PLANT DESIGN PROJECT FOR PRODUCTION OF 30,000 MTA MALEIC ANHYDRIDE	GROUP NO	16
	DESIGNER	SYED OSAMA
	MATRIC	13490

DETAILED DESIGN OF
MAJOR EQUIPMENT
MULTI TUBULAR FIXED-BED REACTOR (R-1)
&
DESIGN OF MINOR EQUIPMENTS
HEAT EXCHANGER (E-2)
MIXER (M-1)

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CHAPTER 1 INTRODUCTION

1.1 Process Description

The process involves catalytic oxidation of n-butane to produce Maleic anhydride (MA) as the main product. By products include carbon dioxide, water, acrylic acid and formic acid. The process is carried out in gas phase, where the reactions are catalyzed by heterogeneous Vanadium Phosphorus Oxide (VPO) catalyst. The catalytic oxidation reaction of n-butane is highly exothermic, thus, it is necessary to control the maintain temperature within the catalytic bed. Following are the reactions for this process:

$$C_6H_6 + \frac{9}{2}O_2 \rightarrow C_4H_2O_3 + 2CO_2 + 2H_2O \quad (1)$$

$$C_4H_2O_3 + 3O_2 \rightarrow 4CO_2 + H_2O \quad (2)$$

$$C_6H_6 + \frac{15}{2}O_2 \rightarrow 6CO_2 + 3H_2O \quad (3)$$

As for this project, a single vertical multitubular fixed-bed reactor type has been chosen to carry out the exothermic oxidation reaction. The MA Reactor, R-1 is designed as single shell and tube heat exchanger wherein the catalyst is packed in bundles of tubes to achieve an optimum conversion of 85 %. Molten salt is used as heat-exchange medium for temperature control.

1.2 Design Methodology

1. Select the major equipment – Reactor, R-1
2. Justification of selection of the type of reactor.
3. Calculations for the volume of the reactor.
4. Sizing of reactor.
5. Obtain the necessary parameters from reactor sizing calculation.
6. Proceed with equipment mechanical design.
7. Costing is done on the reactor and the utilities needed.
8. Perform technical drawing of the designed reactor.
9. Perform start up and shut down procedures for the reactor.

1.3 Reactor and Operating Conditions

Commercially there are two reactor technologies preferred in industry for the the production of Maleic anhydride. These are Huntsman fixed-bed and Alusiusse-

Lummus (ALMA) fluidized bed. For this plat we have selected a fixed-bed reactor because the process is a heterogeneous catalysis process where the catalyst and reacting species are of different phases [Timmerhaus et al, 2003]. The solid catalyst is present as a bed of relatively small individual particles, randomly oriented and fixed in position. The gas moves by convective flow through the spaces between the particles. The advantages using fixed bed reactor as compared to fluidized bed is summarized in the table below.

Table 1.1: The advantages and disadvantages of Fluidized bed and Multi-tubular Packed Bed reactor

	Fluidized bed reactor	Multi- tubular Fix bed reactor
Advantages	1. internal cooling coils for heat removal- effective temperature control- avoid hot spot	1. efficient contacting in the reactor – flow in PFR manner
	2. internal or external cyclones to minimize catalyst carry over	2. gives higher conversion per weight of catalyst
	3. usually use for liquid phase- assure intimate contact between feed & product vapors, catalyst and heat transfer surface	3. suitable liquid and gas phase
		4. No catalyst stickiness and highly efficient over many years of operation
Disadvantages	1. agglomeration – catalyst carry over downstream- copper contaminated	1. not effective in temperature control- Hot spots - overcome this problem by putting the cooling medium on the shell side
	2. reduce heat transfer capability in the reactor and reduce reaction rates	2. Besides, temperature control by multiple reactors in series- but increase cost
	3. inherent back mixing- difficult to achieve total conversion of limiting feed	
	4. High cost of the reactor and catalyst regeneration equipment	

1.4 Flow and Thermal Bed Arrangement

From the figure above, the first division is with respect to flow arrangement. Most fixed-bed reactors are operated with axial flow of fluid down the bed of solid particles. In this case, R-1 is operated with axial flow of gas.

In production of EDC, the operation is non-adiabatic. Heat transfer for control of temperature is accomplished within the bed itself. Thus the reactors are multitubular reactors and not multistage reactors. Molten salt is used as heat transfer medium for temperature control. However Reactor is assumed to be isothermal for simplicity.

1.5 Optimum Operating Conditions and Stream Data

The optimum operating conditions are obtained from literature review of Felthouse, T. R. *et al.* (2001). These conditions are as below:

Operating Temperature: 410°C

Operating Pressure: 5atm

Optimum single-pass conversion: 85 %

Type of catalyst: Vanadium Phosphorus Oxide (VPO)

Operation mode: Continuous

CHAPTER 2 PROCESS DESIGN

2.1 Reactor Volume, Space Time and Amount of Catalyst

The size of the reactor can be estimated from the volume of the bulk catalyst. Since *multitubular* fixed bed reactor is employed, the catalyst is packed in bundles of tubes to achieve a desired conversion of 85%. According to Fogler (2006), the design equation for a fixed-bed reactor is analogous to that for a plug-flow reactor (PFR) with catalyst. Thus, for a specified conversion, we determine the weight of catalyst required by solving the design equation. The differential equation for a PFR with catalyst is:

$$\frac{dX}{dW} = \frac{-r'_A}{F_{Ao}}$$

This equation deals with catalyst weight. To find the catalyst volume we can divide the the catalyst bulk density by the catalyst weight. First of all, we determine the kinetics of the catalytic oxidation of *n*-butane over VPO. This has been a subject of numerous investigations as reported in the literature. However we select the kinetic model by Centi et al. (1985) where the reaction mechanism was derived from the data generated in an isothermal, steady state tubular fixed bed reactor in which the oxygen was supplied to the catalyst through the gas phase. In this kinetic model, the desired reaction to produce Maleic anhydride is described as below:

$$C_4H_{10} + 3.5O_2 \xrightarrow{k_1} C_4H_2O_3 + 4H_2O$$

For this reaction, the rate law proposed by Centi et al. (1985) is as follows:

$$r_1 = \frac{k_1 K_B C_{O_2}^{\alpha} C_B}{(1 + K_B C_B)}$$

Where r_1 is the rate of Maleic anhydride formation from *n*-butane, α=0.2298 and K_B=2.616 m^3/mol. The rate constant k_1 obey the Arrhenius equation as:

$$k_i = k_{0,i} \exp\left[-\frac{E_i}{R}\left(\frac{1}{T} - \frac{1}{T_0}\right)\right]$$

Where E_i = 45.1 x 10^3 J/mol, $k_{0,i}$ = 2.191 x 10^{-4} , R = 8.314 $Jmol^{-1}$ K^{-1} and T_0 *(K)* = 653

$$K_1 = 3.155 \times 10^{-4}$$

Secondly, we determine the stoichiometric expressions for the gas phase continuous reactions from Fogler (2006)

For n-butane,

$$C_A = C_{Ao}\left(\frac{1-X}{1+\varepsilon X}\right)\frac{T_0}{T}\left(\frac{P}{P_o}\right)$$

For Oxygen,

$$C_{O2} = C_{B0}\left(\frac{\emptyset_B - (\frac{b}{a})X}{1+\varepsilon X}\right)\frac{T_0}{T}\left(\frac{P}{P_o}\right)$$

Where $C_{B0} = (y_{B0} \times P_0) / (RT_0) = 2.68$ mol/m^{3,} X = 0.85, b = 3.5, a = 1

$\varepsilon = y_{B0} \times 0.5 = 0.017 \times 0.5 = 0.0085$,

$\emptyset_B = 57.8$

Assuming no pressure drop and isothermal conditions, the temperature and pressure terms are eliminated. Hence to calculate the weight of catalyst, the following equation is used:

$$W = F_{B0} / (k_1 C_{B0}) \int^{0.85} dX / ((K_B C_{02} C_B)/ (1 + K_B C_B))$$

Given that the mass flow rate of n - butane is F_{B0} = 11.02 Kg/s. The above integral can be solved using definite integral calculator at www.solvemyamath.com

Weight of Catalyst, W = 177,102 Kg

Density of Vanadium Phosphorous Oxide (VPO) = 900Kg/ m^3

Therefore the Volume of bulk catalyst = 177102/900 = 196.78 m^3

Assuming Void fraction = 0.5,

Volume of Reactor, V = 196.78/0.5 = 393.6m^3

The volumetric flow rate, v_o into R-1 obtained from ICON simulation = 20.3 m^3/s. Thus, the space time is calculated as:

Space time = $\tau = V/v_o$ = 196.78/20.3 = 9.7s

2.2 Shell and Tube Design for Reactor

Stainless steel type 304(18Cr/8Ni) is a suitable material for the construction of reactor tubes because of its good corrosion resistance and mechanical properties. It is usually used for heat exchanger tubing. This multi tube reactor can be designed with close approximation to a shell and tube heat exchanger [Timmerhaus, 2003]. The catalyst volume must be equal to the inside volume of the tubes bundle. Hence by selecting a standard tube diameter and length, the number of tubes and shell inside diameter can be determined.

