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Contents

Restriction Orifice [RO] Sizing
Jayanthi Vijay Sarathy
..... 2

FCC Technologies as Flexible Strategy to Maximize Refining Margins under a Hostile Scenario for Fossil Fuels
Dr. Marcio Wagner da Silva
.....20

All About Downcomers and Weirs
Dr. -Ing. Volker Engel
.....38

Design and Revamp Guidelines for Palm Oil Stripper Columns
Karl Kolmetz
.....55

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Restriction Orifice [RO] Sizing

Jayanthi Vijay Sarathy

Restriction Orifices [RO] are specialized non-moving plate devices inserted into piping to permanently reduce fluid pressure, restrict flow rates, and/or minimize noise/vibrations. Their key application in the oil and gas industry lies in reducing pressure to protect downstream equipment from excessive pressure, regulating flow to other units such as the flare stack, controlling pump minimum flows and enabling safe depressurization. Restriction orifices can be single stage or multistage. In the current undertaking, a single stage orifice is considered and the fluid properties are evaluated using Soave-Redlich-Kwong (SRK) Equation of State (EoS) and ISO 5167-2:2003 standard.

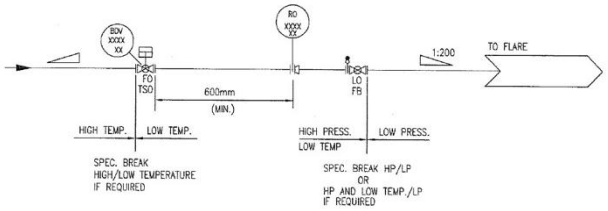

Figure 1. Blowdown line with RO arrangement [1]

General Notes

1. Although ISO 5167-2 is written primarily for flow measurement for high accuracy, it is also applicable for pure pressure reduction applications to ensure more predictable behaviour in high ΔP and multiphase flow applications.
2. Unlike general orifice sizing methods, ISO 5167-2 takes into consideration a more rigorous discharge coefficient [C] and expansion factor [ε] estimation. As a result, the discharge coefficient [C] can be lower than 0.6 also. The ISO sizing method is applicable for a beta [β] of $0.2 \leq \beta \leq 0.75$
3. In the current undertaking, for single stage orifice, the orifice type chosen is a standard sharp edged orifice.

Methodology – SRK EoS

To estimate properties of the hydrocarbon mixture, the Soave-Redlich-Kwong (SRK) Equation of State (EoS) can be expressed as,

$$P = \frac{RT}{[V_m - b_m]} + \frac{a_m}{V_m[V_m + b_m]} \quad (1)$$

Where,

P = Pressure [bar(a)]

R = Gas Constant = 0.08314 [m^3.bar/kmol.K]

T = Temperature [K]

a_m = Mixture attraction parameter [m^6.bar/$kmol^2$]

b_m = Mixture co-volume parameter [m^3/kmol]

v_m = Mixture molar volume [m^3/kmol]

To estimate the values of a_m and b_m of the mixture, Van der Waals mixing rules can be considered as follows,

$$a_m = \sum_i \sum_j x_i x_j a_{ij} \quad (2)$$

$$a_m = \sum_i \sum_j x_i x_j \times \sqrt{a_i a_j} \times \left(1 - k_{ij}\right) \quad (3)$$

$$b_m = \sum_i x_i b_i \quad (4)$$

Where,

$$a_i = a_j = \frac{0.42474 \times R^2 \times T_{c,i}^2}{P_{c,i}} \times \alpha_i(T) \quad (5)$$

$$b_i = b_j = \frac{0.08664 \times R \times T_{c,i}}{P_{c,i}} \quad (6)$$

$$x_i = x_j \quad (7)$$

$$\alpha_i(T) = \left[1 + \left[m_i\left(1 - \sqrt{T_{r,i}}\right)\right]\right]^2 \quad (8)$$

$$m_i = 0.48505 + 1.55171\omega_i - 0.15613\omega_i^2 \quad (9)$$

The binary interaction parameters [BIP] [k_{ij}] are calculated based on Cheu-Prausnitz (1967) correlation using critical volume [V_c] values for non-polar pairs as follows,

$$k_{ij} = 1 - \left[\frac{2\left[V_{c,i} V_{c,j}\right]^{1/6}}{\left[V_{c,i}^{\ 1/3} + V_{c,j}^{\ 1/3}\right]}\right]^3 \quad (10)$$

$x_i = x_j$ = Mole fractions [mol frac]

$P_{c,i}$ = Critical Pressure [bara]

$T_{c,i}$ = Critical Temperature [K]

$V_{c,i} = V_{c,j}$ = Critical Volume [m^3/kmol]

k_{ij} = Binary Interaction Parameter [-]

ω = Acentric Factor [-]

α_i(T) = Graboski and Daubert formulation [-]

Based on the above expressions, the molar volume of the mixture [V_m] can be calculated iteratively as,

$$P - \frac{RT}{[V_m - b_m]} - \frac{a_m}{V_m[V_m + b_m]} = 0 \quad (11)$$

The mixture density [ρ] is computed as,

$$\rho = \frac{MW_m}{V_m} \tag{12}$$

Where,

ρ = mixture density [kg/m^3]

MW_m = Mixture molecular weight [kg/kmol]

To estimate the compressibility factor of the mixture, the cubic form is,

$$Z^3 - Z^2 + [A - B - B^2]Z - AB \tag{13}$$

$$A = \frac{a_m P}{R^2 T^2} \tag{14}$$

$$B = \frac{b_m P}{RT} \tag{15}$$

To identify the phases, the largest positive root represents the vapour phase compressibility factor, and the smallest positive root represents the liquid phase compressibility factor. Two phase if two positive roots.

The molar enthalpy can be calculated using the residual properties as follows,

$$H_m = H_{ideal} + \int_{T_{ref}}^{T} C_{p,m} dT + H^R \tag{16}$$

Where,

$$H_{ideal} = \sum x_i H_{ideal,i} \tag{17}$$

$$\int_{T_{ref}}^{T} C_{p,m} dT = \left[\sum x_i C_{p,i}\right] \times \left(T - T_{ref}\right) \tag{18}$$

$$H^R = RT[Z-1] + \left[\frac{T\left(\frac{da}{dT}\right) - a_m}{b_m} \times ln\left[\frac{Z+B}{Z}\right]\right] \tag{19}$$

$$\frac{da}{dT} = \sum_i \sum_j \frac{x_i x_j \times \sqrt{a_i a_j} \times \left(1 - k_{ij}\right)}{2} \left[\frac{d\alpha_i}{dT} + \frac{d\alpha_j}{dT}\right] \tag{20}$$

$$\frac{d\alpha_i}{dT} = \frac{d\alpha_j}{dT} = -\frac{m_i\left[m_i \times \left(1 + \sqrt{T_r}\right)\right]}{\sqrt{T_r}} \tag{21}$$

Adding all the terms, molar enthalpy [H_m] is

$$H_m = \sum x_i H_{ideal,i} + \left[\sum x_i C_{p,i}\right]\left(T - T_{ref}\right) + RT[Z-1] + \left[\frac{T\left(\frac{da}{dT}\right) - a_m}{b_m} \times ln\left[\frac{Z+B}{Z}\right]\right] \tag{22}$$

The gas viscosity can be computed using Lee et al, 1966 [3], as follows,

$$\mu_g = K_1 e^{X \rho^Y} \tag{23}$$

Where,

$$K_1 = \frac{\left[0.00094 + \left[2 \times 10^{-6} MW_g\right]\right] \times T^{1.5}}{\left[209 + 19MW_g + T\right]} \tag{24}$$

$$X = 3.5 + \frac{986}{T} + 0.01MW_g \quad (25)$$

$$Y = 2.4 - 0.2X \quad (26)$$

μ_g = Gas Viscosity [cP]

ρ = Gas density [g/cm^3]

P = Pressure [psia]

T = Temperature [^{0}Rankine]

MW_g = Gas Molecular weight [-]

The Lee gas viscosity correlation is valid for

- 100 psia [6.9 bara]< P <8000 psia [551 bara]

- 100^{0}F (37.8^{0}C) < T < 340^{0}F (171^{0}C)

