23.25 in				
47	Baffle-long - Seal Type				Inlet 35.625 in
48	Supports-tube U-bend 0 Type				
49	Bypass seal Tube-tubesheet joint Expanded only (2 grooves)(App.A 'i')				
50	Expansion joint - Type None				
51	RhoV2-Inlet nozzle 1512 Bundle entrance 954 Bundle exit 7 lb/(ft-s²)				
52	Gaskets - Shell side - Tube side Flat Metal Jacket Fibe				
53	Floating head -				
54	Code requirements ASME Code Sec VIII Div 1 TEMA class R - refinery service				
55	Weight/Shell 11898.4 Filled with water 16335.6 Bundle 7432.1 lb				
56	Remarks				
57					
58					



1	Size	26	X	240	in	Type	BEM	Hor		Connected in	1	parallel	1	series
2	Surf/Unit (gross/eff/finned)	2069.5	/	2017.8	/				ft²	Shells/unit	1			
3	Surf/Shell (gross/eff/finned)	2069.5	/	2017.8	/				ft²					
4	Design (Sizing)					PERFORMANCE OF ONE UNIT								

			Shell Side		Tube Side	
5						
6	**Process Data**		In	Out	In	Out
7	Total flow	lb/h	54127		507055	
8	Vapor	lb/h	54127	0	0	0
9	Liquid	lb/h	0	54127	507055	507055
10	Noncondensable	lb/h	0		0	
11	Cond./Evap.	lb/h	54127		0	
12	Temperature	°F	305.6	305.56	172.4	284
13	Bubble Point	°F	305.56	305.56		
14	Dew Point	°F	305.56	305.56		
15	Vapor mass fraction		1	0	0	0
16	Pressure (abs)	psi	72.52	68.87	580.15	575.8
17	DeltaP allow/cal	psi	3.75	3.65	7.25	4.35
18	Velocity	ft/s	78.2	0.22	6.22	7.02
19	**Liquid Properties**					
20	Density	lb/ft³		57.115	45.754	41.34
21	Viscosity	cp		0.1935	0.4634	0.2195
22	Specific heat	BTU/(lb-F)		1.0146	0.7514	0.9812
23	Therm. cond.	BTU/(ft-h-F)		0.398	0.094	0.085
24	Surface tension	lbf/ft				
25	Molecular weight			18.01	45.36	45.36
26	**Vapor Properties**					
27	Density	lb/ft³	0.163			
28	Viscosity	cp	0.0142			
29	Specific heat	BTU/(lb-F)	0.5693			
30	Therm. cond.	BTU/(ft-h-F)	0.017			
31	Molecular weight		18.01			
32	**Two-Phase Properties**					
33	Latent heat	BTU/lb	903	903		

Heat Transfer Parameters				
Total heat load		BTU/h	48882100	
Eff. MTD/ 1 pass MTD		°F 58.97	/	59.01
Actual/Reqd area ratio - fouled/clean		1.05	/	1.05
Coef./Resist.	BTU/(h-ft²-F)	ft²-h-F/BTU		%
Overall fouled	430.58	0.0023		
Overall clean	430.58	0.0023		
Tube side film	580.95	0.0017		74.12
Tube side fouling		0		0
Tube wall	3706.16	0.0003		11.62
Outside fouling		0		0
Outside film	3017.99	0.0003		14.27
Shell Side Pressure Drop		psi		%
Inlet nozzle		0.26		6.86
InletspaceXflow		0.59		15.67
Baffle Xflow		2.14		56.58
Baffle window		0.59		15.6
OutletspaceXflow		0.05		1.42
Outlet nozzle		0.15		3.87
Intermediate nozzles				
Tube Side Pressure Drop		psi		%
Inlet nozzle		0.37		8.69
Entering tubes		0.2		4.7
Inside tubes		3.19		73.93
Exiting tubes		0.33		7.65
Outlet nozzle		0.22		5.03
Intermediate nozzles				



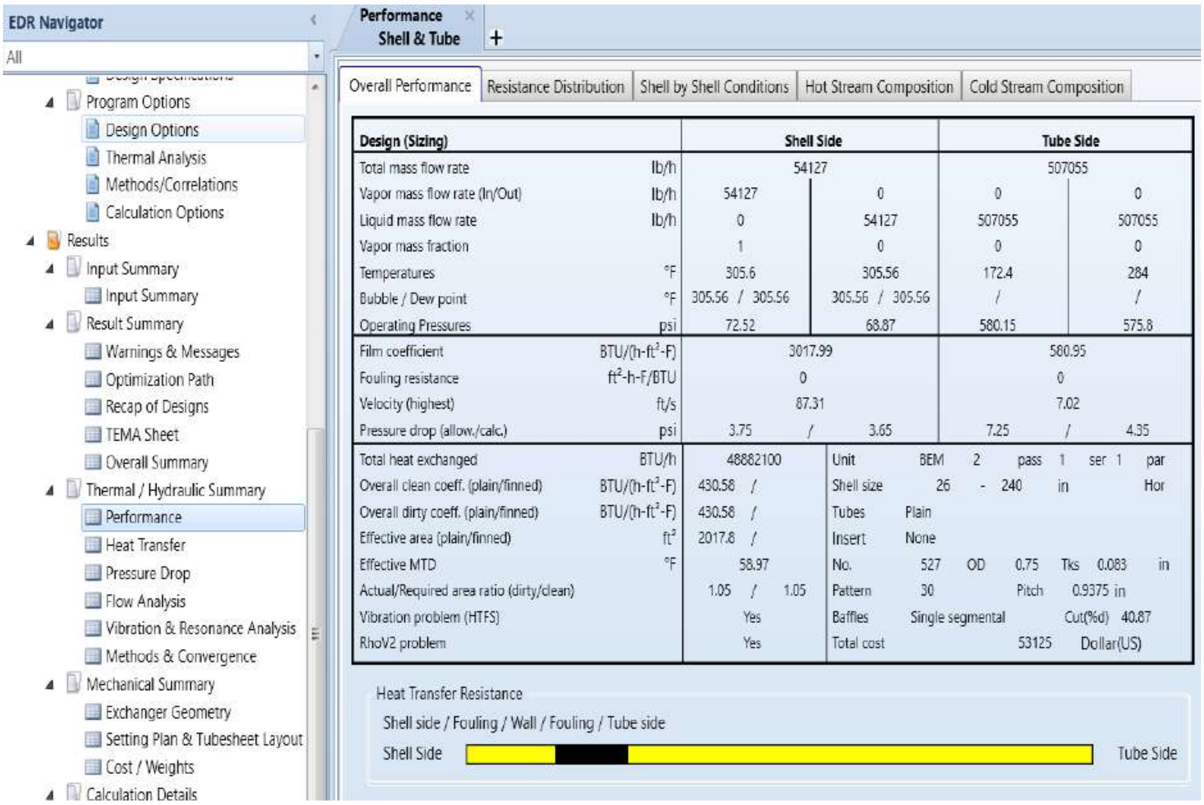
34	**Heat Transfer Parameters**					
35	Reynolds No. vapor	83268.28				
36	Reynolds No. liquid		6110.12	44486.24	95738.59	
37	Prandtl No. vapor	1.13				
38	Prandtl No. liquid		1.19	8.98	6.15	
39	**Heat Load**		BTU/h		BTU/h	
40	Vapor only		-6150		0	
41	2-Phase vapor		0		0	
42	Latent heat		-48875960		0	
43	2-Phase liquid		0		0	
44	Liquid only		0		48882100	

Velocity / Rho*V2		ft/s		lb/(ft-s²)
Shell nozzle inlet		96.27		1512
Shell bundle Xflow	78.2		0.22	
Shell baffle window	87.31		0.25	
Shell nozzle outlet		5.13		1502
Shell nozzle interm				
		ft/s		lb/(ft-s²)
Tube nozzle inlet		8.86		3593
Tubes	6.22		7.02	
Tube nozzle outlet		9.81		3976
Tube nozzle interm				

45	**Tubes**				
46	Type				Plain
47	ID/OD	in	0.584	/	0.75
48	Length act/eff	ft	20	/	19.5
49	Tube passes		2		
50	Tube No.		527		
51	Tube pattern		30		
52	Tube pitch	in	0.9375		
53	Insert				None
54	Vibration problem (HTFS / TEMA)		Yes	/	

Baffles		
Type		Single segmental
Number		8
Cut(%d)		40.87
Cut orientation		H
Spacing: c/c	in	23.25
Spacing at inlet	in	35.625
Spacing at outlet	in	35.625

		Shell Side			Tube Side		
Nozzles: (No./OD)							
Inlet	in	1	/	14	1	/	8.625
Outlet		1	/	3.5	1	/	8.625
Intermediate			/			/	
Impingement protection		None					
RhoV2 violation							Yes



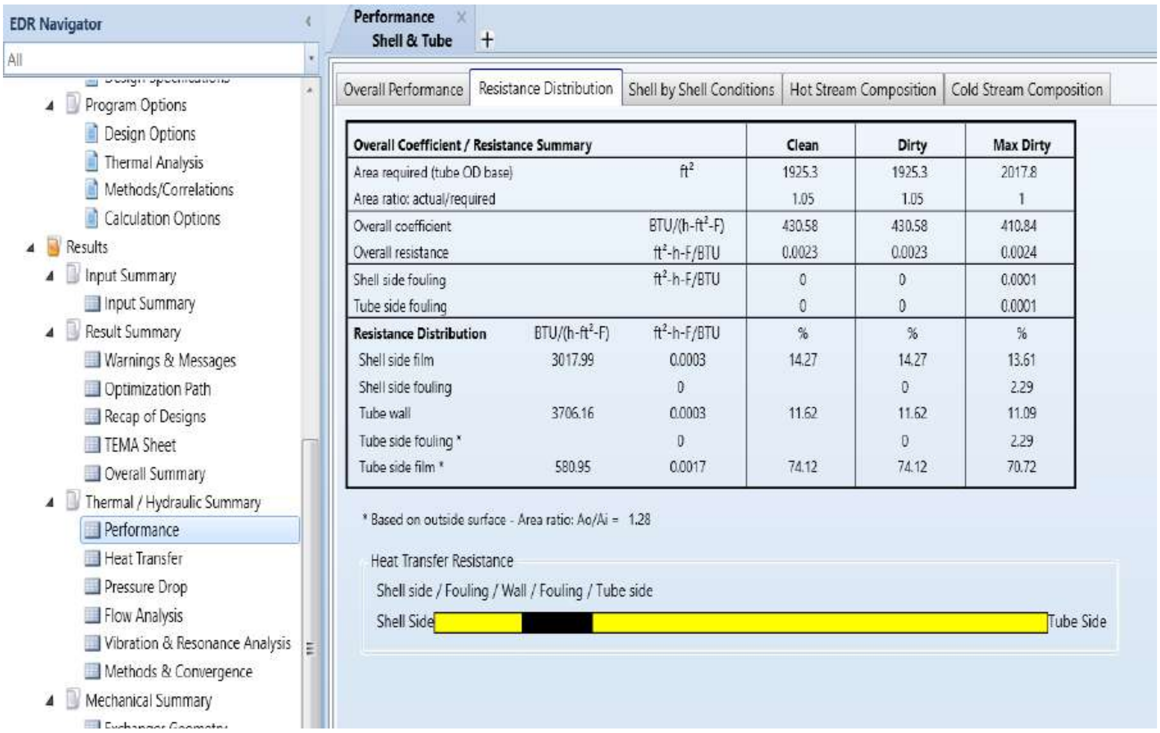
Design (Sizing)		Shell Side		Tube Side	
Total mass flow rate	lb/h	54127		507055	
Vapor mass flow rate (In/Out)	lb/h	54127	0	0	0
Liquid mass flow rate	lb/h	0	54127	507055	507055
Vapor mass fraction		1	0	0	0
Temperatures	°F	305.6	305.56	172.4	284
Bubble / Dew point	°F	305.56 / 305.56	305.56 / 305.56	/	/
Operating Pressures	psi	72.52	68.87	580.15	575.8
Film coefficient	BTU/(h-ft²-F)	3017.99		580.95	
Fouling resistance	ft²-h-F/BTU	0		0	
Velocity (highest)	ft/s	87.31		7.02	
Pressure drop (allow./calc.)	psi	3.75 /	3.65	7.25 /	4.35

Total heat exchanged	BTU/h	48882100	Unit BEM 2 pass 1 ser 1 par
Overall clean coeff. (plain/finned)	BTU/(h-ft²-F)	430.58 /	Shell size 26 - 240 in Hor
Overall dirty coeff. (plain/finned)	BTU/(h-ft²-F)	430.58 /	Tubes Plain
Effective area (plain/finned)	ft²	2017.8 /	Insert None
Effective MTD	°F	58.97	No. 527 OD 0.75 Tks 0.083 in
Actual/Required area ratio (dirty/clean)		1.05 / 1.05	Pattern 30 Pitch 0.9375 in
Vibration problem (HTFS)		Yes	Baffles Single segmental Cut(%d) 40.87
RhoV2 problem		Yes	Total cost 53125 Dollar(US)

Heat Transfer Resistance

Shell side / Fouling / Wall / Fouling / Tube side

Shell Side ▬ Tube Side



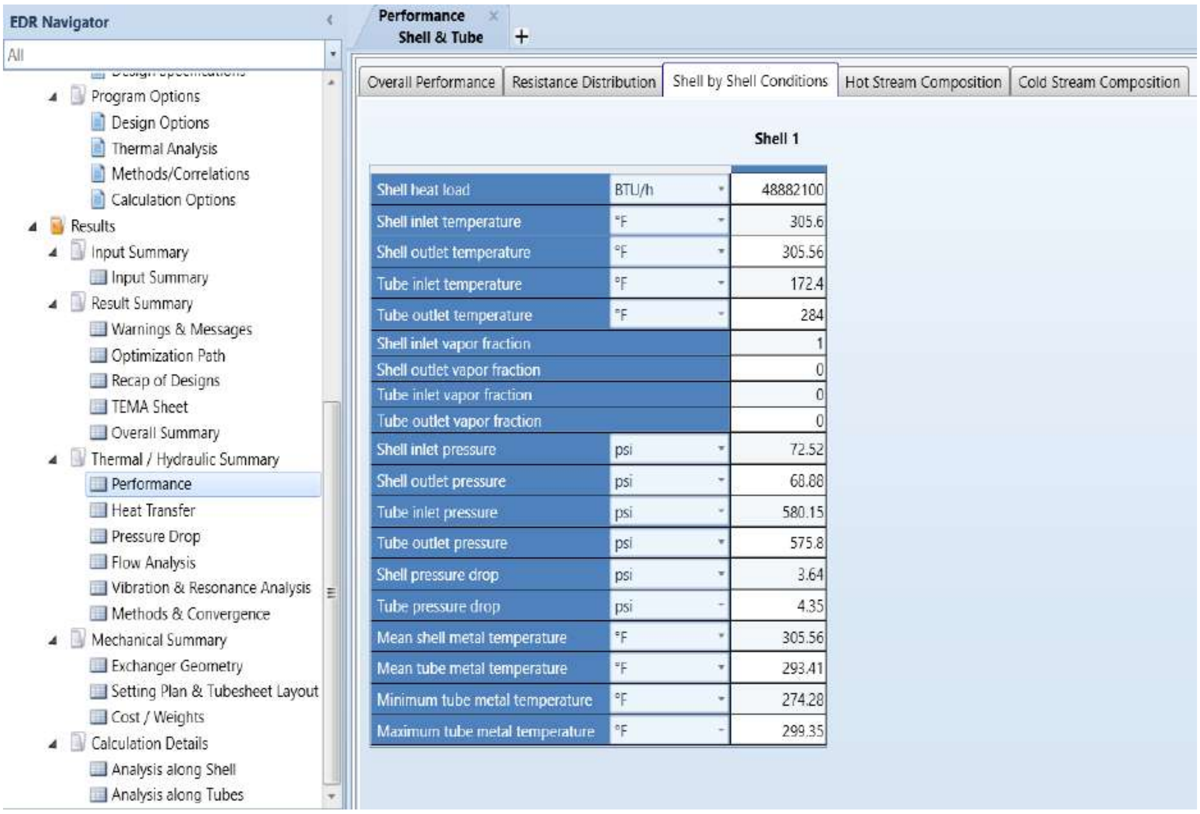
Overall Coefficient / Resistance Summary			Clean	Dirty	Max Dirty
Area required (tube OD base)		ft²	1925.3	1925.3	2017.8
Area ratio: actual/required			1.05	1.05	1
Overall coefficient		BTU/(h-ft²-F)	430.58	430.58	410.84
Overall resistance		ft²-h-F/BTU	0.0023	0.0023	0.0024
Shell side fouling		ft²-h-F/BTU	0	0	0.0001
Tube side fouling			0	0	0.0001
Resistance Distribution	BTU/(h-ft²-F)	ft²-h-F/BTU	%	%	%
Shell side film	3017.99	0.0003	14.27	14.27	13.61
Shell side fouling		0		0	2.29
Tube wall	3706.16	0.0003	11.62	11.62	11.09
Tube side fouling *		0		0	2.29
Tube side film *	580.95	0.0017	74.12	74.12	70.72

* Based on outside surface - Area ratio: Ao/Ai = 1.28

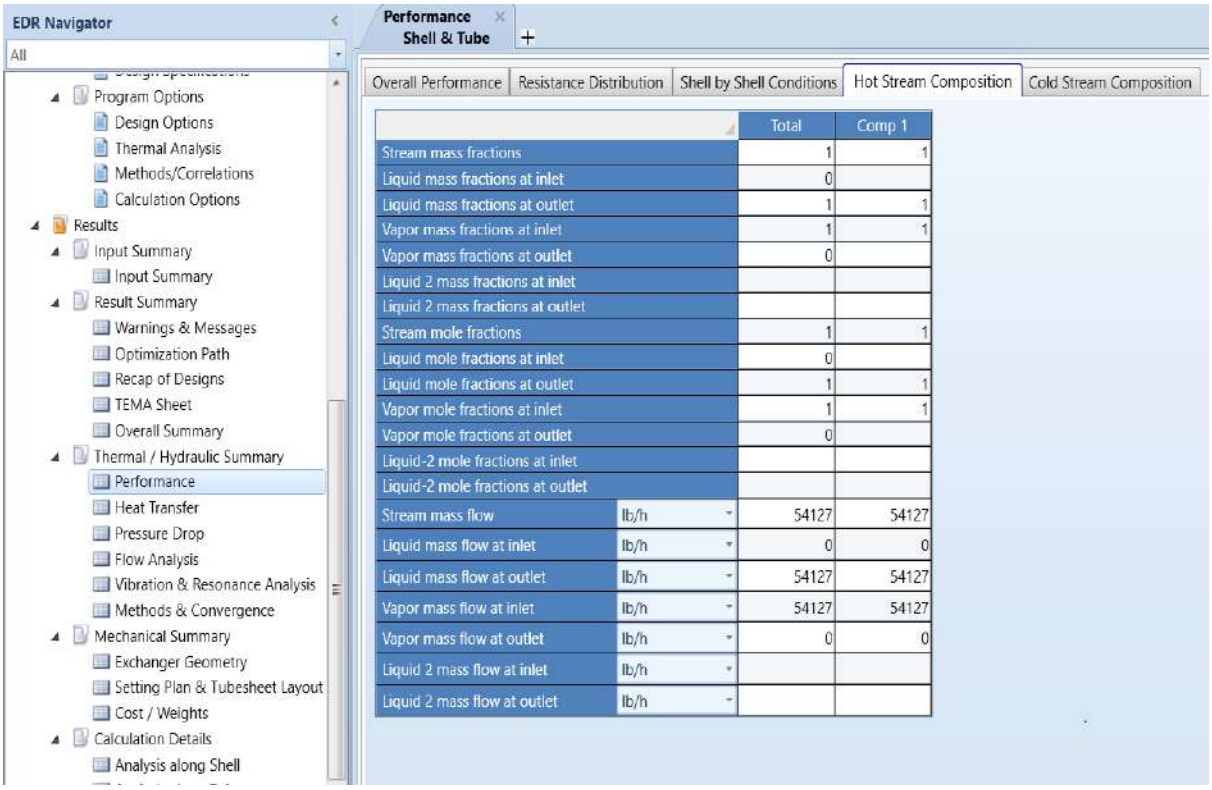
Heat Transfer Resistance

Shell side / Fouling / Wall / Fouling / Tube side

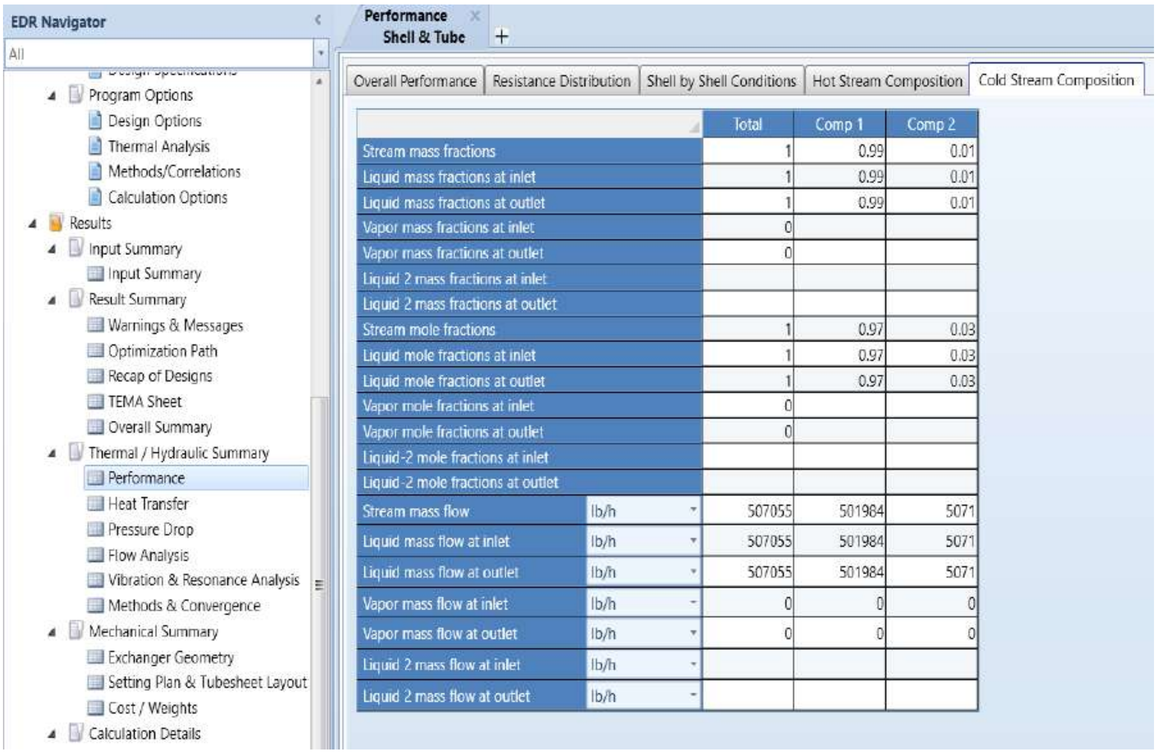
Shell Side ▬ Tube Side



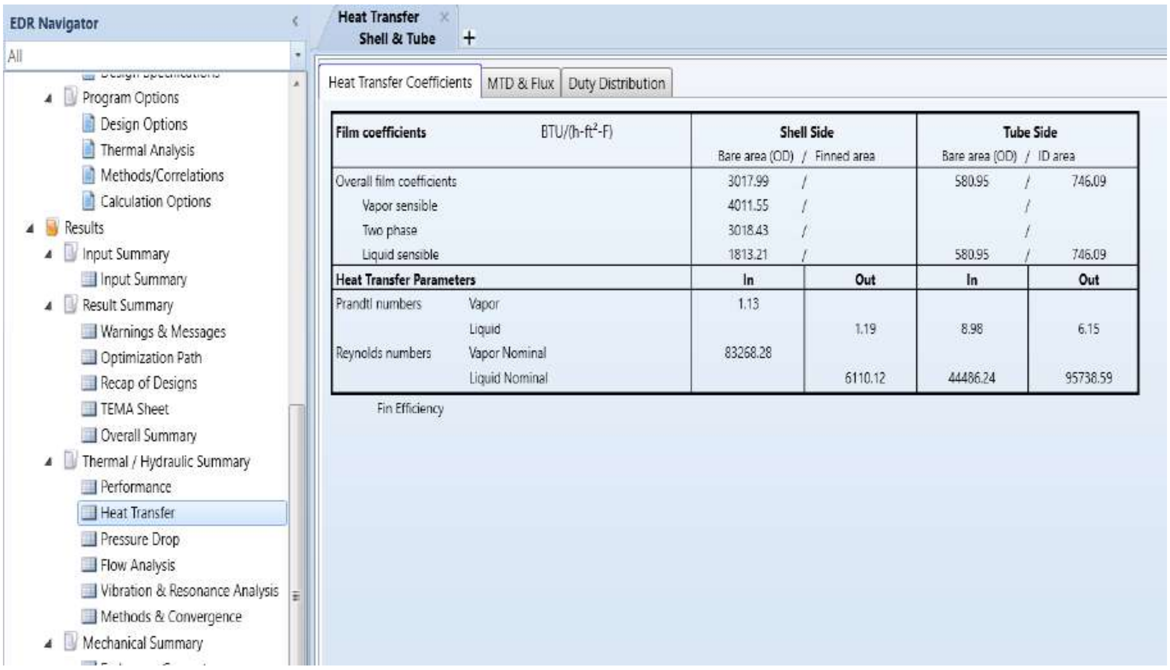
		Shell 1
Shell heat load	BTU/h	48882100
Shell inlet temperature	°F	305.6
Shell outlet temperature	°F	305.56
Tube inlet temperature	°F	172.4
Tube outlet temperature	°F	284
Shell inlet vapor fraction		1
Shell outlet vapor fraction		0
Tube inlet vapor fraction		0
Tube outlet vapor fraction		0
Shell inlet pressure	psi	72.52
Shell outlet pressure	psi	68.88
Tube inlet pressure	psi	580.15
Tube outlet pressure	psi	575.8
Shell pressure drop	psi	3.64
Tube pressure drop	psi	4.35
Mean shell metal temperature	°F	305.56
Mean tube metal temperature	°F	293.41
Minimum tube metal temperature	°F	274.28
Maximum tube metal temperature	°F	299.35



		Total	Comp 1
Stream mass fractions		1	1
Liquid mass fractions at inlet		0	
Liquid mass fractions at outlet		1	1
Vapor mass fractions at inlet		1	1
Vapor mass fractions at outlet		0	
Liquid 2 mass fractions at inlet			
Liquid 2 mass fractions at outlet			
Stream mole fractions		1	1
Liquid mole fractions at inlet		0	
Liquid mole fractions at outlet		1	1
Vapor mole fractions at inlet		1	1
Vapor mole fractions at outlet		0	
Liquid-2 mole fractions at inlet			
Liquid-2 mole fractions at outlet			
Stream mass flow	lb/h	54127	54127
Liquid mass flow at inlet	lb/h	0	0
Liquid mass flow at outlet	lb/h	54127	54127
Vapor mass flow at inlet	lb/h	54127	54127
Vapor mass flow at outlet	lb/h	0	0
Liquid 2 mass flow at inlet	lb/h		
Liquid 2 mass flow at outlet	lb/h		

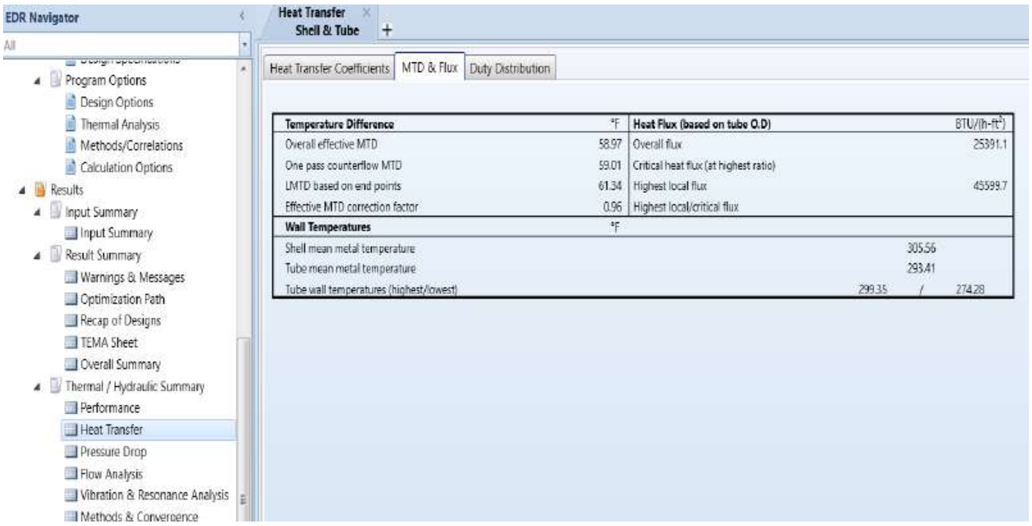


		Total	Comp 1	Comp 2
Stream mass fractions		1	0.99	0.01
Liquid mass fractions at inlet		1	0.99	0.01
Liquid mass fractions at outlet		1	0.99	0.01
Vapor mass fractions at inlet		0		
Vapor mass fractions at outlet		0		
Liquid 2 mass fractions at inlet				
Liquid 2 mass fractions at outlet				
Stream mole fractions		1	0.97	0.03
Liquid mole fractions at inlet		1	0.97	0.03
Liquid mole fractions at outlet		1	0.97	0.03
Vapor mole fractions at inlet		0		
Vapor mole fractions at outlet		0		
Liquid-2 mole fractions at inlet				
Liquid-2 mole fractions at outlet				
Stream mass flow	lb/h	507055	501984	5071
Liquid mass flow at inlet	lb/h	507055	501984	5071
Liquid mass flow at outlet	lb/h	507055	501984	5071
Vapor mass flow at inlet	lb/h	0	0	0
Vapor mass flow at outlet	lb/h	0	0	0
Liquid 2 mass flow at inlet	lb/h			
Liquid 2 mass flow at outlet	lb/h			



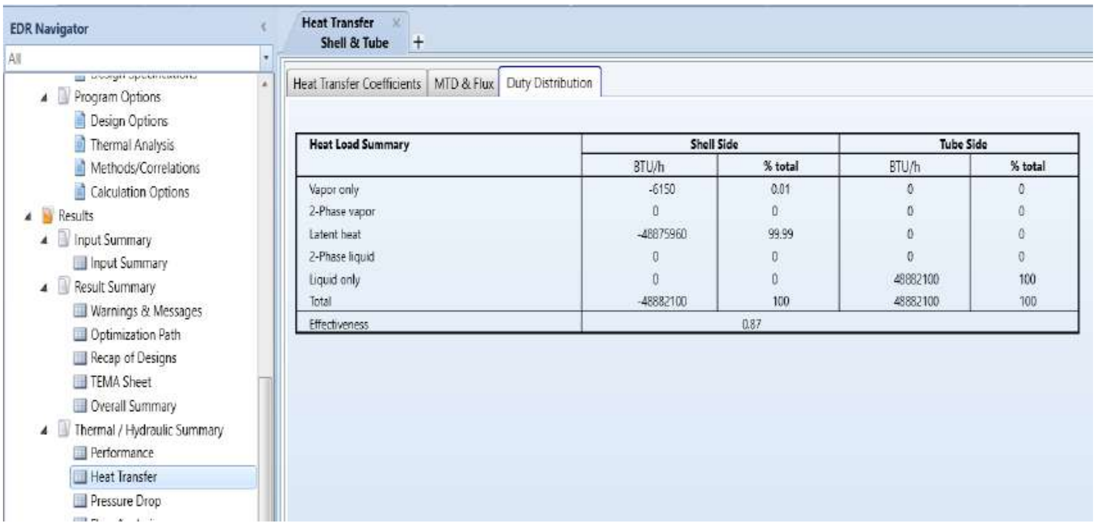
Film coefficients	BTU/(h-ft^2-F)	**Shell Side** Bare area (OD) / Finned area		**Tube Side** Bare area (OD) / ID area	
Overall film coefficients		3017.99 /		580.95 /	746.09
Vapor sensible		4011.55 /		/	
Two phase		3018.43 /		/	
Liquid sensible		1813.21 /		580.95 /	746.09
Heat Transfer Parameters		**In**	**Out**	**In**	**Out**
Prandtl numbers	Vapor	1.13			
	Liquid		1.19	8.98	6.15
Reynolds numbers	Vapor Nominal	83268.28			
	Liquid Nominal		6110.12	44486.24	95738.59