According to Perry's chemical engineer's handbook, hundreds of tubes of a few centimetres (cm) in diameter maybe required in the fixed-bed multi-tube reactor. In this reactor design, standard tubes of **2.375 in. (0.06033 m) outside diameter** of stainless steel 304 pipe, with **6.1 m length** are selected. Although this is a large size for heat exchanger tubing, a large size is desirable for good catalyst distribution and minimal wall effects. The properties of the pipe are stated in Table below wherein the values are taken from Geankoplis (2003).

Table2.1 Standard dimensions for Stainless Steel tubes

Nominal pipe size	: 2 in (0.0508 m)
Outside diameter	: 2.375 in (0.06033 m)
Inside diameter	: 2.067 in (0.0525 m)
Wall thickness	: 0.154 in (0.00391 m)
Length	: 20 ft (6.1 m)
Cross sectional area	: 3.356 in^2 (0.00216 m^2)

By using the tube dimensions given above,

Volume of one tube = cross sectional area x tube length

$$= 0.00216\ m^2 \times 6.1\ m$$

$$= 0.0132\ m^3$$

$$\text{Number of tubes, } Nt = \frac{\text{Actual volume of catalyst}}{\text{Volume of one tube}} = \frac{196.78\ m^3}{0.0132\ m^3} = \textbf{14908 tubes}$$

Given that the Maleic anhydride reaction is exothermic, a high heat transfer rate is required. Therefore equivalent triangular pattern for tube arrangement is selected. The recommended tube pitch (distance between tube centre) is 1.25 time the outside diameter of the tube (Sinnott, 2000).

Tube Pitch, $d_o = 0.06033m$, $P_t = 1.25 \times 0.06033 = 0.0754m =$ **75.4mm**

For estimation of tube layout or bundle diameter we use the following equation:

$$D_b = d_0 \left(\frac{N_t}{K_1} \right)^{\frac{1}{n_1}}$$

Where N_t = Number of tubes

D_b = Bundle diameter, m

d_0 = Tube outside diameter, m

The value of K_1 and n_1 are taken from table below, Chemical Engineering Vol. 6.

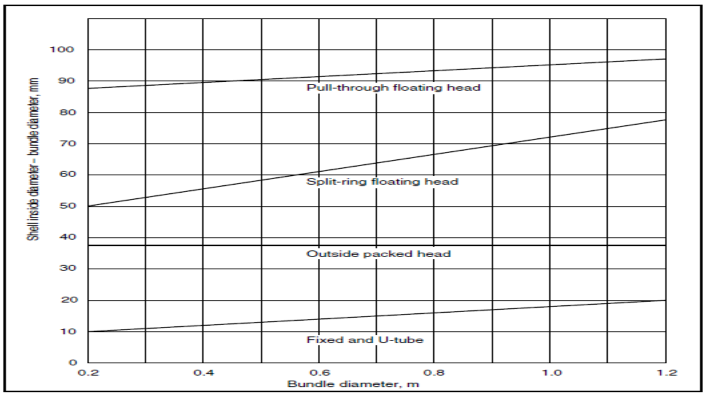
Table 12.4. Constants for use in equation 12.3

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
Square pitch, $p_t = 1.25d_o$					
No. passes	1	2	4	6	8
K_1	0.215	0.156	0.158	0.0402	0.0331
n_1	2.207	2.291	2.263	2.617	2.643

The No of passes is 1, therefore K_1 = 0.319 and n_1 = 2.142

Bundle diameter, D_b = 0.06033 (14908 / 0.319)$^{1/2.142}$ = 9.1m

For shell diameter, a method should be selected that gives a close a fit to the tube bundle. This is to reduce bypassing round the outside of the bundle. Typical values of clearance required between the outermost tubes in the bundle and the shell inside diameter can be obtained from Figure 12.10 (Sinnot, 2000) below. Extrapolation on the fixed and U-tube line is performed.



Shell inside diameter – bundle diameter = Clearance

= 10(Bundle diameter-0.2) + 10

= 10 (9.1 - 0.2) + 10 = 99 mm

Shell inside diameter = Clearance + Bundle diameter

D_S = 0.099 m + 9.1 m

D_S = 9.199 m

Baffles are used in the shell to direct the fluid stream across the tubes, and to increase the fluid velocity. Therefore, the rate transfer will increase. The Baffle type used is single segmental baffle. From Table 12.5, Chemical Engineering, Vol. 6, the recommended baffle diameter is:

D_{bf} = Ds - 4.8 mm = 9.2 - 0.0048 = 9.194 m

No of baffles calculation, N_b:

The ideal baffle spacing is between 0.3 to 0.5 times of the shell diameter. An optimum value of 0.3 is used.

Optimum baffle spacing, l_b = 0.3 (9.194)

= **2.76 m**

Number of baffle required = (Total tube length / baffle spacing)-1

= (6.1 / 2.76)-1

= 1.2 ≈ **2 baffles**

2.3 Reactor Heat Removal

Since the reaction in R-201 is exothermic, all excess heat has to be removed to maintain an optimum temperature of 683K for the catalyst. A molten salt composed of 53% potassium nitrate, 40% sodium nitrate and 7% sodium nitrite has been chosen as the heat-transfer fluid. This salt circulates in the shell side of the reactor. Properties of the molten salt are provided in the table below:

Table 2.2 Properties of Molten Salt

Operating Pressure	200kPa
Density at T_{av} = 410°C	1976 kg/m^3
Heat Capacity at T_{av}= 410°C	2.73 kJ/kg°C
Viscosity at T_{av}= 410°C	2.0 cP

From ICON simulation results, 1.846 x 10^8 kJ/hr of heat has to be removed. The heat transfer liquid has temperature is set to 220°C at the inlet and 410°C at the outlet. Heat required to be removed, q = 1.864 x 10^8 kJ/hr = 5.178 x 10^4 kJ/s.

$$q = mC_p\Delta T$$

Hence flow rate of coolant required, m = (5.178 x 10^4) / (2.73 x 190) = 99.8 Kg/s

2.4 Pressure Drop Calculations

The equation used to calculate pressure drop on tube side in a packed porous bed is the Ergun's Equation;

$$\beta_o = \frac{G(1-\phi)}{\rho_o g_c D_p \phi^3}\left[\frac{150(1-\phi)\mu}{D_p} + 1.75G\right]$$

$$\frac{P}{P_o} = \left(1 - \frac{2\beta_o L}{P_o}\right)^{0.5}$$

Where;

P = outlet pressure (kPa), P_o = inlet pressure = 507 kPa, ϕ = porosity = 0.5, g_c = 1.0 (for metric system), D_P = diameter of particle in the bed = 5 mm, μ = viscosity of gas passing through bed, cP (1x 10^{-3} Pa.s) = 3.023 x 10^{-5}Pa.s, gas density ρ = 1.050 kg/m^3
Total number of tubes, N = 14908

From ICON simulation, the mass velocity of entering feed, m = 23.7 kg/s.

Superficial mass velocity, G = Mass velocity / Total cross area of tubes

= 23.7 / (0.00216 × 14908 tubes) = 0.736 kg/s.m^2

$$\beta_o = 976.5 = 0.976 kPa/m$$

$$\frac{P}{P_0} = \left(1 - \frac{2\times 0.976\times 6.1}{507}\right)^{0.5}$$

$$\frac{P}{P_0} = 0.988$$

$$P = 0.988\times 507 = 501\, kPa$$

$$\Delta P = 507 - 501 = 6\, kPa$$

Hence, **pressure drop on tube side is 6 kPa**

For pressure drop on shell side, we use the Kern's method (Coulson Richardson's Vol. 6):

$$\Delta P_s = 8 j_f \left(\frac{D_s}{d_e}\right)\left(\frac{L}{l_B}\right)\frac{\rho u_s^2}{2}\left(\frac{\mu}{\mu_w}\right)^{-0.14}$$

$$\Delta P_s = 8 j_f \left(\frac{D_s}{d_e}\right)\left(\frac{L}{l_B}\right)\frac{\rho u_s^2}{2}$$

L = tube length = 6.10 m

l_B = baffle spacing = 2.76 m

D_s = shell diameter = 9.199 m

d_e = equivalent diameter (triangular pitch arrangement)

$$= \frac{1.10}{d_o}\left(p_t^{\,2} - 0.917 d_o^{\,2}\right) = \frac{1.10}{0.06033}\left(0.0754^2 - 0.917 \times 0.06033^2\right) = 0.043 \text{ m}$$

ρ = molten salt density = 1976.0 kg/m^3 at 410^{o}C

μ = viscosity of molten salt = 2.0 x 10^{-3} Pa.s at 410^{o}C

u_s = molten salt linear velocity, m/s = G_s/ρ

j_{f} = friction factor, read from Fig. 12.30 (CE Vol. 6) by knowing the Reynolds number,

Re = $\rho u_s d_e/\mu$

$$A_s = \frac{(p_t - d_o)D_s l_B}{p_t}$$

$$A_s = \frac{(0.0754 - 0.06033) \times 9.199 \times 2.76}{0.0754} = 5.07\ m^2$$

$$G_s = \frac{m}{A_s}$$

$$G_s = \frac{99.8}{5.07} = 19.68\ kg/m^2.s$$

$$u_s = \frac{G_s}{\rho}$$

$$u_s = \frac{19.68}{1975.0} = 0.01\ m/s$$

Reynolds number, Re:

$$\text{Re} = \frac{\rho v d}{\mu}$$

$$\text{Re} = \frac{1975.0 \times 0.01 \times 0.043}{0.002} = 424.625$$

Since Re < 2100, hence flow is laminar.