- 0.9 mol% < CO_2 < 3.2 mol%

- 0.0 mol% < N_2 < 4.8 mol%

RO Sizing – ISO 5167-2:2003

To size the restriction orifice [RO], the expression provided in ISO 5167 is written as,

$$q_v = \frac{\varepsilon d^2 C}{\sqrt{1-\beta^4}} \times \frac{\pi}{4} \times \sqrt{\frac{2g[P_1 - P_2]}{\rho}} \quad (27)$$

Where,

$$C = 0.5961 + 0.0261\beta^2 - 0.216\beta^8 + \left[0.000521\left(\frac{10^6\beta^2}{Re}\right)^{0.7}\right] + \left[(0.0188 + 0.0063A)\beta^{3.5}\left(\frac{10^6}{Re}\right)^{0.3}\right] + \left[(0.043 + 0.08e^{-10L_1} - 0.123e^{-7L_1})(1 - 0.11A)\left(\frac{\beta^4}{1-\beta^4}\right)\right] - \left[0.031\left(M_2' - 0.8M_2'^{1.1}\right)\beta^{1.3}\right] \quad (28)$$

Where, D < 71.12 mm, the following term should be added as $+ 0.011(0.75 - \beta)\left[2.8 - \frac{D}{25.4}\right]$

$$\beta = \frac{d}{D} \quad (29)$$

$$\varepsilon = 1 - \left[[0.351 + 0.256\beta^4 + 0.93\beta^8]\left[1 - \left(\frac{P_2}{P_1}\right)^{\frac{1}{k}}\right]\right] \quad (30)$$

d = RO diameter [mm]

D = Pipe diameter [mm]

β = Diameter ratio [-]

$k = C_p/C_v$ = Specific heat ratio [-]

Re = Reynolds number w.r.t 'D' [-]

The discharge coefficient [C] expression is referred to as the Reader-Harris/Gallagher (1998) equation and is valid only for the tapping arrangements specified in Sec 5.2.2 and 5.2.3 of ISO 5167-2:2003

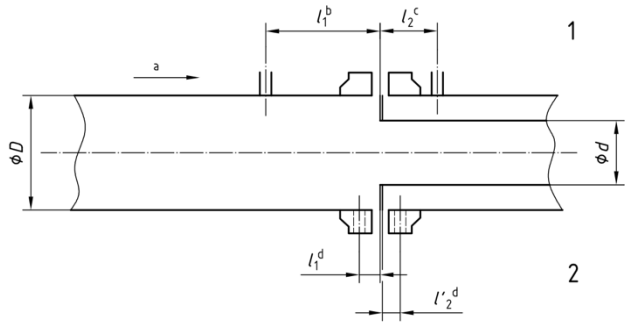

Fig 2. Spacing of Pressure Tappings for Orifice Plates with D and D/2 Tappings or Flange Tappings

$L_1 = l_1/D$ is the quotient of the distance of the upstream tapping from the upstream face of the plate and the pipe diameter

$L'_2 = l'_2/D$ is the quotient of the distance of the downstream tapping from the downstream face of the plate and the pipe diameter (L'_2 denotes the reference of the downstream spacing from the downstream face, while L_2 would denote the reference of the downstream spacing from the upstream face)

$$M'_2 = \frac{2L'_2}{1-\beta} \tag{31}$$

$$A = \left[\frac{19000\beta}{Re}\right]^{0.8} \tag{32}$$

For corner tappings, $L_1 = L'_2 = 0$

For D and D/2 tappings, $L_1 = 1$; $L'_2 = 0.47$

For flange tappings, $L_1 = L'_2 = 25.4/D$ [mm]

The limits of use of ISO 5167 expressions are,

For orifice plates with corner or with D and D/2 pressure tappings,

$$d \geq 12.5\ mm \tag{33}$$

$$50\ mm \leq D \leq 1000\ mm \tag{34}$$

$$0.1 \leq \beta \leq 0.75 \tag{35}$$

$$Re \geq 5{,}000\ for\ 0.1\ \leq \beta \leq 0.75 \tag{36}$$

$$Re \geq 16{,}000\beta^2\ for\ \beta > 0.56 \tag{37}$$

For orifice plates with flange tappings,

$$d \geq 12.5\ mm \tag{38}$$

$$50\ mm \leq D \leq 1000\ mm \quad (39)$$

$$0.1 \leq \beta \leq 0.75 \quad (40)$$

$$\text{Both } Re \geq 5000 \text{ and } Re \geq 170\beta^2 D \quad (41)$$

Case Study

A restriction orifice [RO] is to be sized for hydrocarbon service operating at a pressure and temperature of 40 bara and 60^{0}C. The flow rate is 1,000 kmol/h and is let down to 4 bara. The reference temperature is 25^{0}C and gas constant is 0.08314 m^3.bar/kmol.K. The fluid composition and properties are,

Table 1. Stream Composition

Component	MW	Mol%	$C_{p,1\ atm,25^0C}$
	kg/kmol	%	kJ/kmol.K
Methane	16.043	76.23	35.97
Ethane	30.070	10.00	53.12
Propane	44.097	5.00	75.42
i-Butane	58.124	1.00	98.30
n-Butane	58.124	1.00	98.91
i-Pentane	72.151	0.30	165.53
n-Pentane	72.151	0.10	167.79
n-Hexane	86.178	0.05	191.94
H_2O	18.015	0.25	79.93
CO_2	44.010	3.00	38.50
H_2S	34.081	0.07	34.54
N_2	28.013	3.00	29.19
Total	21.17	100	$\sum C_{p,mix}$ = 41.49

The critical properties are,

Table 2. Critical properties

Component	T_c	P_c	V_c	ω
	K	bara	m^3/kmol	m^3/kmol
Methane	190.7	46.41	0.0990	0.0115
Ethane	305.4	48.84	0.1480	0.0986
Propane	369.9	42.57	0.2000	0.1524
i-Butane	408.1	36.48	0.2630	0.1848
n-Butane	425.2	37.97	0.2550	0.2010
i-Pentane	460.4	33.34	0.3080	0.2222
n-Pentane	469.6	33.75	0.3110	0.2539
n-Hexane	507.9	30.32	0.3680	0.3007
H_2O	647.3	221.20	0.0571	0.3440
CO_2	304.1	73.70	0.0939	0.2389
H_2S	373.6	90.08	0.0980	0.0810
N_2	126.2	33.94	0.0900	0.0400

The ideal gas enthalpy [H_{id}] at 1 atm, 25^{0}C is taken to be the same as heat of formation [ΔH_f] at 1 atm, 25^{0}C. To estimate the mixture ideal gas enthalpy, the ideal gas enthalpy of each component is multiplied by the mole fraction and added as follows,

Table 3. Ideal Gas Enthalpy of Hydrocarbon mixture

Component	Mol%	$H_{id,1\ atm,\ 25^0C}$	x_i . $H_{id,1\ atm,25^0C}$
	%	kJ/kmol	kJ/kmol
Methane	76.23	-74,900	-57,096
Ethane	10.00	-84,738	-8,474
Propane	5.00	-103,890	-5,195
i-Butane	1.00	-134,590	-1,346
n-Butane	1.00	-126,190	-1,262
i-Pentane	0.30	-154,590	-464
n-Pentane	0.10	-146,490	-146
n-Hexane	0.05	-167,290	-84
H_2O	0.25	-241,814	-605
CO_2	3.00	-393,790	-11,814
H_2S	0.07	-20,179	-14
N_2	3.00	0	0
Total	100	$\sum H_{id,mix}$	-86,499

The reduced temperature [T_r], m_i, α(T), a_i, b_i are as follows,

$$T_{r,CH_4} = \frac{T}{T_c} = \frac{60+273.15}{190.7} = 1.747 \quad (42)$$

$$\sqrt{T_{r,CH_4}} = \sqrt{1.747} = 1.322 \quad (43)$$

$$m_{CH_4} = 0.48505 + [1.55171 \times 0.0115] - [0.15613 \times 0.0115^2] = 0.5029 \quad (44)$$

$$\alpha_{CH_4} = [1 + [0.5029(1 - 1.322)]]^2 = 0.7026 \quad (45)$$

$$a_{CH_4} = \frac{0.42474 \times 0.08314^2 \times 190.7^2}{46.41} \times 0.7026 \quad (46)$$

$$a_{CH_4} = 1.627 \quad (47)$$

$$b_{CH_4} = \frac{0.08664 \times 0.08314 \times 190.7}{46.41} = 0.0296 \quad (48)$$

$$\frac{d\alpha_{CH_4}}{dT} = \frac{-0.5029[0.5029 \times (1+1.322)]}{1.322} = -0.4443 \quad (49)$$

The above set of calculations are similarly performed for all other components.