Fin Efficiency

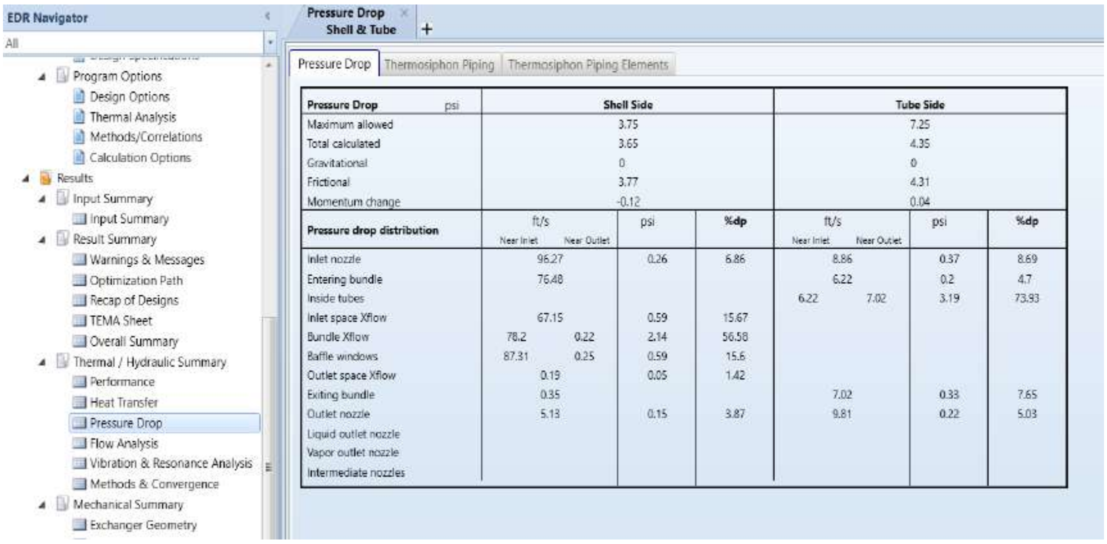


Temperature Difference	°F	**Heat Flux (based on tube O.D)**	BTU/(h-ft²)
Overall effective MTD	58.97	Overall flux	25391.1
One pass counterflow MTD	59.01	Critical heat flux (at highest ratio)	
LMTD based on end points	61.34	Highest local flux	45599.7
Effective MTD correction factor	0.96	Highest local/critical flux	

Wall Temperatures	°F		
Shell mean metal temperature		305.56	
Tube mean metal temperature		293.41	
Tube wall temperatures (highest/lowest)	299.35	/	274.28

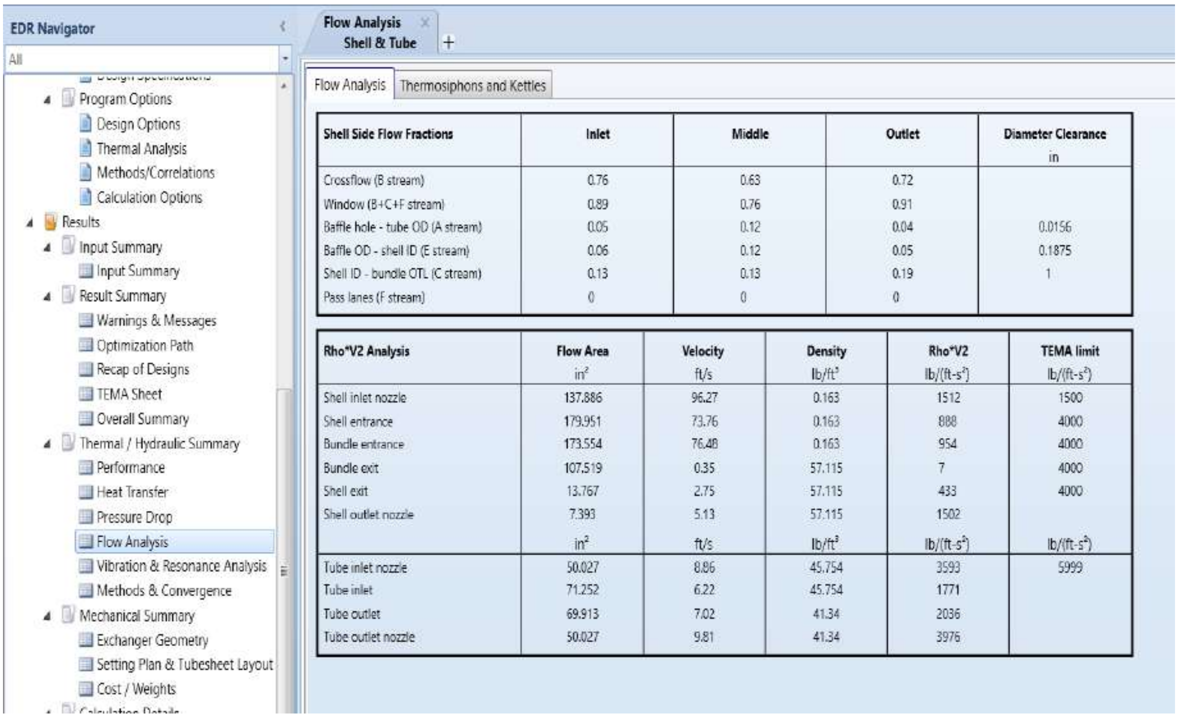


Heat Load Summary	**Shell Side**		**Tube Side**	
	BTU/h	**% total**	BTU/h	**% total**
Vapor only	-6150	0.01	0	0
2-Phase vapor	0	0	0	0
Latent heat	-48875960	99.99	0	0
2-Phase liquid	0	0	0	0
Liquid only	0	0	48882100	100
Total	-48882100	100	48882100	100
Effectiveness	0.87			



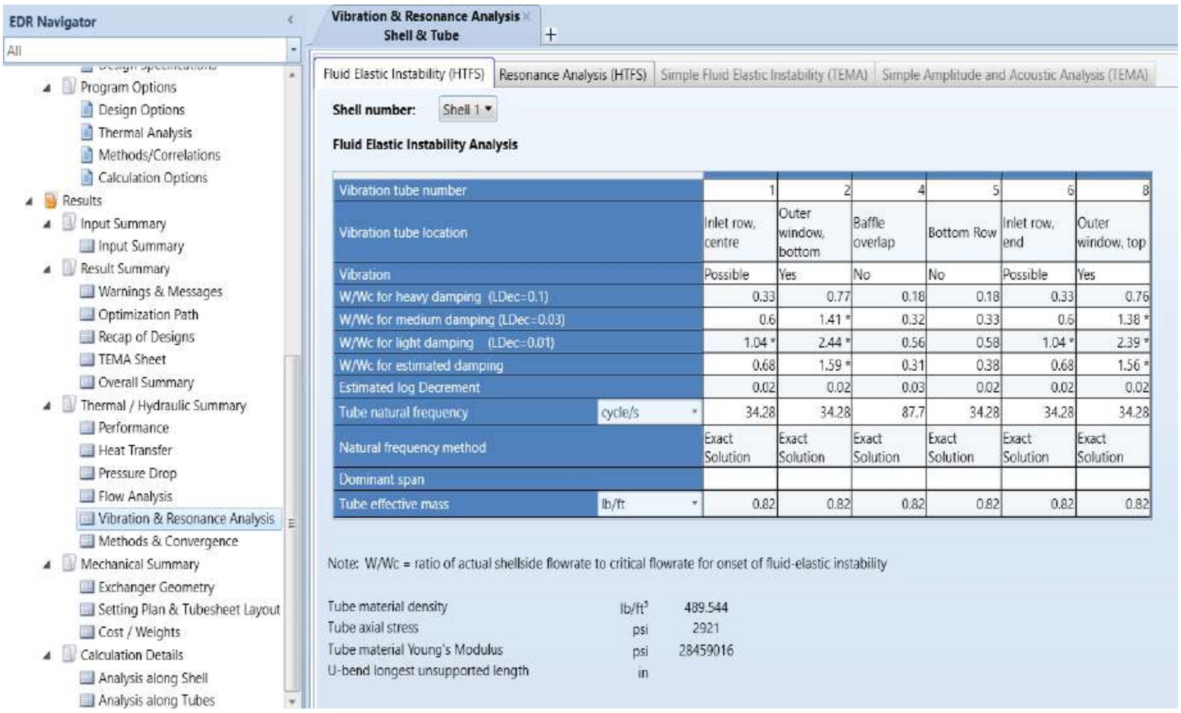
Pressure Drop psi	**Shell Side**	**Tube Side**
Maximum allowed	3.75	7.25
Total calculated	3.65	4.35
Gravitational	0	0
Frictional	3.77	4.31
Momentum change	-0.12	0.04

Pressure drop distribution	ft/s		psi	**%dp**	ft/s		psi	**%dp**
	Near Inlet	Near Outlet			Near Inlet	Near Outlet		
Inlet nozzle	96.27		0.26	6.86	8.86		0.37	8.69
Entering bundle	76.48				6.22		0.2	4.7
Inside tubes					6.22	7.02	3.19	73.93
Inlet space Xflow	67.15		0.59	15.67				
Bundle Xflow	78.2	0.22	2.14	56.58				
Baffle windows	87.31	0.25	0.59	15.6				
Outlet space Xflow	0.19		0.05	1.42				
Exiting bundle	0.35				7.02		0.33	7.65
Outlet nozzle	5.13		0.15	3.87	9.81		0.22	5.03
Liquid outlet nozzle								
Vapor outlet nozzle								
Intermediate nozzles								



Shell Side Flow Fractions	Inlet	Middle	Outlet	Diameter Clearance in
Crossflow (B stream)	0.76	0.63	0.72	
Window (B+C+F stream)	0.89	0.76	0.91	
Baffle hole - tube OD (A stream)	0.05	0.12	0.04	0.0156
Baffle OD - shell ID (E stream)	0.06	0.12	0.05	0.1875
Shell ID - bundle OTL (C stream)	0.13	0.13	0.19	1
Pass lanes (F stream)	0	0	0	

Rho*V2 Analysis	Flow Area	Velocity	Density	Rho*V2	TEMA limit
	in^2	ft/s	lb/ft^3	$\text{lb/(ft-s}^2)$	$\text{lb/(ft-s}^2)$
Shell inlet nozzle	137.886	96.27	0.163	1512	1500
Shell entrance	179.951	73.76	0.163	888	4000
Bundle entrance	173.554	76.48	0.163	954	4000
Bundle exit	107.519	0.35	57.115	7	4000
Shell exit	13.767	2.75	57.115	433	4000
Shell outlet nozzle	7.393	5.13	57.115	1502	
	in^2	ft/s	lb/ft^3	$\text{lb/(ft-s}^2)$	$\text{lb/(ft-s}^2)$
Tube inlet nozzle	50.027	8.86	45.754	3593	5999
Tube inlet	71.252	6.22	45.754	1771	
Tube outlet	69.913	7.02	41.34	2036	
Tube outlet nozzle	50.027	9.81	41.34	3976	



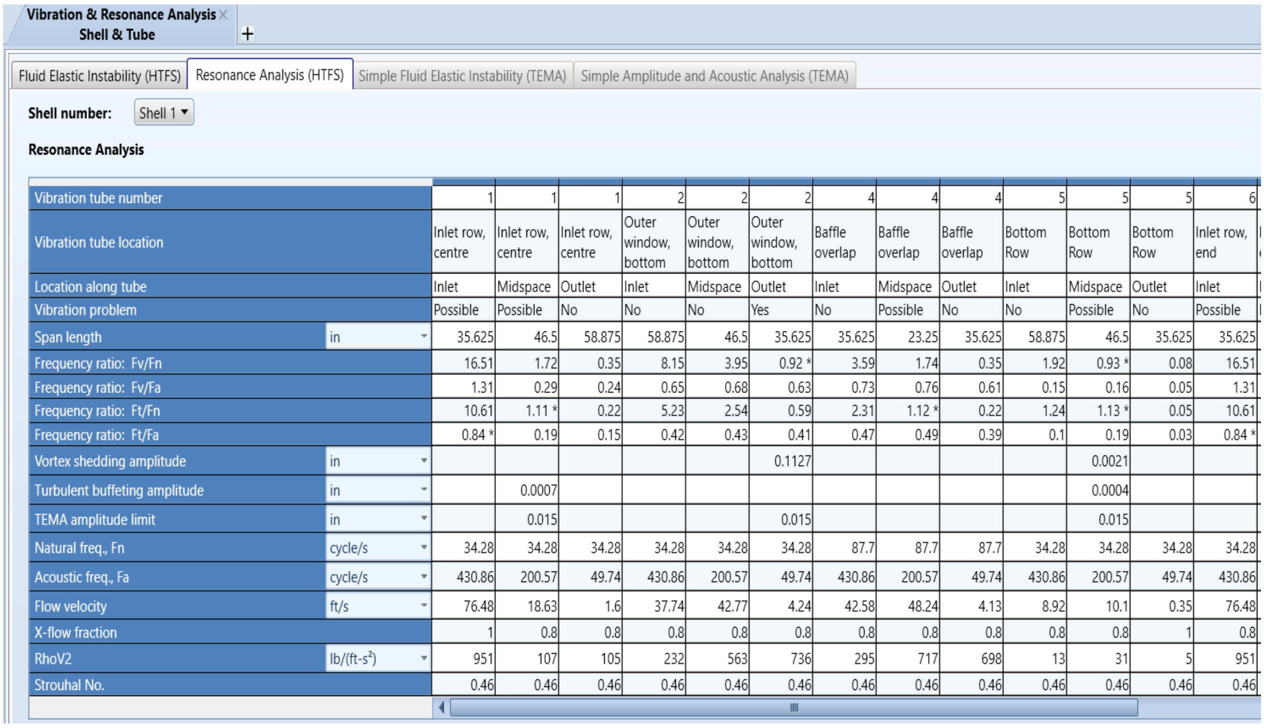
Shell number: Shell 1

Fluid Elastic Instability Analysis

Vibration tube number		1	2	4	5	6	8
Vibration tube location		Inlet row, centre	Outer window, bottom	Baffle overlap	Bottom Row	Inlet row, end	Outer window, top
Vibration		Possible	Yes	No	No	Possible	Yes
W/Wc for heavy damping (LDec=0.1)		0.33	0.77	0.18	0.18	0.33	0.76
W/Wc for medium damping (LDec=0.03)		0.6	1.41 *	0.32	0.33	0.6	1.38 *
W/Wc for light damping (LDec=0.01)		1.04 *	2.44 *	0.56	0.58	1.04 *	2.39 *
W/Wc for estimated damping		0.68	1.59 *	0.31	0.38	0.68	1.56 *
Estimated log Decrement		0.02	0.02	0.03	0.02	0.02	0.02
Tube natural frequency	cycle/s	34.28	34.28	87.7	34.28	34.28	34.28
Natural frequency method		Exact Solution	Exact Solution	Exact Solution	Exact Solution	Exact Solution	Exact Solution
Dominant span							
Tube effective mass	lb/ft	0.82	0.82	0.82	0.82	0.82	0.82

Note: W/Wc = ratio of actual shellside flowrate to critical flowrate for onset of fluid-elastic instability

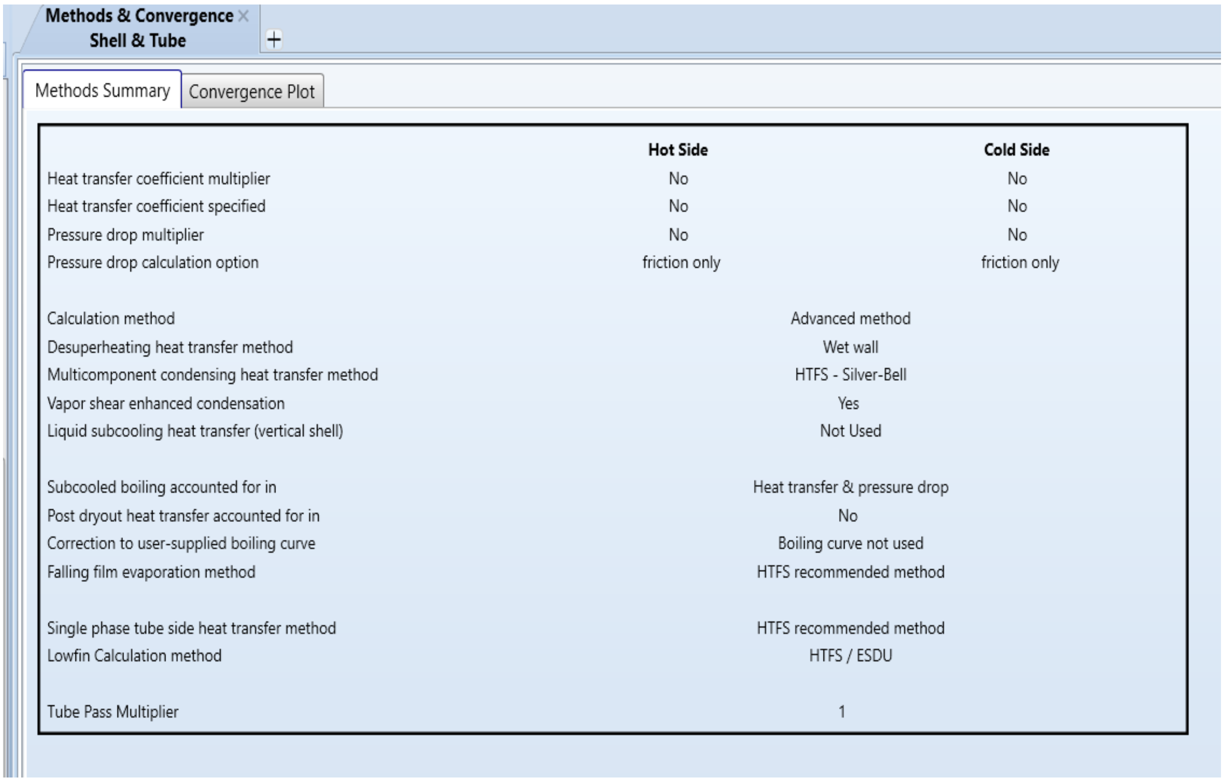
Tube material density	lb/ft^3	489.544
Tube axial stress	psi	2921
Tube material Young's Modulus	psi	28459016
U-bend longest unsupported length	in	



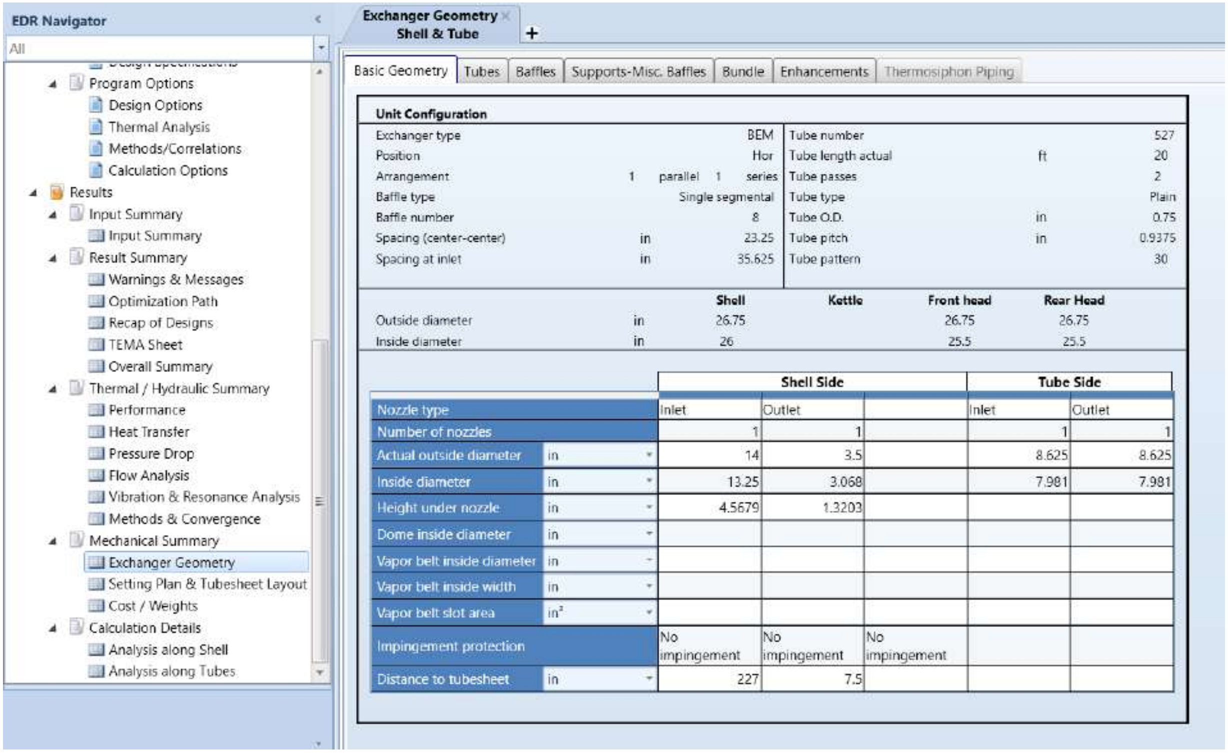
Shell number: Shell 1

Resonance Analysis

Vibration tube number		1	1	1	2	2	2	4	4	4	5	5	5	6
Vibration tube location		Inlet row, centre	Inlet row, centre	Inlet row, centre	Outer window, bottom	Outer window, bottom	Outer window, bottom	Baffle overlap	Baffle overlap	Baffle overlap	Bottom Row	Bottom Row	Bottom Row	Inlet row, end
Location along tube		Inlet	Midspace	Outlet	Inlet	Midspace	Outlet	Inlet	Midspace	Outlet	Inlet	Midspace	Outlet	Inlet
Vibration problem		Possible	Possible	No	No	No	Yes	No	Possible	No	No	Possible	No	Possible
Span length	in	35.625	46.5	58.875	58.875	46.5	35.625	35.625	23.25	35.625	58.875	46.5	35.625	35.625
Frequency ratio: Fv/Fn		16.51	1.72	0.35	8.15	3.95	0.92 *	3.59	1.74	0.35	1.92	0.93 *	0.08	16.51
Frequency ratio: Fv/Fa		1.31	0.29	0.24	0.65	0.68	0.63	0.73	0.76	0.61	0.15	0.16	0.05	1.31
Frequency ratio: Ft/Fn		10.61	1.11 *	0.22	5.23	2.54	0.59	2.31	1.12 *	0.22	1.24	1.13 *	0.05	10.61
Frequency ratio: Ft/Fa		0.84 *	0.19	0.15	0.42	0.43	0.41	0.47	0.49	0.39	0.1	0.19	0.03	0.84 *
Vortex shedding amplitude	in						0.1127					0.0021		
Turbulent buffeting amplitude	in		0.0007									0.0004		
TEMA amplitude limit	in		0.015				0.015					0.015		
Natural freq., Fn	cycle/s	34.28	34.28	34.28	34.28	34.28	34.28	87.7	87.7	87.7	34.28	34.28	34.28	34.28
Acoustic freq., Fa	cycle/s	430.86	200.57	49.74	430.86	200.57	49.74	430.86	200.57	49.74	430.86	200.57	49.74	430.86
Flow velocity	ft/s	76.48	18.63	1.6	37.74	42.77	4.24	42.58	48.24	4.13	8.92	10.1	0.35	76.48
X-flow fraction		1	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	1	0.8
RhoV2	lb/(ft-s²)	951	107	105	232	563	736	295	717	698	13	31	5	951
Strouhal No.		0.46	0.46	0.46	0.46	0.46	0.46	0.46	0.46	0.46	0.46	0.46	0.46	0.46



	Hot Side	Cold Side
Heat transfer coefficient multiplier	No	No
Heat transfer coefficient specified	No	No
Pressure drop multiplier	No	No
Pressure drop calculation option	friction only	friction only
Calculation method	Advanced method	
Desuperheating heat transfer method	Wet wall	
Multicomponent condensing heat transfer method	HTFS - Silver-Bell	
Vapor shear enhanced condensation	Yes	
Liquid subcooling heat transfer (vertical shell)	Not Used	
Subcooled boiling accounted for in	Heat transfer & pressure drop	
Post dryout heat transfer accounted for in	No	
Correction to user-supplied boiling curve	Boiling curve not used	
Falling film evaporation method	HTFS recommended method	
Single phase tube side heat transfer method	HTFS recommended method	
Lowfin Calculation method	HTFS / ESDU	
Tube Pass Multiplier	1	

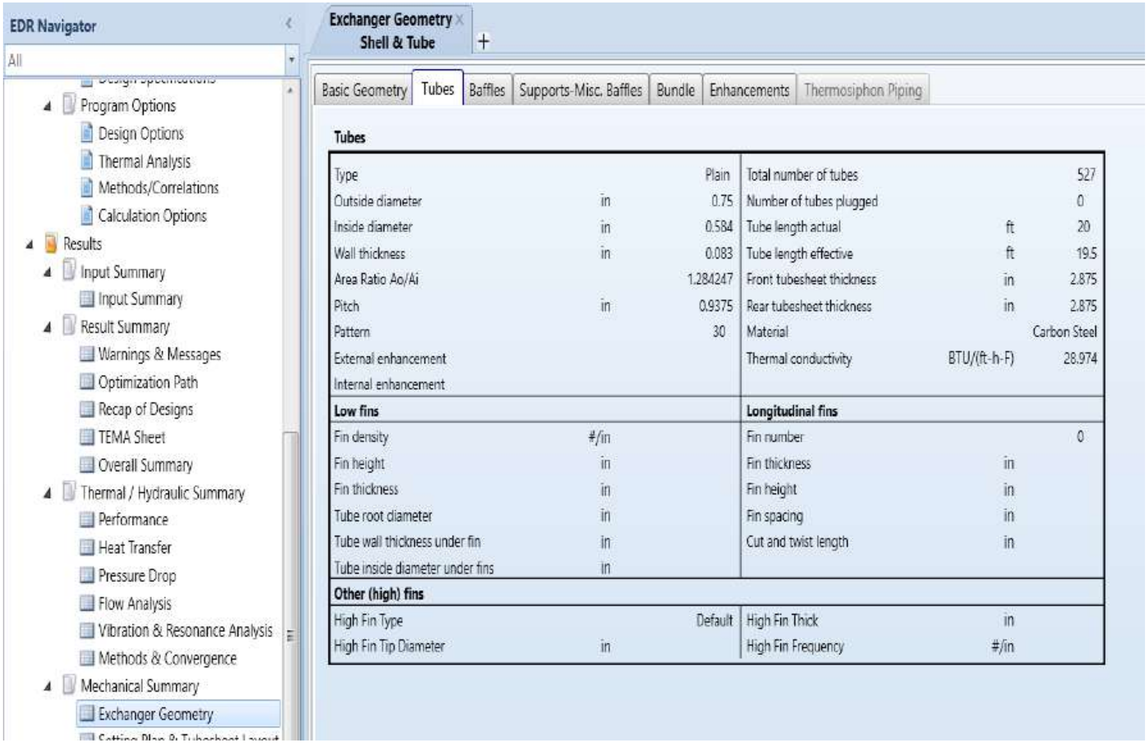


Unit Configuration

Exchanger type		BEM	Tube number		527
Position		Hor	Tube length actual	ft	20
Arrangement		1 parallel 1 series	Tube passes		2
Baffle type		Single segmental	Tube type		Plain
Baffle number		8	Tube O.D.	in	0.75
Spacing (center-center)	in	23.25	Tube pitch	in	0.9375
Spacing at inlet	in	35.625	Tube pattern		30

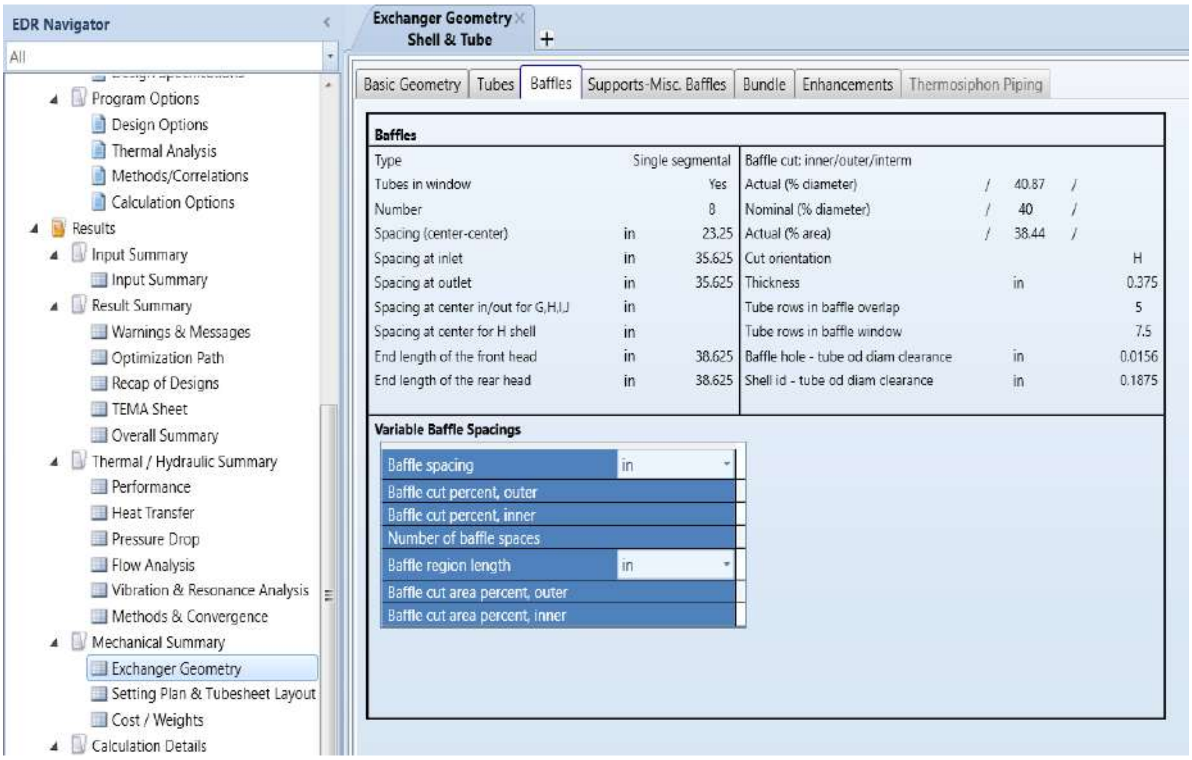
		Shell	Kettle	Front head	Rear Head
Outside diameter	in	26.75		26.75	26.75
Inside diameter	in	26		25.5	25.5

		Shell Side			Tube Side	
Nozzle type		Inlet	Outlet		Inlet	Outlet
Number of nozzles		1	1		1	1
Actual outside diameter	in	14	3.5		8.625	8.625
Inside diameter	in	13.25	3.068		7.981	7.981
Height under nozzle	in	4.5679	1.3203			
Dome inside diameter	in					
Vapor belt inside diameter	in					
Vapor belt inside width	in					
Vapor belt slot area	in^2					
Impingement protection		No impingement	No impingement	No impingement		
Distance to tubesheet	in	227	7.5			



Tubes

Type		Plain	Total number of tubes		527
Outside diameter	in	0.75	Number of tubes plugged		0
Inside diameter	in	0.584	Tube length actual	ft	20
Wall thickness	in	0.083	Tube length effective	ft	19.5
Area Ratio Ao/Ai		1.284247	Front tubesheet thickness	in	2.875
Pitch	in	0.9375	Rear tubesheet thickness	in	2.875
Pattern		30	Material		Carbon Steel
External enhancement			Thermal conductivity	BTU/(ft-h-F)	28.974
Internal enhancement					
Low fins			**Longitudinal fins**		
Fin density	#/in		Fin number		0
Fin height	in		Fin thickness	in	
Fin thickness	in		Fin height	in	
Tube root diameter	in		Fin spacing	in	
Tube wall thickness under fin	in		Cut and twist length	in	
Tube inside diameter under fins	in				
Other (high) fins					
High Fin Type		Default	High Fin Thick	in	
High Fin Tip Diameter	in		High Fin Frequency	#/in	