From Figure 12.30, Chemical Engineering Vol. 6, baffle cut is taken as 25%.

For Re = 424.625, jf = 0.087

Hence, shell side pressure drop is

$$\Delta P_s = 8 j_f \left(\frac{L}{l_B} \right) \left(\frac{D_s}{d_e} \right) \frac{\rho u_s^{\,2}}{2}$$

$$= 8(0.087) \left(\frac{6.10}{2.76} \right) \left(\frac{9.199}{0.043} \right) \frac{1975.0 \times 0.01^2}{2}$$

= 32.5 Pa

= 0.0325 kPa

CHAPTER 3 MECHANICAL DESIGN

3.1 Reactor Design Pressure & Temperature

Shell-side operating pressure, P = 200kPa (absolute)

Thus design pressure, $\mathbf{P_{ds}}$ = (200) x 1.05 = **210 kPa**

Tube side operating pressure, P = 500kPa (absolute)

Thus design pressure, $\mathbf{P_{dt}}$ = (500) x 1.05 = **525 kPa**

The above pressures are calculated after considering 5% safety factor for internal pressure. Similarly, the reactor design temperature is calculated by considering 5% safety factor.

$\mathbf{T_d}$ = T_{op} (1.05) = 410 (1.05) = **430.05 ^{o}C**

3.2 Reactor Cylindrical Shell Thickness

On the shell side, molten salt at an average temperature of 315°C will be circulating to control the temperature of the reactor. Since molten salt is rather corrosive, stainless steel 304 (18Cr/8Ni) is selected as the material of fabrication for the shell of reactor. Minimum shell thickness is obtained as:

$$e = \frac{P_i D_i}{(2fJ - P_i)} + c$$

Where

e	=	Minimum thickness required, mm
D_i	=	Internal diameter, 9.199 m
f	=	Design stress, N/mm^2
P_i	=	Design pressure, N/mm^2
J	=	Welding efficiency, 0.9 For Class 1: Single welded butt joint with bonding strips
c	=	Corrosion allowance, mm

Design stress of Stainless Steel (304) from Mechanical Design Data Handbook,

Design stress, $f_{410^\circ C}$ = 108 N/mm^2

Since the process fluid is corrosive, 4 mm corrosion allowance shall be used.

$$e = \frac{(210\times10^3)\times(9.199\times10^3)}{2(1.08x10^8\times0.9)-(210\times10^3)} + 4$$

$$e = 13.95\,mm$$

Shell thickness equals to the higher wall thickness obtained, e = 13.95 mm. From the nearest standard steel sheet available, shell thickness = 16 mm.

$$D_0 = D_i + t = 9.22\ m$$

$$\frac{D_o}{D_i} = \frac{9.22\ m}{9.199\ m} = 1.002 \quad < 1.5 \quad (Acceptable)$$

$$\frac{t}{D_i} = \frac{16\ mm}{9199\ mm} = 0.0017 < 0.25 (Acceptable)$$

3.3 Reactor Head and Closure

A torispherical flanged standard dished head is chosen for this design. The advantages of using this head are that it can be used for application of higher pressure up to 15 bar and it has less stress concentration as compared to flat plate. This head is usually used for vertical pressure vessels up to 15 bars and suitable for Maleic anhydride reactor application. The minimum thickness required is:

$$t = \frac{PRcCs}{2fJ + P(Cs - 0.2)}$$

$$Cs = \frac{1}{4}(3 + \sqrt{\frac{Rc}{Rk}})$$

Where,

J	=	Weld joint factor, 0.9
f	=	Design stress, 100 N/mm^2
P	=	210 x 10^3 Pa
R_c	=	Crown radius = D_o
R_k	=	Knuckle radius = $0.06R_c$

R_k/R_c should not be less than 0.06

Internal diameter, D_i	=	9.199 m
Shell thickness, t	=	16 mm
Outer diameter, D_o	=	$D_i + t = 9.215m$
Crown radius, R_c	=	D_o = 9.215m

$$C_s = \frac{1}{4}\left[3 + \sqrt{\frac{1}{0.06}}\right] = 1.7706$$

$$t = \frac{(210 \times 10^3) \times 9.199 \times 1.7706}{2(0.9 \times 1.0 \times 10^8) + 210 \times 10^3 \times (1.7706 - 0.2)}$$

$$t = 0.019\ m$$

$$= 19\ mm$$

Thickness of closure, t_c = t + thinning of torus during fabrication (6% of thickness)

t_c = 19 + 0.06(19) = **20.14 mm**

Volume of dish, V = 0.0847 Di^3

= 0.0847 × $(9.199)^3$

= **65.9 m^3**

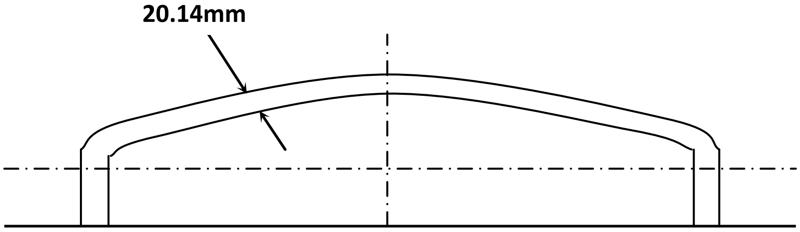


Figure 3.1Torispherical flanged standard dished head

3.4 Reactor Height

Height of straight flange, s_f = 0.1 m

Height of closure, h_o = 2 m (assumed)

The allowance of closure height is for internal fittings and maintenance.

Total height of the reactor = height of closures + tube length + height of flange

= 2(2) + 6.1 + 0.1

= **10.2 m**

3.5 Gasket Design

Gaskets are used to make a leak-tight joint between two surfaces. For reactor temperatures between 260 to 450 ^{0}C, metal reinforced gasket is recommended. Gasket specification is obtained from Table 13 of Data Hand Book of Mechanical Design of Process Equipment (ECB 5233).

Gasket material = Corrugated metal (Stainless steel, asbestos inserted)

Gasket factor, m = 3.5

Min design seating stress, y = 45 MN/m^2

Min actual gasket width = 10 mm

Design pressure, P_D = 210 kPa

Gasket diameter ratio,

$$\frac{do}{di} = \sqrt{\frac{y - P_D m}{y - P_D(m+1)}}. = 1.0024$$

di	=	D_o + 10 mm
D_o	=	9.2 m
Inner diameter of gasket, d_i	=	9.21 m
Outer diameter of gasket, d_o	=	9.23 m
Minimum gasket width, *N*	=	0.5 × (d_o – d_i)
	=	0.01 m
Actual total width, N_{total}	=	2 × 0.01
	=	0.02 m
Actual outer diameter, d_o	=	d_i + 2*N* = 9.25 m

3.6 Bolt Sizing

Type of flange facing = Plain face

Basic gasket seating width, b_o= 0.5*N* = 5 mm

Gasket load reaction, *G* = d_i + *N* = 9.2 + 0.01= 9.21 m

Allowable stress of bolting material (18Cr2Ni), $S_o = S_g$ = 144 MN/m^2

Load due to design pressure, *H*

$$H = \frac{\pi G^2}{4} \times P_D$$

$$H = \frac{\pi \times (9.21)^2}{4} \times (210 \times 10^{-3})$$

$$= 14\ MN$$

Load to keep the joint tight under operating condition, H_p

$$H_p = \pi\ G\ (2b)\ mp$$

$$H_p = \pi \times 9.21\left(2 \times 5 \times 10^{-3}\right) \times 3.5 \times \left(210 \times 10^{-3}\right) = 0.213\ MN$$

Total load under operating condition, W_o

W_o =force due to pressure + load to achieve minimum sealing

$$W_o = H + H_p = 14.213\ MN$$

Area under operating condition, A_o

$$A_o = \frac{W_o}{S_o} = 0.099\ m^2$$

Load to seat gasket under bolting up condition, W_g

$$W_g = \pi\ G\ (by)$$

$$W_g = \pi \times 9.21 \times (0.005 \times 45) = 6.51 MN$$

Area under bolting up condition, A_g

$$A_g = \frac{W_g}{S_g} = 0.045\ m^2$$

Since $W_o > W_g$; hence controlling load, W = 14.213 MN

Since $A_o > A_g$; hence minimum bolt area, A_{min} = 0.099 m^2

3.7 Design of Flange Ring

(i) Moment about Flange,

W_1 = Hydrostatic end force on area inside flange

W_2 =Unbalanced forces due to pressure acting on downward direction

W_3 = Gasket load

$$W_1 = \frac{\pi B^2}{4} \times P_{D1} = \frac{\pi \times 9.199^2}{4} \times \left(210 \times 10^{-3}\right) = 13.96\ MN$$
$$W_2 = H - W_1 = 14 - 13.96 = 0.04\ MN$$
$$W_3 = H_p = 0.213\ MN$$

