SLURRY HYDROCRACKER PROJECT

Appendix E - Detailed Equipment Sizing

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Table of contents

E.1 Summary

This appendix contains detailed sizing of major equipment used in the process. Section E2 shows the type, material, specifications, as well as the operating conditions of each of the equipment. Section E3 provides the line sizing table with the pipe diameter, material, and pressure drop of major streams.

E.2 Equipment Sizing

Table E1. Equipment sizing specifications and operating conditions.

Equipment Number	Equipment Description	Equipment Sub-Type	Material	Capacity / Size / Duty Specifications	Temperature (°C)	Pressure (bar)	Mass Flow Rate (tonnes/hr)	Number of Trains	Number of Equipment per Train	Total Number of Equipment
C-01	Waste Heat Boiler	Floating Head	CS/SS	A = 924.5 m^2	380.0	168.0	646.0	2	4	8
C-02	Cooler	Air fin	Stainless Steel	A = 1827 m^2	455.3	79.0	13.0	2	1	2
C-03	Cooler	Air fin	Stainless Steel	A = 84 m^2	319.0	1.0	9.3.0	1	1	1
C-04	Cooler	Air fin	Stainless Steel	A = 1638 m^2	173.0	20.0	630.0	1	1	1
CP-01	Compressor	Centrifugal	Stainless Steel	w_s = 480 kW	70.0	167.0	140.0	1	1	1
CP-02	Compressor	Centrifugal	Stainless Steel	w_s = 9,581 kW	76.0	0.9	33.0	1	1	1
CT-C01	Cooler	Air fin	Carbon Steel	A = 161 m^2	245.3	60.0	40.0	1	1	1
CT-CP01	Compressor	Centrifugal	Stainless Steel	w_s = 26,777 kW	80.0	21.0	40.0	1	1	1
CT-CP02	Compressor	Centrifugal	Stainless Steel	w_s = 27,957 kW	80.0	59.0	40.0	1	1	1
D-01	Flash Drum	Vertical	Stainless Clad	D = 4 m L = 20 m	450.0	169.0	211.0	2	2	4
D-02	Flash Drum	Vertical	Stainless Clad	D = 4 m L = 20 m	380.0	168.0	208.0	2	2	4
D-03	3 Phase Separator	Horizontal	Stainless Clad	D = 3 m L = 7.08 m	70.0	167.0	646.0	1	1	1
D-04	Flash Drum	Vertical	Stainless Clad	D = 1.5 m L = 7.5 m	350.0	78.0	12.7.0	1	1	1
D-05	Flash Drum	Vertical	Stainless Clad	D = 4 m L = 20 m	80.0	20.0	156.0	1	4	4

Equipment Number	Equipment Description	Equipment Sub-Type	Material	Capacity / Size / Duty Specifications	Temperature (°C)	Pressure (bar)	Mass Flow Rate (tonnes/hr)	Number of Trains	Number of Equipment per Train	Total Number of Equipment
D-06	Flash Drum	Vertical	Stainless Clad	D = 4 m L = 20 m	78.2	4.0	305.0	1	2	2
D-07	3 Phase Separator	Horizontal	Stainless Clad	D = 1 m L = 1.1 m	80.0	0.9	93.0	1	1	1
E-01	Heat Exchanger	Floating Head	CS/SS	A = 895 m^2	250	171.0	665.5	2	1	2
E-02	Heat Exchanger	U Tube	CS/SS	A = 30 m^2	72.7	171.0	140.0	2	1	2
EX-01	Gas Expander	Axial	Stainless Steel	w_s = 2640 kW	70.0	167.0	60.0	1	1	1
F-01	Heater	Fired Heater	Carbon Steel	Q = 31627 kW	250.0	171.0	665.5	2	1	2
F-02	Heater	Fired Heater	Carbon Steel	Q = 56431 kW	72.7	171.0	140.0	2	1	2
P-01A	Pump	Centrifugal	Cast Steel	w_s = 769 kW Q = 0.104 m^3/s	80.0	5.0	333.0	2	1	2
P-01B	Pump	Centrifugal	Cast Steel	w_s = 769 kW Q = 0.104 m^3/s	80.0	61.0	333.0	2	1	2
P-01C	Pump	Centrifugal	Cast Steel	w_s = 769 kW Q = 0.104 m^3/s	80.0	116.0	333.0	2	1	2
P-02	Pump	Centrifugal	Cast Steel	w_s = 0.85 kW Q = 0.0016 m^3/s	80.0	0.9	47.0	1	1	1
R-01	Main Reactor	Vertical	Stainless Steel	D = 4 m L = 30.8 m	466.9	170.0	106.0	2	4	8
R-05, R-06	Conversion Reactor	Vertical	Stainless Steel	D = 4 m L = 19.7 m	119.0	55.0	93.0	1	2	1
T-02	Stripping Column	Vertical	Stainless Steel	D = 1.5 m L = 4.04 m	350.0	78.0	12.6	1	1	1

E.2.1 Two-Phase Separator Sizing Sample Calculations

Two - Phase Separators Sizing

ex) Flash Drum D-01

According to S13 & S14 of the stream table:

$T = 450°C$, $P = 169$ bar

Total flow rates: $Q_L = 16.323\ m^3/h$, $Q_G = 12005.798\ m^3/h$

$Q_G > Q_L$ → Gas phase is the continuous phase.
Liquid phase is the dispersed phase.