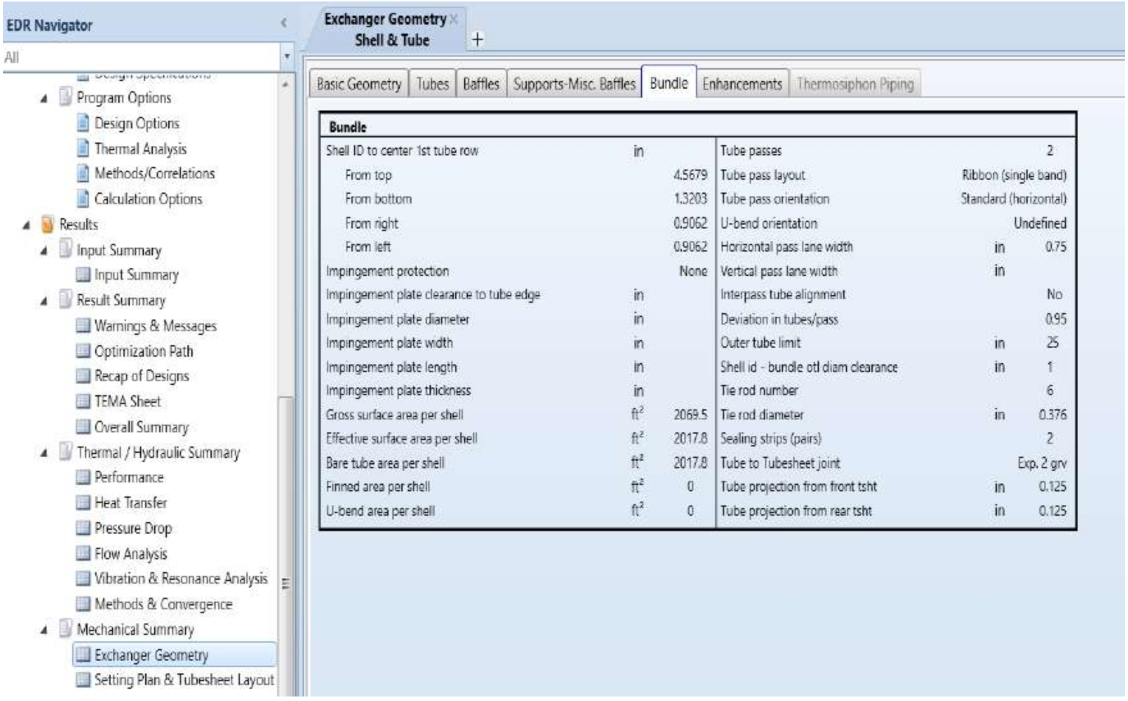


Baffles

Type		Single segmental	Baffle cut: inner/outer/interm		
Tubes in window		Yes	Actual (% diameter)		/ 40.87 /
Number		8	Nominal (% diameter)		/ 40 /
Spacing (center-center)	in	23.25	Actual (% area)		/ 38.44 /
Spacing at inlet	in	35.625	Cut orientation		H
Spacing at outlet	in	35.625	Thickness	in	0.375
Spacing at center in/out for G,H,I,J	in		Tube rows in baffle overlap		5
Spacing at center for H shell	in		Tube rows in baffle window		7.5
End length of the front head	in	38.625	Baffle hole - tube od diam clearance	in	0.0156
End length of the rear head	in	38.625	Shell id - tube od diam clearance	in	0.1875

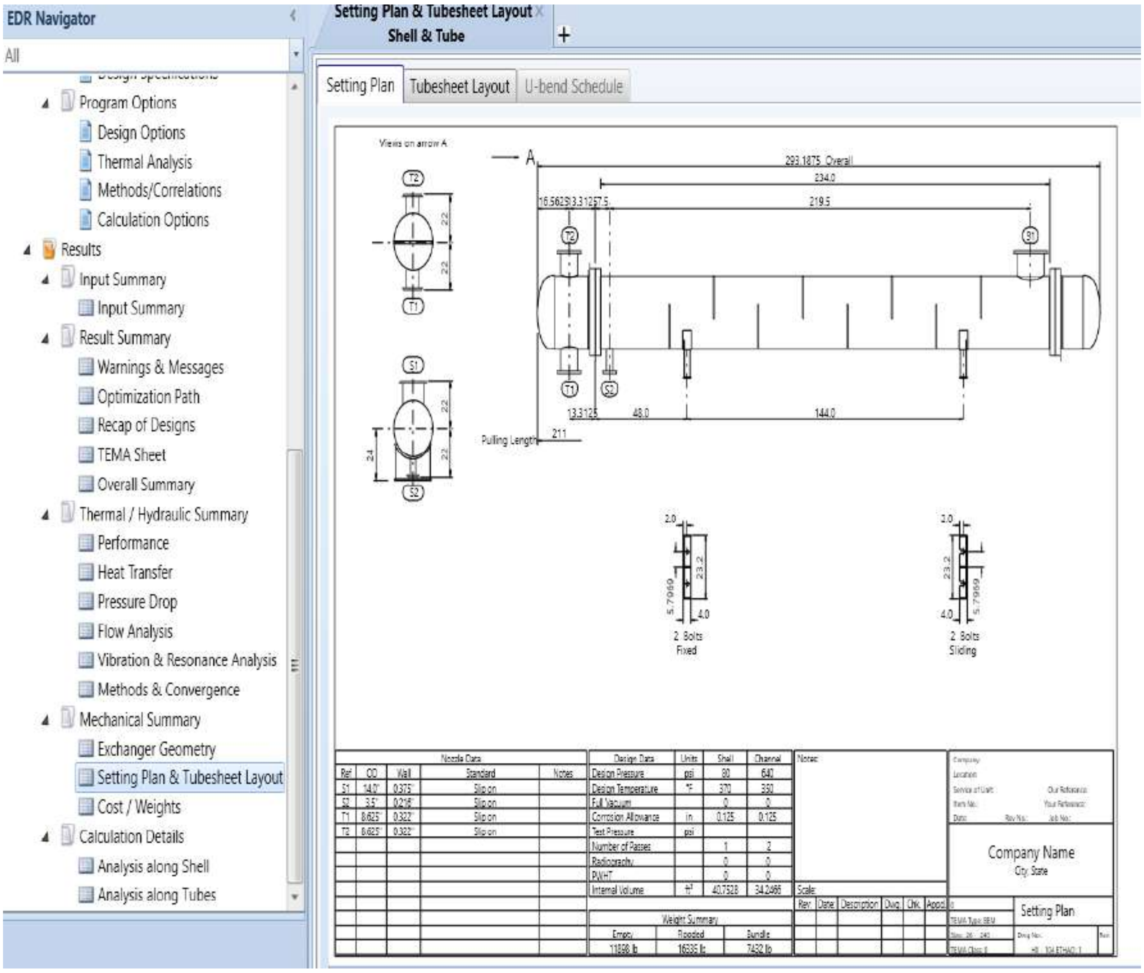
Variable Baffle Spacings

Baffle spacing	in
Baffle cut percent, outer	
Baffle cut percent, inner	
Number of baffle spaces	
Baffle region length	in
Baffle cut area percent, outer	
Baffle cut area percent, inner	



Bundle

Shell ID to center 1st tube row	in		Tube passes		2
From top		4.5679	Tube pass layout		Ribbon (single band)
From bottom		1.3203	Tube pass orientation		Standard (horizontal)
From right		0.9062	U-bend orientation		Undefined
From left		0.9062	Horizontal pass lane width	in	0.75
Impingement protection		None	Vertical pass lane width	in	
Impingement plate clearance to tube edge	in		Interpass tube alignment		No
Impingement plate diameter	in		Deviation in tubes/pass		0.95
Impingement plate width	in		Outer tube limit	in	25
Impingement plate length	in		Shell id - bundle otl diam clearance	in	1
Impingement plate thickness	in		Tie rod number		6
Gross surface area per shell	ft^2	2069.5	Tie rod diameter	in	0.376
Effective surface area per shell	ft^2	2017.8	Sealing strips (pairs)		2
Bare tube area per shell	ft^2	2017.8	Tube to Tubesheet joint		Exp. 2 grv
Finned area per shell	ft^2	0	Tube projection from front tsht	in	0.125
U-bend area per shell	ft^2	0	Tube projection from rear tsht	in	0.125



Nozzle Data				
Ref	OD	Wall	Standard	Notes
S1	14.0"	0.375"	Slip on	
S2	3.5"	0.216"	Slip on	
T1	8.625"	0.322"	Slip on	
T2	8.625"	0.322"	Slip on	

Design Data	Units	Shell	Channel
Design Pressure	psi	80	640
Design Temperature	°F	370	350
Full Vacuum		0	0
Corrosion Allowance	in	0.125	0.125
Test Pressure	psi		
Number of Passes		1	2
Radiography		0	0
PWHT		0	0
Internal Volume	ft^3	40.7528	34.2466

Weight Summary		
Empty	Flooded	Bundle
11898 lb	16535 lb	7432 lb

Notes:

Scale:

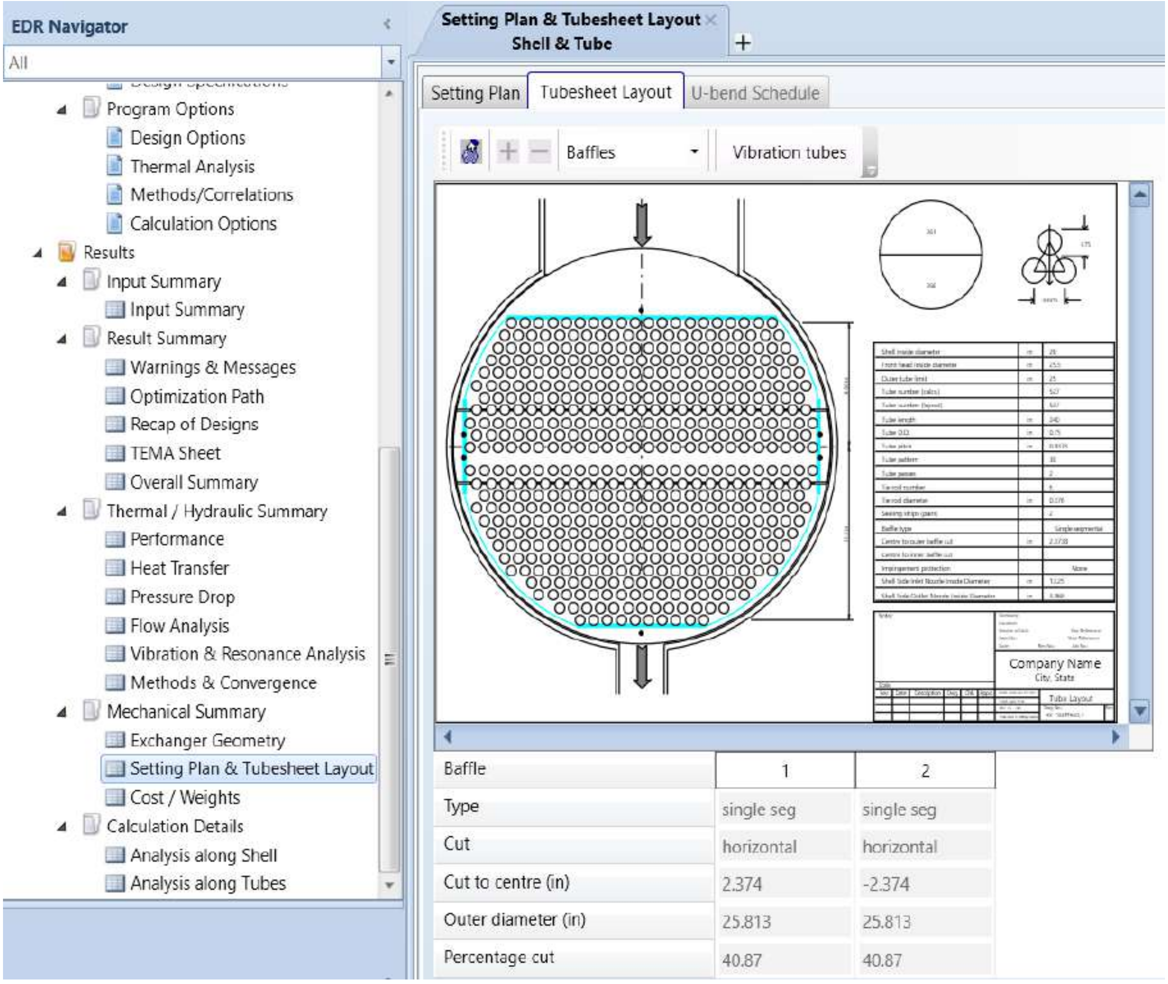
Rev.	Date	Description	Dwg.	Chk.	Appd.

Company:
Location:
Service of Unit: Our Reference:
Item No.: Your Reference:
Date: Rev No.: Job No.:

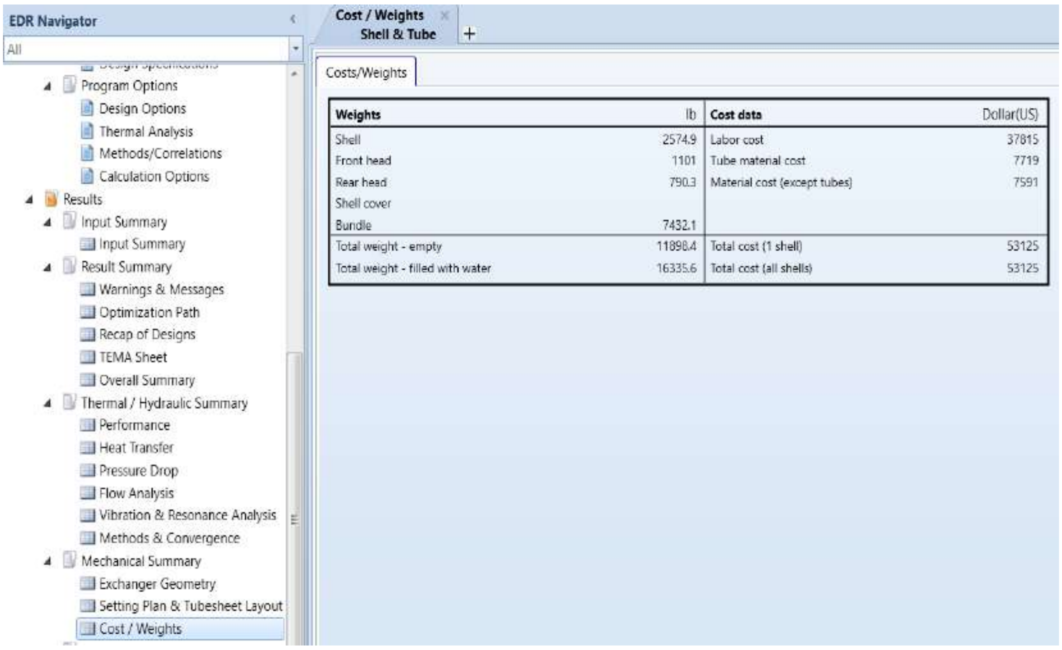
Company Name
City, State

Setting Plan

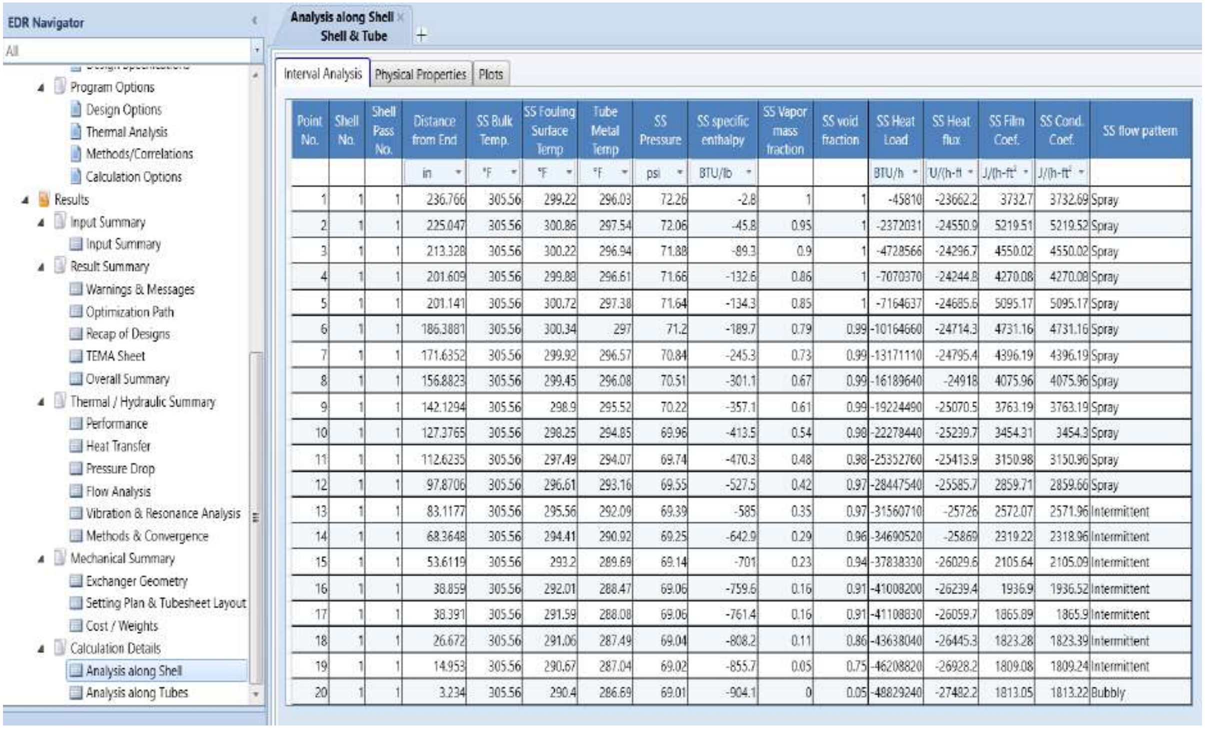
TEMA Type: BEM
Size: 26 - 240
TEMA Class: 8
Dwg No.:
HE - 104 ETHAD 1
Rev



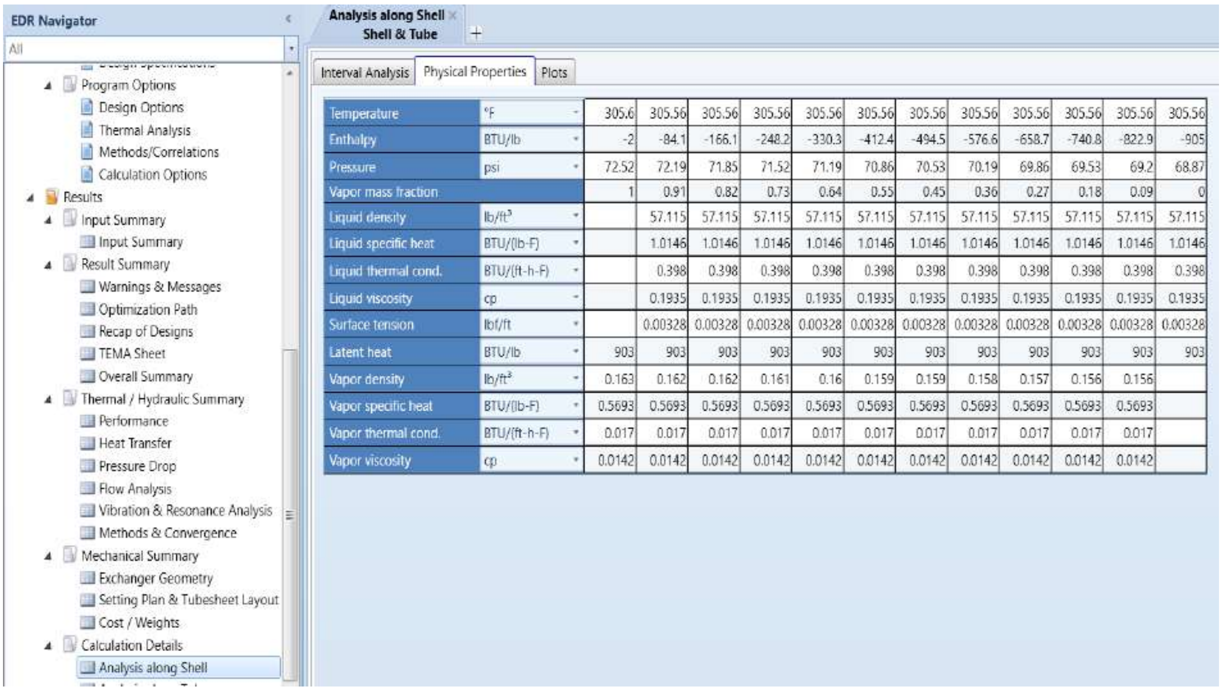
Baffle	1	2
Type	single seg	single seg
Cut	horizontal	horizontal
Cut to centre (in)	2.374	-2.374
Outer diameter (in)	25.813	25.813
Percentage cut	40.87	40.87



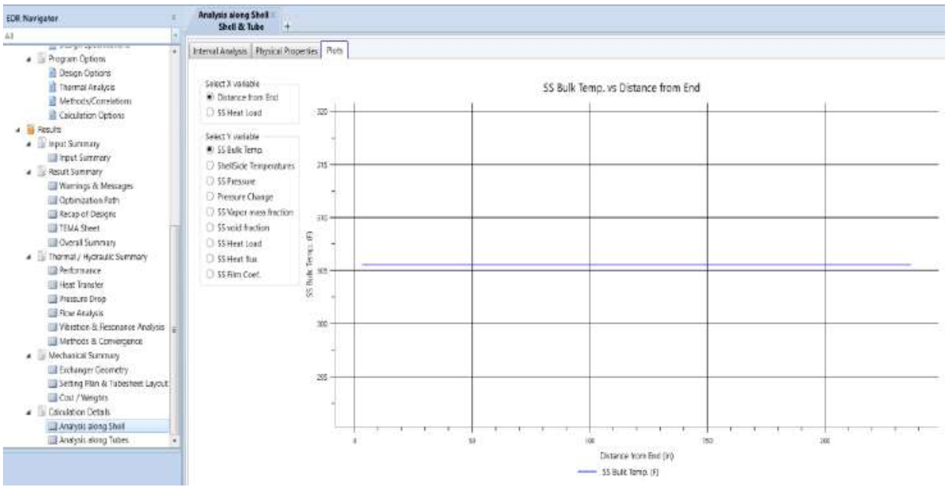
Weights	lb	**Cost data**	Dollar(US)
Shell	2574.9	Labor cost	37815
Front head	1101	Tube material cost	7719
Rear head	790.3	Material cost (except tubes)	7591
Shell cover			
Bundle	7432.1		
Total weight - empty	11898.4	Total cost (1 shell)	53125
Total weight - filled with water	16335.6	Total cost (all shells)	53125



Point No.	Shell No.	Shell Pass No.	Distance from End	SS Bulk Temp.	SS Fouling Surface Temp	Tube Metal Temp	SS Pressure	SS specific enthalpy	SS Vapor mass fraction	SS void fraction	SS Heat Load	SS Heat flux	SS Film Coef.	SS Cond. Coef.	SS flow pattern
			in	°F	°F	°F	psi	BTU/lb			BTU/h	'U/(h-ft	J/(h-ft²	J/(h-ft²	
1	1	1	236.766	305.56	299.22	296.03	72.26	-2.8	1	1	-45810	-23662.2	3732.7	3732.69	Spray
2	1	1	225.047	305.56	300.86	297.54	72.06	-45.8	0.95	1	-2372031	-24550.9	5219.51	5219.52	Spray
3	1	1	213.328	305.56	300.22	296.94	71.88	-89.3	0.9	1	-4728566	-24296.7	4550.02	4550.02	Spray
4	1	1	201.609	305.56	299.88	296.61	71.66	-132.6	0.86	1	-7070370	-24244.8	4270.08	4270.08	Spray
5	1	1	201.141	305.56	300.72	297.38	71.64	-134.3	0.85	1	-7164637	-24685.6	5095.17	5095.17	Spray
6	1	1	186.3881	305.56	300.34	297	71.2	-189.7	0.79	0.99	-10164660	-24714.3	4731.16	4731.16	Spray
7	1	1	171.6352	305.56	299.92	296.57	70.84	-245.3	0.73	0.99	-13171110	-24795.4	4396.19	4396.19	Spray
8	1	1	156.8823	305.56	299.45	296.08	70.51	-301.1	0.67	0.99	-16189640	-24918	4075.96	4075.96	Spray
9	1	1	142.1294	305.56	298.9	295.52	70.22	-357.1	0.61	0.99	-19224490	-25070.5	3763.19	3763.19	Spray
10	1	1	127.3765	305.56	298.25	294.85	69.96	-413.5	0.54	0.98	-22278440	-25239.7	3454.31	3454.3	Spray
11	1	1	112.6235	305.56	297.49	294.07	69.74	-470.3	0.48	0.98	-25352760	-25413.9	3150.98	3150.96	Spray
12	1	1	97.8706	305.56	296.61	293.16	69.55	-527.5	0.42	0.97	-28447540	-25585.7	2859.71	2859.66	Spray
13	1	1	83.1177	305.56	295.56	292.09	69.39	-585	0.35	0.97	-31560710	-25726	2572.07	2571.96	Intermittent
14	1	1	68.3648	305.56	294.41	290.92	69.25	-642.9	0.29	0.96	-34690520	-25869	2319.22	2318.96	Intermittent
15	1	1	53.6119	305.56	293.2	289.69	69.14	-701	0.23	0.94	-37838330	-26029.6	2105.64	2105.09	Intermittent
16	1	1	38.859	305.56	292.01	288.47	69.06	-759.6	0.16	0.91	-41008200	-26239.4	1936.9	1936.52	Intermittent
17	1	1	38.391	305.56	291.59	288.08	69.06	-761.4	0.16	0.91	-41108830	-26059.7	1865.89	1865.9	Intermittent
18	1	1	26.672	305.56	291.06	287.49	69.04	-808.2	0.11	0.86	-43638040	-26445.3	1823.28	1823.39	Intermittent
19	1	1	14.953	305.56	290.67	287.04	69.02	-855.7	0.05	0.75	-46208820	-26928.2	1809.08	1809.24	Intermittent
20	1	1	3.234	305.56	290.4	286.69	69.01	-904.1	0	0.05	-48829240	-27482.2	1813.05	1813.22	Bubbly

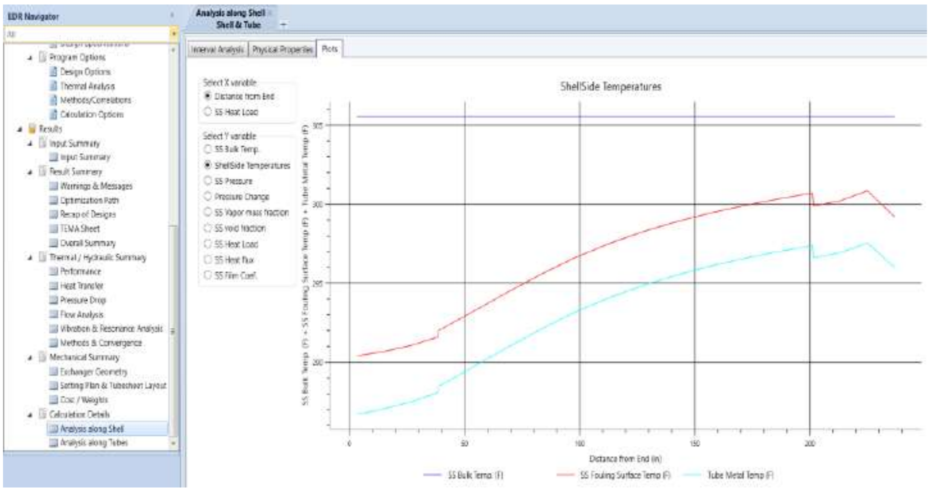


Temperature	°F	305.6	305.56	305.56	305.56	305.56	305.56	305.56	305.56	305.56	305.56	305.56	305.56
Enthalpy	BTU/lb	-2	-84.1	-166.1	-248.2	-330.3	-412.4	-494.5	-576.6	-658.7	-740.8	-822.9	-905
Pressure	psi	72.52	72.19	71.85	71.52	71.19	70.86	70.53	70.19	69.86	69.53	69.2	68.87
Vapor mass fraction		1	0.91	0.82	0.73	0.64	0.55	0.45	0.36	0.27	0.18	0.09	0
Liquid density	lb/ft³		57.115	57.115	57.115	57.115	57.115	57.115	57.115	57.115	57.115	57.115	57.115
Liquid specific heat	BTU/(lb-F)		1.0146	1.0146	1.0146	1.0146	1.0146	1.0146	1.0146	1.0146	1.0146	1.0146	1.0146
Liquid thermal cond.	BTU/(ft-h-F)		0.398	0.398	0.398	0.398	0.398	0.398	0.398	0.398	0.398	0.398	0.398
Liquid viscosity	cp		0.1935	0.1935	0.1935	0.1935	0.1935	0.1935	0.1935	0.1935	0.1935	0.1935	0.1935
Surface tension	lbf/ft		0.00328	0.00328	0.00328	0.00328	0.00328	0.00328	0.00328	0.00328	0.00328	0.00328	0.00328
Latent heat	BTU/lb	903	903	903	903	903	903	903	903	903	903	903	903
Vapor density	lb/ft³	0.163	0.162	0.162	0.161	0.16	0.159	0.159	0.158	0.157	0.156	0.156	
Vapor specific heat	BTU/(lb-F)	0.5693	0.5693	0.5693	0.5693	0.5693	0.5693	0.5693	0.5693	0.5693	0.5693	0.5693	
Vapor thermal cond.	BTU/(ft-h-F)	0.017	0.017	0.017	0.017	0.017	0.017	0.017	0.017	0.017	0.017	0.017	
Vapor viscosity	cp	0.0142	0.0142	0.0142	0.0142	0.0142	0.0142	0.0142	0.0142	0.0142	0.0142	0.0142	



Select X variable
● Distance from End
○ SS Heat Load

Select Y variable
● SS Bulk Temp.
○ ShellSide Temperatures
○ SS Pressure
○ Pressure Change
○ SS Vapor mass fraction
○ SS void fraction
○ SS Heat Load
○ SS Heat flux
○ SS Film Coef.