Table 4. SRK Parameters [T_r, m_i, $\alpha(T)$]

Component	$\sqrt{T_r}$	m_i	α(T)
	-	-	-
Methane	1.3217	0.5029	0.7026
Ethane	1.0444	0.6366	0.9443
Propane	0.9490	0.7179	1.0745
i-Butane	0.9035	0.7665	1.1534
n-Butane	0.8852	0.7907	1.1898
i-Pentane	0.8507	0.8222	1.2607
n-Pentane	0.8423	0.8690	1.2929
n-Hexane	0.8099	0.9376	1.3882
H_2O	0.7174	1.0004	1.6453
CO_2	1.0467	0.8469	0.9225
H_2S	0.9443	0.6097	1.0691
N_2	1.6248	0.5469	0.4334

Table 5. SRK Parameters [a_i, b_i]

Component	a_i	b_i
	$m^6.bar/kmol^2$	$m^3/kmol$
Methane	1.6270	0.0296
Ethane	5.3300	0.0451
Propane	10.2067	0.0626
i-Butane	15.5617	0.0806
n-Butane	16.7436	0.0807
i-Pentane	23.6881	0.0995
n-Pentane	24.9635	0.1002
n-Hexane	34.9070	0.1207
H_2O	9.2098	0.0211
CO_2	3.4206	0.0297
H_2S	4.8952	0.0299
N_2	0.6008	0.0268

Table 6. SRK Parameters [b_m, $d\alpha(T)/dT$]

Component	b_i	$d\alpha(T)/dT$
	$m^3/kmol$	-
CH_4	0.0226	-0.4443
C_2H_6	0.0045	-0.7932
C_3H_8	0.0031	-1.0585
i-C_4H_{10}	0.0008	-1.2377
n-C_4H_{10}	0.0008	-1.3314
i-C_5H_{12}	0.0003	-1.4708
n- C_5H_{12}	0.0001	-1.6517
n-C_6H_{14}	0.0001	-1.9644
H_2O	0.0001	-2.3958
CO_2	0.0009	-1.4026
H_2S	0.0000	-0.7655
N_2	0.0008	-0.4832
Total	0.0340	

The binary interaction parameters [k_{ij}] for methane-ethane combination is estimated as,

$$k_{ij} = 1 - \left[\frac{2[0.099\times0.148]^{1/6}}{\left[0.099^{1/3}+0.148^{1/3}\right]}\right]^3 = 0.0067 \quad (50)$$

The value of a_{ij} for methane-ethane combination is estimated as,

$$a_{ij} = \sqrt{a_{CH_4} \times a_{C_2H_6}} \times \left[1 - k_{CH_4-C_2H_6}\right] \quad (51)$$

$$a_{ij} = \sqrt{1.627 \times 5.33} \times [1 - 0.0067] = 2.925 \quad (52)$$

$$a_m = x_i x_j a_{ij} = 0.7623 \times 0.1 \times 2.925 \quad (53)$$

$$a_m = 0.223 \quad (54)$$

The above calculations are similarly performed for all other components, and the values of a_m for all combinations are added.

$$a_m = \sum_i \sum_j x_i x_j \times \sqrt{a_i a_j} \times \left(1 - k_{ij}\right) = 2.393 \quad (55)$$

The SRK A and B factors are computed as,

$$A = \frac{0.223\times40}{0.08314^2\times(60+273.15)^2} = 0.1248 \quad (56)$$

$$B = \frac{0.03404\times40}{0.08314\times(60+273.15)} = 0.04916 \quad (57)$$

The gas compressibility factor[Z] of the mixture is calculated as,

$$Z^3 - Z^2 + [0.1248 - 0.04916 - 0.04916^2]Z - (0.1248 \times 0.04916) = 0 \quad (58)$$

$$Z^3 - Z^2 + 0.0732Z - 0.0061 = 0 \quad (59)$$

Solving for Z using Goal seek in MS-Excel,

$$Z = 0.9283 \quad (60)$$

The molar volume [V_m] can be estimated as,

$$P - \frac{0.08314\times333.15}{[V_m-0.03404]} - \frac{2.393}{V_m[V_m+0.03404]} = 0 \quad (61)$$

Solving for V_m using Goal seek in MS-Excel,

$$V_m = 0.6428\ m^3/kmol \quad (62)$$

The mixture density [ρ] is computed as,

$$\rho = \frac{21.17}{0.6428} = 32.92\ kg/m^3 \quad (63)$$

The enthalpy of the hydrocarbon mixture is computed based on dα(T)/dT as,

$$\frac{d\alpha_{CH_4}}{dT} = \frac{-0.5029[0.5029\times(1+1.3217)]}{1.3217} = -0.4443 \quad (64)$$

$$\frac{d\alpha_{C_2H_6}}{dT} = \frac{-0.6366[0.6366\times(1+1.0444)]}{1.0444} = -0.7932 \quad (65)$$

$$\frac{da_m}{dT} = \frac{0.2233}{2}[-0.4443 - 0.7932] = -0.138 \quad (66)$$

The above calculations are similarly performed for all other components, and the value of da_m/dT for all combinations are added.

$$\frac{da}{dT} = \sum_i \sum_j \frac{da_m}{dT} = -1.5842 \frac{m^6 bar}{kmol^2.K} \quad (67)$$

The mixture specific heat is calculated as,

$$\left[\sum x_i C_{p,i}\right](T - T_{ref}) = 41.49 \times [60 - 25] \quad (68)$$

$$\left[\sum x_i C_{p,i}\right](T - T_{ref}) = 1452.1\ kJ/kmol \quad (69)$$

The residual molar enthalpy is calculated as,

$$H^R = \left[[0.08314 \times (60 + 273.15)] \times [0.9283 - 1]\right] + \left[\frac{[(60+273.15)\times(-1.5842)]-2.393}{0.03404} \times ln\left[\frac{0.9283+0.04916}{0.9283}\right]\right] = -805.66\ kJ/kmol \quad (70)$$

Adding all the terms, molar enthalpy [H_m] is

$$H_m = -86{,}499 + 1452.1 - 805.66 \quad (71)$$

$$H_m = -85{,}852.2\ kJ/kmol \quad (72)$$

The mass enthalpy [H_{mass}] becomes,

$$H_{mass} = \frac{-85{,}852.2}{21.17} = -4{,}056\ kJ/kg \quad (73)$$

The gas viscosity is estimated as,

ρ = 32.92 kg/m^3 =0.03292 g/cm^3

T = 60^{0}C ≃ 600^{0}R

MW_g = 21.17

$$K_1 = \frac{\left[0.00094+\left[(2\times10^{-6})\times21.17\right]\right]\times600^{1.5}}{[209+(19\times21.17)+600]} \quad (74)$$

$$K_1 = 0.0119204 \quad (75)$$

$$X = 3.5 + \frac{986}{600} + (0.01 \times 21.17) = 5.355 \quad (76)$$

$$Y = 2.4 - [0.2 \times 5.355] = 1.329 \quad (77)$$

$$\mu_g = 0.0119204 \times e^{5.355\times0.03292^{1.329}} \quad (78)$$

$$\mu_g = 0.01262\ cP \quad (79)$$

Restriction Orifice [RO] Sizing

The restriction orifice [RO] is installed in a pipe of ID 154.10 mm with a flange tapping. For a flow rate of 1,000 kmol/h, the required RO size is,

$$A_{pipe} = \frac{\pi}{4} \times \left(\frac{154.10}{1000}\right)^2 = 0.01865\ m^2 \quad (80)$$

The required volumetric flow rate [$q_{v,req}$] is,

$$q_{v,req} = \frac{1000 \times 21.17}{32.92} = 642.8\ m^3/h \quad (81)$$

The fluid velocity is,

$$V = \frac{q_{v,req}}{A_{pipe}} = \frac{642.8}{0.01865 \times 3600} = 9.6\ m/s \quad (82)$$