$\rho_d = 622.690\ kg/m^3$, $\rho_c = 69.388\ kg/m^3$

After being splitted into two trains and two vessels per train:

$$Q_d = \frac{16.323\ m^3/h}{4} \frac{h}{3600\,s} = 0.83374\ m^3/s$$

$$Q_c = \frac{12005.798\ m^3/h}{4} \frac{h}{3600\,s} = 0.00\ 1134\ m^3/s$$

Assuming a droplet size of 100 µm, at P = 169 bar

$k_s = 0.025\ m/s$ (Campbell, 2014)

$$V_{G,max} = k_s \sqrt{\frac{\rho_d - \rho_c}{\rho_c}} \quad \text{(Campbell. 2014)}$$

$$V_{G.max} = (0.025\,m/s)\sqrt{\frac{622.690 - 69.388}{69.388}}$$

$$V_{G,max} = 0.0706\ m/s$$

$$D_{min} = \sqrt{\frac{(4/\pi)\,Q_c}{F_G\,V_{Gmax}}} \quad \text{(Campbell, 2014)}$$

$F_G = 1$ for vertical vessels. (Campbell, 2014)

$$D_{min} = \sqrt{\frac{(4/\pi)(0.001134\ m^3/s)}{(1)(0.0706\ m/s)}} = 3.8778\ m$$

$$\Delta P = P - P_{atm} = 169\ bar - 1\ bar = 168\ bar$$

According to Ulrich & Vasudevan, (2004), Table 4.24:

$\frac{L}{D} = 5.0$ for vessels with an internal pressure > 35 bar.

$$L_{min} = (5.0)(D_{min}) = (5.0)(3.8778 m) = 19.3888\ m$$

Vessels must have a liquid hold-up time of at least 20 mins. (Rehm et. al, 2012)

$$V_{hold\text{-}up} = (Q_d)(20\ mins)\left(\frac{min}{60s}\right)$$

$$= (0.83374\ m^3/s)(20\ mins)\left(\frac{min}{60s}\right)$$

$$= 1.360\ m^3$$

$$V_{min} = \frac{\pi}{4} D_{min}^2 L_{min} + V_{hold\text{-}up}$$

$$= \left(\frac{\pi}{4}\right)(3.8778\ m)^2(19.3888\ m) + 1.360\ m^3$$

$$= 230.342\ m^3$$

To satisfy all the limits D_{min}, L_{min}, V_{min} & $\frac{L}{D}$, the actual size of the vessel is determined to be:

$D = 4\ m$. $L = 20\ m$.

The rest of the two phase separators are sized in the same manner.

E.2.2 Pump Sizing Sample Calculations

Pump Sizing

ex) Pump P-02 (centrifugal)

According to S49 & S50 of the stream table:

$T = 80\,°C$, $P_s = 90$ kPa, $P_d = 490$ kPa

$$Q = (5.715\ m^3/h)\left(\frac{h}{3600\,s}\right) = 1.5875 \times 10^{-3}\ m^3/s$$

$$\Delta P = P_d - P_s = 490\ kPa - 90\ kPa = 40\ kPa$$

Centrifugal pumps typically have an efficiency

$\varepsilon_i = 50 - 85\ \%$. (Ulrich & Vasudevan, 2004)

Assuming $\varepsilon_i = 0.75$,

$$\dot{W}_s = \varepsilon_i \frac{Q}{\Delta P} \quad \text{(Ulrich \& Vasudevan, 2004)}$$

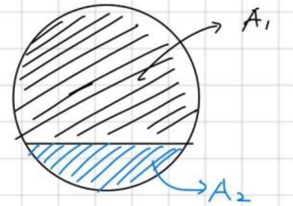
$$= (0.75)\left(\frac{1.5875 \times 10^{-3}\ m^3/s}{40,000\ Pa}\right)$$

$$= 846.7\ W$$

The rest of the pumps are sized in the same manner

E.2.3 Three-Phase Separator Sample Calculations

3 phase separator sizing



- the black shaded area is A_1, it represents the cross-sectional area occupied by gas phase
- the blue shaded area is A_2, it represents the cross-sectional area occupied by the two liquid phase.