Net bolt load, $W_o = W_1 + W_2 + W_3 =$ **14.213 MN**

(ii) Moment Arms on Flange

Shell outside diameter, B = 9.199 m

Bolt circle diameter, C = 9.268 m

Gasket load reaction, G = 9.21 m

$$a_1 = \frac{C - B}{2} = 0.0345\ m$$
$$a_3 = \frac{C - G}{2} = 0.029\ m$$
$$a_2 = \frac{a_1 + a_3}{2} = 0.03175\ m$$

(iii) Moment of Force

Moment of force about bolt circle diameter (BCD) under operating condition, M_o

$$M_o = W_1 a_1 + W_2 a_2 + W_3 a_3$$
$$M_o = (13.96 \times 0.0345) + (0.04 \times 0.03175) + (0.213 \times 0.029) = 0.489\ MN.m = 0.489\ MJ$$

Moment of force about bolt circle diameter under bolting up condition, M_g

$$W_g = \frac{A_{\min} + A_b}{2} \times S_g$$
$$W_g = \frac{0.099 + 0.089}{2} \times 144 = 13.536\ MN$$

$M_g = W_g a_3$

$M_g = 13.536 \times 0.029$

$= 0.393\ MN.m = 0.393\ MJ$

Since $M_o > M_g$, thus a larger moment of M_o is used, M = **0.489 MJ**

(iv) Flange Thickness, t

$$t = \sqrt{\frac{M.C_F y}{BS_t}}$$

Initial assumption of bolt pitch correction factor, C_f = 1.00

Correction coefficient, y = 18.50

Allowable stress of flange material, S_t = 100 MN/m^2

Thus, t = 99.17 mm

The nearest standard steel sheet has a thickness of 100 mm.

Thus, flange thickness, t = **100mm**

Flange outside diameter, A = BCD + bolt diameter + 20 mm = **9.30**

3.8 Reactor Weight

Weight of Shell: For cylindrical vessel with domed ends, and uniform wall thickness, the approximate weight can be estimated from: $W_v = 240\, C_v D_m \left(H_v + 0.8 D_m\right) t$

Where

W_v = total weight of shell, excluding internal fittings (such as plates)

C_v = factor to account for the weight of nozzles, manways, internal support

1.15 for distillation column or similar vessels with fittings.

D_m = mean diameter of vessel = (D_i + t ×10^{-3}) = (9.199 + 16 × 10^{-3}) = 9.215m

H_v = height or length between tangent lines (cylindrical length) = 6.1 m

Weight of shell, W_v =240 × 1.15 × 9.215 (6.1 + 0.8 × 9.215)16 = **548.22 kN**

For Weight of Baffles,

Number of baffles	=	2
Baffle diameter, D_{bf}	=	9.194 m
Baffle area, A_b	=	$\pi(9.194)^2/4$
	=	66.39 m^2
Typical weight of baffle	=	1.2 kN/m^2 plate area
Weight per baffle, W	=	66.39 × 1.2 kN/m^2 = 79.67 kN/m^2
Total weight of baffles	**=**	**2 × 79.67 = 159.34 kN**

For Weight of Tubes,

Number of tubes	=	14908
Weight per feet of tube	=	2.72 lb/ft = 39.7090 N/m
Length of tube	=	6.1 m
Total weight of tubes	**=**	**14908 × 6.1 × 39.0790 = 3553.8 kN**

For Weight of Fluid,

Total weight of fluid comprises of the weight of process fluid, catalyst and coolant.

Volume of process fluid	=	196.78 m^3
Density of process fluid	=	0.5242 kg/m^3
Weight of process fluid	=	196.78 × 0.5242 × 9.81 = **1.01 kN**
Density of catalyst	=	1000 kg/m^3
Weight of catalyst	=	196.78 × 1000 × 9.81 = **1930.4 kN**

Assuming maximum volume of coolant in the reactor,

Volume of shell = $(\pi\ 9.199^2)/4 \times 6.1$ = 405.4 m^3

Volume occupied by tubes = 14908 × $[\pi/4 \times (0.06033)^2 \times 6.1]$ = 260 m^3

Volume of molten salt = Volume of shell - volume occupied by tubes = 145.4 m^3

Density, $\rho_{molten\ salt}$ = 1975.0 kg/m^3

Weight of molten salt = 145.4 × 1975.0 × 9.81= **2817 kN**

Total weight of fluid = 1.01 + 1930.4 + 2817 = 4748.4 kN

Hence total weight of reactor = 9009.8 kN

3.9 Reactor Support- Skirt Support

A skirt support consists of a cylindrical or conical shell welded to the base of the vessel. For this design, a straight cylindrical (θ = 90˚) skirt support is used. A flange at the bottom of the skirt transmits the load to the foundations. Openings must be provided in the skirt for access and for any connecting pipes, the openings are normally reinforced. The skirt may be welded to the bottom head of the vessel shell. Skirt supports are recommended for vertical vessels as they do not impose concentrated loads on the vessel shell. They are particularly suitable for use with tall columns subjected to wind loading.

Carbon steel has been chosen as the material for skirt with the design stress = 135 N/mm2 and Young Modulus, E = 200,000 N/mm2 at ambient temperature.

W_v = 548.22 kN

For safety purpose, add 10 kN to Wv, = 558.22 kN

Bending moment, at the base of the skirt, $M_s = F_w \dfrac{(H_v + H_{skirt})^2}{2}$

Where; Wind loading, F_w = 11.533 kN

H_v = 6.1 m

H_{skirt} = 3 m (assumed height of skirt)

M_s = 509.5 kN

Bending stress in skirt,

$$\sigma_{bs} = \frac{4M_s}{\pi(D_s + t_s)t_s D_s}$$

Where; Skirt thickness, t_s = 0.01 m

$D_s = D_o$ = 9.2 m

So, σ_{bs} = 833007.45 N/m^2

Dead weight stress in skirt,

$$\sigma_{ws} = \frac{W}{\pi(D_s + t_s)t_s}$$

= 1889307.96 N/m^2

Resultant stresses;

$\sigma_s(tensile) = \sigma_{bs} - \sigma_{ws}$ = -1.0563 N/mm^2

$\sigma_s(compressive) = \sigma_{bs} + \sigma_{ws}$ = 2.7223 N/mm^2

The following conditions must be satisfied for the design to be valid:

1) $\sigma_s(tensile) \leq f_s J \sin\theta_s$

2) $\sigma_s(compressive) \leq 0.125E\left(\frac{t_s}{D_s}\right)\sin\theta_s$

Where

f_s = maximum allowable design stress for the skirt material at ambient temperature

= 135 N/mm^2

J = weld joint factor = 1.0

θ_s = base angle of a conical skirt, 90^o

E = Young modulus of the material = 200,000 N/mm^2 for plain carbon steel

After calculation:

1) $\sigma_s(tensile) \leq f_s J \sin\theta_s$

$-2.435 N/mm^2 \leq 135 N/mm^2$

2) $\sigma_s(compressive) \leq 0.125E\left(\frac{t_s}{D_s}\right)\sin\theta_s$

$17.231 N/mm^2 \leq 28.345 N/mm^2$

$\therefore$ **Both conditions are satisfied.**

3.10 Nozzle Sizing

Four nozzles are designed according to each stream specifications, namely: feed stream nozzle, reactor product outlet nozzle, molten salt (coolant) inlet, and molten salt outlet.