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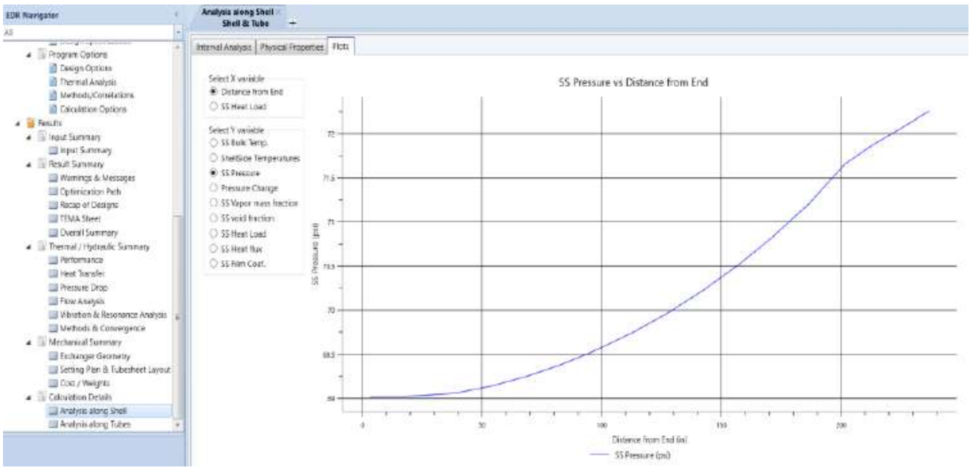

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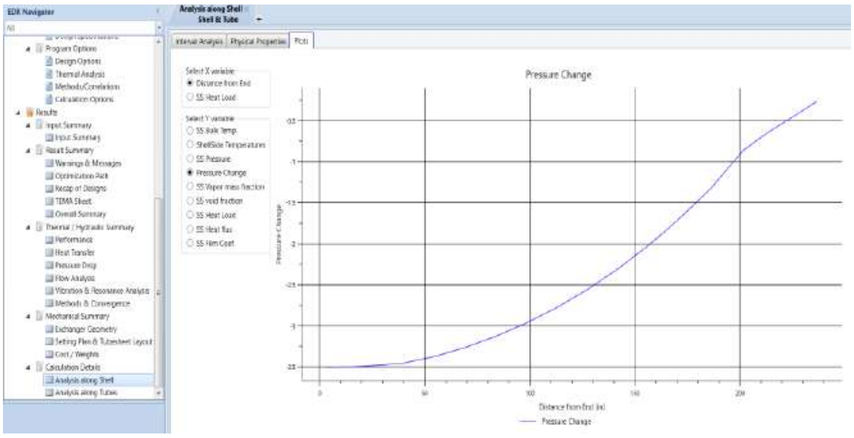

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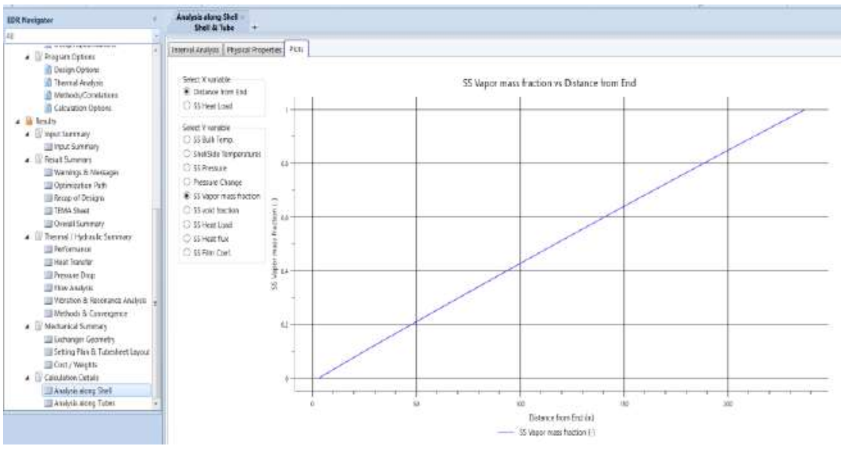

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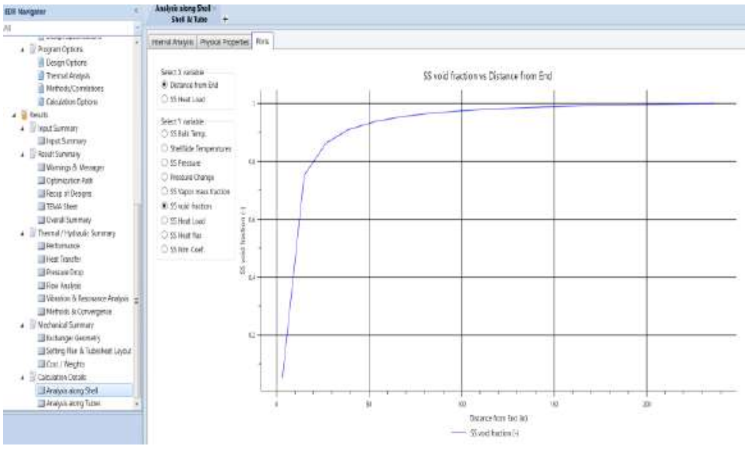

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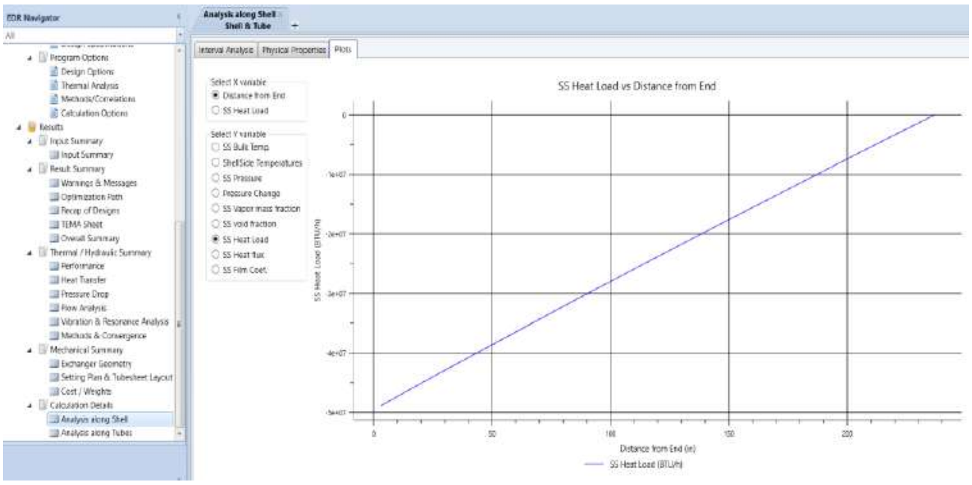

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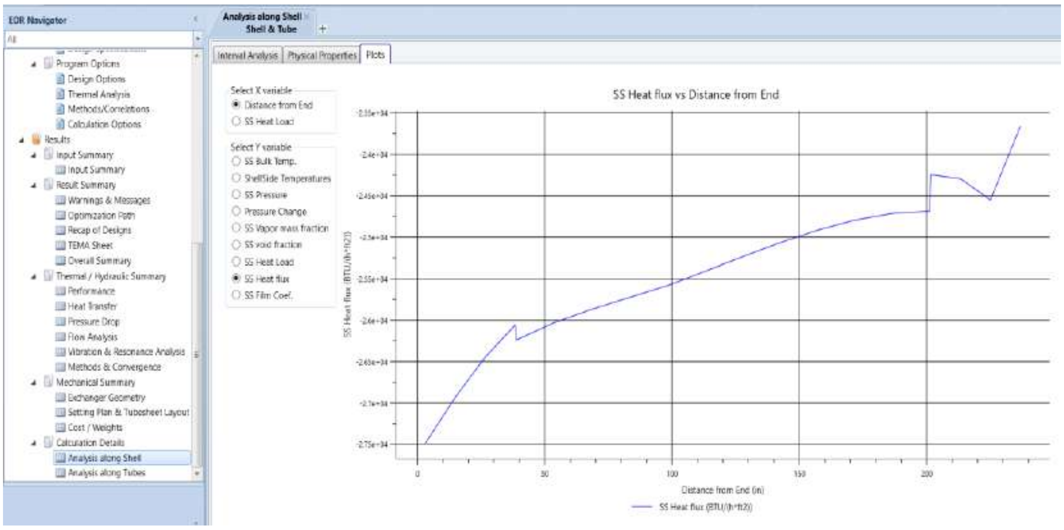

Select X variable
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○ SS Heat Load

Select Y variable
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○ ShellSide Temperatures
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○ Pressure Change
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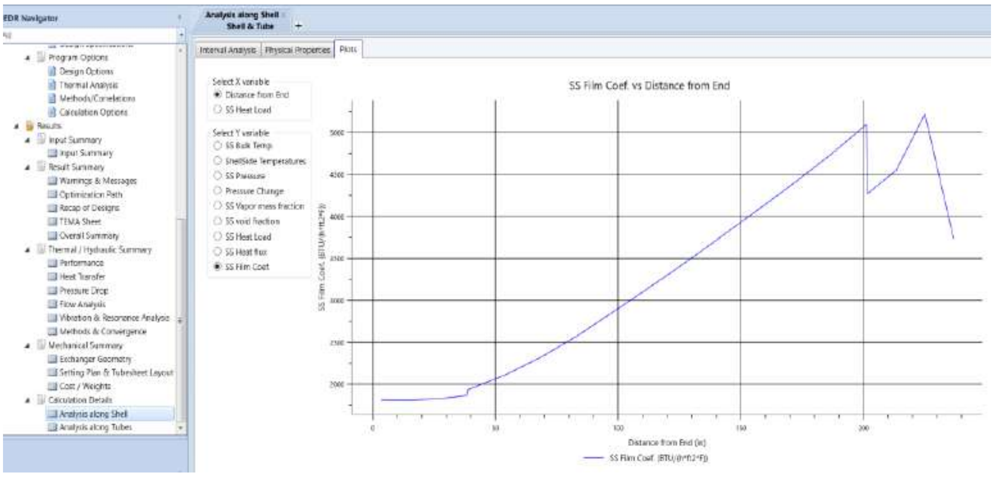
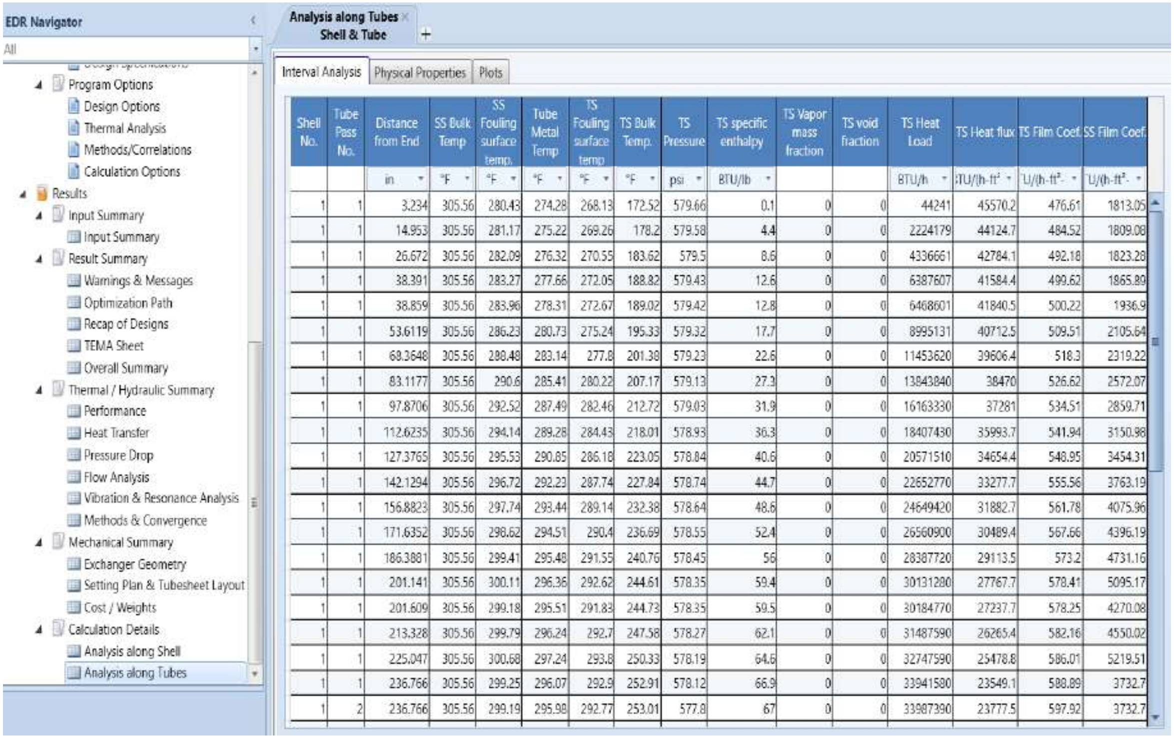
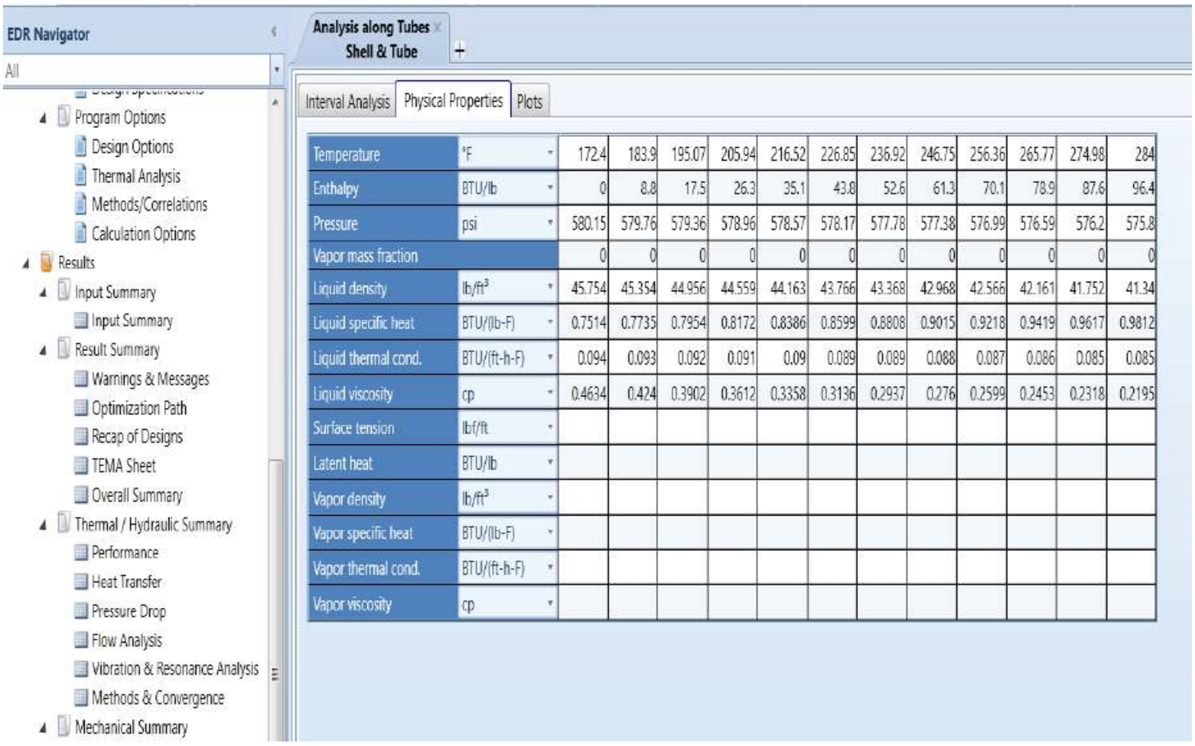


Select X variable
● Distance from End
○ SS Heat Load

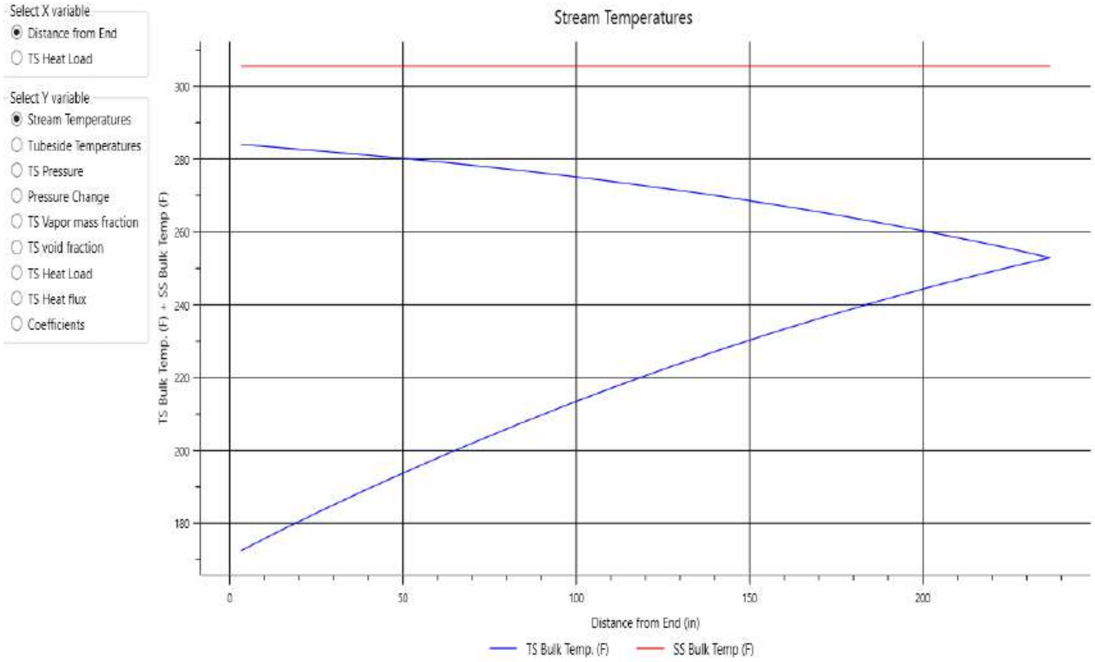
Select Y variable
○ SS Bulk Temp.
○ ShellSide Temperatures
○ SS Pressure
○ Pressure Change
○ SS Vapor mass fraction
○ SS void fraction
○ SS Heat Load
○ SS Heat flux
● SS Film Coef.

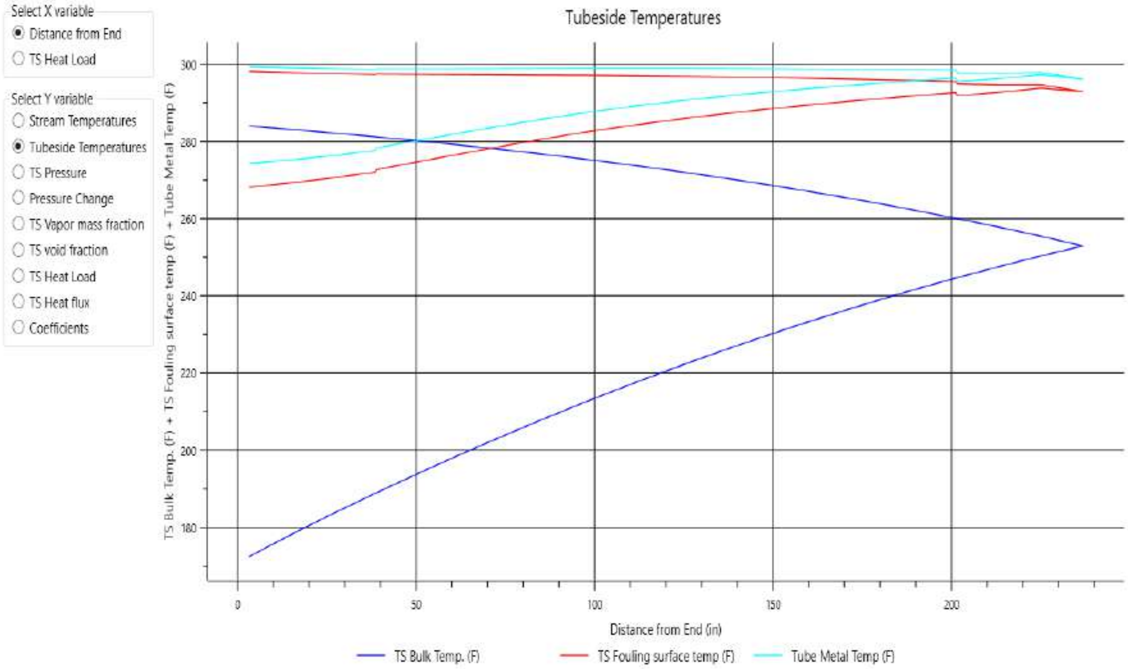


Shell No.	Tube Pass No.	Distance from End	SS Bulk Temp	SS Fouling surface temp.	Tube Metal Temp	TS Fouling surface temp	TS Bulk Temp.	TS Pressure	TS specific enthalpy	TS Vapor mass fraction	TS void fraction	TS Heat Load	TS Heat flux	TS Film Coef.	SS Film Coef.
		in	°F	°F	°F	°F	°F	psi	BTU/lb			BTU/h	TU/(h-ft^2	U/(h-ft^2-	U/(h-ft^2-
1	1	3.234	305.56	280.43	274.28	268.13	172.52	579.66	0.1	0	0	44241	45570.2	476.61	1813.05
1	1	14.953	305.56	281.17	275.22	269.26	178.2	579.58	4.4	0	0	2224179	44124.7	484.52	1809.08
1	1	26.672	305.56	282.09	276.32	270.55	183.62	579.5	8.6	0	0	4336661	42784.1	492.18	1823.28
1	1	38.391	305.56	283.27	277.66	272.05	188.82	579.43	12.6	0	0	6387607	41584.4	499.62	1865.89
1	1	38.859	305.56	283.96	278.31	272.67	189.02	579.42	12.8	0	0	6468601	41840.5	500.22	1936.9
1	1	53.6119	305.56	286.23	280.73	275.24	195.33	579.32	17.7	0	0	8995131	40712.5	509.51	2105.64
1	1	68.3648	305.56	288.48	283.14	277.8	201.38	579.23	22.6	0	0	11453620	39606.4	518.3	2319.22
1	1	83.1177	305.56	290.6	285.41	280.22	207.17	579.13	27.3	0	0	13843840	38470	526.62	2572.07
1	1	97.8706	305.56	292.52	287.49	282.46	212.72	579.03	31.9	0	0	16163330	37281	534.51	2859.71
1	1	112.6235	305.56	294.14	289.28	284.43	218.01	578.93	36.3	0	0	18407430	35993.7	541.94	3150.98
1	1	127.3765	305.56	295.53	290.85	286.18	223.05	578.84	40.6	0	0	20571510	34654.4	548.95	3454.31
1	1	142.1294	305.56	296.72	292.23	287.74	227.84	578.74	44.7	0	0	22652770	33277.7	555.56	3763.19
1	1	156.8823	305.56	297.74	293.44	289.14	232.38	578.64	48.6	0	0	24649420	31882.7	561.78	4075.96
1	1	171.6352	305.56	298.62	294.51	290.4	236.69	578.55	52.4	0	0	26560900	30489.4	567.66	4396.19
1	1	186.3881	305.56	299.41	295.48	291.55	240.76	578.45	56	0	0	28387720	29113.5	573.2	4731.16
1	1	201.141	305.56	300.11	296.36	292.62	244.61	578.35	59.4	0	0	30131280	27767.7	578.41	5095.17
1	1	201.609	305.56	299.18	295.51	291.83	244.73	578.35	59.5	0	0	30184770	27237.7	578.25	4270.08
1	1	213.328	305.56	299.79	296.24	292.7	247.58	578.27	62.1	0	0	31487590	26265.4	582.16	4550.02
1	1	225.047	305.56	300.68	297.24	293.8	250.33	578.19	64.6	0	0	32747590	25478.8	586.01	5219.51
1	1	236.766	305.56	299.25	296.07	292.9	252.91	578.12	66.9	0	0	33941580	23549.1	588.89	3732.7
1	2	236.766	305.56	299.19	295.98	292.77	253.01	577.8	67	0	0	33987390	23777.5	597.92	3732.7



Temperature	°F	172.4	183.9	195.07	205.94	216.52	226.85	236.92	246.75	256.36	265.77	274.98	284
Enthalpy	BTU/lb	0	8.8	17.5	26.3	35.1	43.8	52.6	61.3	70.1	78.9	87.6	96.4
Pressure	psi	580.15	579.76	579.36	578.96	578.57	578.17	577.78	577.38	576.99	576.59	576.2	575.8
Vapor mass fraction		0	0	0	0	0	0	0	0	0	0	0	0
Liquid density	lb/ft^3	45.754	45.354	44.956	44.559	44.163	43.766	43.368	42.968	42.566	42.161	41.752	41.34
Liquid specific heat	BTU/(lb-F)	0.7514	0.7735	0.7954	0.8172	0.8386	0.8599	0.8808	0.9015	0.9218	0.9419	0.9617	0.9812
Liquid thermal cond.	BTU/(ft-h-F)	0.094	0.093	0.092	0.091	0.09	0.089	0.089	0.088	0.087	0.086	0.085	0.085
Liquid viscosity	cp	0.4634	0.424	0.3902	0.3612	0.3358	0.3136	0.2937	0.276	0.2599	0.2453	0.2318	0.2195
Surface tension	lbf/ft												
Latent heat	BTU/lb												
Vapor density	lb/ft^3												
Vapor specific heat	BTU/(lb-F)												
Vapor thermal cond.	BTU/(ft-h-F)												
Vapor viscosity	cp												


Select X variable
● Distance from End
○ TS Heat Load
Select Y variable
● Stream Temperatures
○ Tubeside Temperatures
○ TS Pressure
○ Pressure Change
○ TS Vapor mass fraction
○ TS void fraction
○ TS Heat Load
○ TS Heat flux
○ Coefficients
Stream Temperatures
TS Bulk Temp. (F) + SS Bulk Temp (F)
300
280
260
240
220
200
180
0
50
100
150
200
Distance from End (in)
TS Bulk Temp. (F)
SS Bulk Temp (F)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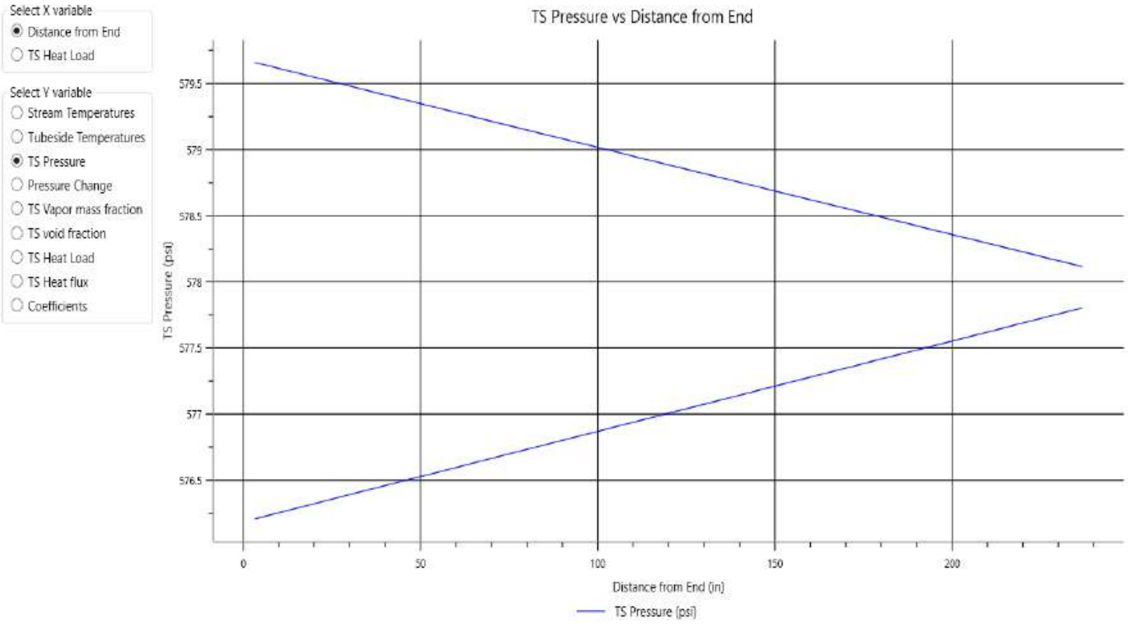
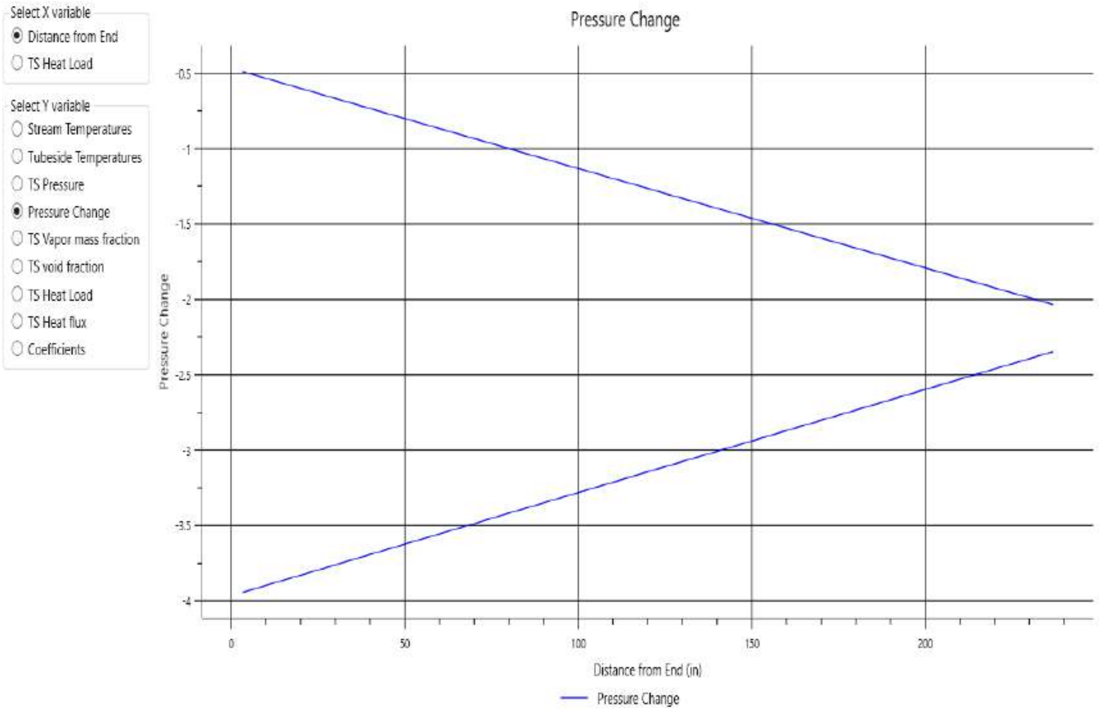
○ TS Heat Load
Select Y variable
○ Stream Temperatures
● Tubeside Temperatures
○ TS Pressure
○ Pressure Change
○ TS Vapor mass fraction
○ TS void fraction
○ TS Heat Load
○ TS Heat flux
○ Coefficients
Tubeside Temperatures
TS Bulk Temp. (F) + TS Fouling surface temp (F) + Tube Metal Temp (F)
300
280
260
240
220
200
180
0
50
100
150
200
Distance from End (in)
TS Bulk Temp. (F)
TS Fouling surface temp (F)
Tube Metal Temp (F)



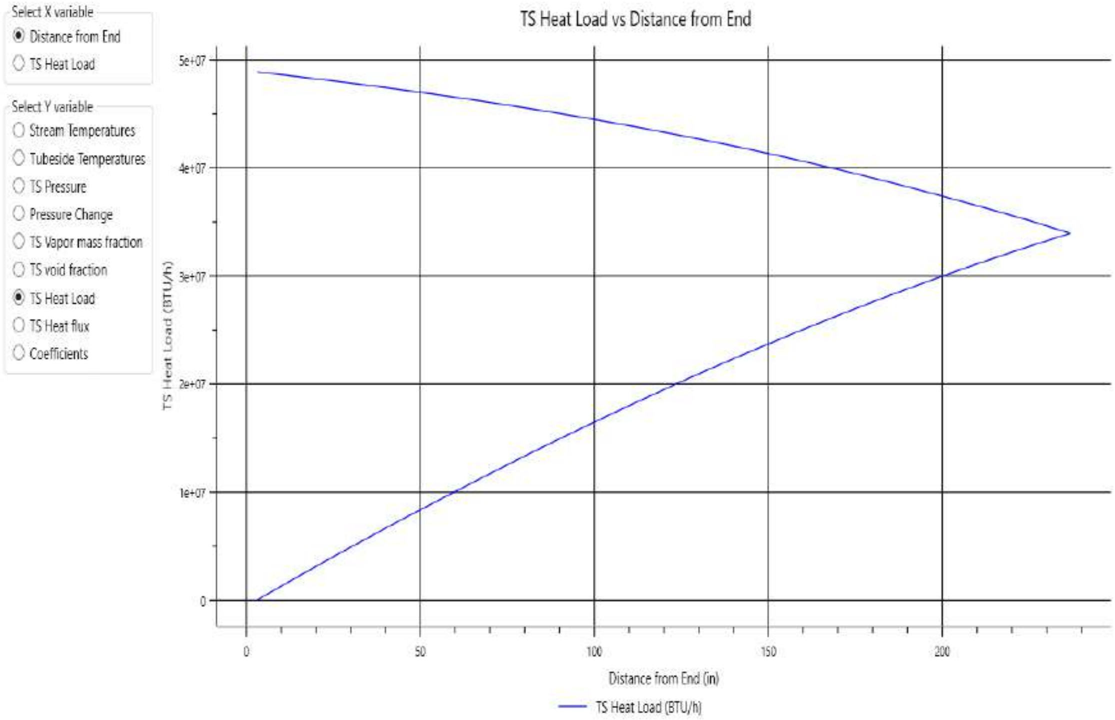
Select X variable
● Distance from End
○ TS Heat Load