The fluid mass specific heat is assumed to change very little between RO inlet and outlet

$$C_{p,1\ atm,25℃} = 41.49 \frac{kJ}{kmol}.K \quad (83)$$

The Reynolds number is calculated as,

$$Re = \frac{ID_{Pipe} \times V \times \rho}{\mu_g} = \frac{0.1541 \times 9.6 \times 32.92}{(0.0126/1000)} \quad (84)$$

$$Re = 3{,}849{,}878 \quad (85)$$

The specific heat ratio is computed as,

$$k = \frac{C_p}{C_v} = \frac{C_p}{C_p - R} = \frac{41.49}{41.49 - 8.314} = 1.2506 \quad (86)$$

To estimate the orifice diameter [d], an iterative approach is taken up. With a guess value of 'd', the equations become

The diameter ratio [β] is calculated as,

$$\beta = \frac{d}{D} = \frac{d}{0.1541} \quad (87)$$

$$\varepsilon = 1 - \left[\left[0.351 + 0.256\left(\frac{d}{0.1541}\right)^4 + 0.93\left(\frac{d}{0.1541}\right)^8\right] \times \left[1 - \left(\frac{4}{40}\right)^{\frac{1}{1.2506}}\right]\right] \quad (88)$$

For flange tappings, $L_1 = L'_2 = 25.4/D$ [mm]

$$L_1 = L'_2 = \frac{25.4}{154.1} = 0.1648 \quad (89)$$

$$M'_2 = \frac{2 \times 0.1648}{1 - \frac{d}{0.1541}} \quad (90)$$

$$A = \left[\frac{19000 \times \frac{d}{0.1541}}{3{,}849{,}878}\right]^{0.8} \quad (91)$$

$$C = 0.5961 + 0.0261\left(\frac{d}{0.1541}\right)^2 - 0.216\left(\frac{d}{0.1541}\right)^8 + \left[0.000521\left(\frac{10^6\left(\frac{d}{0.1541}\right)^2}{3{,}849{,}878}\right)^{0.7}\right] + \left[\left(0.0188 + 0.0063\left[\frac{19000\times\frac{d}{0.1541}}{3{,}849{,}878}\right]^{0.8}\right)\frac{d}{0.1541}^{3.5}\left(\frac{10^6}{3{,}849{,}878}\right)^{0.3}\right] + \left[(0.043 + 0.08e^{-1.648} - 0.123e^{-1.1536})\left(1 - 0.11\left[\frac{19000\times\frac{d}{0.1541}}{3{,}849{,}878}\right]^{0.8}\right)\left(\frac{\frac{d}{0.1541}^4}{1-\frac{d}{0.1541}^4}\right)\right] - \left[0.031\left(\frac{2\times0.1648}{1-\frac{d}{0.1541}} - 0.8\times\frac{2\times0.1648}{1-\frac{d}{0.1541}}^{1.1}\right)\frac{d}{0.1541}^{1.3}\right] \quad (92)$$

$$q_v - \frac{\varepsilon d^2 C}{\sqrt{1-\beta^4}} \times \frac{\pi}{4} \times \sqrt{\frac{2g[P_1 - P_2]}{\rho}} = 0 \quad (93)$$

Solving the above set of equations iteratively, the orifice diameter [d] and β for a volumetric flow rate of 642.8 m^3/h [1000 kmol/h] is,

$$d = 34.3\ mm \quad (94)$$

$$\beta = \frac{34.3}{0.1541} = 0.2226 \quad (95)$$

The discharge temperature of the hydrocarbon fluid exiting the restriction orifice is computed for a discharge pressure of 4 bara using SRK EoS. The discharge temperature at exit of the orifice comes to 42.5^0C

References

1. https://www.linkedin.com/pulse/600mm-sufficient-keep-blowdown-valve-functional-saeid-rahimi-mofrad/
2. "The Properties of Gases and Liquids", Bruce Poling, John M. Prausnitz, John P. O'Connell, 5th Edition
3. https://wiki.pengtools.com/index.php?title=Lee_correlation
4. "Measurement of fluid flow by means of pressure differential devices inserted in circular cross-section conduits running full" – Part 2: Orifice Plates, ISO 5167-2:2003(E)

Author

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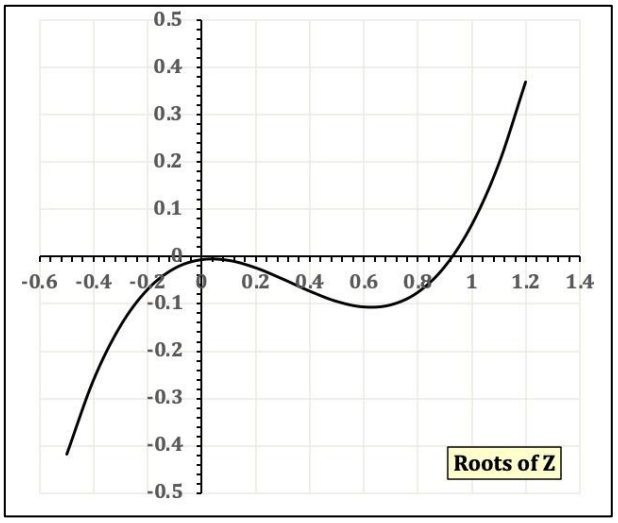
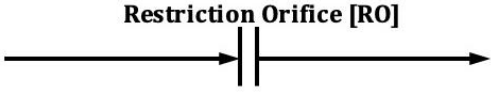
Natural Gas Conditions		
Parameter	**Value**	**Units**
Gas Constant [R]	0.08314	m^3.bar/kmol.K
Standard Pressure [P_{std}]	1.01325	bar(a)
Reference Temperature [T_{ref}]	25.0	^{0}C
Flowing Pressure [P]	**40.0**	bar(a)
Flowing Temperature [T]	**60.0**	^{0}C
	333.15	K
Exit Pressure [T_{exit}]	**4.0**	bara
Molar Flow Rate [q_m]	**1,000**	kmol/h
Soave Redlich Kwong (SRK) EoS Calculations		
Mixture Attraction Parameter [a_m]	2.3930	m^6.bar/$kmol^2$
Mixture Co-volume Parameter [b_m]	0.03404	m^3/kmol
SRK A Factor	0.1248	-
SRK B Factor	0.04916	-
[A-B-B^2]	0.0732	-
AB	0.0061	-
Z^3-Z^2+(A-B-B^2)Z-AB =0	0.0000	-
Compressibility Factor [Z] [-]	**0.9283**	**Solve for Z, V_m, T_{exit}**
P - [RT/(V_m-b_m)] - [a_m/(V_m(V_m+b)] = 0	0.0000	
Molar Volume [V_m] [m^3/kmol]	**0.6428**	
Density [ρ_m]	**32.92**	kg / m^3
∑ [da/dT]	-1.5842	m^6.bar/$kmol^2$.K
∑ C_p.dT [T_{ref} = 298.15 K]	1452.1	kJ/kmol
Residual Molar Enthalpy [$H_{,R}$]	-805.66	kJ/kmol
Molar Enthalpy [H] = H_{id} + ∑C_p.dT + $H_{,R}$	-85,852	kJ/kmol
Mass Enthalpy [$H_{,mass}$]	**-4,056**	kJ/kg
Viscosity [**μ**]	**0.0126**	cP



Restriction Orifice [RO]



Inlet Pressure [P_{inlet}]	40.0	bar(a)	Exit Pressure [P_{exit}]	4.0	bar(a)
Inlet Temperature [T_{inlet}]	60.0	^{0}C	**Exit Temperature [T_{exit}]**	**42.5**	^{0}C
Inlet Mass Enthalpy [$H_{,inlet,mass}$]	-4,056	kJ/kg	Exit Mass Enthalpy [$H_{,exit,mass}$]	-4,056	kJ/kg
Molar Flow Rate [q_m]	1,000	kmol/h	Molar Flow Rate [q_m]	1,000	kmol/h

Soave Redlich Kwong [SRK] Equation of State [EoS]