Feed Nozzle,

Optimum duct diameter is given as, $d_{opt} = 226G^{0.5}\rho^{-0.35}$

Flow rate, *G*	=	17.2389 kg/s (ICON)
Density, ρ	=	0.5242 kg/m^3 (ICON)
d_{opt}	=	1176.357 mm

Nozzle thickness, $e = \frac{P_i D_i}{2f - P_i}$

Design pressure, P_i = 210 kPa

Material of construction = Stainless Steel 304

Design stress, f = 1×10^8 N/m^2

Nozzle thickness, e = 1.236 mm

Adding corrosion allowance of 4 mm, thickness of feed nozzle = 5.236 mm

Outlet Product Nozzle,

Flow rate, G = 34.472 kg/s (ICON)

Density, ρ = 1.050 kg/m^3 (ICON)

d_{opt} = 1304.44 mm

Design pressure, P_i = 210 kPa

Material of construction = Stainless Steel 304

Design stress, f = 1×10^8 N/m^2

Nozzle thickness, e = 1.371 m m

Adding corrosion allowance of 4 mm, thickness of output nozzle = 5.371 mm

Molten Salt Inlet Nozzle,

Optimum duct diameter is given as, $d_{opt} = 226G^{0.5}\rho^{-0.35}$

Flow rate, G = 122.43 kg/s (ICON)

Density, ρ = 1975.0 kg/m^3 (ICON)

d_{opt} = 175.632 mm

Nozzle thickness, $e = \frac{P_i D_i}{2f - P_i}$

Design pressure, P_i = 210 kPa

Material of construction = Stainless Steel 304

Design stress, f = 1×10^8 N/m^2

Nozzle thickness, e = 0.184 mm

Adding corrosion allowance of 4 mm, thickness of feed nozzle = 4.184 mm

Molten Salt Outlet Nozzle,

Flow rate, G = 122.43 kg/s (ICON)

Density, ρ = 1975.0 kg/m^3 (ICON)

d_{opt}	=	175.632 mm
Design pressure, P_i	=	210 kPa
Material of construction	=	Stainless Steel 304
Design stress, f	=	1×10^8 N/m^2
Nozzle thickness, e	=	0.184 mm

Adding corrosion allowance of 4 mm, thickness of feed nozzle = 4.184 mm

CHAPTER 4 REACTOR COSTING

For this section, Guthrie's cost correlation is applied to determine the purchase and installation cost of the reactor. The Guthrie's cost approximation corresponds to conservative cost estimate, however it is sufficient for the preliminary design application and it is updated by using a ratio of the Marshall and Swift Indices (M&S).

Based on Douglas (1988), the purchased cost of a reactor is

Purchased Cost, $ = $\left(\frac{M\ \&\ S}{280}\right)\left(101.9D^{1.066}H^{0.82}F_C\right)$

$= (101.9)\ (28.937^{1.066}27.887^{0.82})\ (1)\ (\frac{1433.5}{280})$

= $ 288770.46

Installed Cost, $C_{Inst} = 101.9D^{1.066}H^{0.802}(2.18+F_c) \times (\frac{M\ \&\ S}{280})$

$= 101.9(28.937^{1.066}27.887^{0.802})(2.18+1) \times (\frac{1433.5}{280})$

= $ 864893.5

Total Cost, $C_T = C_p + C_{Inst}$

= $ 288770.46 + $ 864893.53

= $ 1153663.98

* 1$ = RM 3.2 (Exchange Rate)

Thus, the total cost of the reactor is = **RM 3.69 Million**

CHAPTER 5 OPERATING MANUAL (START UP / SHUTDOWN)

5.1 Reactor Pre Start-up Procedure

No.	Procedures
1	Ensure that ample inventory is available at each unit operation. Note that reactor will be last unit operation to start up.
2	Verify Vessel Readiness for start up, i.e., all maintenance and I&E works completed, the reactor is clean and rinse with process water if necessary, man way closed, all blinds are removed and proper gasket are installed.
3	Line up molten salt (MS) into the shell-side of R-1.
4	Line up all transmitters and stroke all control valves.
5	Close, plug and cap all bleeders.
6	Place the reactor temperature indicator (TI) and pressure indicator (PI) in service.
7	Purge reactor with high pressure N_2 until the vent of O_2 is lower than 6%.
8	Pressure up the reactor with high-pressure nitrogen to 400kPa and perform leak check on all flanges.
10	Pressure up the reactor to 800kPa and check flanges for leaks.

5.2 Initiation of Reactor

No.	Procedures
1	Set reactor operating condition at 200 kPa and 410°C.
2	Stop circulation. Allow reactor effluent to pass.
3	Adjust the n-butane feed with air at a ratio of 1.7%
5	The system is stabilized, after air, steam and n-butane feed are heated up to the standard operating condition.

5.3 Hot Hold & Shut-Down of Reactor

No.	Procedures Steps
1	Notify Wastewater Unit, Utilities Unit and Shipping Unit.
2	Reduce the rate of n-butane and air entering reactor to 70% of feed rate.
3	Shut down E-2 by gradually reducing the hot stream flow rate.
4	Reduce reactor feed further to 50% of feed rate, then to 30%.
5	Isolate n-butane feed to the reactor.
6	To **HOT HOLD** the reactor, block all isolation valves, control valves. Manual block valves for molten salt, air and n-butane feed. Verify at field that all isolation and control valves are closed. This is to put reactor on **HOT HOLD**.
7	To **SHUT DOWN** the reactor, block all isolation valves, control valves. Manual block valves for air and n-butane feed at inlet reactor. Verify at field that all shutoff and control valves are closed.
8	Open both man ways of the reactor and inspect the cleanliness inside the reactor. Access the need of washing.
9	Access the condition of catalyst inside reactor. Check if there is any occurrence of coking or crash powder of catalyst.
10	Prepare the reactor for washing with process water if required.
11	If reactor is clean, then proceed with blinding and prepare vessel for maintenance.
12	If reactor is not clean, then close the man ways and perform washing.

5.4 Safety Procedures

When Reactor (R-1) is placed on "Hot Hold", the reactor is isolated in an attempt to maintain reactor pressure and temperature so that the feed stays at optimum temperature. Molten salt is not allowed to cool down and solidify. An electric heater with backup power supply (or generator) may be utilized to keep molten salt temperature from dropping.

Loss of electrical power is the primary reason for placing the reactor on Hot Hold. For power outage of short duration (< 10 minutes), the reactor shall be placed on Hot Hold during power outage.

CHAPTER 6 MINOR EQUIPMENT- HEAT EXCHANGER (E-1)

6.1 Introduction

Heater in the industries is actually just an alias for a heat exchanger. In a heat exchanger two fluids at different temperatures come in indirect contact with one another and exchange heat, it is usually constructed from stainless steel sheet and is divided into two parts; tube side and shell side. The objective of a heater basically is to bring increase the temperature of a stream to a desired level. There are various types of heat exchanger. However the most common are fixed tube, shell and tube and external floating head types.

For Exchanger E-2, we have chosen a shell and tube type heat exchanger due to the following advantages:

1. The configuration provide large heat transfer area in a small volume
2. Provide a good mechanical layout in the form of pressure operation
3. Capable of using various materials for construction
4. Easy maintenance and cleaning

6.2 Process Design

The hot stream for E-2 is S-11 from at 490 ^{0}C. This stream is on the tube side and exits exchanger at 76.9 ^{0}C after giving up heat to the cold stream in shell side. The cold stream is S-9 which enters at 103.4^{0}C and leaves at 410 ^{0}C.

Process simulation by ICON has provides several important parameters as below:

Table 6.1 Inlet and Outlet Streams Data for Shell and Tube

Parameters	**Tube**	**Shell**
Stream	S-11	S-9
Temperature in, ^{o}C	490	103.4
Temperature out, ^{o}C	76.9	410
Thermal conductivity, k (W/m.^{o}C)	0.125	0.138
Heat capacity, C_p (kJ/kg.^{o}C)	429.93	457.54

6.3 Design Method

In designing this heat exchanger, the Kern's method is chosen since this method was based on experimental work on commercial heat exchangers with standard tolerances and will give a reasonably satisfactory prediction of the heat transfer coefficient for standard designs. Although Kern's method does not take account of the bypass and leakage streams, it is simple to apply and is accurate enough for preliminary design calculations and for designs where uncertainty in other design parameters is such that the use of more elaborate methods is not justified. The design methodology is as follow:

1. Define the duty: Heat transfer rate, fluid flow rate, temperatures
2. Collect together the fluid physical properties required: density, viscosity, thermal conductivity.
3. Decide the type of heat exchanger to be used.
4. Select a trial value of overall coefficient, U.
5. Calculate the mean temperature difference, ΔT_m.
6. Calculate the heat exchange area required.
7. Decide the heat exchanger layout.
8. Calculate the individual coefficients.
9. Calculate the overall coefficient and compare with the trial value. If the calculated value differs significantly from the estimated value, substitute the calculated for the estimated value and return to step 6.
10. Calculate the heat exchanger pressure drop; if unsatisfactory return to step 7 or 4 or 3, in that order of preference.
11. Optimize the design: repeat step 4 to 10, as necessary, to determine the cheapest exchanger that will satisfy the duty. Usually this will be the one with the smallest area.