Select Y variable
○ Stream Temperatures
○ Tubeside Temperatures
● TS Pressure
○ Pressure Change
○ TS Vapor mass fraction
○ TS void fraction
○ TS Heat Load
○ TS Heat flux
○ Coefficients



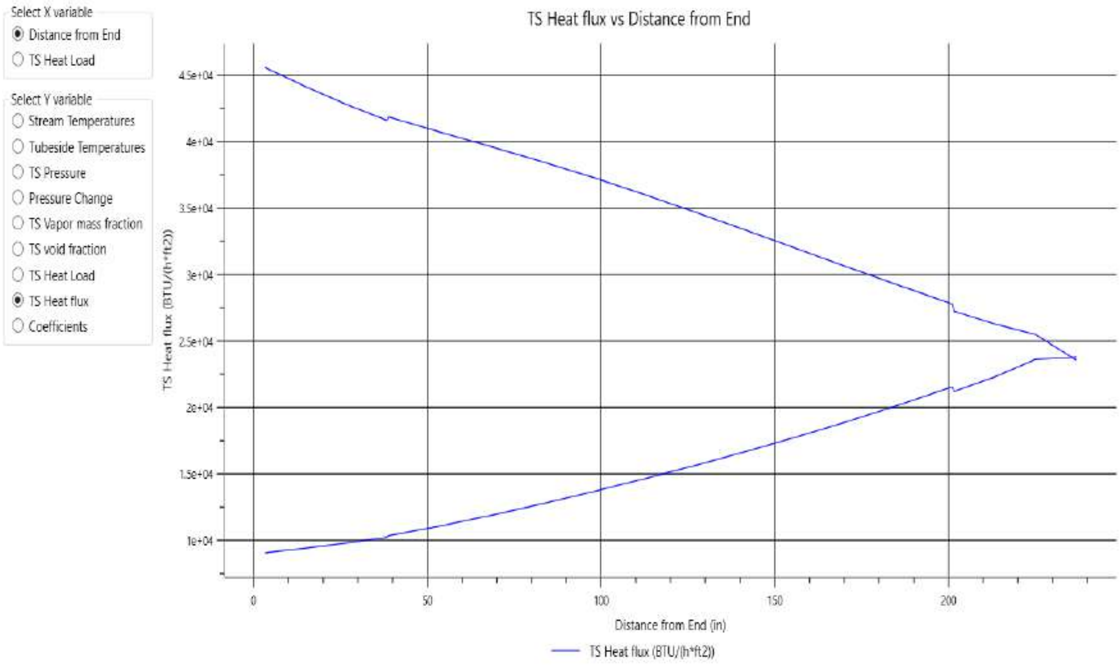
Select X variable
● Distance from End
○ TS Heat Load

Select Y variable
○ Stream Temperatures
○ Tubeside Temperatures
○ TS Pressure
● Pressure Change
○ TS Vapor mass fraction
○ TS void fraction
○ TS Heat Load
○ TS Heat flux
○ Coefficients



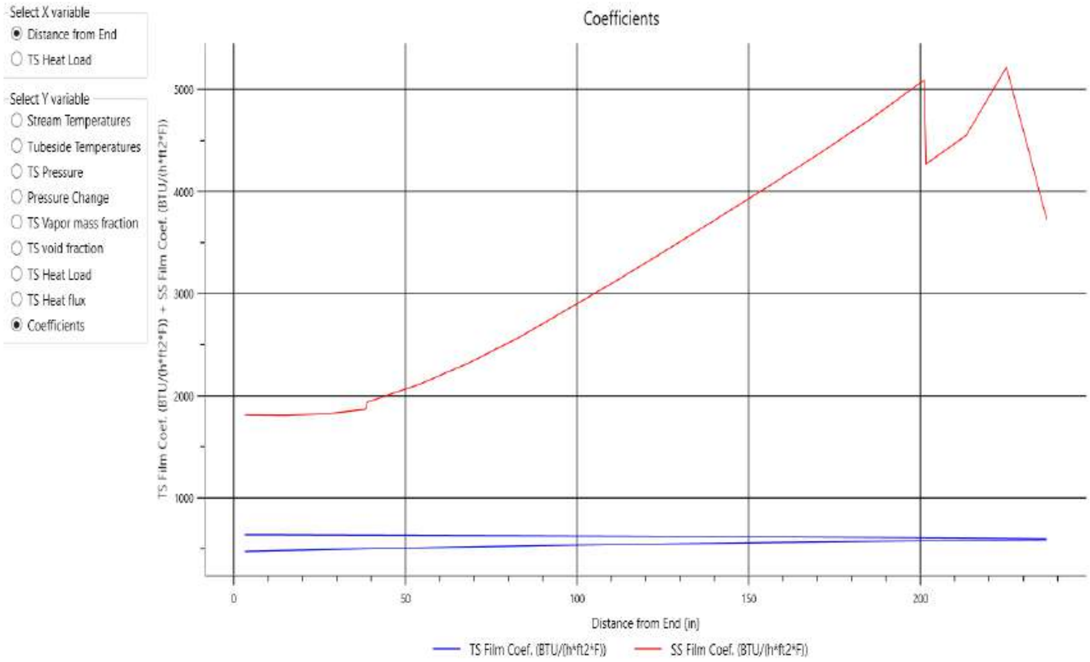
Select X variable
● Distance from End
○ TS Heat Load

Select Y variable
○ Stream Temperatures
○ Tubeside Temperatures
○ TS Pressure
○ Pressure Change
○ TS Vapor mass fraction
○ TS void fraction
● TS Heat Load
○ TS Heat flux
○ Coefficients



Select X variable
● Distance from End
○ TS Heat Load

Select Y variable
○ Stream Temperatures
○ Tubeside Temperatures
○ TS Pressure
○ Pressure Change
○ TS Vapor mass fraction
○ TS void fraction
○ TS Heat Load
● TS Heat flux
○ Coefficients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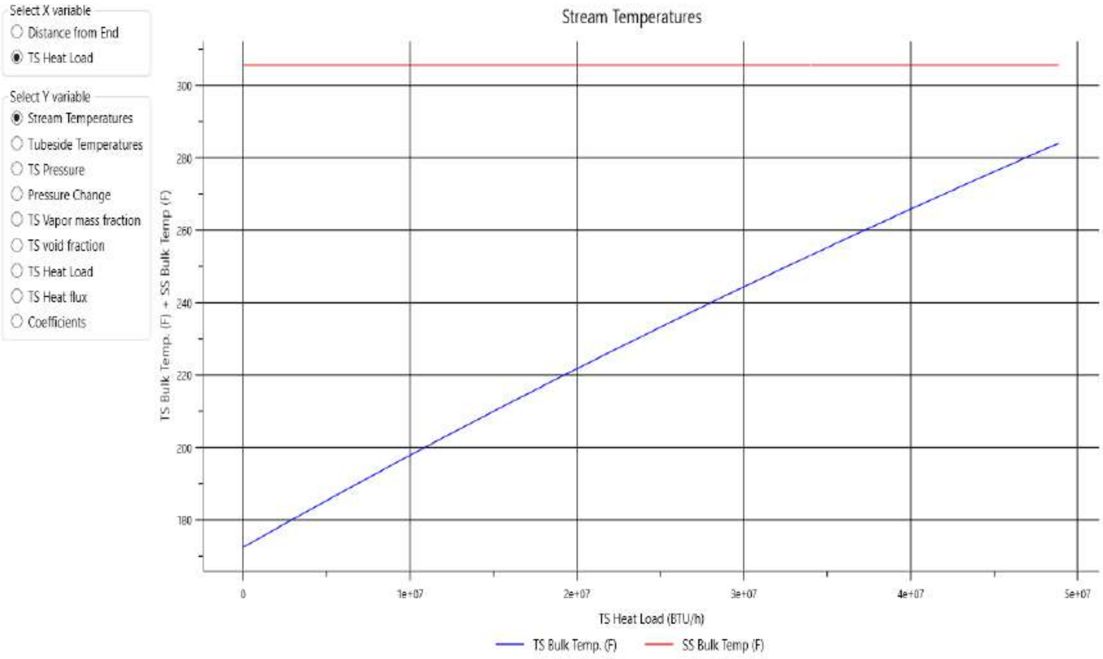
○ TS Heat Load
Select Y variable
○ Stream Temperatures
○ Tubeside Temperatures
○ TS Pressure
○ Pressure Change
○ TS Vapor mass fraction
○ TS void fraction
○ TS Heat Load
○ TS Heat flux
● Coefficients
Coefficients
TS Film Coef. (BTU/(h*ft2*F)) + SS Film Coef. (BTU/(h*ft2*F))
5000
4000
3000
2000
1000
0
50
100
150
200
Distance from End (in)
TS Film Coef. (BTU/(h*ft2*F))
SS Film Coef. (BTU/(h*ft2*F))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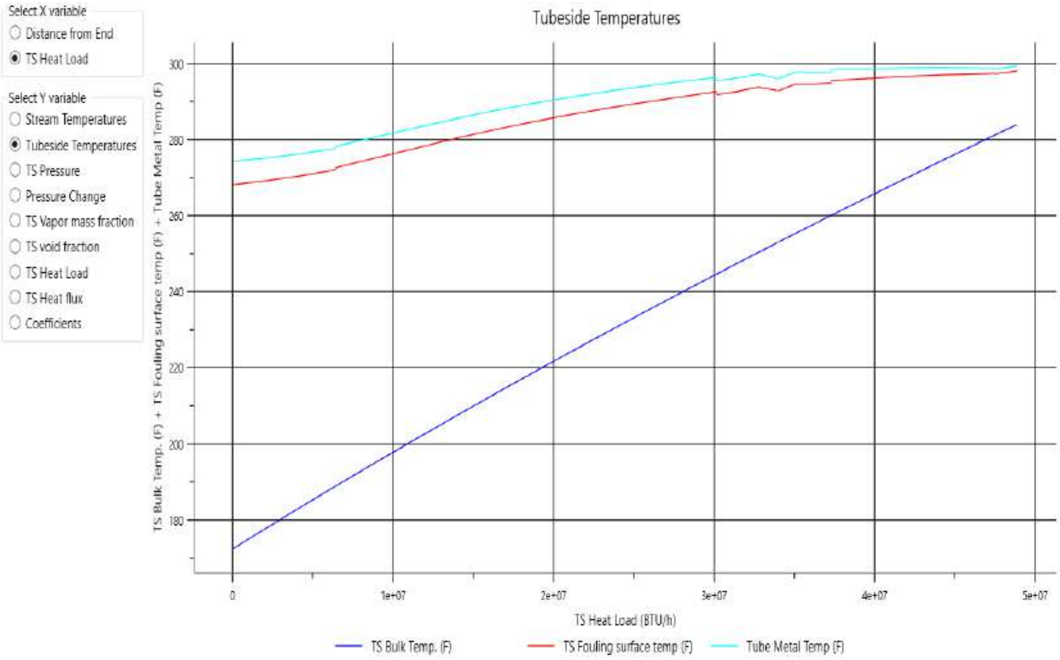
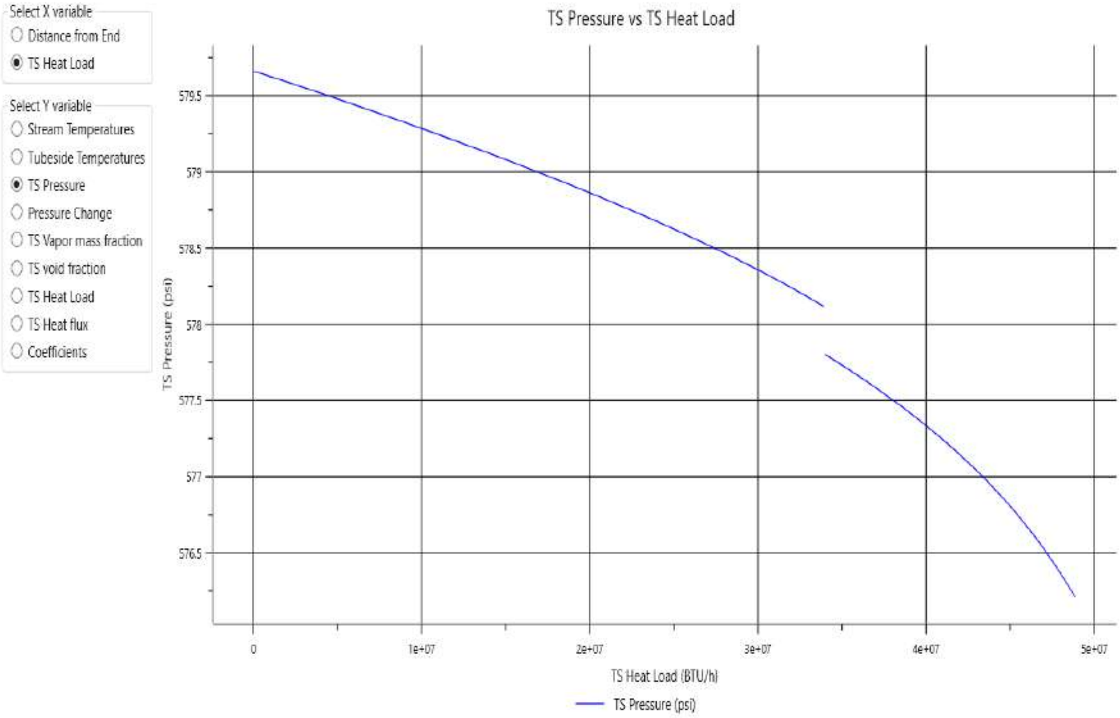
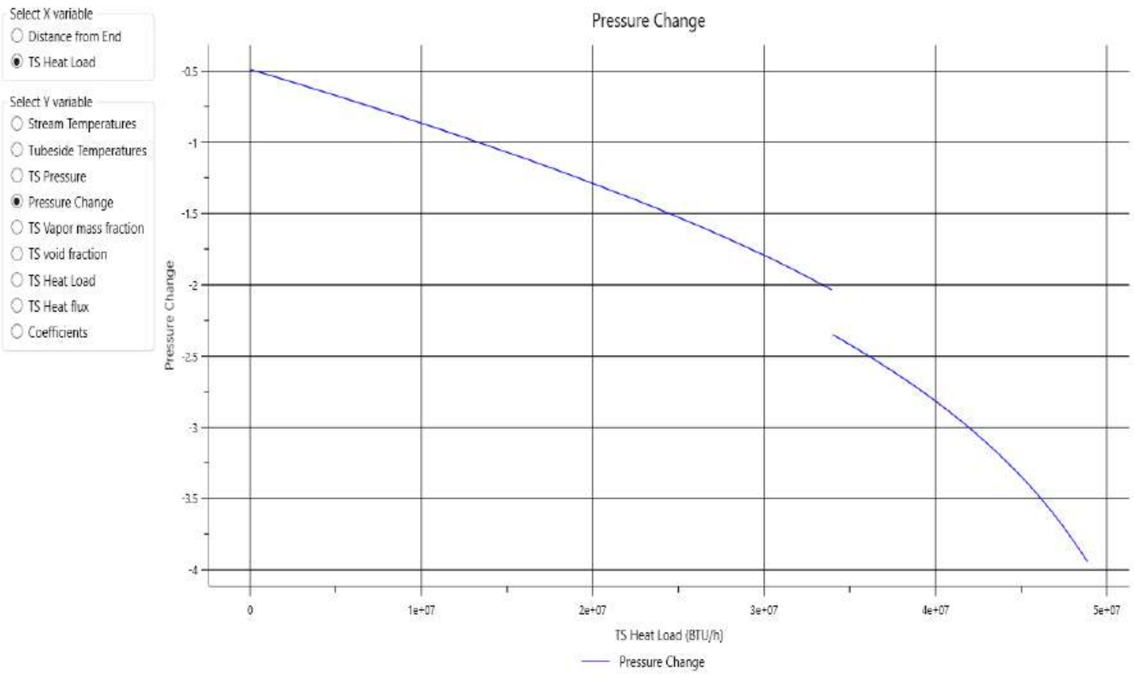
○ Distance from End
● TS Heat Load
Select Y variable
● Stream Temperatures
○ Tubeside Temperatures
○ TS Pressure
○ Pressure Change
○ TS Vapor mass fraction
○ TS void fraction
○ TS Heat Load
○ TS Heat flux
○ Coefficients
Stream Temperatures
TS Bulk Temp. (F) + SS Bulk Temp (F)
300
280
260
240
220
200
180
0
1e+07
2e+07
3e+07
4e+07
5e+07
TS Heat Load (BTU/h)
TS Bulk Temp. (F)
SS Bulk Temp (F)


Select X variable
○ Distance from End
● TS Heat Load
Select Y variable
○ Stream Temperatures
● Tubeside Temperatures
○ TS Pressure
○ Pressure Change
○ TS Vapor mass fraction
○ TS void fraction
○ TS Heat Load
○ TS Heat flux
○ Coefficients
Tubeside Temperatures
TS Bulk Temp. (F) + TS Fouling surface temp (F) + Tube Metal Temp (F)
300
280
260
240
220
200
180
0
1e+07
2e+07
3e+07
4e+07
5e+07
TS Heat Load (BTU/h)
TS Bulk Temp. (F)
TS Fouling surface temp (F)
Tube Metal Temp (F)


Select X variable
○ Distance from End
● TS Heat Load
Select Y variable
○ Stream Temperatures
○ Tubeside Temperatures
● TS Pressure
○ Pressure Change
○ TS Vapor mass fraction
○ TS void fraction
○ TS Heat Load
○ TS Heat flux
○ Coefficients
TS Pressure vs TS Heat Load
TS Pressure (psi)
579.5
579
578.5
578
577.5
577
576.5
0
1e+07
2e+07
3e+07
4e+07
5e+07
TS Heat Load (BTU/h)
TS Pressure (psi)


Select X variable
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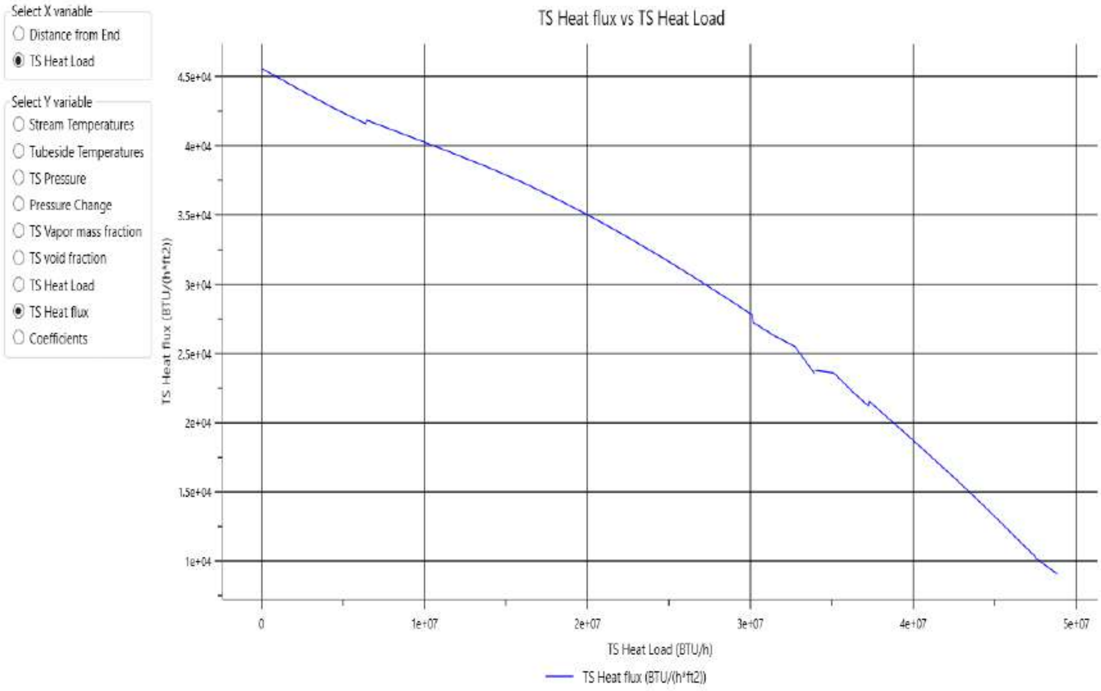
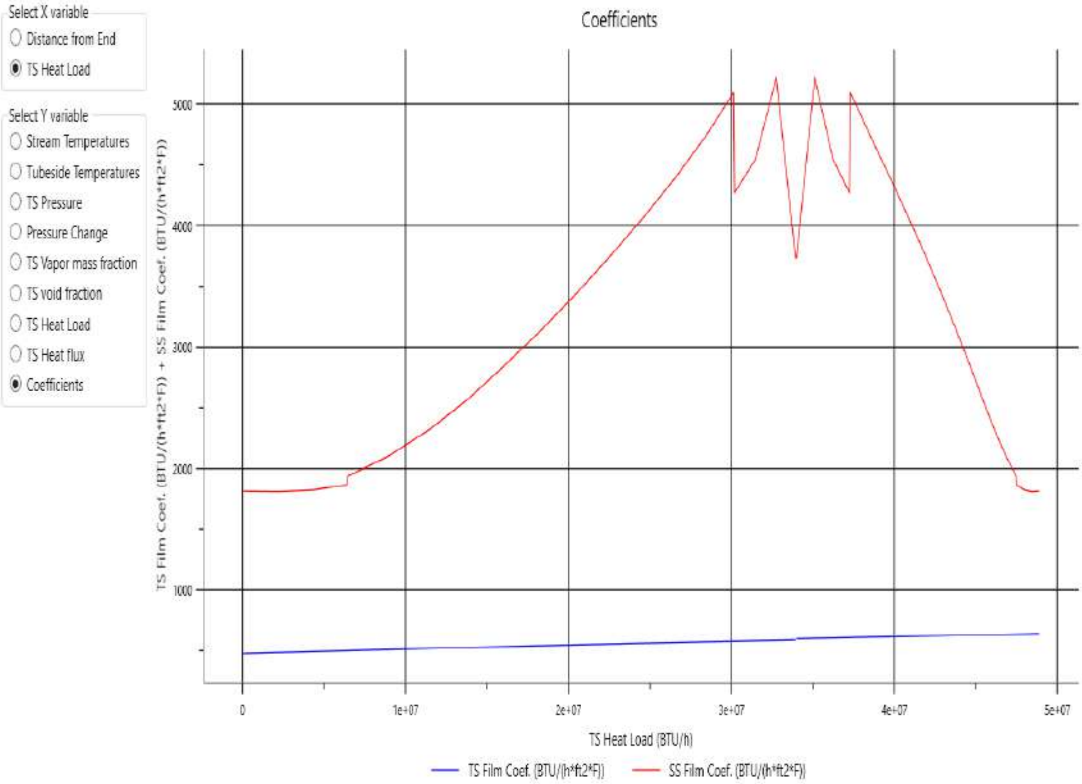
● TS Heat Load
Select Y variable
○ Stream Temperatures
○ Tubeside Temperatures
○ TS Pressure
● Pressure Change
○ TS Vapor mass fraction
○ TS void fraction
○ TS Heat Load
○ TS Heat flux
○ Coefficients
Pressure Change
Pressure Change
-0.5
-1
-1.5
-2
-2.5
-3
-3.5
-4
0
1e+07
2e+07
3e+07
4e+07
5e+07
TS Heat Load (BTU/h)
Pressure Change


Select X variable
○ Distance from End
● TS Heat Load
Select Y variable
○ Stream Temperatures
○ Tubeside Temperatures
○ TS Pressure
○ Pressure Change
○ TS Vapor mass fraction
○ TS void fraction
○ TS Heat Load
● TS Heat flux
○ Coefficients
TS Heat flux vs TS Heat Load
TS Heat flux (BTU/(h*ft2))
4.5e+04
4e+04
3.5e+04
3e+04
2.5e+04
2e+04
1.5e+04
1e+04
0
1e+07
2e+07
3e+07
4e+07
5e+07
TS Heat Load (BTU/h)
TS Heat flux (BTU/(h*ft2))


Select X variable
○ Distance from End
● TS Heat Load
Select Y variable
○ Stream Temperatures
○ Tubeside Temperatures
○ TS Pressure
○ Pressure Change
○ TS Vapor mass fraction
○ TS void fraction
○ TS Heat Load
○ TS Heat flux
● Coefficients
Coefficients
TS Film Coef. (BTU/(h*ft2*F)) + SS Film Coef. (BTU/(h*ft2*F))
5000
4000
3000
2000
1000
0
1e+07
2e+07
3e+07
4e+07
5e+07
TS Heat Load (BTU/h)
TS Film Coef. (BTU/(h*ft2*F))
SS Film Coef. (BTU/(h*ft2*F))

CHAPTER # 09

COST ESTIMATION

9.1 Economic Evaluation:

The economic analysis is the crucial step in any design. A solid project should guarantee a speedy return on the investment and excellent profitability during the predicted lifetime, similar to other industrial or financial ventures. It is both a specific field and a profession to estimate costs. To choose between different designs and for project evaluation, the design engineer must be able to quickly and roughly estimate costs. Chemical plants are constructed with the intention of making a profit, thus before the profitability of a project can be evaluated, an estimate of the investment necessary and the cost of production is needed. In this project, a variety of factors that make up a plant's capital cost and its operational costs are taken into account, and the cost estimation techniques are assessed. [10].

9.2 Total Capital Investment:

Working and fixed capital investments make up the total capital investment. The whole cost of the plant in its startup state is known as fixed capital. The price that the contractors were charged.

It Covers the Price of:

- Construct and design other engineering and construction supervision.
- All products of equipment and their installation.

Working capital is the additional investment required, above and beyond fixed capital, to start and run the plant until revenue is generated. It consists of the price of:

- Raw materials and process intermediates.
- Inventory count for Finished goods.

Cost of Conveyor:

Screw (stainless steel):

$C = 0.85\ L0.78$ *(9.1)*

$C = \$\ 38$

Cost of Crusher:

Hammer Mill:

$C = 2.97\ W\ 0.78$

$C = \$\ 1124$

Cost of Mixers:

$C = FMCb + Ca$ (Vessel)

$C = 1.218 \exp[a+b \ln HP+c(\ln HP)2]$ *(9.2)*

$C = \$\ 46669$

Cost of Storage Tank:

$C = 1.218Fm \exp[2.631+1.3673(\ln V)-0.06309(\ln V)2]$ *(9.3)*

C = \$ 56671

Cost of Reactors:

$C = F_M * Cb + C_a$ (Vessel)

C = 1.218 exp[a+b ln HP+c(ln HP)2] *(9.4)*

C = \$ 46669

Cost of Flash Tank:

C = 1.218Fm exp[2.631+1.3673(ln V) - 0.06309(ln V)2] *(9.5)*

C = \$ 195820

Cost of Pumps:

$C = F_M F_T C_b$ *(9.6)*

C = \$ 599148

Cost of Heat Exchangers:

C = 1.218×(fd × fm× fp× Cb) *(9.7)*

C = 1.218 ×(0.6002) ×(1.9) ×(1.00) ×(18468)

C = \$ 25651

Cost of Distillation Column:

$Ct = 1.218[f_1 C_b + N f_2 f_3 f_4 C_t + C_{pt}] = Ct = \$ 259662$

9.2.1 Purchased Equipment Cost:

Table 9.1: Purchased Equipment Cost for Plant

Purchased Equipment Cost	
Reactors	
No. Required	06
Purchased Cost	\$ 4359040
Heat Exchangers	
No. Required	06
Purchased Cost	\$ 1480130
Distillation Column	
No. Required	02
Purchased Cost	\$ 2819862
Mixers	
No. Required	02
Purchased Cost	\$ 93338
Flash Vessel	

No. Required	02
Purchased Cost	$ 391640
Pumps	
No. Required	06
Purchased Cost	$ 599148
Crushers	
No. Required	03
Purchased Cost	$ 1124
Conveyors	
No. Required	03
Purchased Cost	$ 114
Mechanical Separator	
No. Required	01
Purchased Cost	$ 20765
Total Purchased Equipment Cost	**$ 10069592**

9.2.2 Direct Cost:

Table 9.2: Direct Cost of Plant

Direct Cost		
Item	**% (Purchased Equipment)**	**Cost $**
Purchased equipment	100%	10069592
Installation	47%	4732708
Instrument and Control	18%	1812527
Piping	66%	6645931
Electricity	11%	1107655
Building	18%	1812527
land	6%	604175.5
Service facility	70%	7048714
Yard Improvement	10%	1006959
Total		**34840788**

9.2.3 Indirect Cost:

Table 9.3: Indirect Cost of Plant

Indirect Cost		
Items	**% (Direct Cost)**	**Cost $**
Engg. & Supervision	33%	3322965.36
Contractor fee	19%	1913222.48
Construction Expenses	41%	4128532.72
Contingences	37%	3725749.04
Total		**13090469.6**

Fixed capital = direct cost + indirect Cost *(9.8)*

Fixed capital = $47931258

Working capital investment = 15% of fixed capital investment

Working capital investment = $7189688

Total capital investment = Fixed capital investment + Working capital investment

Total capital investment = $ 55120946

9.3 Operating Cost:

For the purpose of evaluating a project's viability and selecting amongst potential alternative processing techniques, an estimate of the operational expenses, or the cost of producing the product, is mandatory.

The items on the following list will be included in the price of making a chemical product. There are two groups formed from them.

1. Variable operating costs: expenses that change according to the volume of output..
2. Fixed operating costs: expenses that don't change based on the rate of production. No matter how much is generated, these bills must be paid.

9.4 Total Production Cost:

9.4.1 Variable Cost:

Raw Material Cost

Flow rate of corn stover = 562120 kg/hr

For 330 days = 4451990400 kg/year

Price of corn stover per kg = $0.0585/kg

Total price of corn stover = $260441438/year

Catalyst Cost

Price of catalyst (Al2O3) = $1.2/kg

Weight of catalyst = 1760 kg

Price of catalyst = $2112/year

Miscellaneous Material

Maintenance cost = 7% of FCI Maintenance cost = $ 335518

Miscellaneous Material = 10% of maintenance cost

Miscellaneous Material = $ 33551

Steam Cost

Price of steam in 2022 = $0.014/kg

Total steam required = 432000 kg/hr

For 330 days = 3421440000 kg/year

Total price of steam per year = $ 47900160/year

Cooling water

Cooling Water price = $0.00001/kg

Cooling water required = 51780960000 kg/year

Total price of cooling water = $517809/year

Variable cost = raw material cost + miscellaneous cost + utilities cost

Variable cost = $ 305502483/year

9.4.2 Fixed Cost:

Table 9.4: Fixed Operating Cost for Process Plant

Fixed Operating Cost		
Item	**% (FCI)**	**Cost $**
Maintenance	7%	3355188
Operating cost of labor	10%	4793126
Laboratory Cost	20%	9586252
Supervision Cost	15%	7189689
Plant Overheads	50%	23965629
Capital Charges	10%	4793126
Insurance	1%	47931.26
Local Taxes	2%	95862.52
Royalties	1%	47931.26
Total		**34840788**

9.4.3 Direct Production Cost:

Direct Production Cost = variable cost + fixed cost *(9.9)*

Direct Production Cost = \$ 305502483 + \$ 34840788

Direct Production Cost = \$ 340343271

9.4.4 Overhead Charges:

30% of direct production cost

Overhead charges = (0.3)(340343271) = \$ 102102981

Total Production Cost = Direct Production Cost + Overhead Charges

Total Production Cost = \$442446252/year

Total Production Rate = 990000000kg/year

Production Cost (\$/kg) = Total Production Cost / Total Production Rate

Production Cost (\$/kg) = \$ 0.47/kg

9.5 Profitability Analysis:

Total Income

Selling Price = \$600/ton

Total Production per year = 990000 ton/year

Total Income = \$594000000/year

Gross Profit

Gross Profit = Total Income - Total Production Cost *(9.10)*

= \$ 594000000 /year – \$442446252/year

= \$151553748/year

Depreciation:

Machinery and equipment = 20% of FCI

= \$ 9586251

Building = 4% of Building cost

= \$ 1812527

Total Depreciation = Machinery and equipment + Building *(9.11)*

= \$ 11398778

Profit before Taxation:

Net Profit before Taxation = Gross profit – Depreciation

= \$ 151553748 /year - \$ 11398778/year

= \$ 140154970/year

Net Profit after Taxation:

Net Profit after Taxation = (1-0.40)* Profit before taxation *(9.12)*

= $ 15335449/year

Rate of Return

$$= \frac{\text{Net Profit}}{\text{Total Capital Investment}} \times 100 \qquad (9.13)$$

$$= \quad 27.8\ \%$$

Pay Back Period

$$= \frac{1}{\text{Rate of Return}} \qquad (9.14)$$

$$= \quad 3.5 \text{ years}$$

9.6 Feasibility Analysis:

9.6.1 Discounted Cash Flow:

Total Capital Cost = C_{FC} + C_L +C_{WC} *(9.15)*

C_{FC} = Fixed Capital

C_L = Land Cost

C_{WC} = Working Capital

Annual Expense = Cost of manufacturing

COM = 0.304FCI+2.73COL+1.23(C_{UT}+C_{RM})

COL = Cost of Labor

C_{UT} = Utilities Cost

C_{RM} = Raw Material Cost

9.6.2 Net Present Worth:

The future profitability of an investment, project, or business is assessed using net present value. The NPV of an investment is essentially the total discounted to present value of all future cash flows during the investment's lifetime.