Component	MW	Mol%	Cp	T_c	P_c	V_c	Omega [ω]	Ideal gas Enthalpy [H_{id}] at 25 °C [ΔH_f]		$\sqrt{T_r}$	m_i	$\alpha(T)$	a_i	b_i	a_m	b_m	$d\alpha(T)/dT$
	[kg/kmol]	[%]	[kJ/kmol.K]	[K]	[bara]	[m^3/kmol]	[-]	[kJ/kmol]		-	[-]	[-]	[m^6.bar/$kmol^2$]	[m^3/kmol]	[m^6.bar/$Kmol^2$]	[m^3/kmol]	[-]
Methane	16.043	76.23	35.97	190.7	46.41	0.0990	0.0115	-74,900	-57,096	1.3217	0.5029	0.7026	1.6270	0.0296	-	0.0226	-0.4443
Ethane	30.070	10.00	53.12	305.4	48.84	0.1480	0.0986	-84,738	-8,474	1.0444	0.6366	0.9443	5.3300	0.0451	-	0.0045	-0.7932
Propane	44.097	5.00	75.42	369.9	42.57	0.2000	0.1524	-103,890	-5,195	0.9490	0.7179	1.0745	10.2067	0.0626	-	0.0031	-1.0585
i-Butane	58.124	1.00	98.30	408.1	36.48	0.2630	0.1848	-134,590	-1,346	0.9035	0.7665	1.1534	15.5617	0.0806	-	0.0008	-1.2377
n-Butane	58.124	1.00	98.91	425.2	37.97	0.2550	0.2010	-126,190	-1,262	0.8852	0.7907	1.1898	16.7436	0.0807	-	0.0008	-1.3314
i-Pentane	72.151	0.30	165.53	460.4	33.34	0.3080	0.2222	-154,590	-464	0.8507	0.8222	1.2607	23.6881	0.0995	-	0.0003	-1.4708
n-Pentane	72.151	0.10	167.79	469.6	33.75	0.3110	0.2539	-146,490	-146	0.8423	0.8690	1.2929	24.9635	0.1002	-	0.0001	-1.6517
n-Hexane	86.178	0.05	191.94	507.9	30.32	0.3680	0.3007	-167,290	-84	0.8099	0.9376	1.3882	34.9070	0.1207	-	0.0001	-1.9644
H_2O	18.015	0.25	79.93	647.3	221.20	0.0571	0.3440	-241,814	-605	0.7174	1.0004	1.6453	9.2098	0.0211	-	0.0001	-2.3958
CO_2	44.010	3.00	38.50	304.1	73.70	0.0939	0.2389	-393,790	-11,814	1.0467	0.8469	0.9225	3.4206	0.0297	-	0.0009	-1.4026
H_2S	34.081	0.07	34.54	373.6	90.08	0.0980	0.0810	-20,179	-14	0.9443	0.6097	1.0691	4.8952	0.0299	-	0.0000	-0.7655
N_2	28.013	3.00	29.19	126.2	33.94	0.0900	0.0400	0	0	1.6248	0.5469	0.4334	0.6008	0.0268	-	0.0008	-0.4832
Avg MW	**21.17**	**100.00**	**41.49**					**Total**	**-86,499**					**Total**	**2.3930**	**0.0340**	

Binary Interaction Factors [BIP] [K_{ij}] [Cheu-Prausnitz Correlation]

BIP [K_{ij}] i↓\ j→		CH_4	C_2H_6	C_3H_8	i-C_4	n-C_4	i-C_5	n-C_5	n-C_6	H_2O	CO_2	H_2S	N_2
		0.0990	0.1480	0.2000	0.2630	0.2550	0.3080	0.3110	0.3680	0.0571	0.0939	0.0980	0.0900
CH_4	0.0990	0.000	0.007	0.020	0.039	0.036	0.052	0.053	0.069	0.013	0.000	0.000	0.000
C_2H_6	0.1480	0.007	0.000	0.004	0.014	0.012	0.022	0.023	0.034	0.037	0.009	0.007	0.010
C_3H_8	0.2000	0.020	0.004	0.000	0.003	0.002	0.008	0.008	0.015	0.063	0.023	0.021	0.026
i-C_4	0.2630	0.039	0.014	0.003	0.000	0.000	0.001	0.001	0.005	0.092	0.043	0.040	0.047
n-C_4	0.2550	0.036	0.012	0.002	0.000	0.000	0.001	0.002	0.006	0.088	0.041	0.037	0.044
i-C_5	0.3080	0.052	0.022	0.008	0.001	0.001	0.000	0.000	0.001	0.110	0.057	0.053	0.061
n-C_5	0.3110	0.053	0.023	0.008	0.001	0.002	0.000	0.000	0.001	0.111	0.058	0.054	0.062
n-C_6	0.3680	0.069	0.034	0.015	0.005	0.006	0.001	0.001	0.000	0.133	0.074	0.070	0.079
H_2O	0.0571	0.013	0.037	0.063	0.092	0.088	0.110	0.111	0.133	0.000	0.010	0.012	0.009
CO_2	0.0939	0.000	0.009	0.023	0.043	0.041	0.057	0.058	0.074	0.010	0.000	0.000	0.000
H_2S	0.0980	0.000	0.007	0.021	0.040	0.037	0.053	0.054	0.070	0.012	0.000	0.000	0.000
N_2	0.0900	0.000	0.010	0.026	0.047	0.044	0.061	0.062	0.079	0.009	0.000	0.000	0.000

a_{ij} Table [$a_{ij} = (\sqrt{a_i.a_j}) . (1-k_{ij})$]

$a_{ij} = (\sqrt{a_i.a_j}) \times (1-k_{ij})$ $a_i \rightarrow$ $a_j \downarrow$		CH_4	C_2H_6	C_3H_8	i-C_4	n-C_4	i-C_5	n-C_5	n-C_6	H_2O	CO_2	H_2S	N_2
		1.6270	5.3300	10.2067	15.5617	16.7436	23.6881	24.9635	34.9070	9.2098	3.4206	4.8952	0.6008
CH_4	1.6270	1.627	2.925	3.992	4.836	5.029	5.886	6.036	7.018	3.822	2.359	2.822	0.988
C_2H_6	5.3300	2.925	5.330	7.348	8.983	9.331	10.988	11.274	13.178	6.748	4.233	5.072	1.771
C_3H_8	10.2067	3.992	7.348	10.207	12.564	13.041	15.429	15.833	18.586	9.085	5.770	6.921	2.412
i-C_4	15.5617	4.836	8.983	12.564	15.562	16.141	19.180	19.687	23.198	10.874	6.982	8.382	2.915
n-C_4	16.7436	5.029	9.331	13.041	16.141	16.744	19.886	20.411	24.041	11.322	7.261	8.716	3.032
i-C_5	23.6881	5.886	10.988	15.429	19.180	19.886	23.688	24.317	28.718	13.142	8.491	10.199	3.544
n-C_5	24.9635	6.036	11.274	15.833	19.687	20.411	24.317	24.963	29.485	13.473	8.708	10.461	3.634
n-C_6	34.9070	7.018	13.178	18.586	23.198	24.041	28.718	29.485	34.907	15.550	10.117	12.160	4.219
H_2O	9.2098	3.822	6.748	9.085	10.874	11.322	13.142	13.473	15.550	9.210	5.555	6.633	2.332
CO_2	3.4206	2.359	4.233	5.770	6.982	7.261	8.491	8.708	10.117	5.555	3.421	4.092	1.433
H_2S	4.8952	2.822	5.072	6.921	8.382	8.716	10.199	10.461	12.160	6.633	4.092	4.895	1.714
N_2	0.6008	0.988	1.771	2.412	2.915	3.032	3.544	3.634	4.219	2.332	1.433	1.714	0.601

$a_m = x_i . x_j . a_{ij}$

$x_i . x_j . a_{ij}$ [i↓\ j→]		CH_4	C_2H_6	C_3H_8	i-C_4	n-C_4	i-C_5	n-C_5	n-C_6	H_2O	CO_2	H_2S	N_2
		0.7623	0.1000	0.0500	0.0100	0.0100	0.0030	0.0010	0.0005	0.0025	0.0300	0.0007	0.0300
CH_4	0.7623	0.945	0.223	0.152	0.037	0.038	0.013	0.005	0.003	0.007	0.054	0.002	0.023
C_2H_6	0.1000	0.223	0.053	0.037	0.009	0.009	0.003	0.001	0.001	0.002	0.013	0.000	0.005
C_3H_8	0.0500	0.152	0.037	0.026	0.006	0.007	0.002	0.001	0.000	0.001	0.009	0.000	0.004
i-C_4	0.0100	0.037	0.009	0.006	0.002	0.002	0.001	0.000	0.000	0.000	0.002	0.000	0.001
n-C_4	0.0100	0.038	0.009	0.007	0.002	0.002	0.001	0.000	0.000	0.000	0.002	0.000	0.001
i-C_5	0.0030	0.013	0.003	0.002	0.001	0.001	0.000	0.000	0.000	0.000	0.001	0.000	0.000
n-C_5	0.0010	0.005	0.001	0.001	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
n-C_6	0.0005	0.003	0.001	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
H_2O	0.0025	0.007	0.002	0.001	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
CO_2	0.0300	0.054	0.013	0.009	0.002	0.002	0.001	0.000	0.000	0.000	0.003	0.000	0.001
H_2S	0.0007	0.002	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
N_2	0.0300	0.023	0.005	0.004	0.001	0.001	0.000	0.000	0.000	0.000	0.001	0.000	0.001