- for a gas-liquid phase separation, it requires more space than liquid-liquid separation. Thus the sizing will be done based on gas-liquid separation.
- Assume 5 min = 300 s of retention time is needed for both separation and operation requirements.
- If Q represents the volumetric flow rate in m^3/s. Then the ratio beteen A_1 and A_2 will depends on the ratio between Q_{gas} and Q_{liquid}.
- We choose a diameter of $D_v = 3m$ for this separator, under overall consideration to the scale of this plant.
- Thus $A = A_1 + A_2 = \frac{\pi}{4} D_v^2 = \frac{\pi}{4} 3^2 = 7.068583\ m^2$

$$A_2 = A\left(\frac{Q_L}{Q_G+Q_L}\right)$$

- where from simulation:

For oil phase, $\left.\begin{array}{l} \dot{m}_{oil} = 442967.15\ kg/h \\ \rho_{oil} = 665.9887\ kg/m^3 \end{array}\right\}$ $Q_{oil} = \frac{\dot{m}_{oil}}{\rho_{oil}} = 665.13\ m^3/h$

For water phase, $\left.\begin{array}{l} \dot{m}_{H_2O} = 2608.63\ kg/h \\ \rho_{H_2O} = 938.1176\ kg/m^3 \end{array}\right\}$ $Q_{H_2O} = 2.78\ m^3/h$

For gas phase $\left.\begin{array}{l} \dot{m}_G = 200074.56\ kg/h \\ \rho_G = 42.8839\ kg/m^3 \end{array}\right\}$ $Q_G = 4665.5\ m^3/h$

$$\Rightarrow Q_L = Q_{oil} + Q_{H_2O} = 665.13\ m^3/h + 2.78\ m^3/h$$
$$= 667.91\ m^3/h$$
$$Q_G = 4665.5\ m^3/h$$

$- A_2 = A\left(\frac{Q_L}{Q_G+Q_L}\right) = 7.068583\ m^2 \times \left(\frac{667.90}{667.90+4665.50}\right)$

$\Rightarrow A_2 = 0.885207\ m^2$

- The flow velocity is

$$V_2 = \frac{Q_L}{A_2} = \frac{667.90\ m^3/h}{0.885207\ m^2} \times \frac{1h}{3600s} = 0.026247\ m/s$$

- In order to have a retention time for 300 s, that is, the liquid has to flow in the separator for 300s.

The length of the separator is then

$$L = V_2 \times t_R = 0.026247\ m/s \times 300\ s = 7.874\ m$$

- Finally the $\frac{L}{D_V}$ ratio is thereby:

$$\frac{L}{D_V} = \frac{7.874m}{3m} = 2.62$$

which is a normal number.

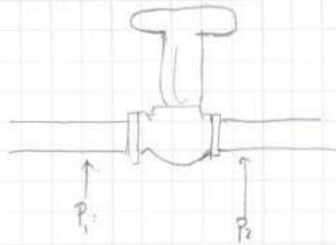
The sizing for another 3-phase separator is done in the same manner.

E.2.4 Control Valve Sizing Sample Calculations

Control Valve sizing

Stream 21

$T = 70°C$
$Q = 665 m^3/h$
$SG = 0.665$



Assuming the maximum allowable pressure drop is 2000 kPa across the valve

$\Delta P = 2000 kPa$

The valve characteristic C_v is described as:

$$C_v = 11.6\, Q \sqrt{\frac{SG}{\Delta P}}$$

$$C_v = 139.1$$

Based on valve selection guidelines[5], linear opening valves is appropriate with level control with increasing pressure drop as load increases

For proper design, the C_v at normal flow should be at 35% to 45% opening

From catalogue 12 of Fisher Control Valves Page ED-4 (June 2014)

The most appropriate valve has

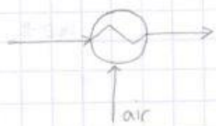
Valve NPS = 6 in ED valve with linear cage
Trim = 7 in port

% travel	C_v
10	46.3
20	107
30	171
40	228
50	279
60	327
70	367
80	402
90	420
100	433

E.2.5 Heat Exchanger Sizing Sample Calculations

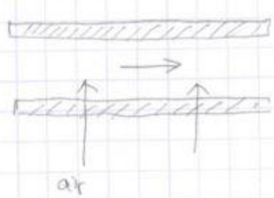
1.) Detailed Sizing: Heat Exchanger CA-02

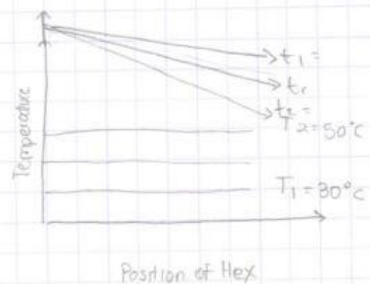
Step 1 Specifications



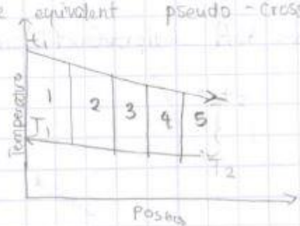
$Q_{TOTAL} = \sum Q_n$, n= Zone #

Cross Flow Exchanger





The equivalent pseudo-cross-flow temperature-position diagram is shown below



Were $t_1 = 30°C$, $t_2 = 50°C$ } assumptions

$T_1 = 240°C$

$T_2 = 70°C$

The overall mean temperature difference can be calculated using the equation

$$LMTD_i = \frac{\Delta T_i - \Delta T_{i+1}}{\ln\left[\frac{\Delta T_i}{\Delta T_{i+1}}\right]} = 97.20°C$$

$$CMTD = LMTD \times F$$

According to Hudson Manufacturer's Specifications for MTD correction (reference ...)