9.6.3 Cash Flow:

Cash flows include all money produced or spent for the benefit of the investment, such as interest and loan repayments as well as capital outlays. The cash flow for each period comprises both inflows for profits, revenues, and dividends as well as outflows for costs.

9.6.4 Internal Rate of Return:

An internal rate of return computation is used in capital planning to determine whether projects or investments are worthy of funding and to rank them. The discount rate at which the net present value (NPV) is zero is known as the IRR. (when time-adjusted future cash flows equal the initial investment). An indicator of actual investment performance is the annual rate of return, or IRR.

Table 9.5: Discounted Cash Flow Analysis for Process Plant

Years	Gross Income	Annual Expense	Cash Flow Before Depreciation	Depreciation	Taxable Income	Tax	Cash Flow after Tax	Cumulative Cash Flow
	$	$	$	$	$	$	$	$
-1			-47931258					-47931258
-1			-1917250					-49848508
-1			-7189689					-57038197
1	594000000	562529739	31470261	1195870	30274391	13623476	16650915	-40387282
2	594000000	562529739	31470261	2265859	29204402	13141981	16062421	-24324861
3	594000000	562529739	31470261	2108507	29361754	13212789	16148964	-8175896
4	594000000	562529739	31470261	1951156	29519105	13283597	16235508	8059611.5
5	594000000	562529739	31470261	1793805	29676456	13354405	16322051	24381662
6	594000000	562529739	31470261	1667924	29802337	13411052	16391285	40772948
7	594000000	562529739	31470261	1416162	30054099	13524345	16529755	57302703
8	594000000	562529739	31470261	1416162	30054099	13524345	16529755	73832457
9	594000000	562529739	31470261	1416162	30054099	13524345	16529755	90362212
10	594000000	562529739	31470261	1416162	30054099	13524345	16529755	106891966
11	594000000	562529739	31470261	1416162	30054099	13524345	16529755	123421721
12	594000000	562529739	31470261	1416162	30054099	13524345	16529755	139951476
13	594000000	562529739	31470261	1416162	30054099	13524345	16529755	156481230
14	594000000	562529739	31470261	1416162	30054099	13524345	16529755	173010985
15	594000000	562529739	31470261	1416162	30054099	13524345	16529755	189540740
16	594000000	562529739	31470261	1416162	30054099	13524345	16529755	206070494
17	594000000	562529739	31470261	1416162	30054099	13524345	16529755	222600249
18	594000000	562529739	31470261	1416162	30054099	13524345	16529755	239130004
19	594000000	562529739	31470261	1416162	30054099	13524345	16529755	255659758
20	594000000	562529739	31470261	1416162	30054099	13524345	16529755	272189513

9.7 Cumulative Cash Flow Diagram:

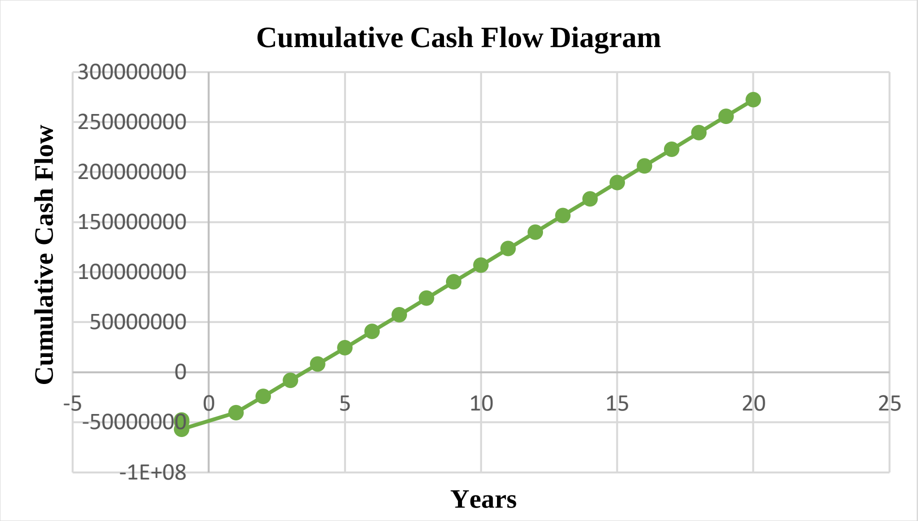


Figure 9.1: Cumulative Cash Flow Diagram for Breakeven

Minimum Acceptable Rate of Return (MAAR) = 15%

Internal Rate of Return = i %

$$NPW = \frac{net\ profit(1+\mathrm{i})^{\mathrm{n}}}{i(1+\mathrm{i})^{\mathrm{n}}} - TCI = 0 \qquad (9.16)$$

By hit and trial method

IRR = 31%

NPW = $11984107

IRR > MAAR

31% > 15%

It proves that our investment is economically safe and feasible.

CHAPTER # 10

INSTRUMENTATION AND PROCESS CONTROL

10.1 Introduction:

Control in process industries refers to the regulation of all aspects of the process. Precise control of level, temperature, pressure and flow is important in many process applications. This module introduces you to control in process industries, explains why control is important, and identifies different ways in which precise control is ensured. The objective of an automatic process control is to use the manipulated variable to maintain the controlled variable at its set point in spite of disturbances. Instruments are provided to monitor the key process variables during plant operations. Instruments monitoring critical process variables will be fitted with automatic alarms to alert, the operations to critical and hazardous situations. Pneumatic instruments are used in this plant. The main process parameters are all indicated in the control room where automatic or remote control is carried out centrally. The process parameters e.g. temperatures, pressure flow, liquid level etc. are converted to signals with transducers and then indicated, recorded and controlled with secondary instruments.

10.2 Importance of Process Control:

Refining, combining, handling, and otherwise manipulating fluids to profitably produce end products can be a precise, demanding, and potentially hazardous process. Small changes in a process can have a large impact on the result. Variations in proportions, temperature, flow, turbulence, and many other factors must be carefully and consistently controlled to produce the desired product with a minimum of raw materials and energy. Process control technology is the tool that enables manufacturers to keep their operations running within specified limits and to set more precise limits to maximize profitability, ensure quality and safety.

10.3 Process:

Process as used in the terms process control and process industry, refers to the methods of changing or refining raw materials to create end products. The raw materials, which either pass through or remain in a liquid, gaseous, or slurry (a mix of solids and liquids) state during the process, are transferred, measured, mixed, heated or cooled, filtered, stored, or handled in some other way to produce the product. Process industries include the chemical industry, the oil and gas industry, the food and beverage industry, the pharmaceutical industry, the water treatment industry, and the power industry. Process Control: Process control refers to the methods that are used to control process variables when manufacturing a product. For example, factors such as the proportion of one ingredient to another, the temperature of the materials, how well the ingredients are mixed, and the pressure under which the materials are held can significantly impact the quality of an end product. Manufacturers control the production process for three reasons:

- ✓ Reduce variability.
- ✓ Increase efficiency.
- ✓ Ensure safety.

10.3.1 Reduce Variability:

Process control can reduce variability in the end product, which ensures a consistently high quality product. Manufacturers can also save money by reducing variability. For example, in a gasoline blending process, as many as 12 or more different components may be blended to make a specific grade of gasoline. If the refinery does not have precise control over the flow of the separate components, the gasoline may get too much of the high-octane components. As a result, customers would receive a higher grade and more expensive gasoline than they paid for, and the refinery would lose money. The opposite situation would be customers receiving a lower grade at a higher price.

10.3.2 Increase Efficiency:

If manufacturers do not retain exact control over all of the processing factors, a run-away process, such as an out-of-control nuclear or chemical reaction, may result occurs. A process that runs amok might have disastrous results. To maintain safety, precise process control might also be necessary. Plant safety operations:

- ✓ To maintain process variables within predetermined safe operating ranges.
- ✓ To recognize potentially dangerous circumstances as they arise and to set up alarms and automatic shut-down mechanisms.
- ✓ To offer alerts and interlocks to prevent risky operating methods

10.4 Process Control Terms:

10.4.1 Process Variable:

A process variable is a property of the fluid that are in processing that has the potential to alter the manufacturing process. The process variable in the illustration of you relaxing by the fire was temperature. These are typical given below,

- ✓ Pressure
- ✓ Flow
- ✓ Level
- ✓ Temperature

10.4.2 Set Point:

The set point is a value for a process variable that is desired to be maintained. For example, if a process temperature needs to keep within 5 °C of 100 °C, then the set point is 100 °C. A temperature sensor can be used to help maintain the temperature at set point. The sensor is inserted into the process, and a controller compares the temperature reading from the sensor to the set point. The burner's fuel valve is instructed to close slightly until the process cools to 100 °C if the temperature reading is 110 °C, which indicates that the process is above set point. Additionally, set points may be maximum or minimum values.

10.4.3 Measured Variables:

The process fluid's state, which may be maintained at the predetermined set point, is the measured variable.

10.4.4 Manipulated Variable:

The varying variable that can be used to maintain the control variable at the correct value

10.5 Hardware Elements of Control System:

10.5.1 The Measuring Instruments or Sensors:

These are the tools that are used to measure controlled variables and disturbances.

10.5.2 Transducers:

An apparatus that changes one form of energy into another is called a transducer. A transducer mainly transforms a signal from one form of energy to another.

10.5.3 Transmission Line:

The measurement signal is transferred from the measuring equipment to the controller using it.

10.5.4 Controller:

This gets the data from the measuring tools that determine if the information is accurate or not.

10.5.5 The Final Controller Element:

The Final control element is a device controlled by a controller to change the operating conditions of a process. Final control elements require energy to operate against the process. It is the hardware element that implements the decision taken by controller.

10.6 Classification of Control Systems:

The following control loops are the most frequently utilized for instrumenting and controlling various plant areas and equipment.

- ✓ Feed-back control loop
- ✓ Feed forward control loop
- ✓ Ratio control loop

10.6.1 Backward Feed Control Loop:

A method of control in which a measured value of a process variable is cross-pond with the desired value of the process variable and any important action is taken. Feedback control is considered as the basic control loops system. Its disadvantage lies in its operational procedure. For example if a certain quantity is entering in a process, then a monitor will be there at the process to note its value. Any changes from the set point will be sent to the final control element through the controller so that to adjust the incoming quantity according to desired value (set point). But in fact changes have already occurred and only corrective action can be taken while using feedback control system[1].

10.6.2 Forward Feed Control Loop:

A method of control in which the value of disturbance is measured than action is taken to prevent the disturbance by changing the value of a process variable. This is a control method designed to prevent errors from occurring in a process variable. This control system is better than feedback control because it anticipates the change in the process variable before it enters the process and takes the preventive action. While in feedback control system action is taken after the change has occurred.

10.6.3 Ratio Control:

A control loop in which, the controlling element maintains a predetermined ratio of one variable to another. Usually this control loop is attached to such a system where two different systems enter a vessel for reaction that may be of any kind. To maintain the stoichiometric quantities of different streams, this loop is used so that to ensure proper process going on in the process vessel.

10.6.4 Split Range Loop:

In this loop controller is preset with different values corresponding to different actions to be taken at different conditions. The advantage of this loop is to maintain the proper conditions and avoid abnormalities at very differential levels.

10.6.5 Cascade Control Loop:

This is a control in which two or more control loops are arranged so that the output of one controlling element adjusts the set point of another controlling element. This control loop is used where proper and quick control is difficult by simple feed forward or feed backward control. Normally first loop is a feedback control loop.

10.6.6 Control Schemes of Distillation Column:

In distillation column control any of following may be the goals to achieve.

- ✓ Composition of bottom
- ✓ Constant bottom product rate.

The objectives in distillation column control could be any of the following:

- ✓ Steam flow rate to Re-boiler.
- ✓ Rate of Reflux.

10.7 Control Scheme:

Since the bottom product rate might change, any variation in rate of entering the material also absorbed by the fixed overhead by rate of utput. Since the vapor rate remains almost constant while the feed rate increases, the purity of the top product also rises. The dynamics of the system that adjusts it for level control determine how the overhead reflux changes.

10.8 Instrumentation and Control for Distillation Column:

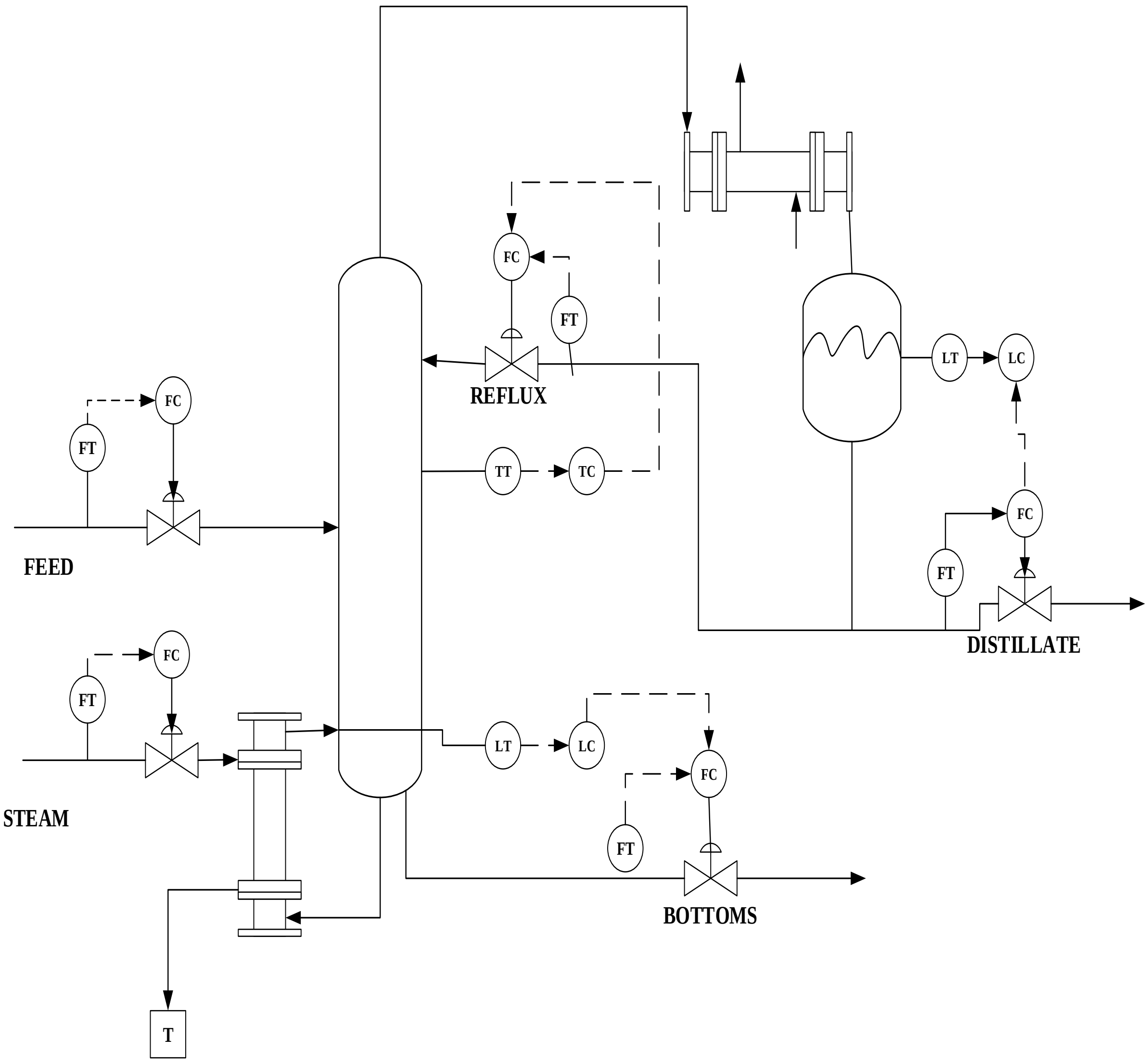


Figure 10.1: Instrumentation and Control for Distillation Column

10.8.1 Description:

The LC is the level controller. The sensing element (level measuring device) measure the level and sends signal to the controller.

Which has a preset value feed to it. It compares the set point and the measured value and send signal to the valve. Which in turn act on the instruction provided by the controller.

The temperature sensor senses the temperature of the feed and immediately send signal to the controller TC-2 which in turn sends signal to the flow controller.

FC-4 that will immediately control the reflux rate.-

10.8.2 Action:

If the rate is less than the set point, increase the feed flow rate; if it is greater than the set point, decrease the feed flow rate.

Table 10.1: Elements of Control Loop for Feed of Distillation Column

Process	Flow rate of feed stream
Controller	Automatic(PIC)
Controlled variable	Flow rate
Measuring element	Orifice
Manipulated element	Valve
Regulating element	Valve
Load variable	Leakage and value characteristics friction in pipe

10.8.3 Operation:

If the feed flow increases the (FC) will send signal to the controller. The controller also send signal to the valve which will drive the control valve to close. If the feed flow decreases the (FC) will send signal by (FT) to the controller. The controller also send signal to the valve which will drive the control valve to open.

10.8.4 Top Pressure Control:

The efficiency of separation will be impacted by any change in operating pressure, which is why distillation column pressure needs to be controlled.

Set point: 1 bar

Table 10.2: Elements of Control Loop for Distillation Column Top Section Pressure

Process	Distillation column
Controller	Automatic(P)
Controlled variable	Pressure
Measuring element	Manometer
Regulating element	Valve(pneumatic)
Manipulated element	Vapor Flow rate
Load variable	Feed flow rate of fed temperature, change in the ratio of gas to liquid, valve characteristic

10.8.5 Operation:

The transmitter sends the measured value to the controller, which compares it to the set point.

If the pressure is higher than 1 bar, the flow rate of the cooling water in the condenser must be increased; if it is lower than 1 bar, the flow rate of the cooling water must be decreased..

10.8.6 Drum level control:

Using a valve, the distillate flow rate can be changed to adjust the drum level in the top part. 85% of the holdup volume is the set point.

10.8.7 Operation:

The distillate flow rate lowers if the drum level exceeds the predetermined point. If it falls below the specified point, the distillate flow is increased.

Table 10.3: Control Loop Elements for Drum Level Control for Distillation Column

Process	Distillation
Controller	Automatic PID
Controlled variable	Drum level
Measuring element	Orifice
Regulating element	Valve

10.9 Instrumentation and Control on Reactor:

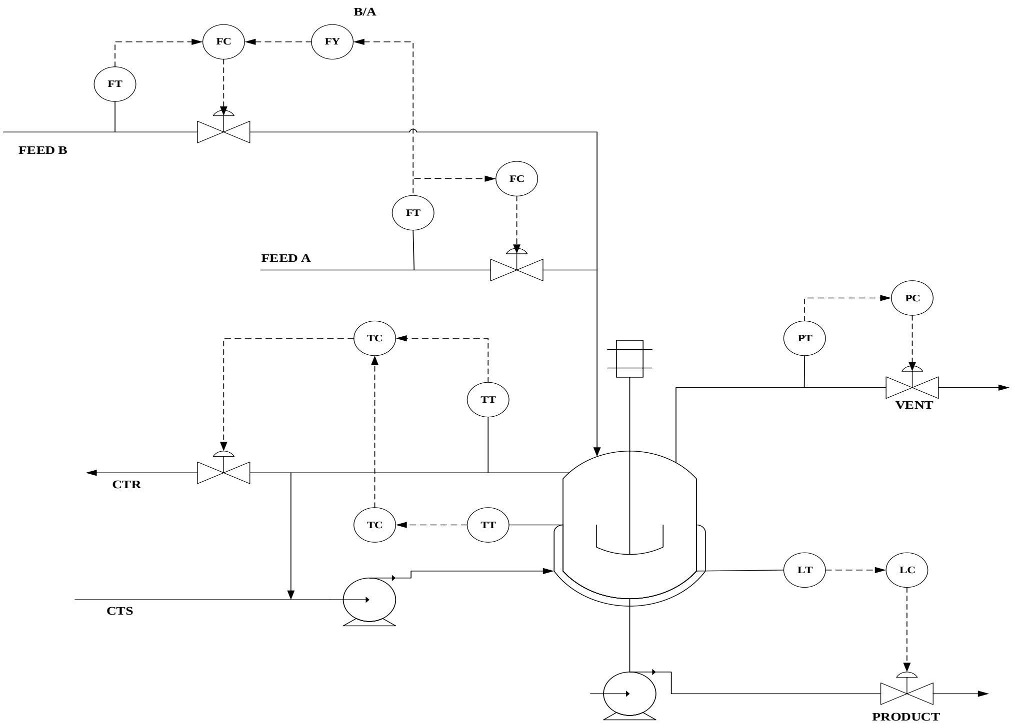


Figure 10.2: Instrumentation and Control on Reactor

10.9.1 Temperature Control:

The reason of temperature control in reactor is to avoid the runaway reaction. This can be done by controlling the water flow rate of cooling water. Set Point =190℃

10.9.2 Operation:

The temperature sensor measures the temperature inside the reactor and send an electrical signal analog to the value of the temperature. The controller receives this signal and compare it, then gives corrective action. (If temperature is increased than 190℃ the controller increases the flow rate of the water and vice versa). The transducer converts the value to pneumatic signal to the regulating element. (Water valve)

10.9.3 Pressure Control:

The operating pressure of the vinyl acetate reactor is 8 bar. Controlling this pressure is highly significant process to maintain the plant safety. We can control the pressure inside the reactor by controlling the pressure of the inlet feed.

10.9.4 Operation:

The pressure sensor measures the pressure inside the reactor, send an electrical signal analog to the value of the temperature. The controller receives this signal and compares it, then gives corrective action. The transducer converts value of the pneumatic signal to the electrical signal.

CHAPTER # 11

SAFETY AND HAZOP STUDY

11.1 Introduction:

Chemical industry procedures that involve industrial biology are frequently used in the production of food, pharmaceuticals, and energy. Interest in occupational health and safety issues is growing because of the expansion of bioprocessing facilities. However, there are still gaps in the assessment of potential difficulties with occupational health and safety. While being handled industrially, the biomass may expose workers to biological and chemical agents that could be released into the work environment. Particular risks associated with the production of biofuels (bioethanol, biogas, bio ethylene, etc.) include the development of potentially explosive atmospheres and pool fires. These unintentional discharges may also result from the breakdown of equipment in processing plants, such as tanks, flanges, valves, pumps, and compressors.

11.2 Hazard and Operability Study:

A HAZOP analysis is a meticulous method used to determine how a process could stray from its intended course. This approach can differentiate between hazard (any operation that could result in a catastrophic leak or harm to personnel) and operability (any project-related procedure that could result in a plant closure with potential effects on safety or earnings). It is described as the process of applying a formal, methodical, and critical assessment of the procedure and the engineering goals of new or existing facilities to analyze the outcomes of potential operating failures of specific equipment parts and the ensuing repercussions on the facility.

Since it has shown to be effective in identifying environmental, safety, and health concerns, the HAZOP safety-analysis technique is used globally and is recognized by law. The HAZOP system was created in the 1970s when chemical factories grew and a preventive strategy became necessary. This technology was later used in continuous chemical operations, the nuclear, pharmaceutical, and transportation industries.

Numerous attempts have been made to automate the analysis, and HAZOP has been frequently used in conjunction with other analysis methodologies.

All installation types and stages are covered, but new installations in particular benefit because they lack operational experience. This examination systematically identifies the dangers and weaknesses.

11.3 When is HAZOP Carried Out?

- ✓ Checking that operating and emergency measures are sufficient before starting up.
- ✓ Analyzing the safety implications of maintenance procedures or any installation modifications while the installation is in use.
- ✓ Ensuring that safety objectives are properly satisfied during the design phase.

11.4 Why is HAZOP Carried Out?

- ✓ To confirm the project's safety
- ✓ Examine operating and safety procedures
- ✓ Boost that our system is safely working

11.5 Benefits of HAZOP Study:

- ✓ Creating operating guidelines.
- ✓ Verification of design parameters, processes, and prospective changes.
- ✓ Knowledge that can be used to assess and manage the risk posed by the discovered unintentional incidents.

11.6 Success or Failure of the HAZOP:

The following elements affect whether the HAZOP is successful or unsuccessful:

- ✓ Team's technological expertise and perspectives.
- ✓ Team's capacity to visualize variances, causes, and effects by using the approach as a creative tool.
- ✓ The team's capacity to concentrate on the identified more critical dangers.

11.7 Systematic Terms:

11.7.1 Study Nodes:

The sites where deviations from the process parameters are looked for (on pipe and instrumentation drawings and procedures).

11.7.2 Intention:

The intention defines how the plant is expected to operate in the absence of deviations at the study nodes. This can take a number of forms and can either be descriptive or diagrammatic; e.g., flow sheets, line diagrams, P&IDs.

11.7.3 Deviations:

These are changes from the purpose that are found by repeatedly using the helping phrases (for example, apply additional pressure).

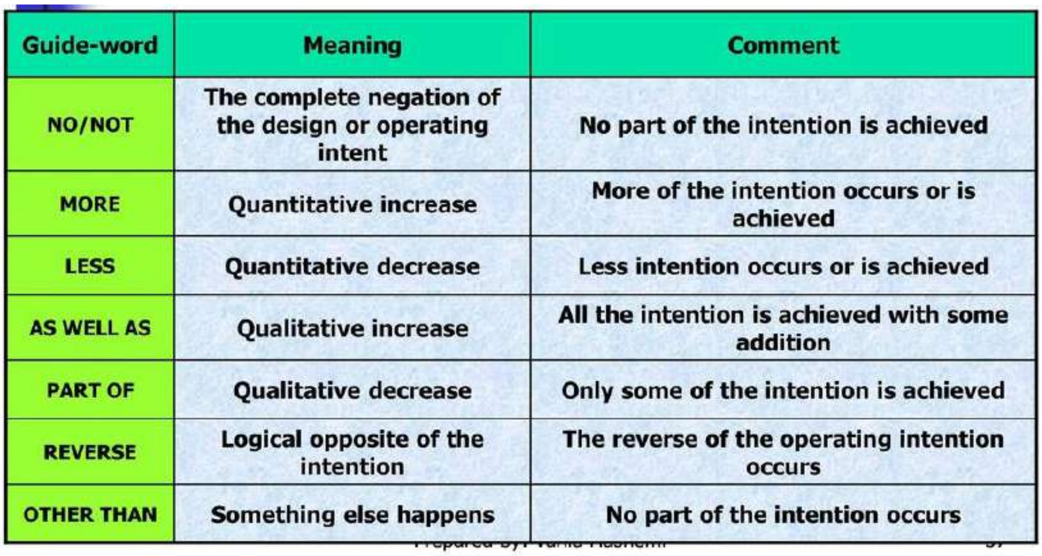
11.7.4 Foundation:

These are the reasons why deviations might occur. Once a deviation has been shown to have a credible cause, it can be treated as a meaningful deviation. This cause can be hardware failures, human errors, an unanticipated process state (e.g., change of composition), external disruptions (e.g., loss of power), etc.

11.7.5 Consequences:

These are the results of the deviations should they occur. Trivial consequences, relative to the study objective, are dropped.

11.7.6 Guide Words:



Guide-word	Meaning	Comment
NO/NOT	The complete negation of the design or operating intent	No part of the intention is achieved
MORE	Quantitative increase	More of the intention occurs or is achieved
LESS	Quantitative decrease	Less intention occurs or is achieved
AS WELL AS	Qualitative increase	All the intention is achieved with some addition
PART OF	Qualitative decrease	Only some of the intention is achieved
REVERSE	Logical opposite of the intention	The reverse of the operating intention occurs
OTHER THAN	Something else happens	No part of the intention occurs

Figure 11.1: Guide Words for HAZOP [10]

These are straightforward terms that are used to qualify or quantify the aim in purpose to direct and inspire brainstorming so that change can be found. At the research node (place in the plant) under investigation, each guidance word is necessary to the process system

11.8 Steps for HAZOP Study:

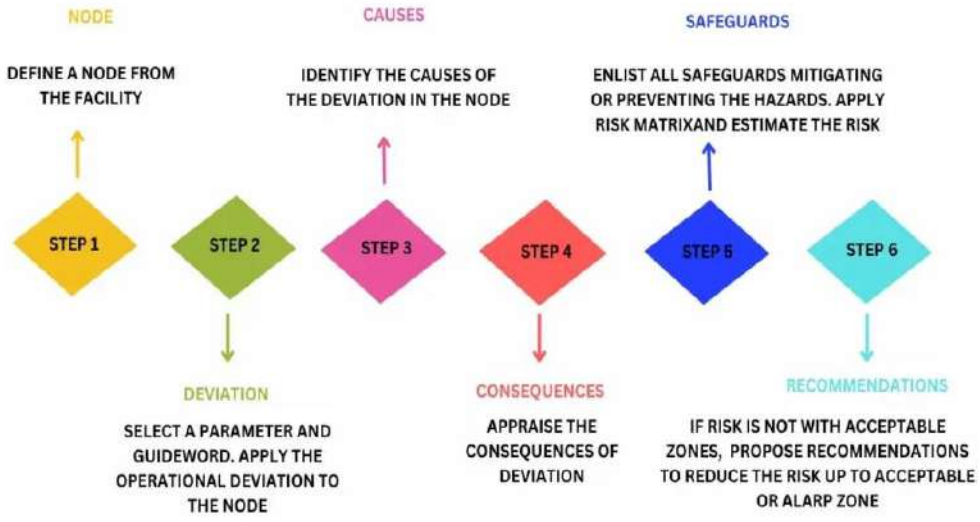


Figure 11.2: Steps for HAZOP Study [3]

11.8.1 Scope of Work:

Specify the purpose, objective, and scope of the study. The purpose may the analysis of a yet to be built plant or a review of the risk of un-existing unit. Given the purpose and the circumstances of the study, the objectives listed above can he made more specific. The scope of the study is the boundaries of the physical unit, and also the range of events and variables considered. For example, at one time HAZOP's were mainly focused on fire and explosion endpoints, while now the scope usually includes toxic release, offensive odor, and environmental end-points. The initial establishment of purpose, objectives, and scope is very important and should be precisely set down so that it will be clear, now and in the future, what was and was not included in the study. These decisions need to be made by an appropriate level of responsible management.

11.8.2 HAZOP Team:

Select the HAZOP study team. The team leader should be skilled in HAZOP and in interpersonal techniques to facilitate successful group interaction. As many other experts should be included in the team to cover all aspects of design, operation, process chemistry, and safety. The team leader should instruct the team in the HAZOP procedure and should emphasize that the end objective of a HAZOP survey is hazard identification; solutions to problems are a separate effort.