6.4 Exchanger Sizing

The table below provides the for exchanger sizing

Table 6.2: Data for Exchanger Sizing

Data	Description
Heat exchanger type	Shell and Tube
Heat exchanger orientation	Horizontal
Tube inlet direction	Horizontal
Heat duty (W)	147036.7252
Overall coefficient $W/m^2\,^oC$	200.00

To start sizing for the heat exchanger, a temperature correcting factor must be identified first in order to determine the true temperature difference, ΔT_m. So, these two dimensionless temperature ratios, *R* and *S* have to be computed first where:

$S = (t_2-t_1) / (T_1-T_2)$ $\quad\quad$ $R = (T_1-T_2) / (t_2-t_1)$

$=(410-103.4) / (490 – 76.9)$ $\quad\quad$ $= 1.3$

$= 0.74$

Based on these data, the temperature correcting factor (F_t) is obtained by using this chart:

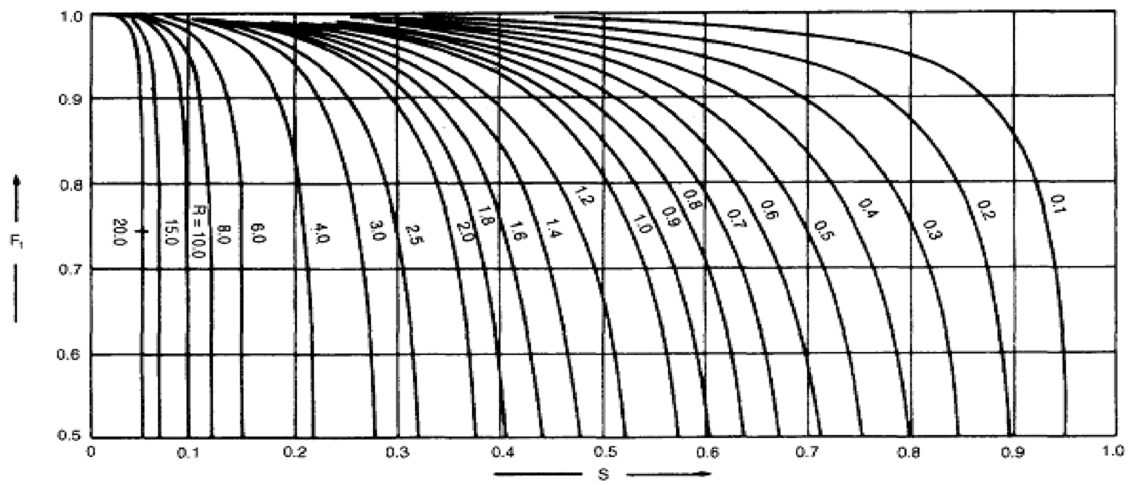


Figure 6.1: Temperature correction factor

Based on these chart, at R = 1.3 and S = 0.74, the corresponding correction factor, Ft = 0.72. So, the actual temperature difference is:

$\Delta T_{lm} = [(T_1 – t_2) – (T_2 – t_1)] / \ln [(T_1 – t_2) – (T_2 – t_1)]$

$= [(490 – 410) – (76.9 – 103.4)] / \ln [(490 – 410) – (76.9 – 103.4) = 22.8\ ^oC$

So, the actual temperature difference is:

$F_t\Delta T_{lm} = 22.8\ ^oC \times 0.72$

$= 16.4\ ^oC$

Hence, the area of heat exchanger is:

$Q = UA \Delta T_{lm}$

$A = Q / U\Delta T_{lm} = 147036.72 / (350)(16.4) =25.6\ m^2$

Tube Rating,

From the book of transport unit operation and unit by Christie J. Geankoplis, the material suitable for a heat exchanger is carbon steel and the suitable inner and outer diameter of the tube are as follow:

Table 6.3: Material Data

Data	**Details**
Material	Carbon Steel
Length of tube L_t (m)	2.5
Outer Diameter, D_{to} (mm)	31.75
Inner diameter, D_{ti} (mm)	27.53

Based on these data, the corresponding heat transfer area for the tube, A_t:

$A_t = L_t \pi D_t$ = (2.5)(3.1416)(31.75E-3 = $0.25m^2$

So, number of tube, N_t:

$N_t = A / A_t$

$= 25.6 / 0.25$

$= 102.3 \approx 103$ tubes

The tube pitch, P_t:

$P_t = 1.25D_{to}$

$= 1.25 \times 31.75$

$= 39.6875$mm

Next, to calculate the bundle diameter, D_b:

$D_b = D_{to}(N_t/K)^{1/n}$

$= (31.75)(103/ 0.249)^{1/2.207}$

$= 486.8$mm $= 0.487$m

In order to get the shell bundle clearance and shell internal diameter, the value of bundle diameter for fixed and U tube was taken from the following graph:

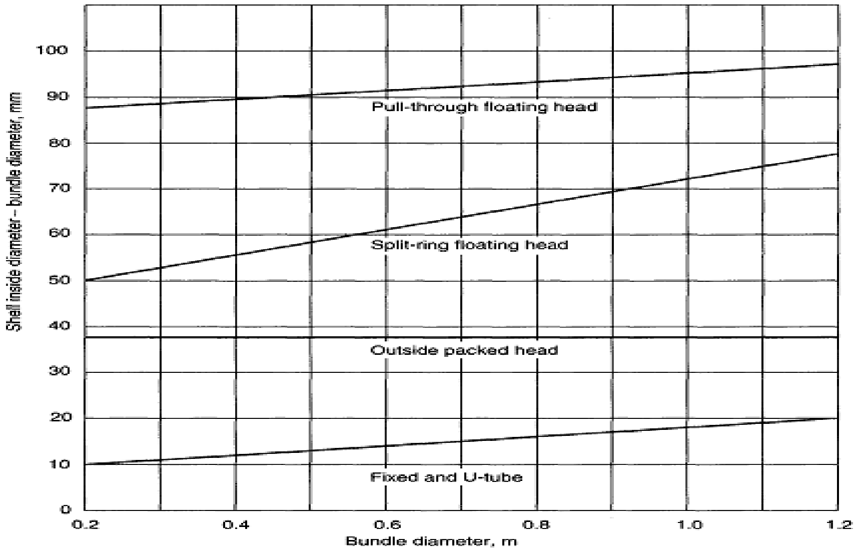


Figure 6.2: Shell inside diameter

Based on the graph, at bundle diameter of 0.487m, the corresponding shell bundle clearance is 11.25mm equal to 0.01125m. So, the inner diameter for the shell side,

Shell side diameter, Ds = Db + Shell bundle clearance

= 0.487m + 0.01125 = 0.498m

Tube Side Coefficient,

The mean temperature, $T_{mean:}$

T_{mean} = ($T_{c\ in}$ + $T_{c\ out}$) / 2

= (763 + 349.9) / 2

= 556.45K

Tube cross-sectional area, A_t = $\pi D_{ti}^2 / 4$

= $(3.1416)(27.53)^2 / 4$

= $595.26mm^2 = 0.59526m^2$

Tube per pass = N_t /2 = 103 /2 =52 tubes

Total flow area, A_t = N_tA_t

= (103) (0.59526)

= $61.3\ m^2$

Based on simulation, the mass flow rate inside the tube, m =20128.85 kg/h

So, fluid velocity, V_f = m/A_t

= (20128.85 kg/hr) / ($61.3m^2$)

= 328.4 kg/m^2hr

The linear velocity, μ = V_f / ρ

= 328.4 / 997.10 = 0.33 m/s

The Reynolds number, Re = $\rho \mu D_{ti} / \eta$

= (997.10)(0.33)(27.53) / (0.798)

= 11351.6

The Prandtl number, Pr = $C_p \eta / K_f$ = 5.43

L/D_{ti} = 2.5m / 0.02773m = 90.16

Now, to identify the heat transfer factor this chart is use:

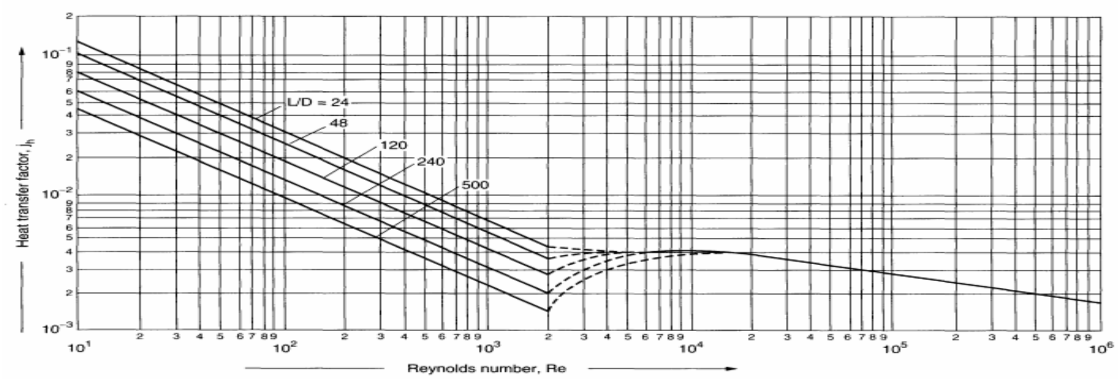


Figure 6.3: Heat transfer coefficient

Based on the graph, the heat transfer factor, $j_h = 4.2x10^{-3}$. So, the tube side heat transfer coefficient, $h_i = [K_f j_h Re Pr (0.33) / D_{ti}](\eta/\eta_w)^{0.1}$ = 604.7229 W/m^2K

Tube Side Pressure Drop

In order to find the tube side pressure drop, a friction factor must be obtained first. The friction factor can be found using this chart:

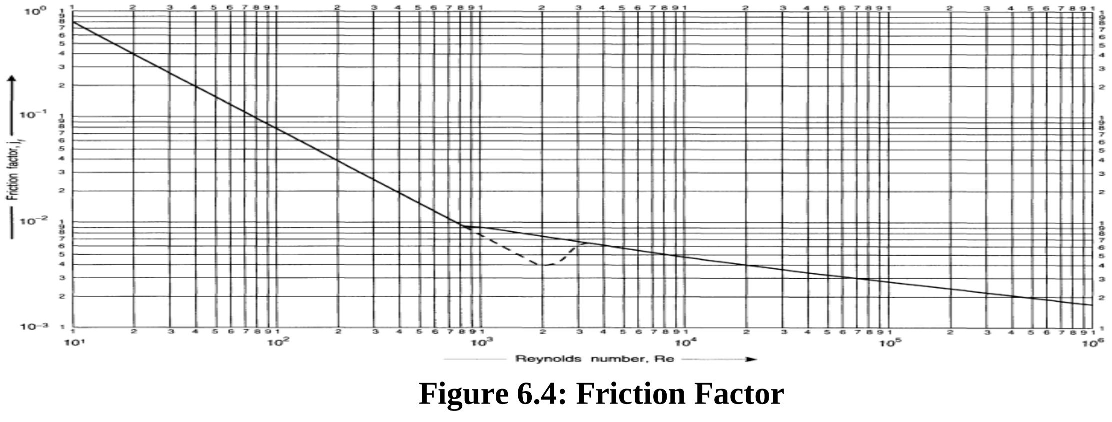


Figure 6.4: Friction Factor

Based on this chart, at Reynolds number of 4.8643E-4, the corresponding friction factor, j_f = 5.61E-3. So, the tube side pressure drop is:

$\Delta P_s = N_p[8\ j_f\ (L/D_{ti})(\mu/\mu_w)^{-m} + 2.5\ ]\ [\rho u^2/2]$ = 121.956Pa = 0.121956kPa

Shell Side Heat Transfer Coefficient,

Based on the previous calculation, the diameter for the shell side, Ds is 0.498m. So, the corresponding baffle spacing :

Baffle Spacing, l_B $= 0.5D_s$

$= 0.5 \times 0.498 = 0.249$m

The Baffler Diameter $= D_s - 0.0016$

$= 0.498 - 0.0016 = 0.4964$m

Tube pitch, $P_t = 1.75D_o = 0.0555625$m

So, the cross flow area, $A_s = (p_t - D_{to})\ D_s\ L_b\ /\ p_t = 0.00155\text{m}^2$

Based on the area obtained, the shell mass velocity, $G_s = w_s\ /\ A_s$

= 4.6828 / 0.00155

= 3021.1613 kg/m^{2}hr

So, the shell side equivalent diameter, $D_e = (1.10/D_{to})(P_t^2 - 0.917D_{to}^2)$ = 0.07308m

The Reynolds number, Re $= G_sD_e\ /\ \mu$

= (3021.1613)(0.07308) / (0.00089929)

= 245511.9792

The Prandtl number, $Pr = C_p\mu\ /\ K_f$ = (429.938)(0.00089929) / (0.1253) = 3085.67

In order to identify the shell side heat transfer coefficient, h_s this chart is use:

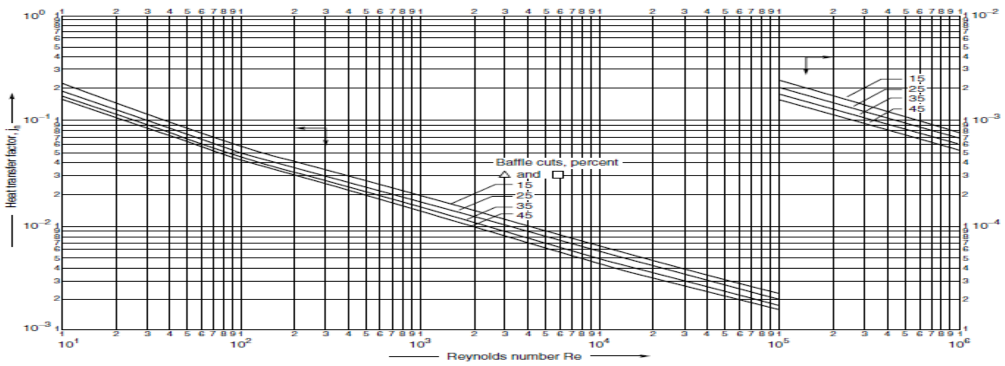


Figure 6.5: Shell side heat transfer coefficient

So, the corresponding heat transfer factor j_h is 0.00137. Based on that, the heat transfer coefficient, $h_s = [K_f\, j_h\, Re\, Pr^{1/3} / De][\mu/\mu_w]^{0.1}$ = 3245.85 W/m^2.k

Shell Side Pressure Drop,

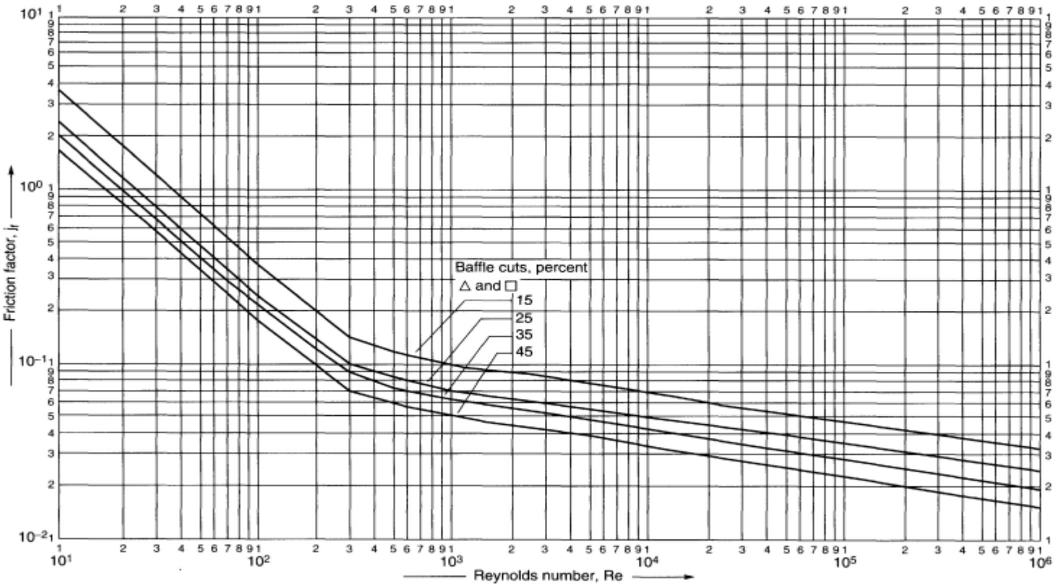


Figure 6.6: Shell side friction factor

For the shell side, the linear velocity, U_s $= G_s / \rho$

= (3021.1613) / (11.8013) = 256.00 m/hr

= 0.0711m/s

Based on the graph provided, at Reynolds number of 245511.9792, the friction factor, j_f value is 3.13E-2. So, the pressure drop at the shell side is:

Shell side pressure drop, $\Delta P_s = 8\, j_f\, (D_s/D_e)(L/L_B)(\rho u_s^2/2)(\mu/\mu_w)^{-0.14}$ = 1.2855kPa

Overall Coefficient

The calculation that has been done so far covers for the outside fluid film coefficient, inside fluid film coefficient, tube inside diameter, tube outside diameter and thermal conductivity of the tube wall material. However, to enable the calculation of the overall heat transfer, two more parameters are needed which are the outside dirt coefficient and inside dirt coefficient where this two parameter's value can be obtain from a heat exchanger design book. Based on the book, the value of outside dirt coefficient and inside dirt coefficient are:

1. Outside dirt coefficient (fouling factor), h_{od} = 4500 W/m$^2.^o$C

2. Inside dirt coefficient, h_{id} = 5000 W/m$^2.^o$C

So, the overall heat transfer coefficient:

$$\frac{1}{U_o} = \frac{1}{h_s} + \frac{1}{h_{od}} + \frac{d_{to}\ln(d_{to}/d_{ti})}{2k_w} + \frac{d_{to}}{d_{ti}} \times \frac{1}{h_{id}} + \frac{d_{to}}{d_{ti}} \times \frac{1}{h_i}$$

$(1/U_o) = (1/h_s) + (1/h_{od}) + [D_{to} \ln (D_{to}/D_{ti}) / 2(K_w)] + (D_{to}/D_{ti}) (1/h_{id}) + (D_{to}/D_{ti})(1/h_i)$

= (1/3245.85) + (1/4500) + [(0.03175) ln (0.03175/0.02753)]/2(36) +

(0.03175/0.02753)(1/5000) + (0.03175/0.02753)(1/604.7229)

= 0.002731

Hence, the overall heat transfer coefficient, U_o = 1/0.002731 = **366.17 W/m.^{0}C**

CHAPTER 7 MINOR EQUIPMENT – MIXER (M-1)

7.1 Introduction

Mixer in general is a vessel where two or more substances are mixed together forming a homogeneous stream. A simple example that demonstrates the principles behind a mixer is pumping of the water in a swimming pool to homogenize the water temperature, and the stirring of pancake batter to eliminate lumps. In this design we mix n- butane and air to form a mixture that can be sent into the reactor. The type of mixer selected is Magnetic Drive Mixer. This mixer is well suited for continuous processes and has the following advantages:

1. Simple design. The compact mixing head is the only moving part inside the mixing vessel, offering reliable maintenance-free operation.
2. Easy to use and can be cleaned and sterilized in-place.
3. Transferable drive unit. The external drive unit is attached to the mixing vessel pads and mixing heads.