$$R = \frac{T_1 - T_2}{t_2 - t_1} = \frac{240°C - 70°C}{50°C - 20°C} = 5.66 \qquad r = \frac{t_2 - t_1}{T_1 - t_1} = \frac{50 - 30°}{240 - 30}°C$$

for a 1 pass cross flow heat exchanger

$F \approx 0.95$

for a 2 pass cross flow

$F \approx 1$

$\therefore\ CMTD_T = LMTD \times (F)$
$= LMTD$
$= 97.20°C$

Overall heat transfer coefficient

From Table 4.15b from Ulrich and Vasudevan (2004)

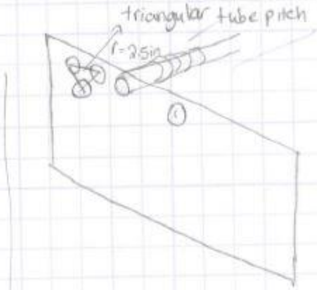
$U_{air/medium\ hydrocarbons\ steam} = 310\ J/m^2 \cdot s \cdot K$

Heat Transfer Area

$$A = \frac{Q}{U_A\ LMTD} = \frac{(1.10 \times 10^6\ W)}{\left(\frac{310\ J}{m^2 s\ K}\right)(97.20°C)} = 36.50\ .6\ m^2$$

Finned Tube Dimensions

The selected tube sizing is



To find # of tube rows calculate

$$Z = \left[\frac{T_1 - T_2}{T_1 - t_1}\right] = \frac{240-70}{240-30} = \frac{170}{210} = 0.8091$$

From Table 1 in Hudson (2...

# of rows = $\boxed{8\ rows}$
with face velocity = 400 ft/min
= 2.03 m/s air velocity

↳ tranversable pitch of 2.5 in = 0.0635m

n (number of tubes per row) $= \frac{12}{2.5} \approx 5$ tubes per row

a (area per ft of tube) $= \frac{\pi}{12} \cdot OD = 0.2618\ ft^2/ft = 0.08\ \frac{m^2}{m}$

The heat transfer rate of of fluid is

$$C_f = \frac{2.75 \times 10^7\ W}{240 - 70} = 161\ 764\ \frac{W}{K}$$

$$C_{air} = \frac{2.75 \times 10^7\ W}{50 - 30} = 1\ 375\ 000\ \frac{W}{K}$$

To size the ACHE the values of R and k must be calculated to refer to Figure (12) (Hudson's)

$$R = \frac{C_{min}}{C_{max}} = \frac{C_{hot}}{C_{air}} = \frac{161\ 764 W/K}{1\ 375\ 000} = 0.118$$

$$NTU = \frac{n \cdot N \cdot a \cdot W \cdot L}{\left[\frac{Q}{T_1 - T_2}\right] \cdot U}$$

since $R = \dfrac{Q}{FV \cdot L \cdot W \cdot 1.08 (T_1 - T_2)}$ (Hudson

$$k = R \cdot NTU = \frac{n \cdot N \cdot a}{1.08 \cdot FV \cdot (1/U)} =$$

n = 5 tubes/row
N = 8 tubes rows
a = 0.2618 ft^2/ft tube
FV = 400 scfm
U = 54.5 BTU = 310 W/m^2k

$k = R\ NTU = 1.320$

$$Z = \frac{T_1 - T_2}{T_1 - t_1} = 0.77$$

Reading from Figure 13 for 2 pass cross flow ACHE

$R = 0.5$

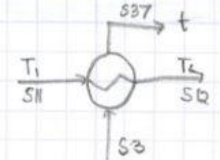
The face Area FA is

$$FA = \frac{R \cdot Q}{(T_1 - T_2) 1.08 \cdot FV} = \boxed{187 ft^2}$$

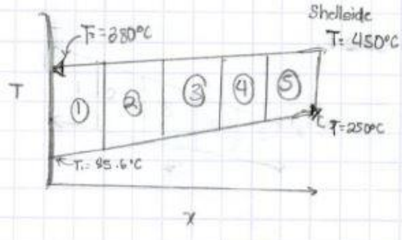
Heat Exchanger Sizing (E-02)

Step 1: Determine the Q

$Q_n = \dot{m}\Delta H = \dot{m}(H_2 - H_1)$



Streams	H(kJ/kmol)	Q
1	39415	1463
2	40879	1463
3	42343	1463
4	43807	1463
5	45270	1463



LMTD

$$LMTD = \frac{\Delta T_i - \Delta T_{i+1}}{\ln\left[\frac{\Delta T_i}{\Delta T_{i+1}}\right]}$$

i.e. $$LMTD_1 = \frac{(395-122.3)-(380-85.6)}{\ln\left[\frac{395-122.3}{380-85.6}\right]}$$

$= 260.23°C$

Stream	LMTD (°C)
1	282.45
2	260.23
3	240.73
4	223.08
5	207.40

Overall U Value Coefficient

- From Table 9.15a, shell and tube exchange overall coefficient for "asphalt" and condensing vapor stream hydrocarbons with inert gas.