11.8.3 Data Collection:

- ✓ Process flow sheets.
- ✓ Process layout
- ✓ Control Diagram

11.8.4 Operating Procedures:

- ✓ Maintenance procedures
- ✓ Emergency response procedures
- ✓ Safety and training manuals

11.8.5 Conduct the Study:

Conduct the study. Using the information collected, the unit is divided into study "nodes" and the sequence diagrammed in Figure, is followed for each node. Nodes are points in the process where process parameters (pressure, temperature, composition, etc.) Have known and intended values. These values change between nodes as a result of the operation of various pieces of equipment' such as distillation columns, heat exchanges, or pumps. Various forms and work sheets have been developed to help organize the node process parameters and control logic information.

When the nodes are identified and the parameters are identified, each node is studied by applying the specialized guide words to each parameter. These guide words and their meanings are key elements of the HAZOP procedure.

Repeated cycling through this process, which considers how and why each parameter might vary from the intended and the consequence, is the substance of the HAZOP study.

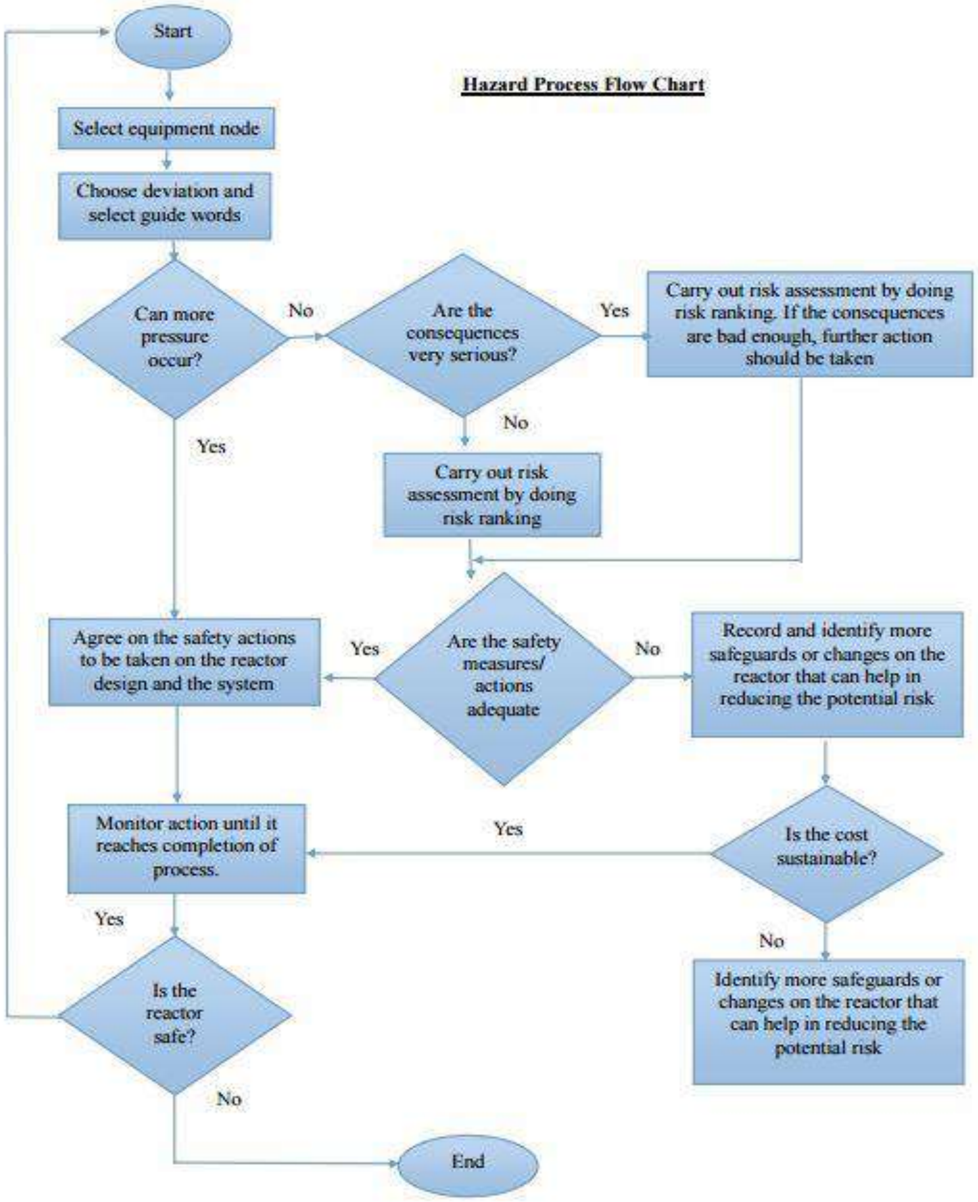


Figure 11.3: HAZOP Process Flow Chart [11]

11.9 Procedure:

HAZOP studies are normally carried out by a team of experienced people, who have complementary skills and knowledge, led by team leader who is experienced in the technique. The team examines the process vessel by vessel, line by line using the guide words to detect any hazard. The information required for the study will depend on the extent of investigation. A preliminary study can be made from a description of the process and the process flow sheets. For detailed, final, the study of design, flow sheets, piping and instrumentation diagrams, equipment specifications and layout drawings would be needed. In general, the procedure steps involving HAZOP are:

1. Divide the system into sections (i.e., reactor, storage).
2. Choose a study node (i.e., line, vessel, pump, operating instructions).
3. Describe the design intent.

4. Selecting a process parameter.
5. Used a guide word.
6. Determine causes.

11.10 HAZOP on Reactor:

Exothermic is the reaction. For the method of removing reaction's excess energy, a cooling system is offered. The temperature of the reactor would rise if the cooling function were to fail. This would cause a faster reaction rate, which would release more energy. A run-away reaction that generates pressures higher than the reactor's bursting pressure could be the outcome. A valve measures the temperature within the reactor and uses that information to regulate the flow rate of the cooling water.

Table 11.1: HAZOP Study on Reactor

HAZOP on Reactor				
Guide Word	**Deviation**	**Causes**	**Consequences**	**Action**
No	No cooling	faulty cooling water valve	The temperature in the reactor rises	Set up a high temperature alarm.
Reverse	Reverse cooling flow	Backward flow is the outcome of a water supply failure.	less cooling, perhaps explosive reaction	Set up the check valve
More	More cooling flow	Control valve malfunction, and the operator ignores the alarm	excessive cooling	Installing an operator on the steps
As Well As	Reactor product in coils	increased reactor pressure	Unorthodox product	Examine maintenance schedules and processes.
Other Than	Another material besides cooling water	source of contaminated water	There could be a response and inadequate cooling.	If cooling is less, TAH will notice. Isolate the water source if it is found.

11.11 HAZOP on Heat Exchanger:

Table 11.2: HAZOP Study on Heat Exchanger

HAZOP on Heat Exchanger				
Guide Word	**Deviation**	**Causes**	**Consequences**	**Action**
Less	Less flow of cooling water	pipe obstruction	Process fluid's temperature remains constant.	temperature warning
More	More cooling water	cooling water valve malfunction	reduction in process fluid temperature	Alarm for low temperature
More Of	More Pressure on Process fluid line	process fluid valve failure	rupture of the tube	mounting a high pressure alarm
Contamination	Contamination of process fluid line	A tube leaks, and cooling water enters	Pollution of the process fluid	proper upkeep and attentive operation
Corrosion	Corrosion of tube	Water used for cooling is hard.	less cooling and tube cracking	Suitable upkeep and
None	No Cooling water flow	failing to open the water valve	Temperature must not be reduced correspondingly.	Install a temperature gauge prior to and following the fluid line.
Reverse	Flow is reverse	inlet valve failure	offset by output material	Set the check valve
More	More cooling water flow	the input cooling water valve failing to close	Low fluid temperature	Install a temperature gauge prior to and following the fluid line.
Less	Less cooling water	Leakage in pipe	Process fluid temperature may be low	flow meter is necessary
Contamination	Fluid contamination	cooling water Contamination	Outlet temperature too low	Appropriate upkeep

11.12 HAZOP on Distillation Column:

The following aims are recommended by studies on distillation columns:

1. Maintaining the overhead or bottom composition at a predetermined level through product quality management.
2. Material balance control is used to keep the top and bottom inventories, as well as the column holdup, within the permitted ranges.
3. The column must not overflow. Pressure needs to be high enough to sustain efficient column operation, and the reboiler's temperature difference needs to be below the critical threshold.

Table 11.3: HAZOP Study on Distillation Column

HAZOP on Distillation Column				
Guide Word	**Deviation**	**Possible Causes**	**Consequences**	**Action record**
NO	No flow	• Blockage in packing • Control valve shut • Failure of valve • Pump failure	• Column dry out possible dangerous concentration	• low level alarm being install • Schedule, maintenance, And check procedure, Make bypass
LESS	Less flow	• Pipe blockage • Failure of valve • Pump failure	• Column dry out • Change in product quality	• Install low level alarm • Emergency plant shut down
MORE	More flow	• Control valve is fully opened • Increasing in pumping capacity • failure of Control valve	• Flooding in column • Change the product quality • Decrease in temperature • Raise in bottom liquid level	• Install low level alarm • Schedule, maintenance, check procedure • Install control valve
HIGH	High pressure	• Valve pressure high • Pressure indicator controller fail	• Low efficiency of separation • Rupture of column other related equipment • Product loss	• Install high Pressure alarm • Install pressure relief valve

	High Temperature	• Instrumentation failure • More steam flow • Exchanger tube failure • Heating medium leak into the process	• Separation cannot be done • Change in product quality • Column flooding • Film boiling in column and reboiler	• Install pressure indicator • Instruct operator on procedure • Attention to heat input and out control
LOW	Low pressure	• Vapor line leakage	• Low efficiency of separation • Loss of product	• Install pressure indicator
	Low Temperature	• Instrumentation failure • less steam flow • loss of heating • less steam temperature and pressure	• pressure change • product loss • change in product quality ineffective separation process • phase effect	• install temperature indicator • install operator on procedure • Isolation is up-graded • attention to heat input and out control

CHAPTER # 12

ENVIRONMENTAL IMPACT ASSESSMENT

12.1 Introduction:

The project description, evaluation of the project's environmental and social implications, mitigation strategies, and associated management and monitoring plans are all included in the EIA Report. The report contain

- ✓ It requested information that is relevant and helpful for making judgments.
- ✓ It results in the precise prediction of the negative effects of the recommended actions and their mitigation using both common and unique solutions.
- ✓ The creation of a thorough EIA is fraught with difficulties. These include the use of out-of-date assessment models, inadequately detailed alternatives and mitigation actions, and incomplete identification of the essential consequences. The types of flaws that could be found in various samples of typical EIA reports are listed in the table below.

12.2 Safety and Environmental Factors:

Although light olefins are colorless gases with a mild odor that is nonirritating to the eyes or respiratory system, these are hydrocarbons and therefore flammable. All vessels must be designed for handling the liquids and gases during operation at the temperature and pressure that exists, and safety and relief (depressurizing) valves must be provided to relive excessive pressure. Releasing the hydrocarbons in the air in large amounts must be avoided because of health and fire hazards.

If hydrocarbons must be released in the air it is done under blanket of steam. To protect the plants and personnel in case of fire, a complete fire-fight system is provided with tanks grouped to minimize fire and provided with foam makers and deluge systems. An olefin plant produces liquid, gaseous, and solid wastes that must be disposed of in an environmentally safe manner. Liquid waste generated within the complex consists of wastewater streams of relatively low organic content, and process waste of high organic contents.

Wastewater from various units and operations are serrated according to the wastewater characteristics, such as type of contaminants, concentration, special treatment, or permanent requirements [8]. A segregated sewer allows for the most efficient treatment. Atmospheric emissions from the facility are either controlled or fugitive in nature. Controlled emissions are released from process venting, waste incineration, decoking operations, and heater firing. Fugitive emission may occur from product loading and storage and equipment and valve leaks. Solid wastes are treated in solid waste disposal area to reduce their volume and/or toxicity prior to final disposal in a secure landfill. Combustible wastes can be incinerated in a slagging rotary kiln to reduce volume and toxicity.

12.2.1 Safety:

When a plant is being designed, the concern for operating safety becomes apparent. Even after the factory is finished being built, the design and specifications are thoroughly examined. This is true of the overall design of the instrumentation and control systems. The caliber and competency of the operating and support staff has a significant impact on plant safety.

In the end, everything safety is summarized in the plant's operating manuals. These guides emphasize

- ✓ The operating mode at design and light loads;
- ✓ Emergency measures, for instance, in the event of a sudden loss of power from the outside, to allow for the swift and secure discharge and disposal of the plant's liquid and vapor stockpiles;
- ✓ The use of steam, foam, and water sprays to avoid ignite by dissipating combustible elements that escape from damaged lines and equipment above ground;

12.3 Environmental Impacts of Ethylene Plant:

Three major categories can be used to group the environmental damage caused by the operation of the hot section of the ethylene plant,

- Air pollution
- Water pollution
- Noise pollution

12.3.1 Air Pollution:

The major source of air pollution in on ethylene plant in the stack gases coming out of the cracking furnace, Gaseous pollutants of environmental concern include carbon monoxide and oxides of nitrogen, the high furnace operating temperature and a slight oxidizing atmosphere, preclude the formation of carbon monoxide in any significant quantity. Larger burner capacity will tend to increase the NO_X level of a particular furnace [11].

12.3.2 Water Pollution:

Oily fractions from leaks, spills, and tank draw-off are included in the liquid waste effluent from the hot part of the ethylene plant. The caustic scrubbing column's spent caustic solution is the primary cause of water contamination, and how it is disposed of can have a significant impact on the plant's profitability. The principal corrective actions for decreasing oily wastes are,

- ✓ The prevention of oil leaks through routine maintenance of the equipment and pipes..
- ✓ Elimination and separate treatment of emulsions where they already exist or prevention of their creation.

12.3.3 Environmental:

Stringent environmental laws requires that nitrogen oxides (NOx), and sulfur oxides emission from furnaces be manually reduced. In some areas of the world, regulations require NOx be reduced to 70 ppm or lower. on a wet basis. Conventional burners usually produce 100 to 120 ppm of NOx.

Since NOx production depends on the flame temperature and quantity of excess air, achieving required limits may not be possible through burner design alone.

Therefore, many new designs incorporate $DENO_X$ units that employ catalytic methods to reduce the NOx limit. Platinum containing monolithic catalysts are used. Each catalyst performs commonly for a specific temperature range, and most of them work properly on 400°.

12.4 Emissions:

12.4.1 GHG Emissions:

Cellulosic ethanol from corn Stover and forest wood residue can avoid a significant amount of GHG emissions. Cellulosic ethanol produced from corn Stover could avoid GHG emissions by 86–89% and ethanol produced from forest residues in 2030 could reduce GHG emissions by 85%. Reductions of GHG in corn ethanol cases are moderate (21–24%). Accounting for lime applications and farming machinery in the assessment of the life cycle of corn ethanol causes an increase in GHG emissions of about 4% and 1%, respectively. Lime application increases CO_2 and therefore GHGs emissions because of its $CaCO_3$ chemistry. With one million Btu of fuel, gasoline emits 98 kg of GHGs from wells to wheels, while corn Stover-derived ethanol emits only 14 kg of GHGs. GHG emissions here are CO_2-equivalent emissions of CO_2, N_2O, and CH_4, weighted with their global warming potentials. The trend for CO_2 is similar to that for GHGs. One million Btu of corn Stover-based ethanol to displace one million Btu of gasoline could avoid 85 kg of CO_2. The CO_2 data are presented here to allow comparison with the results from some other studies, which only estimate CO_2 emissions. Apparently, ignoring N_2O and CH_4 emissions gives fuel ethanol some unwarranted additional benefits [12].

12.4.2 Pollutant Emissions:

The results of criteria pollutant emissions, expressed in grams of emissions per mile driven in FFV fueled with E85 are separated into total and urban emissions. Total emissions are the sum of the urban and rural emissions. Urban emissions have long been an environmental and health concern because the potential of exposing the human population to emissions in that setting is high. In comparison with gasoline, ethanol can achieve net reductions in the urban criteria pollutant emissions of VOCs, CO, NOx, PM, and SOx. This phenomenon can be explained by the location of bio-ethanol plants. Corn and cellulosic ethanol plants are most likely to be built near farms to minimize feedstock transportation costs. Criteria pollutants emitted from the farming, feedstock transportation, and ethanol-production steps contribute to rural emissions only. In contrast, several petroleum refineries (up to 60%) are currently situated in or near urban areas, which results in a high urban share of emissions from petroleum refining. Most significant reductions occur with SOx, where 60% of current SOx emissions due to vehicles fueled with gasoline could be avoided. As a result, there is a shift in the emission of criteria pollutants from urban to rural areas with bio-based ethanol in the near term. While urban emissions of VOCs, CO, NOx, PM, and SOx decrease, total emissions (urban and rural) increase, which means there are increased emissions in the rural area. Because urban area emissions are more of an environmental and health concern, this shift provides at least a positive step toward the reduction of regulated pollutants. Cellulosic ethanol derived from corn Stover could achieve a net reduction of total VOCs, NOx, PM, and SOx emissions in 2030.In the near-term scenario (2021), corn Stover-derived ethanol emits slightly higher VOCs than does corn grain ethanol. This higher emission is caused by VOC emissions from the biomass boiler.

12.4.3 Carbon Dioxide:

The estimation of the precise amount of carbon dioxide created during the production of bioethanol is a difficult and imprecise procedure that heavily depends on the technology used to produce the ethanol and the assumptions used in the calculation.

Calculations must take into account:

- ✓ Costs associated with raising the feedstock
- ✓ The price of delivering the feedstock to the manufacturing facility
- ✓ Costs associated with converting the feedstock into bioethanol
- ✓ These consequences may or may not be taken into account in such a calculation.
- ✓ The price of delivering the feedstock to the manufacturing facility
- ✓ Costs associated with converting the feedstock into bioethanol
- ✓ The following impacts may or may not be taken into account in such a calculation:
- ✓ The price of changing how the land is used in the region where the fuel feedstock is produced.
- ✓ The price of moving bioethanol from the plant to where it will be used.
- ✓ The bioethanol's effectiveness in comparison to regular gasoline

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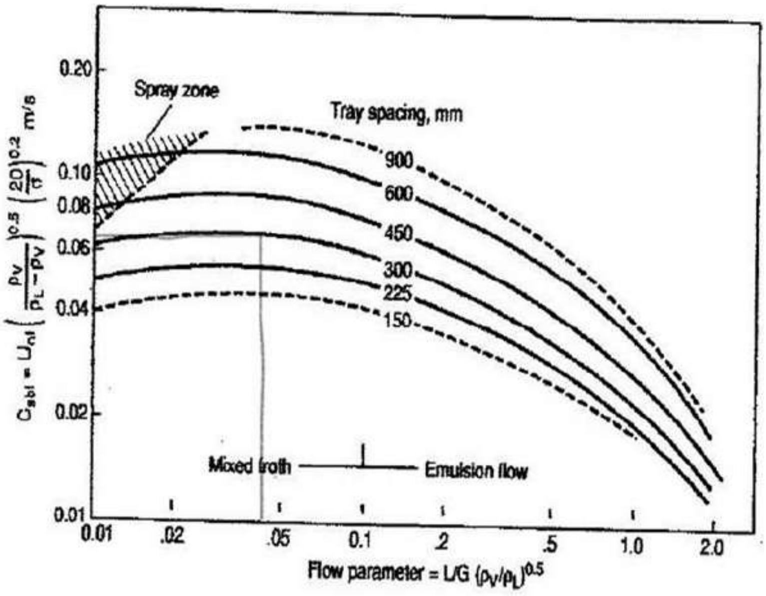
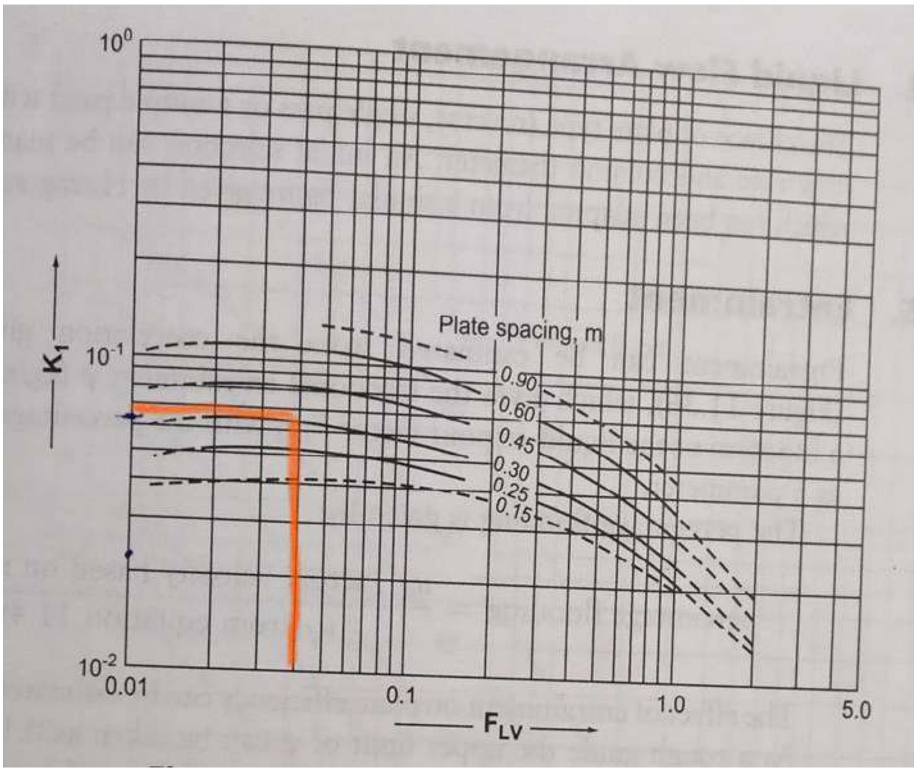
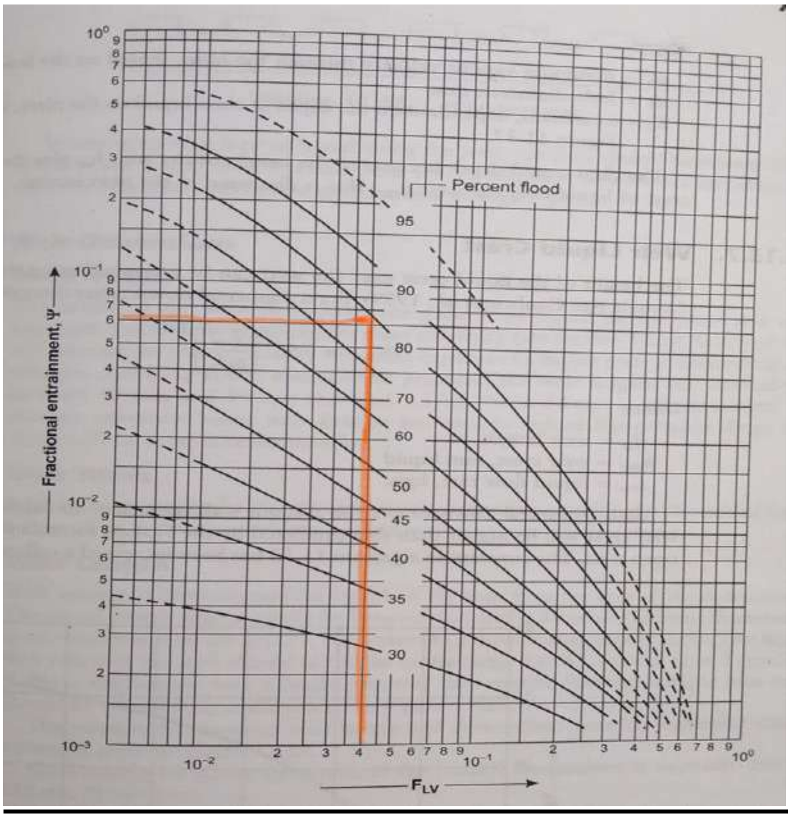
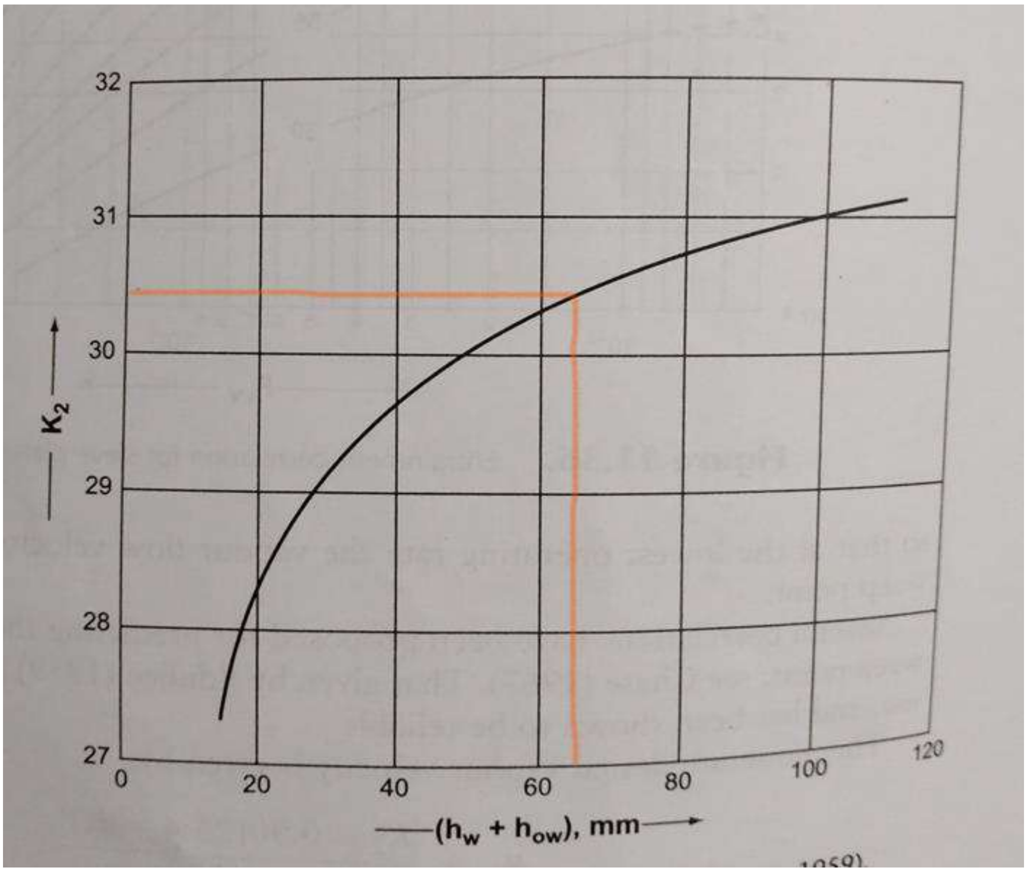
APPENDICES

14.1 Appendix-A:






Fractional entrainment, Ψ
F_{LV}
Percent flood
95
90
80
70
60
50
45
40
35
30
10^0
10^{-1}
10^{-2}
10^{-3}


K_2
$(h_w + h_{ow})$, mm
32
31
30
29
28
27
0
20
40
60
80
100
120

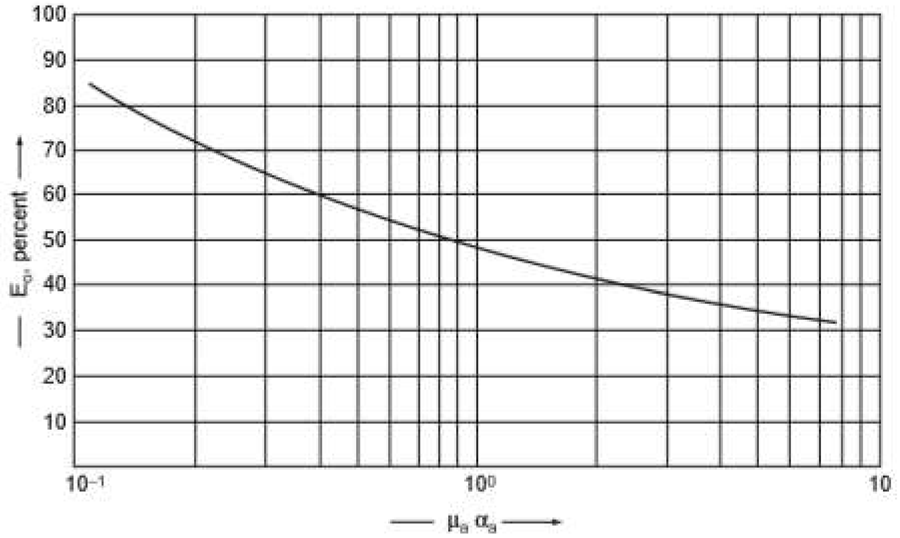


Figure 11.15. Distillation column efficiences (bubble-caps) (after O'Connell 1946).

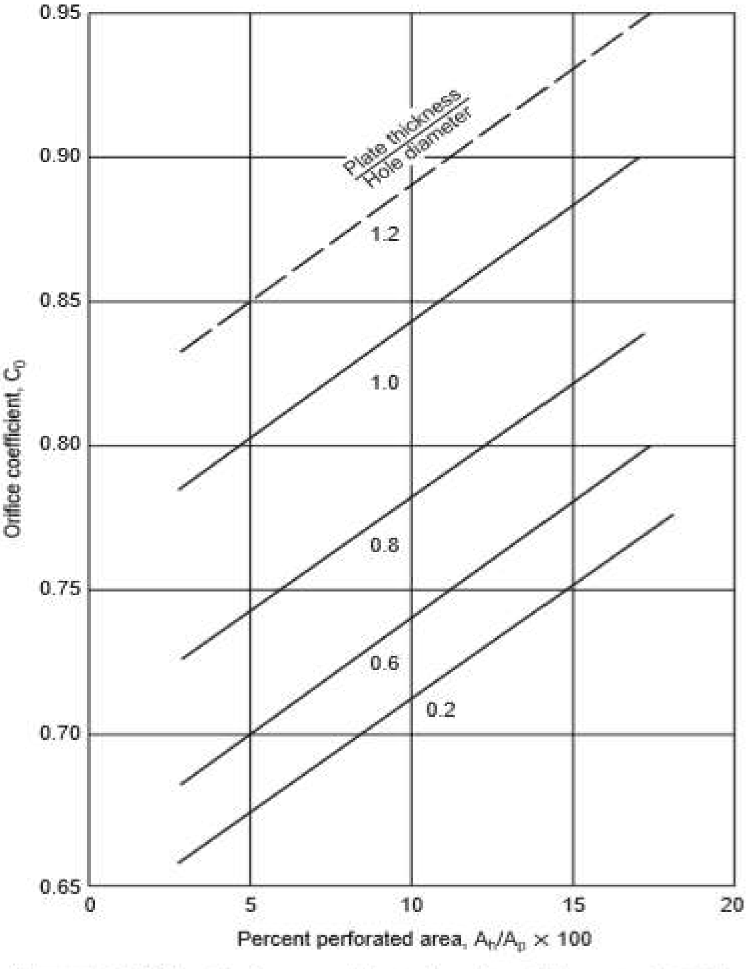


Figure 11.36. Discharge coefficient, sieve plates (Liebson et al., 1957).