7.2 Mixer Sizing

From the process simulation performed by ICON, the properties of mixer are included in the appendix as table 6.1.

To calculate the mixer volume, the mixing time must be identified first. Note that the two stream that being mix using this mixer has the same physical properties since it's basically the same fluid and due to that, the mixing time appropriate for this mixer is 1min.

So**,** the volume of the mixer is:

Mixer Volume = Volumetric Flow Rate Out X Mixing Time

$= 1705.65 m^3/hr \times (60/3600)\ hr$

$= 28.43 m^3$

To attain the mixer's height and diameter, an assumption is made where the mixer is assumed to be in cylindrical shape mixer. Note that the rule of thumb in sizing a mixer, the ratio of diameter to height is 1:1.6.

Height (H)/ Diameter (D) = 1.16

Height, H = 1.16 D

For a cylindrical object, the volume can be calculated using this equation:

Volume, $V = \pi (D/2)^2 H$

$V = 3.1416)(0.5D)^2(1.16D)$

$V = 0.9111D^3 = 28.43m^3$

So, Diameter (D) = $(28.43\ m^3 / 0.9111)^{1/3}$

D = 3.148 m

Hence Height (H) = 1.16D

H = 1.16(3.148m)

H = 3.652m

In terms of the design temperature and pressure, an additional 10% safety factor was added. In other words, this mixer is actually overdesign by 10% of its actual value where this 10% can ensure nothing disastrous happened in case an incident or emergency.

So, the mixer can be overflow or use just about 10% of its actual capacity before its damage. So:

Design Pressure = (Operating Pressure x 10%) + Operating Pressure

Design Pressure = (101.325 kPa x10%) + 101.325 kPa

Design Pressure = 111.457 kPa

Design Temperature = (Operating Temperature x 10%) + Operating Temperature

Design Temperature = (214.15^{O}C x 10%) + 214.15^{O}C

Design Temperature = 235.56 ^{O}C

REACTOR DATA SHEET		
	Equipment No.(Tag)	R-1
	Description	Maleic Anhydride Reactor
	Sheet no.	1/1

OPERATING DATA

NO. REQUIRED	1	**ORIENTATION**	Vertical
TYPE	Multitubular Catalytic Fixed Bed	**JACKETED**	Yes

	SHELL	**TUBE**
CONTENTS	Molten Salt	n-butane, oxygen, VPO catalyst
DIAMETER (OUTER)	9.22 m	0.06033 m
LENGTH	10.2 m	6.1 m
DESIGN CODE	BS 5500	BS 5500
MAX. WORKING PRESSURE	200 kPa	500 kPa
DESIGN PRESSURE	210 kPa	525 kPa
PRESSURE DROP (ALLOWED/CALC)	0.0325 kPa	4.386 kPa
MAX. WORKING TEMP	410 ºC	410 ºC
DESIGN TEMPERATURE	430.5 ºC	430.5 ºC
VELOCITY	122.43 kg/s	34.472 kg/s
NO. OF PASSES	1	1
HEAT EXCHANGED	1.846×10^8 kJ/hr	-1.846×10^8 kJ/hr

MECHANICAL DESIGN OF SHELL

MATERIAL	Stainless Steel 304 (18Cr/8Ni)
JOINT FACTOR	1.0
CORROSION ALLOWANCE	4 mm
THICKNESS	16 mm
TYPE OF SUPPORT	Straight Cylindrical Skirt

NO. OF BOLTS	356	**DIAMETER**	16 mm	**MATERIAL**	5% Cr Mo Steel
END TYPE (TOP)	Torispherical	**THICKNESS**	19.28 mm	**MATERIAL**	Stainless Steel
END TYPE (BOTTOM)	Torispherical	**THICKNESS**	19.28 mm	**MATERIAL**	Stainless Steel
INSULATION		**THICKNESS**	100 mm	**MATERIAL**	Mineral Wool

GASKET	**WIDTH**	20 mm	**MATERIAL**	Asbestos with corrugated SS
REACTOR NOZZLE	**IN DIAMETER**	5.236 mm	**OUT DIAMETER**	5.371 mm
MOLTEN SALT NOZZLE	**IN DIAMETER**	4.184 mm	**OUT DIAMETER**	4.184 mm
NO. OF MAN WAYS		2	**DIAMETER**	600 mm

MECHANICAL DESIGN OF TUBES

MATERIAL	Stainless Steel 304 (18Cr/8Ni)
No. OF TUBES	14908
NOMINAL SIZE	0.0508 m (2 in)
OD	0.06033 m
ID	0.05250 m
THICKNESS	0.00391 m
TUBE PITCH	0.0754 m (triangular)

CATALYST

TYPE	Vanadium Phosphorus Oxide (VPO) [$(VO)_2P_2O_7$]
WEIGHT	177102 kg
SHAPE	Cylindrical
BULK DENSITY	1000 kg/m^3
DIAMETER	5 mm
POROSITY	0.5
LIFE SPAN	5 years
REMARKS AND NOTES :-	- The first layer on top of the tubes is diluent material, ceramic balls
	- Ceramic balls material : Silicon dioxide, clay-based material
	- Man ways are allocated for inspection of reactor, loading and unloading of catalyst.

HEAT EXCHANGER DATA SHEET		Equipment No.		E-2
		Description		Heater
		Sheet No.		1
OPERATING DATA				
Type	Shell and tube	Amount		1
Delta T (C)	16.4	Enthalpy, Q(W)		147036.7252
MECHANICAL DESIGN				
Material	Carbon Steel	Thermal Conductivity		200 $W/m^2.^oC$
Outer Diameter	31.75 mm	Tubes		103
Pitch	39.69 mm	Thickness		2.6mm
Inner Diameter	27.53 mm	Length		2.5 m
THERMODYNAMIC DATA				
Outside Fluid Film Coefficient	3245.85 $W/m^2.^oC$			
Inside Fluid Film Coefficient	604.7229 W/m2.oC			
Outside Dirt Coefficient	4500 W/m2.Oc			
Inside Dirt Coefficient	5000 W/m2.oC			
	SHELL SIDE		TUBE SIDE	
Heat Capacity	4.57.54 $kJ/kg.^oC$		429.938 $kJ/kg.^oC$	
Thermal Conductivity	0.138 $W/m.^oC$		0.125 $W/m.^oC$	
	IN	OUT	IN	OUT
Temperature °C	103.4	410	490	76.9

MIXER DATA SHEET	Equipment No.	M-1
	Description	Mixer
	Sheet No.	1
OPERATING DATA		
Size Of Charge	1705.65 m^3/h	
Time Actually Mixing	600 s	
Type Of Mixing	Severe	
Fluid Viscosity	8.9929×10^{-4} Pa-s	
VESSEL DATA		
Diameter Of Vessel	6.78 m	
Depth Of Vessel	7.86 m	
Depth Of Liquid	6.12 m	
Angle Of Agitator	120.00^o	
TECHNICAL DATA		
Type Of Mixer	Magnetic Drive Mixer	
No Of Blade	3	
Operating Pressure	101.325 kPa	
Design Pressure	111.457 kPa	
Operating Temperature	214.15 ^oC	
Design Temperature	235.56 ^oC	

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APPENDIX

MIXER M-1			
Stream	In (1)	In (2)	Out
VapFrac	0.32	1	1
T [C]	31	121	103
P [kPa]	101.325	101.325	101.325
MoleFlow [kgmole/h]	84.55	38.01	122.56
MassFlow [kg/h]	13886.09	6242.76	20128.85
VolumeFlow [m3/hr]	1199.819	504.436	1705.651
StdLiqVolumeFlow [m3/hr]	10.725	4.821	15.546
StdGasVolumeFlow [SCMD]	4.8072E+4	2.1612E+4	6.9684E+4
Energy [W]	1068900.433	410815.7424	1479716.1754
H [kJ/kmol]	45512.02	38908.04	43463.86
S [kJ/kmol-K]	392.688	375.703	387.485
MolecularWeight	164.235	164.235	164.235
MassDensity [kg/m3]	11.5735	12.3757	11.8013
Cp [kJ/kmol-K]	432.527	424.188	429.938
ThermalConductivity [W/m-K]	0.1243	0.1274	0.1253
Viscosity [Pa-s]	8.3889E-4	1.0529E-3	8.9929E-4
molarV [m3/kmol]	14.191	13.271	13.917
ZFactor	0.3549	0.3418	0.3512
Cv [kJ/kmol-K]	424.213	415.873	421.624
KinematicViscosity [m2/s]	7.30E-06	7.76E-06	7.45E-06