$U = 100 - 200 \frac{W}{m^2 K}$ Choose $150 \frac{W m^2}{K}$

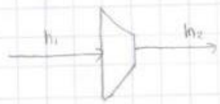
$A = \frac{Q_n}{U\,LMTD}$ $A = \frac{1463}{(80)(282.23)} = 574.86\ m^2$

$\Sigma A_i = 1790\ m^2$

E.2.6 Gas Expander Sizing Sample Calculations

Gas Expander Design

The work outputted by pressure recovery expander can be described by the enthalpy change of inlet and outlet streams



$$W = \dot{m}(h_1 - h_2)$$
$$\dot{m} = 30\,011\ \mathrm{kg/h}$$

Using simulation to retrieve specific enthalpy of hydrocarbon stream at P = 16 MPa and P = 3.5 MPa. The isentropic ideal work is

$$W_i = 30\,011\,\frac{\mathrm{kg}}{\mathrm{h}} \times \left(10218\,\frac{\mathrm{kJ}}{\mathrm{kmol}} - 7435\,\frac{\mathrm{kJ}}{\mathrm{kmol}}\right)\left(\underbrace{\frac{\mathrm{kmol}}{7.48\ \mathrm{kg}}}_{\mathrm{MW}}\right)\frac{1\mathrm{h}}{3600\mathrm{s}}$$

$$= 3.10 \times 10^6\ \mathrm{W}$$

from U & V Figure 4.2 Page 121, the efficiency of gas expander at pressure ratio of 3.02 is approximately 80 - 85% for power production

$$\varepsilon_i = 85\%$$

$\therefore$ actual power converted

$$W_p = \varepsilon_i W_i = 0.85(3.10 \times 10^6\ \mathrm{W})$$
$$= 2.64 \times 10^6\ \mathrm{W}$$

E.2.7 PSV Sizing Sample Calculations

Pressure Safety Valve Sizing

Based on GPSA Section 5 the Critical Flow through PSV

$$A = \frac{100W\sqrt{T_1(Z)}}{C_1 \cdot K \cdot P_1 \cdot K_B \cdot \sqrt{MW}} \qquad C_1 = 387\sqrt{K\left(\frac{2}{K+1}\right)^{\frac{K+1}{K-1}}} \quad (K=0.975)$$

$C_1 = 255$

$W = 200\ 074$ kg/hr

$T_1 = 70°C$

$Z = 1.0215$

$MW = 7.48$ kg/kmol

$P_1 = 18\ 300$ kPa (from MAWP calculation)

$K_B = 1.0$

$K = \frac{C_P}{C_V} = 1.142$ (from Symmetry)

$A = 20.6$ mm = } the designation M with an orfc area of 23.2 mm is sufficient 100 × 150 mm 4" × 6"

The PSV is located top of D-03

E.2.8 Reactor Sizing Sample Calculations

Slurry Bubble Reactor Design and Sizing

In order to complitly design the Slurry Reactor some kinetics would need to be provided, therefore, Mr. Nolte provide the CANMET reactor size specification:

Space velocity $LHSV = 0.21/hr$

Vapor voidage = 40%

From simulation:

$\dot{V}_{in} = 15598\ m^3/day$

$\dot{m}_{in} = 665516\ kg/hr$

$MW = 498.91\ kg/kmol$

Assumptions:

- Reactor conversion 95% : $X_A = 0.95$
- Continuous stirred-tank reactor (CSTR)
- well mixed
- steady-state reaction throughout the reactor
- First-order reaction

$$LHSV = \frac{\text{volumetric feed liquid flow rate}}{\text{reaction volume}} = \frac{\dot{V}}{V}$$

$$V = 15598\ \frac{m^3}{day} \times \frac{1\ day}{24\ hr} \times \frac{1\ hr}{0.21} = 3094.8\ m^3$$

Overall Mole Balance:

$$\text{In} - \text{Out} + \text{Generation} - \text{Consumption} = \cancel{\text{Accumulation}}^{\,0}$$

$$F_{A0} - F_A + \int r_A\, dV = 0$$

$$F_{A0} - F_A + r_A V = 0 \qquad r_A = \frac{F_A - F_{A0}}{V}$$

Mole in:

$$F_{A0} = \frac{\dot{m}}{MW} = 665516 \frac{Kg}{hr} \times \frac{Kmol}{498.91\,Kg} \times \frac{mol}{1000\,Kmol} = 1.33 \frac{mol}{hr}$$

Mole out:

$$F_A = F_{A0}(1 - X_A) = 1.33 \frac{mol}{hr} \times (1 - 0.95) = 0.067 \frac{mol}{hr}$$

Rate law for CSTR 1st order reaction:

$$-r_A = K C_A$$

Initial concentration:

$$C_{A0} = \frac{F_{A0}}{\dot{V}_0} = 1.33 \frac{mol}{hr} \times \frac{day}{15598 m^3} \times \frac{24\,hr}{1\,day} = 2.05 \times 10^{-3} \frac{mol}{m^3}$$