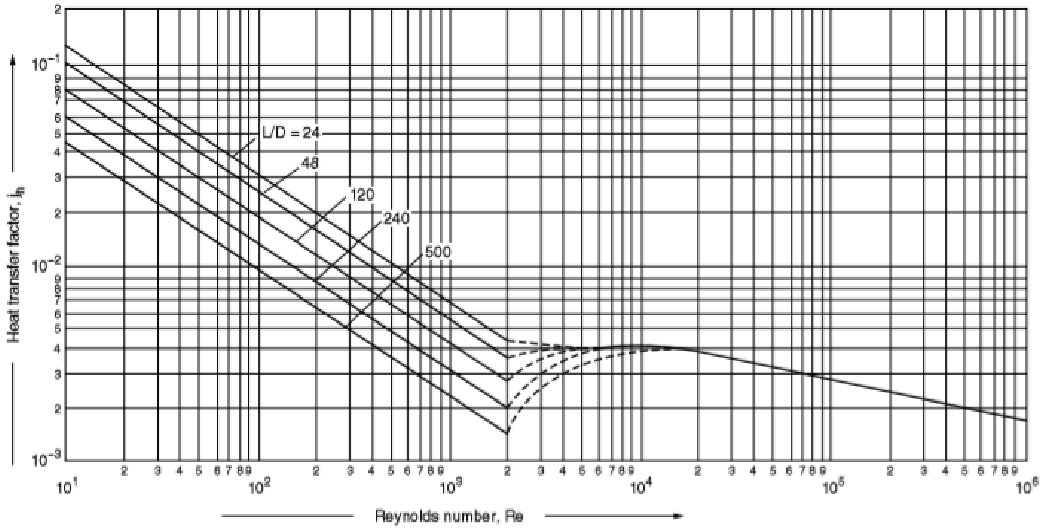


Figure 12.23. Tube-side heat-transfer factor

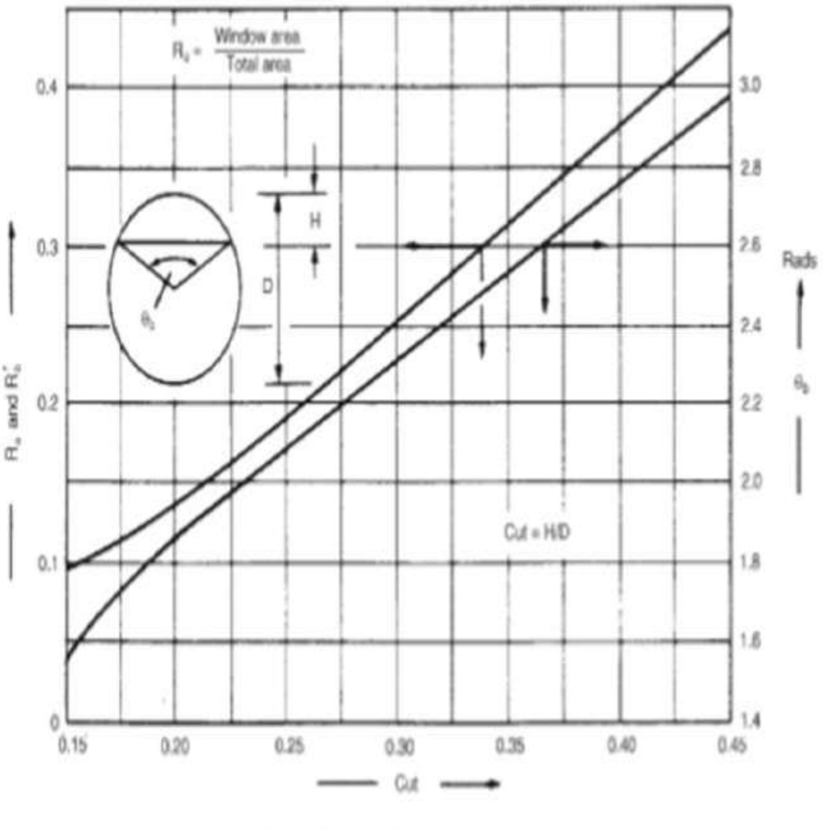


Figure 12.41. Baffle geometrical factors

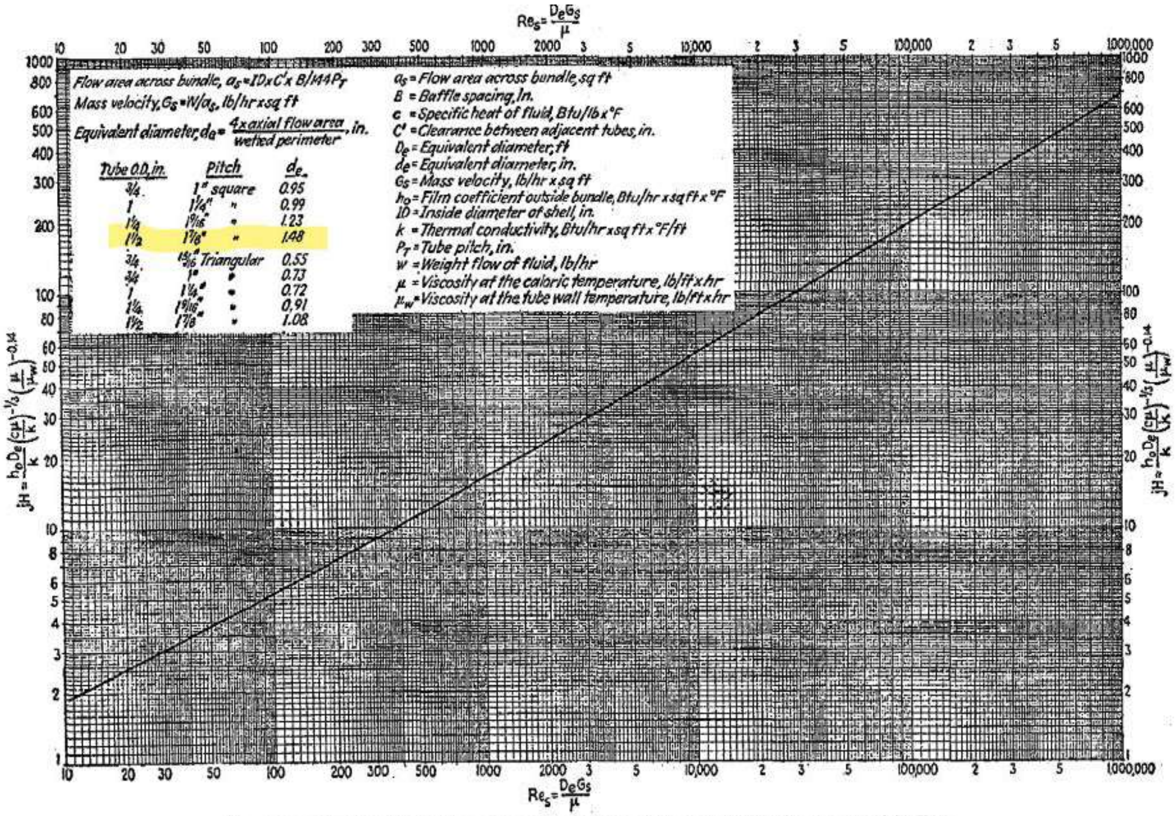


FIG. 28. Shell-side heat-transfer curve for bundles with 25% cut segmental baffles.

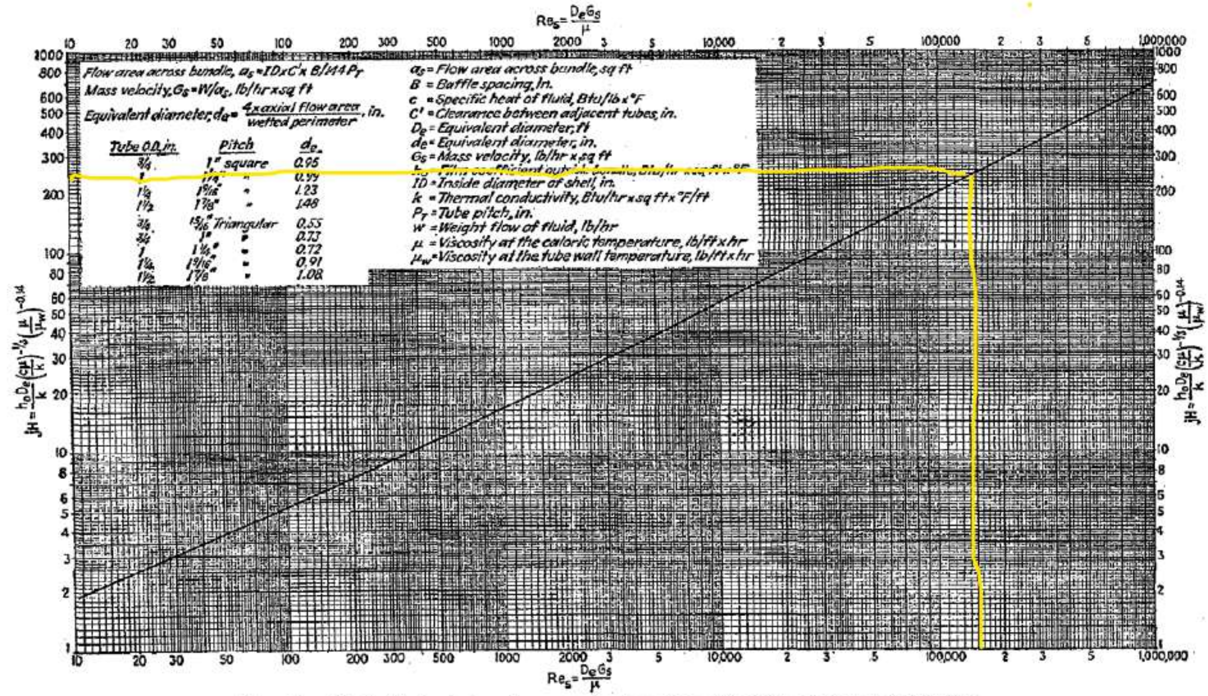


FIG. 28. Shell-side heat-transfer curve for bundles with 25% cut segmental baffles.

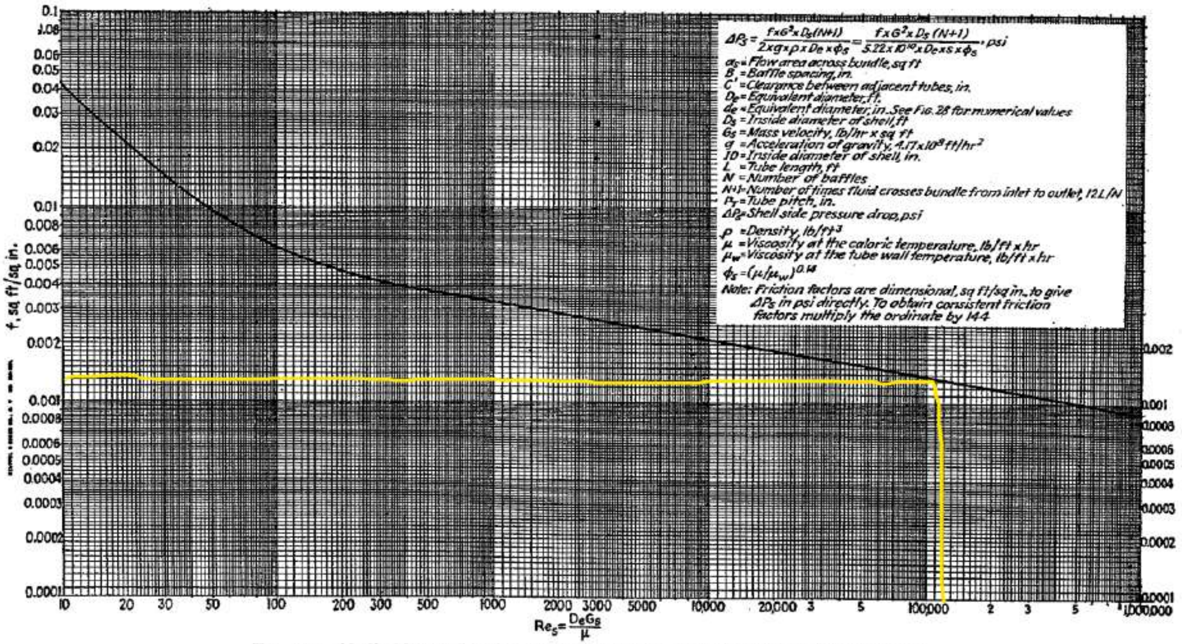


FIG. 29. Shell-side friction factors for bundles with 25% cut segmental baffles.

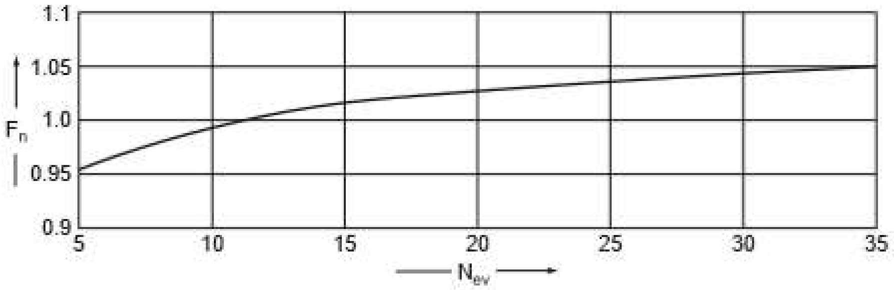


Figure 12.34. Tube row correction factor F_n.

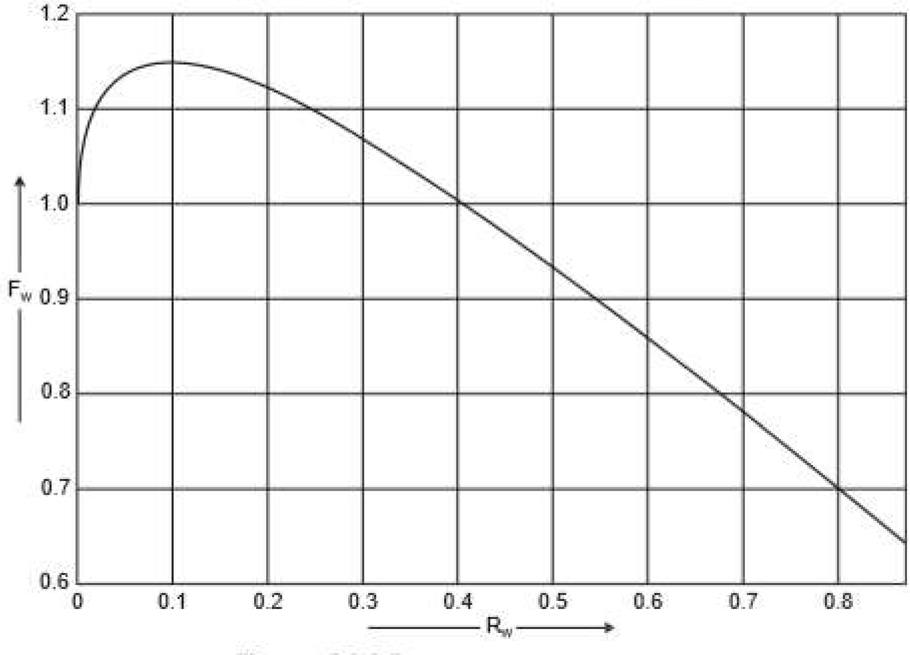
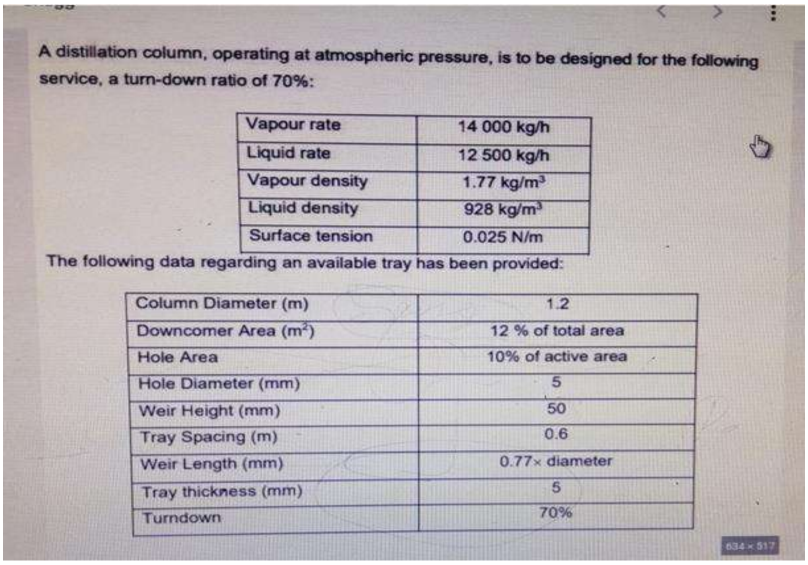


Figure 12.35. Window correction factor.

14.2 Appendix-B:



A distillation column, operating at atmospheric pressure, is to be designed for the following service, a turn-down ratio of 70%:

Vapour rate	14 000 kg/h
Liquid rate	12 500 kg/h
Vapour density	1.77 kg/m³
Liquid density	928 kg/m³
Surface tension	0.025 N/m

The following data regarding an available tray has been provided:

Column Diameter (m)	1.2
Downcomer Area (m²)	12 % of total area
Hole Area	10% of active area
Hole Diameter (mm)	5
Weir Height (mm)	50
Tray Spacing (m)	0.6
Weir Length (mm)	0.77× diameter
Tray thickness (mm)	5
Turndown	70%

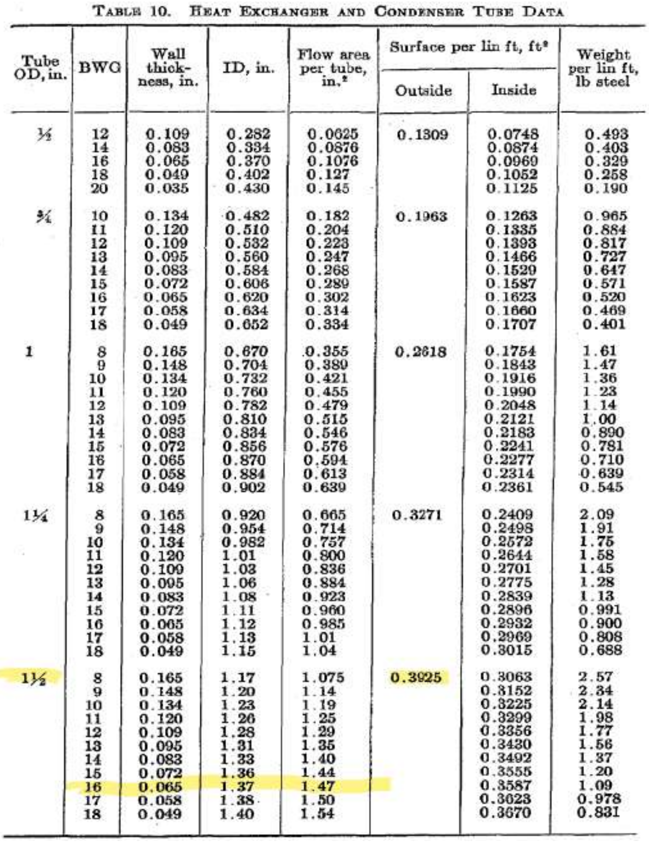
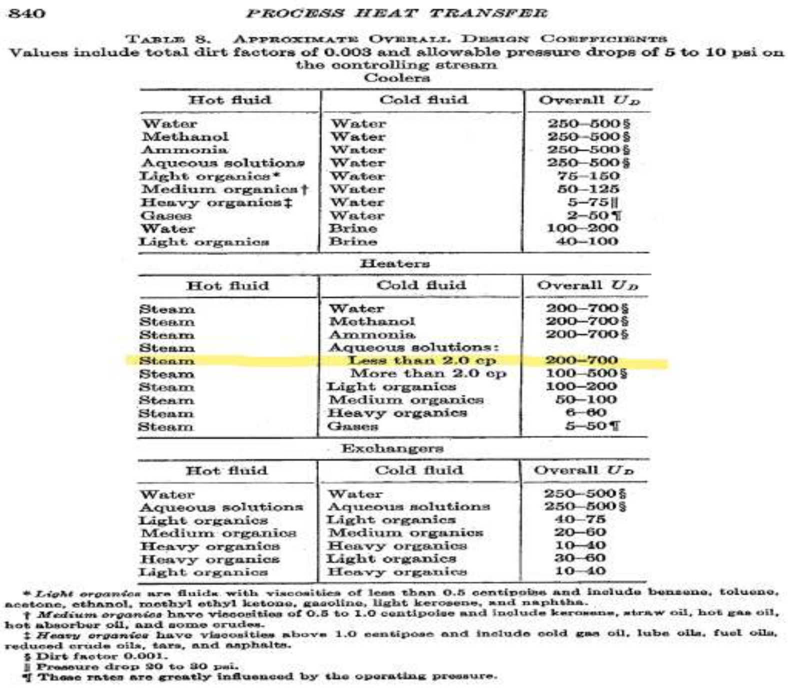


TABLE 10. HEAT EXCHANGER AND CONDENSER TUBE DATA

Tube OD, in.	BWG	Wall thickness, in.	ID, in.	Flow area per tube, in.²	Surface per lin ft, ft²		Weight per lin ft, lb steel
					Outside	Inside	
½	12	0.109	0.282	0.0625	0.1309	0.0748	0.493
	14	0.083	0.334	0.0876		0.0874	0.403
	16	0.065	0.370	0.1076		0.0969	0.329
	18	0.049	0.402	0.127		0.1052	0.258
	20	0.035	0.430	0.145		0.1125	0.190
¾	10	0.134	0.482	0.182	0.1963	0.1263	0.965
	11	0.120	0.510	0.204		0.1335	0.884
	12	0.109	0.532	0.223		0.1393	0.817
	13	0.095	0.560	0.247		0.1466	0.727
	14	0.083	0.584	0.268		0.1529	0.647
	15	0.072	0.606	0.289		0.1587	0.571
	16	0.065	0.620	0.302		0.1623	0.520
	17	0.058	0.634	0.314		0.1660	0.469
	18	0.049	0.652	0.334		0.1707	0.401
1	8	0.165	0.670	0.355	0.2618	0.1754	1.61
	9	0.148	0.704	0.389		0.1843	1.47
	10	0.134	0.732	0.421		0.1916	1.36
	11	0.120	0.760	0.455		0.1990	1.23
	12	0.109	0.782	0.479		0.2048	1.14
	13	0.095	0.810	0.515		0.2121	1.00
	14	0.083	0.834	0.546		0.2183	0.890
	15	0.072	0.856	0.576		0.2241	0.781
	16	0.065	0.870	0.594		0.2277	0.710
	17	0.058	0.884	0.613		0.2314	0.639
	18	0.049	0.902	0.639		0.2361	0.545
1¼	8	0.165	0.920	0.665	0.3271	0.2409	2.09
	9	0.148	0.954	0.714		0.2498	1.91
	10	0.134	0.982	0.757		0.2572	1.75
	11	0.120	1.01	0.800		0.2644	1.58
	12	0.109	1.03	0.836		0.2701	1.45
	13	0.095	1.06	0.884		0.2775	1.28
	14	0.083	1.08	0.923		0.2839	1.13
	15	0.072	1.11	0.960		0.2896	0.991
	16	0.065	1.12	0.985		0.2932	0.900
	17	0.058	1.13	1.01		0.2969	0.808
	18	0.049	1.15	1.04		0.3015	0.688
1½	8	0.165	1.17	1.075	0.3925	0.3063	2.57
	9	0.148	1.20	1.14		0.3152	2.34
	10	0.134	1.23	1.19		0.3225	2.14
	11	0.120	1.26	1.25		0.3299	1.98
	12	0.109	1.28	1.29		0.3356	1.77
	13	0.095	1.31	1.35		0.3430	1.56
	14	0.083	1.33	1.40		0.3492	1.37
	15	0.072	1.36	1.44		0.3555	1.20
	16	0.065	1.37	1.47		0.3587	1.09
	17	0.058	1.38	1.50		0.3623	0.978
	18	0.049	1.40	1.54		0.3670	0.831



840 *PROCESS HEAT TRANSFER*

TABLE 8. APPROXIMATE OVERALL DESIGN COEFFICIENTS
Values include total dirt factors of 0.003 and allowable pressure drops of 5 to 10 psi on the controlling stream

Coolers

Hot fluid	Cold fluid	Overall U_D
Water	Water	250–500§
Methanol	Water	250–500§
Ammonia	Water	250–500§
Aqueous solutions	Water	250–500§
Light organics*	Water	75–150
Medium organics†	Water	50–125
Heavy organics‡	Water	5–75‖
Gases	Water	2–50¶
Water	Brine	100–200
Light organics	Brine	40–100

Heaters

Hot fluid	Cold fluid	Overall U_D
Steam	Water	200–700§
Steam	Methanol	200–700§
Steam	Ammonia	200–700§
Steam	Aqueous solutions:	
Steam	Less than 2.0 cp	200–700
Steam	More than 2.0 cp	100–500§
Steam	Light organics	100–200
Steam	Medium organics	50–100
Steam	Heavy organics	6–60
Steam	Gases	5–50¶

Exchangers

Hot fluid	Cold fluid	Overall U_D
Water	Water	250–500§
Aqueous solutions	Aqueous solutions	250–500§
Light organics	Light organics	40–75
Medium organics	Medium organics	20–60
Heavy organics	Heavy organics	10–40
Heavy organics	Light organics	30–60
Light organics	Heavy organics	10–40

* *Light organics* are fluids with viscosities of less than 0.5 centipoise and include benzene, toluene, acetone, ethanol, methyl ethyl ketone, gasoline, light kerosene, and naphtha.

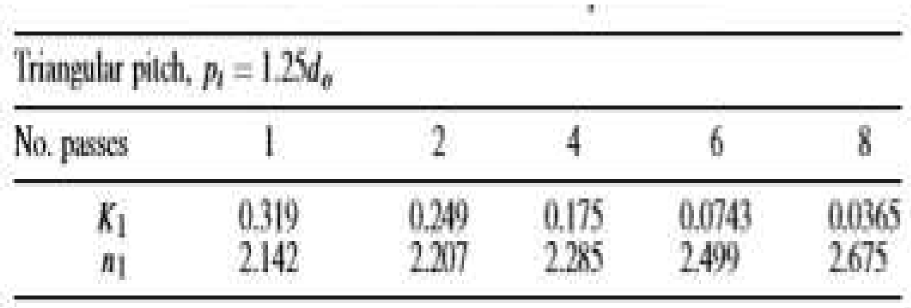
† *Medium organics* have viscosities of 0.5 to 1.0 centipoise and include kerosene, straw oil, hot gas oil, hot absorber oil, and some crudes.

‡ *Heavy organics* have viscosities above 1.0 centipose and include cold gas oil, lube oils, fuel oils, reduced crude oils, tars, and asphalts.

§ Dirt factor 0.001.

‖ Pressure drop 20 to 30 psi.

¶ These rates are greatly influenced by the operating pressure.



Triangular pitch, $p_t = 1.25d_o$

No. passes	1	2	4	6	8
K_1	0.319	0.249	0.175	0.0743	0.0365
n_1	2.142	2.207	2.285	2.499	2.675

ACHIEVEMENTS

Ref No: PEC/PPDC/22/ 334

PAKISTAN ENGINEERING COUNCIL

REGULATING THE ENGINEERING PROFESSION

"Final Year Design Project 2022-23"

Proposal Acceptance Certificate

It is to certify you that Final Year Design Project

Production of 850 X 10 power three TPA of Bio-Ehtylenet
from Zea-Mays Corn Stover.

(Chemical Department, University of Wah, Wah)

of your university has been approved for financial assistance by PEC

CONGRATULATIONS

It is required to complete this Design Project in accordance with the PEC standards so it can be included in the Capstone Expo-2023 and have it uploaded to the PEC E-library.

ENGR. MIR MASOOD RASHID
Convener, PPDC, PEC

ENGR. DR. NASIR M. KHAN
Registrar, PEC

University of Wah

OPEN HOUSE
& JOB FAIR 2023

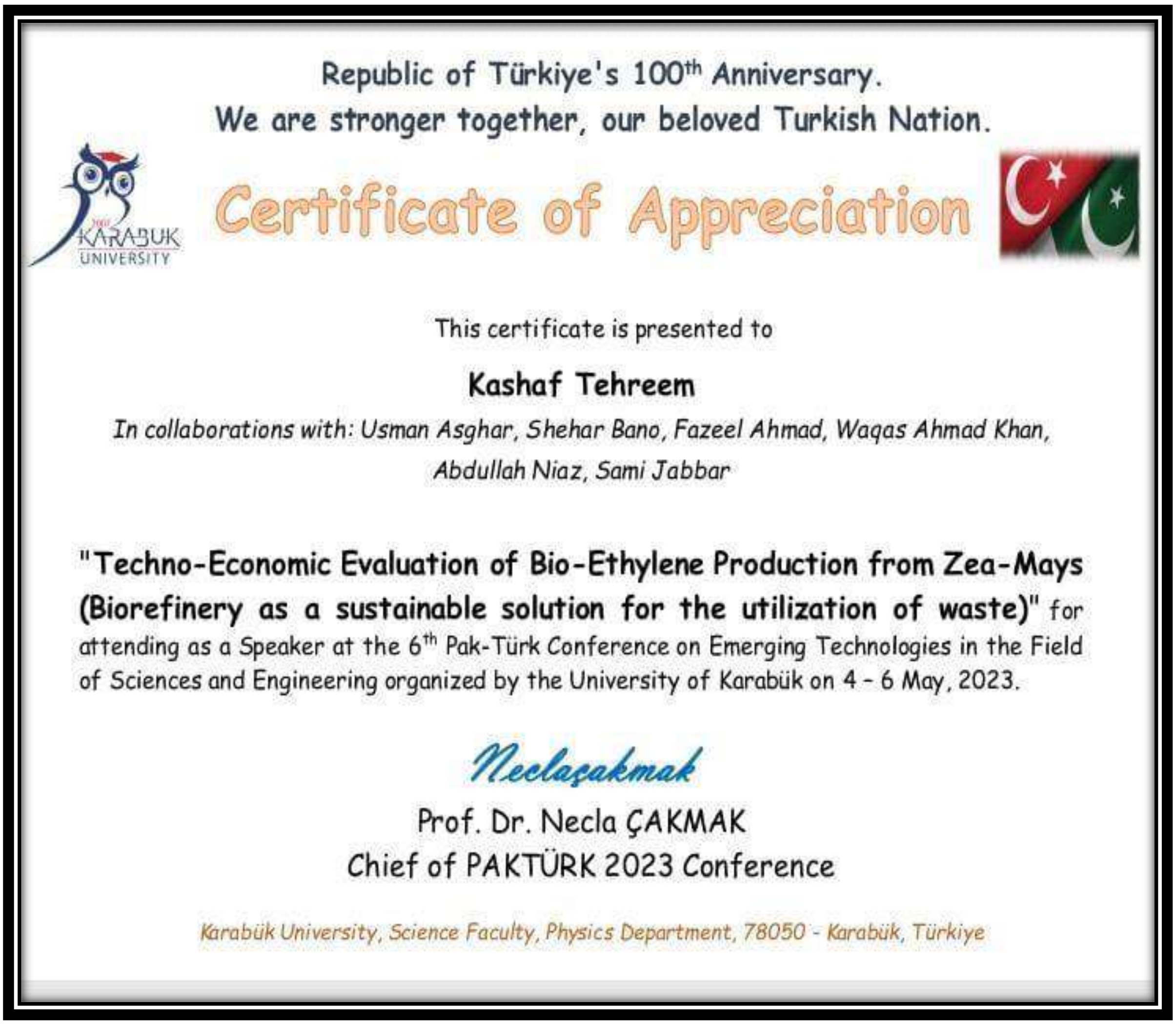
Certificate of Achievement

This certificate is presented to Mr/Ms **Shehar Bano**

for **Production of 3000 Tons/Day of Bio-Ethylene from Zea-Mays**

and securing 1st Position in **Final Year Project 2023**

Prof. Dr. Adnan Tariq
Dean, Faculty of Engineering



Republic of Türkiye's 100th Anniversary.
We are stronger together, our beloved Turkish Nation.

Certificate of Appreciation

This certificate is presented to

Kashaf Tehreem

In collaborations with: Usman Asghar, Shehar Bano, Fazeel Ahmad, Waqas Ahmad Khan, Abdullah Niaz, Sami Jabbar

"Techno-Economic Evaluation of Bio-Ethylene Production from Zea-Mays (Biorefinery as a sustainable solution for the utilization of waste)" for attending as a Speaker at the 6th Pak-Türk Conference on Emerging Technologies in the Field of Sciences and Engineering organized by the University of Karabük on 4 - 6 May, 2023.

Neclaçakmak

Prof. Dr. Necla ÇAKMAK
Chief of PAKTÜRK 2023 Conference

Karabük University, Science Faculty, Physics Department, 78050 - Karabük, Türkiye

CERTIFICATE OF PROOF READING

It is certified that the submitted manuscript titled as "***Production of 3000 TDP of Bio-Ethylene from Zea Mays***" is as per the standard formatting given by the FYP Coordinator.

Supervisor Signature: ----------------------------

FYP Coordinator Signature: ----------------------------

Department of Chemical Engineering

PLAGIARISM REPORT