da/dT

$da/dT = x_i.x_j.0.5.(\sqrt{a_i.a_j}) . (1-k_{ij}) . [(d\alpha_i/dT) + (d\alpha_j/dT)]$ $(d\alpha i/dT) \rightarrow$ $(d\alpha j/dT) \downarrow$		CH_4	C_2H_6	C_3H_8	i-C_4	n-C_4	i-C_5	n-C_5	n-C_6	H_2O	CO_2	H_2S	N_2
		-0.4443	-0.7932	-1.0585	-1.2377	-1.3314	-1.4708	-1.6517	-1.9644	-2.3958	-1.4026	-0.7655	-0.4832
CH_4	-0.4443	-0.420	-0.138	-0.114	-0.031	-0.034	-0.013	-0.005	-0.003	-0.010	-0.050	-0.001	-0.010
C_2H_6	-0.7932	-0.138	-0.042	-0.034	-0.009	-0.010	-0.004	-0.001	-0.001	-0.003	-0.014	0.000	-0.003
C_3H_8	-1.0585	-0.114	-0.034	-0.027	-0.007	-0.008	-0.003	-0.001	-0.001	-0.002	-0.011	0.000	-0.003
i-C_4	-1.2377	-0.031	-0.009	-0.007	-0.002	-0.002	-0.001	0.000	0.000	0.000	-0.003	0.000	-0.001
n-C_4	-1.3314	-0.034	-0.010	-0.008	-0.002	-0.002	-0.001	0.000	0.000	-0.001	-0.003	0.000	-0.001
i-C_5	-1.4708	-0.013	-0.004	-0.003	-0.001	-0.001	0.000	0.000	0.000	0.000	-0.001	0.000	0.000
n-C_5	-1.6517	-0.005	-0.001	-0.001	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
n-C_6	-1.9644	-0.003	-0.001	-0.001	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
H_2O	-2.3958	-0.010	-0.003	-0.002	0.000	-0.001	0.000	0.000	0.000	0.000	-0.001	0.000	0.000
CO_2	-1.4026	-0.050	-0.014	-0.011	-0.003	-0.003	-0.001	0.000	0.000	-0.001	-0.004	0.000	-0.001
H_2S	-0.7655	-0.001	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
N_2	-0.4832	-0.010	-0.003	-0.003	-0.001	-0.001	0.000	0.000	0.000	0.000	-0.001	0.000	0.000

Pag

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FCC Technologies as Flexible Strategy to Maximize Refining Margins under a Hostile Scenario for Fossil Fuels

Dr. Marcio Wagner da Silva

Introduction and Context

The downstream industry faces a transition period where the focus of the players is changing from transportation fuels to petrochemicals aiming to ensure maximum added value to processed crude oils as well as to allow the growth of low carbon energies in the global energetic matrix.

The growing market of petrochemicals have been lead some refiners to look for a closer integration between refining and petrochemicals assets aiming to reach more adherence with the market demand, improve revenues, and reduce operation costs. In this business environment, flexible refining technologies like Fluid Catalytic Cracking (FCC) reach highlighted position in the strategy of the refiners to reach competitiveness in the market.

Facing these challenges, search for alternatives that ensure survival and sustainability of the refining industry became constant by refiners and technology developers. Due to his similarities, better integration between refining and petrochemical production processes appears as an attractive alternative. Although the advantages, it's important take into account that the integration between refining and petrochemical assets increase the complexity, requires capital spending, and affect the interdependency of refineries and petrochemical plants, these facts need to be deeply studied and analyzed case by case. In this business environment, flexible refining technologies like Fluid Catalytic Cracking (FCC) can ensure high competitiveness to refiners once are capable to produce high quality intermediates both to petrochemicals and transportation fuels, in markets with great demand by petrochemicals, the petrochemical FCC technologies can be an attractive option, despite the high capital spending. Refiners with restriction of capital investment can even maximize the profitability of FCC units through optimization actions aiming to maximize the yield of petrochemicals against transportation fuels.

Fluid Catalytic Cracking Technologies – An Overview

Fluid Catalytic Cracking (FCC) is one of the main processes which give higher operational flexibility and profitability to refiners. The catalytic cracking process was widely studied over last decades and became the principal and most employed process dedicated to converting heavy oil fractions in higher economic value streams.

The installation of catalytic cracking units allows the refiners to process heavier crude oils and consequently cheaper, raising the refining margin, mainly in higher crude oil prices scenario or in geopolitics crises that can become difficult the access to light oils. The typical Catalytic Cracking Unit feedstream is gas oils from vacuum distillation process. However, some variations are found in some refineries, like sending heavy coke naphtha, coke gas oils and deasphalted oils from solvent deasphalting units to processing in the FCC unit.

The catalyst normally employed in fluid catalytic cracking units is a solid constituted by small particles of alumina (Al_2O_3) and silica (SiO_2) (zeolite). By the catalyst characteristics and the operational conditions in the catalytic cracking process (temperature higher than 500 ^oC), the process is inefficient to cracking aromatic compounds, therefore, how much more paraffinic is the feedstream, higher is the unit conversion. Figure 1 presents a process scheme for a typical Fluid Catalytic Cracking Unit (FCCU).

In a conventional scheme, the catalyst regeneration process consists in the carbon partial burning deposited over the catalyst, according to chemical reaction below:

$C + ½\ O_2 \rightarrow CO$

The carbon monoxide is burned in a boiler capable of generating higher pressure steam that supplies others process units in the refinery.

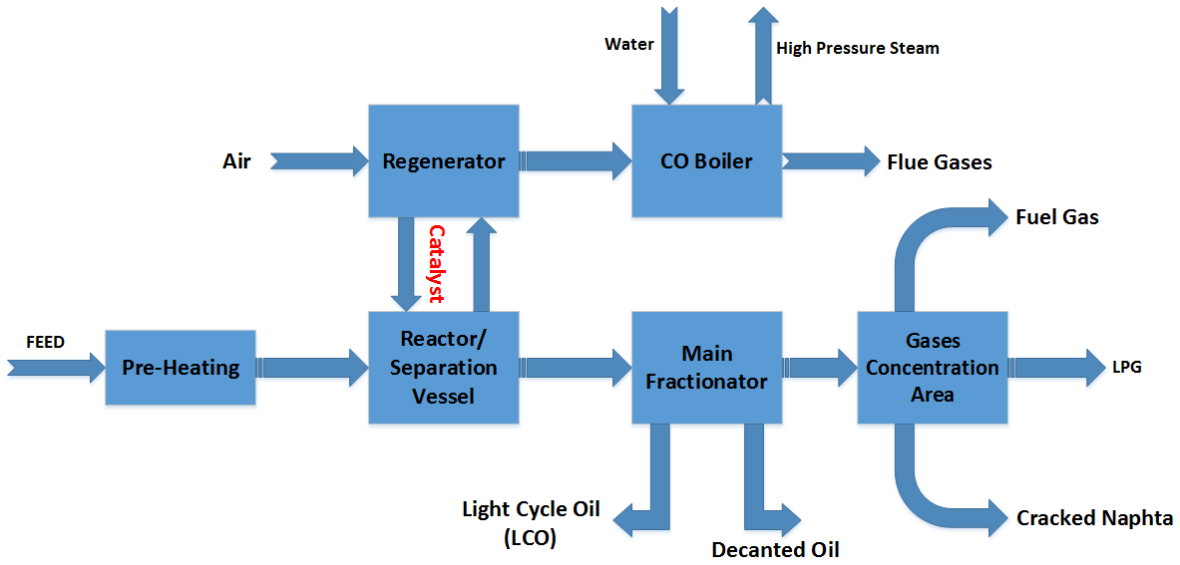

Figure 1 – Schematic Process Flow for a Typical Fluid Catalytic Cracking Process Unit (FCCU)

The principal operational variables in a fluid catalytic cracking unit are reaction temperature, normally considered the temperature in the top of the reactor (called riser), feed stream temperature, feed stream quality (mainly carbon residue), feed stream flow rate and catalyst quality. Feedstock quality is especially relevant, but this variable is a function of the crude oil processed by the refinery, so is difficultly can be changed, but for example, aromatic feedstock's with high metals content are refractory to cracking and conducting to a quick catalyst deactivation.