Final concentration:

$$C_A = C_{A0}(1 - X_A) = 2.05 \times 10^{-3} \frac{mol}{m^3} \times (1 - 0.95) = 1.03 \times 10^{-4} \frac{mol}{m^3}$$

$$r_A = \frac{F_A - F_{A0}}{V} = \frac{0.067 - 1.33}{3094.8} = -4.08 \times 10^{-4} \frac{mol}{m^3 \cdot hr}$$

$$-r_A = K C_A \Rightarrow \text{Rate constant } K = -\frac{r_A}{C_A}$$

$$K = -\frac{(-4.08 \times 10^{-4})}{1.03 \times 10^{-4}} = 3.96\,hr^{-1}$$

Half - life for 1st order

$$t_{1/2} = \frac{0.693}{K} = \frac{0.693}{3.96} = 0.17\,hr$$

Residence time

$$\tau = \frac{C_{A0} V}{F_{A0}} = \frac{2.05 \times 10^{-3} \times 3094.8}{1.33}$$

$$\tau = 4.77\,hr$$

Due to the large amount of volume the residence time is very large. Therefore, the technical decision was to have 2 trains with 4 identical reactors in series. Giving in total 8 Slurry Bubble Reactor operating at 250 bar and 467 °C with a volume of

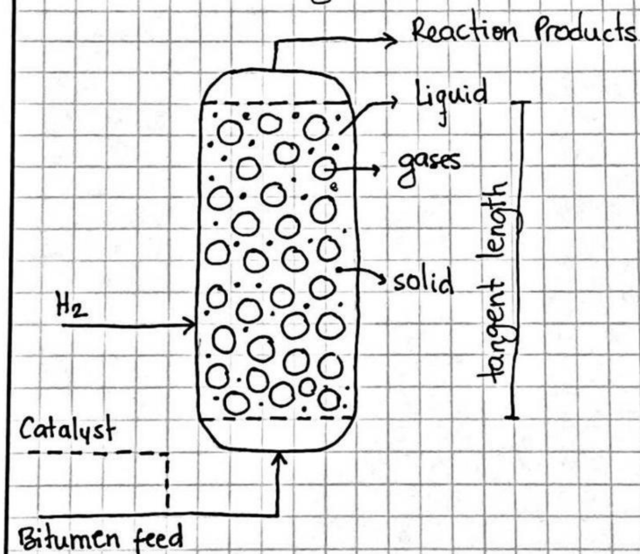
$$V = \frac{3094.8\,m^3}{8} = 386.9\,m^3 \text{ per each reactor.}$$

New Residence time for each reactor

$$\tau = \frac{2.05 \times 10^{-3} \times 386.9}{1.33}$$

$$\tau = 0.6\ hr$$

Reactor Sizing R-01



Darby equation:

$$u_t = \frac{\mu_f}{\rho_f D_p}\left[\left(14.4 + 1.8\sqrt{D_p^3(\rho_p - \rho_f)\rho_f g/\mu_f^2}\right)^{0.5} - 3.8\right]^2$$

Particle size : $D_p = 100\ \mu m = 1\times10^{-4}\ m$

$\rho_{liquid} = 527.02\ kg/m^3$

$\rho_{gas} = 99.86\ kg/m^3$

$\mu_{liquid} = 3.78\times10^{-6}\ Pa.s$

$g = 9.81\ m/s^2$

$$u_t = \frac{3.78\times10^{-6}}{527.02\times1\times10^{-4}}\left[\left(14.4+1.8\sqrt{\frac{(1\times10^{-4})^3(527.02-99.86)\times527.02\times9.81}{(3.78\times10^{-6})^2}}\right)^{0.5} - 3.8\right]^2$$

$u_t = 0.0382\ m/s$

For vertical drum

$$D = \left(\frac{5\dot{q}}{\pi u_t}\right)^{0.5}$$

$$D = \left(5\times\frac{15598\ m^3}{day}\times\frac{1\,day}{86400\,s}\times\frac{s}{0.0382\,m}\times\frac{1}{\pi}\right)^{0.5}$$

$D = 2.74\ m$ ⇒ minimum diameter per each reactor

$V = \pi r^2 L = \frac{\pi}{4}D^2 L \Rightarrow L = \frac{4V}{\pi D^2}$

$$L = \frac{4\times386.9\ m^3}{\pi\times(2.74\ m)^2} = 65.6\ m$$

Check L/D ratio : (Gerunda, 1981)

Based on Pressure of the Drum 250 bar-a $L/D \geq 5$

$$\frac{L}{D} = \frac{65.6\ m}{2.74\ m} = 23.9$$

The L/D ration is too high, therefore increase the diameter in order to obtain a better ratio

$D = 4\,m$

$$L = \frac{4 \times 386.9\,m^3}{\pi \times (4)^2 m^2} = 30.8\,m$$

$$\Rightarrow L/D = \frac{30.8\,m}{4\,m} = 7.7 \quad \Rightarrow \text{more acceptable ratio!}$$

$$\Rightarrow L/D = 7.7 \quad , \quad D = 4\,m$$

Pressure vessel wall thickness

$$t = \frac{Pr}{SE - 0.6P} + C_s$$

$$r = \frac{4m}{2} = 2m$$

S = maximum allowable tensile stress

U&V Fig 4.1 : → Material required : carbon steel

→ T = 467 °C

S = 840 bar

E = welded joint efficiency E = 0.9

C_s = corrosion allowance C_s = 3mm for carbon steel

$$t = \frac{250 \times 2}{(840 \times 0.9) - (0.6 \times 250)} + 0.003 = 0.828\,m$$

$t = 828\,mm$

E.2.9 Stripper Column Sizing Sample Calculations

Stripper Column Sizing

In order to design the tower, the diameter and height must be determined.