An important variation of the fluid catalytic cracking technology is the residue fluid catalytic cracking unit (RFCC). In this case, the feedstock to the process is basically residue from atmospheric distillation column, due to the high carbon residue and contaminants (metals, Sulphur, nitrogen, etc.) are necessary some adaptations in the unit like catalyst with higher resistance to metals and nitrogen and catalyst coolers furthermore, it's necessary apply materials with most noble metallurgy due the higher temperatures reached in the catalyst regeneration step (due the higher coke quantity deposited on the catalyst), that raises significantly the capital investment to the unit installation. Nitrogen is a strong contaminant to the FCC catalyst because they neutralize the acid sites of the catalyst which are responsible for the cracking reactions.

When the residue has high contaminants content, is common the feed stream treatment in hydrotreating units to reduce the metals and heteroatoms concentration to protect the FCC catalyst.

Typically, the average yield in fluid catalytic cracking units is 55% in volume in cracked naphtha and 30 % in LPG. Figure 2 presents a scheme for the main fractionator of the FCC unit with the principal product streams.

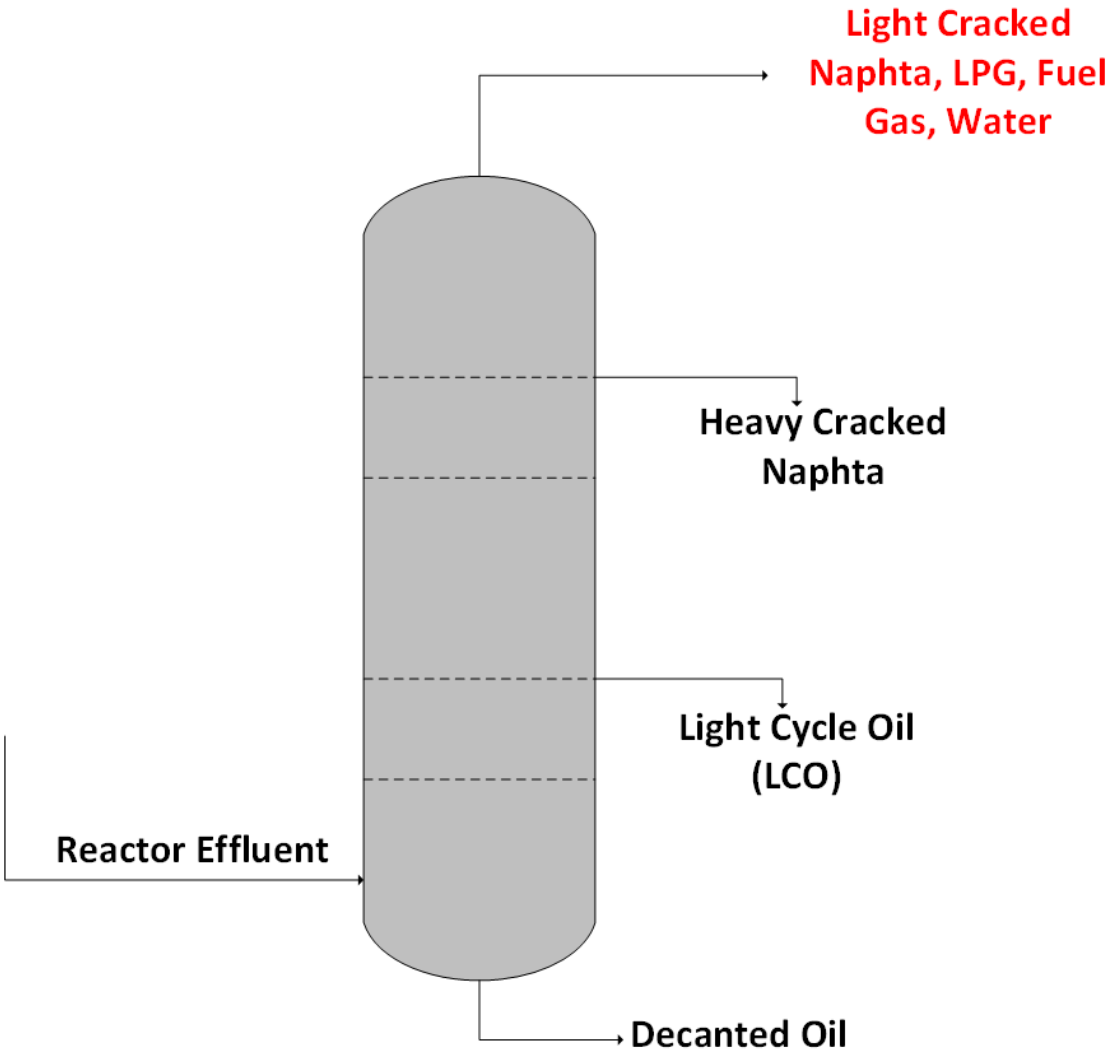

Figure 2 – Main Fractionator Scheme for a Typical Fluid Catalytic Cracking Unit

The decanted oil stream contain the heavier products and have high aromatic content, is common that this product are contaminated with catalyst fines and normally this stream is directed to use like fuel oil diluent, but in some refineries, this stream can be used to produce black carbon.

Light Cycle Oil (LCO) has a distillation range close to diesel and normally this stream is directed to treatment in severe hydrotreating units (due to the high aromaticity), after this treatment the LCO is sent to the refinery diesel pool.

Heavy cracked naphtha is normally directed to refinery gasoline pool, however, in scenarios where the objective is to raise the production of middle distillates, this stream can be sent to hydrotreating units for further diesel production.

The overhead products from main fractionator are still in gaseous phase and are sent to the gas separation section. The fuel gas is sent to the refinery fuel gas ring, after treatment to remove H_2S, where will be burned in fired heaters while the LPG is directed to treatment (MEROX) and further commercialization. The LPG produced by FCC unit have a high content of light olefins (mainly Propylene) so, in some refineries, the LPG stream is processed in a Propylene separation unit to recovery the propylene that has higher added value than LPG.

Cracked naphtha is usually sent to refinery gasoline pool which is formed by naphtha produced by other process units like straight run naphtha, naphtha from the catalytic reforming unit, etc. Due to the production process (deep conversion of residues), the cracked naphtha has high sulfur content and to attend the currently environmental legislation this stream needs to be processed to reducing the contaminants content, mainly sulfur.

The cracked naphtha sulfur removing represents a great technology challenge because is necessary to remove the sulfur components without molecules saturation that gives high octane number for gasoline (mainly olefins).

Over the last decades some technology licensers had developed new processes aiming to reduce the sulfur content in the cracked naphtha with minimum octane number loss, some of the main technologies dedicated for this purpose are technology PRIME G+ ™ from Axens, the processes OCTAGAIN ™ and SCANfining ™ from Exxon Mobil, the process S-Zorb™ from ConocoPhillips and ISAL™ technology from UOP.

Usually, catalytic cracking units are optimized to aiming the production of fuels (mainly gasoline), however, some process units are optimized to maximize the light olefins production (propylene and ethylene). Process units dedicated for this purpose have his project and operational conditions significantly changed once the process severity is strongly raised in this case.

The reaction temperature reaches 600 oC and higher catalyst circulation rate raises the gases production, which requires a scaling up of gas separation section. Figure 3 presents a typical scheme for a gas separation section for a fluid catalytic cracking unit.