From Symmetry

$$N_{theoretical} = 2 \text{ trays}$$

L = Liquid molar flowrate (mol/h)
V = Gas molar flowrate (mol/h)
M_g = molar mass of gas (g/mol)
ρ_L = liquid density (kg/m³)
ρ_V = gas density (kg/m³)

Diameter of the tower

$$D = \sqrt{\frac{4 V_{max} M_g}{\pi \rho_g \cdot U_{sg}}} \quad (1)$$

where

$$U_{sg} = K_{SB} \sqrt{\frac{\rho_L - \rho_V}{\rho_V}} \quad (2)$$

Based on Ulrich and Vasudevan (2004), a value of 0.06 (lower end) applies to gas-liquid systems

$$U_{sg} = 0.06\,m/s \sqrt{\frac{919\,kg/m^3 - 0.7\,kg/m^3}{0.7\,kg/m^3}} = 2.17\,m/s$$

$$D = \sqrt{\frac{4\,(270.34\,mol/h)(34.26\,g/mol)(1h/3600s)}{\pi\,(0.699\,kg/m^3)(2.17\,m/s)}} = 1.5\,m$$

Tray efficiency can be estimated at 80% from (Nag, Ashis)(2016) as conservative option. This was obtained by assuming liquid is similar to HCGO product.

Based on CH E 464 lecture, default spacing of 24 in can be used.

$$H = \frac{N_{theoretical} \times H_t}{\varepsilon_s}$$

Thus,

$$H = \frac{2}{0.3}(24 in)\left(\frac{0.0254 m}{1 in}\right) = 4.04 m$$

The liquid holdup is also part of total tower height

Volumetric flowrate at last tray

$$H_{LH} = \frac{V_{aL}}{A} \times \text{liquid residence time};$$

$$H_{LH} = \frac{(7.4 m^3/h)(3 min)(1/60)}{\frac{(\pi)(1.5 m^2)}{4}} = 0.204 m$$

Dimensions

$D = 1.5 m$
$H_T =$ [illegible] m
Tray spacing $= 24 in$
$N_{actual} = \frac{2}{0.3} = 7$ trays

E.3 Line Sizing

Table E2. Detailed line designation.

Stream Number	Nominal Pipe Diameter (in)	Pipe Schedule	Velocity (m/s)	Pressure Drop (kPa/100m)	Material	Material Grade	Type of Flow	Flow Regime
S3	8	20	3.1	35	ASTM A106 (CS)	B	Liquid	Turbulent
S4	8	100	3.7	55	ASTM A106 (CS)	B	Liquid	Turbulent
S5	8	100	4.6	68	ASTM A106 (CS)	B	Liquid	Turbulent
S6	8	100	5.8	83	ASTM A106 (CS)	B	Liquid	Turbulent
S7	4	160	0.0	n/a	ASTM A106 (CS)	B	Solid	Turbulent
S8	8	160	5.9	84	ASTM A106 (CS)	B	Slurry	Turbulent
S9	14	120	24.2	19	ASTM A376 (SS)	TP304	Vapour	Turbulent
S10	4	140	0.0	n/a	ASTM A376 (SS)	TP304	Slurry	Turbulent
S11	18	140	15.2	26	ASTM A376 (SS)	TP304	Vapour	Turbulent
S12	18	120	14.2	29	ASTM A376 (SS)	TP304	2-Phase	Dispersed
S13	1.5	80	2.0	51	ASTM A376 (SS)	TP304	Liquid	Turbulent
S14	18	140	14.9	25	ASTM A376 (SS)	TP304	Vapour	Turbulent
S15	18	120	12.8	32	ASTM A376 (SS)	TP304	2-Phase	Dispersed
S16	8	120	2.0	5	ASTM A376 (SS)	TP304	Liquid	Turbulent
S17	16	120	15.6	27	ASTM A376 (SS)	TP304	Vapour	Turbulent
S18	16	120	12.1	34	ASTM A376 (SS)	TP304	2-Phase	Dispersed
S19	18	100	11.9	39	ASTM A376 (SS)	TP304	3-Phase	Dispersed
S20	1	40	1.4	71	ASTM A376 (SS)	TP304	Liquid	Turbulent
S21	8	100	6.6	84	ASTM A376 (SS)	TP304	Liquid	Turbulent
S22	12	100	20.9	44	ASTM A376 (SS)	TP304	Vapour	Turbulent
S23	12	100	14.6	21	ASTM A376 (SS)	TP304	Vapour	Turbulent
S24	10	100	10.2	13	ASTM A376 (SS)	TP304	Vapour	Turbulent
S25	10	120	20.3	28	ASTM A376 (SS)	TP304	Vapour	Turbulent

E.3.1 Line Sizing Sample Calculations

LINE SIZING

General Assumptions / Notes:

* at points where streams merge/split between parallel pieces of equipment (e.g. R-A01 – R-A04), the largest combined flow rate is used for line sizing
* 1/16 in corrosion allowance was applied to carbon steel lines; 1/32 in corrosion allowance applied to stainless steel lines
* 100m characteristic length used for all sizing
* line sizing was not performed for S7 (solid catalyst input) or S10 (reactor bottoms slurry line) as they would require more detailed analysis

<u>Line Wall Thickness</u> (Pipe Schedule)

Pipe wall thickness for allowable working pressure determined by ANSI B31.3, "Code for Pressure Piping, Petrollium Refinery Piping"

$$\boxed{t = \frac{P_i d_o}{2(S'E' + P_i Y)}} \qquad \text{(GPSA, Fig 17-23)}$$

* data for allowable material stress, S', obtained from GPSA, Fig 17-25

Example: S6 – set nominal pipe size to 8.0in

$$t = \frac{(17{,}000 - 101.325)(219.0\,mm)}{2[117\,kPa \cdot 1 + (17{,}000 - 101.325)\,0.4]} = 14.8\ mm$$

$$t_m = \underbrace{14.8\,mm}_{\text{design thickness}} + \underbrace{1.59\,mm}_{\text{corrosion allowance}} = \boxed{16.3\,mm \equiv \text{minimum pipe wall thickness}}$$

@ 8.0in pipe, closest schedule pipe is schedule 100, t = 18.3mm

Liquid Flows

Pressure loss due to friction is calculated using the Darcy-Weisbach Equation

$$\boxed{\Delta P_f = \frac{0.5 \rho f_m L V^2}{d}} \qquad \text{(GPSA, Eq 17-2)}$$

* Moody friction factor, f_m, obtained from GPSA, Fig 17-2

Example: 56

$$\Delta P_f = \frac{0.5\,(606.27\ ^{kg}/_{m^3})(0.0146)(100m)(5.835\ ^{m}/_{s})}{(182.4mm)} = \boxed{83\ ^{kPa}/_{100m} = \Delta P_f}$$

Vapour Flows

Pressure loss due to friction (for vapour flows) is calculated using a simplified Darcy-Weisbach formula.

$$\boxed{\Delta P_{100} = \frac{W^2}{\rho}\left[\frac{62,350(10^{2})F}{d^5}\right]} \qquad \text{(GPSA, Eq 17-30)}$$

which can be simplified to:

$$\Delta P_{100} = \frac{C_1 C_2}{\rho} \qquad \text{(GPSA, Eq. 17-31)}$$

where: $C_1 = W^2(10^{-9})$, obtained from GPSA, Fig 17-8

$C_2 = \frac{62.350 \cdot 10^{14} \cdot F}{d^5}$ obtained from GPSA, Fig 17-9

Example: 59

- W (mass flow rate) = 90,114 $^{kg}/_{h}$ ⇒ Fig 17-8 ⇒ $C_1 = 8.4$
- Pipe used: 14.0 in, schedule 120 ⇒ Fig 17-9 ⇒ $C_2 = 33.40$

$$\Delta P_{100} = \frac{(8.4)(33.40)}{14.58\ ^{kg}/_{m^3}} = \boxed{19\ ^{kPa}/_{100m} = \Delta P_{100}}$$

Two-Phase Flow

The Dunkler equation was used for frictional pressure drop calculations

$$\boxed{\Delta P_{fn} = \frac{f_n \, f_{tpr} \, \rho_k \, V_m^2 \, L_m}{2d}} \qquad \text{(GPSA, Eq 17-30)}$$

where $$f_n = 0.0056 + 0.5(Re_y)^{-0.32} \qquad \text{(GPSA, Eq. 17-44)}$$

$$Re_y = \frac{(0.001)\,\rho_k \, V_m \, d}{\mu_n} \qquad \text{(GPSA, Eq. 17-45)}$$

$$f_{tpr} = \frac{y}{1.281 - 0.478y + 0.444y^2 - 0.094y^3 + 0.00843y^4} + 1 \qquad \text{(GPSA, Eq 17-48)}$$

$$y = -\ln(\lambda)$$

Example: S15

$$y = -\ln(0.0343) = 3.37 = y$$

$$f_{tpr} = \frac{3.37}{1.281 - 0.478(3.37) + 0.444(3.37)^2 - 0.094(3.37)^3 + 0.00843(3.37)^4} + 1$$

$$f_{tpr} = 2.53$$

$$Re_y = \frac{(0.001)(76.96\ kg/m^3)(387.2mm)}{1.75 \cdot 10^{-5}\ Pa \cdot s} = 2.17 \cdot 10^7 = Re_y$$

$$f_n = 0.0056 + 0.5(2.17 \cdot 10^7)^{-0.32} = 7.98 \cdot 10^{-3} = f_n$$

$$\Delta P_{fn} = \frac{(7.98 \cdot 10^{-3})(2.53)(76.96\ kg/m^3)(12.77\ m/s)^2(100m)}{2(387.2mm)} = \boxed{32\ kPa/100m = \Delta P_{fn}}$$

E.4 References

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