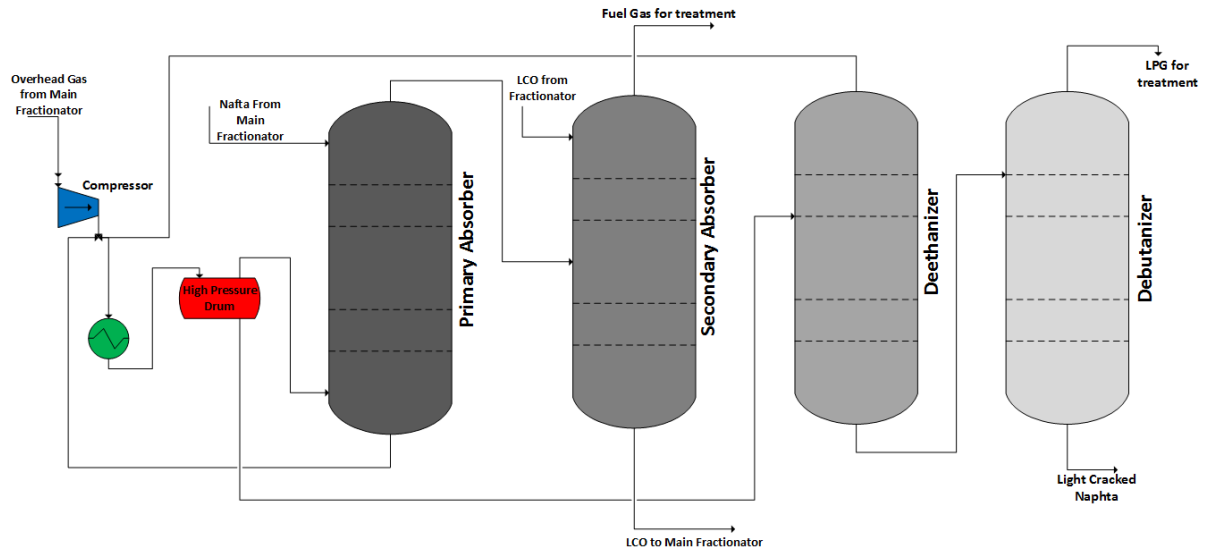

Figure 3 – Basic Process Flow Diagram for a Typical Gas Separation Section from FCC Unit

In several cases, due the higher heat necessity of the unit is advantageous to operate the regenerator with the total combustion of the coke deposited on the catalyst, this arrangement

changes significantly the thermal balance of the refinery once it's no longer possible to resort the steam produced by the CO boiler.

Over last decades, the fluid catalytic cracking technology was intensively studied aiming mainly the development of units capable of producing light olefins (Deep Catalytic Cracking) and to process heavier feedstocks. The main licensers for fluid catalytic cracking technology nowadays are the companies KBR, UOP, STONE & WEBSTER, AXENS, and Lummus.

Residue Fluid Catalytic Cracking (RFCC) Technologies – Dealing with Heavy Feeds

One variation of the fluid catalytic cracking that has been widely applied in the last years is the Residue Fluid Catalytic Cracking (RFCC). In this case, the feed stream to the process is basically the bottom stream from the atmospheric distillation column, called atmospheric residue, that have high carbon residue and higher contaminants content like metals, nitrogen, and sulfur.

Due to the feed stream characteristics, the residue catalytic cracking units require design and optimization changes. The higher levels of residual carbon in the feed stream led to higher temperatures in the catalyst regeneration step and a lower catalyst circulation rate to keep the reactor in constant temperature, this fact reduces the catalyst/oil ratio that leads to a lower conversion and selectivity. To avoid these effects, the RFCC units normally rely on catalyst coolers, as presented in Figure 4.

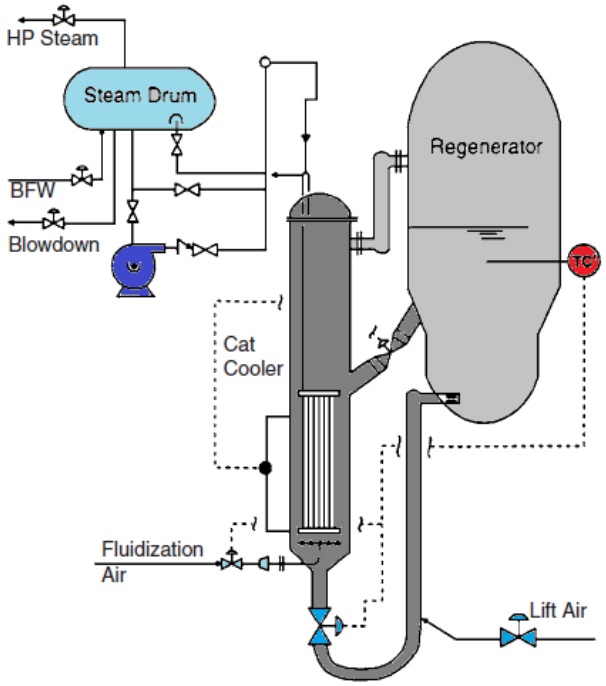

Figure 4 – Catalyst Cooler Process Arrangement for a Typical RFCC Unit (Handbook of Petroleum Refining Processes, 2004)

Installation of catalyst cooler system raises the process unit profitability through the total conversion enhancement and selectivity to noblest products as propylene and naphtha against gases and coke production, furthermore, helps the refinery thermal balance, once produces high-pressure steam. The use of catalyst cooler is also necessary when the unit is designed to operate

under total combustion mode, in this case, the heat release rate is higher due to the total burn of carbon to CO_2, as presented below.

$C + ½\ O_2 \rightarrow CO$ (Partial Combustion) $\Delta H = -27$ kcal/mol

$C + O_2 \rightarrow CO_2$ (Total Combustion) $\Delta H = -94$ kcal/mol

In this case, the temperature of the regeneration vessel can reach values close to 760 oC, leading to higher risks of catalyst damage which is minimized through catalyst cooler installation. The option by the total combustion mode needs to consider the refinery thermal balance, once, in this case, will not the possibility to produce steam in the CO boiler, furthermore, the higher temperature in the regenerator requires materials with noblest metallurgy, this significantly raises the installation costs of these units.

As pointed earlier, the feed streams characteristics to RFCC units require modifications when compared with the conventional fluid catalytic cracking. The presence of higher content of nitrogen compounds leads to an accelerated process of catalyst deactivation through acid sites neutralization, the presence of metals like nickel, sodium, and vanadium raise the coke deposition on the catalyst and lead to a higher production of hydrogen and gases, besides that, reduces the catalyst lifecycle through the zeolitic matrix degradation. Beyond these factors, heavier feed streams normally have high aromatics content that are refractory to the cracking reactions, leading to a higher coke deposition rate and lower conversion.

Due to this operation conditions, the residue fluid catalytic cracking units presents higher catalyst consumption when compared with the conventional process, this fact raises considerably the operational costs of the RFCC units. However, the most modern units have applied specific catalysts to process residual feed streams, in this case, the catalyst has a higher porosity aiming to allow a better adaptation to the high aromatics content, furthermore, the catalyst needs to have a higher metals tolerance.

The control of contaminants content in the feed stream or his effects is a fundamental step to the residue fluid catalytic cracking process. Sodium content can be minimized through an adequate crude oil desalting process and the effects of nickel (dehydrogenation reactions) can be reduced by dosage of antimony compounds that act like neutralizing agent of the nickel dehydrogenation activity, reducing the generation of low added value gases, in its turn, the vanadium effects can be controlled through the addition of rare earth to the catalyst, like cerium compounds. The addition of these compounds needs to be deeply studied once significantly raises the catalyst cost.

The use of visbreaking units to treat the feed streams to RFCC units is a process scheme adopted by some refiners, in these cases, the most significant effect in the reduction in the residual carbon, however, due to his higher effectiveness, the tendency in the last decades is to treat the bottom barrels streams in deep hydrotreating or hydrocracking units before to pump for RFCC units, with this processing scheme it's possible to achieve lower contaminants content, mainly metals, leading to a higher catalyst lifecycle. Furthermore, the hydroprocessing has the advantage of the reduction of the sulfur content in the unit intermediate streams, minimizing the necessity or severity of posterior treatments, a clear disadvantage of this refining scheme is the high hydrogen consumption that significantly raises the operational costs.

Like to the conventional FCC units, the main operational variables to RFCC units are the reaction temperature, normally considered in the highest point in the reactor (also called riser), feed

stream temperature, feed stream quality, feed stream flow rate and catalyst quality. It's relevant to quote that the conventional FCC units can process atmospheric residue as the feed stream, however, it's necessary to control the contaminants content, mainly metals, which requires processing lighter crudes with higher costs that raise the operational costs and reduces the flexibility of the refiner in relation of the crude oil supplier.

Some of the most relevant residue fluid catalytic cracking technologies available commercially are the R2R™ by Axens Company, the INDMAX™ process licensed by Lummus Company and the RxPro™ process developed by the UOP Company. Figure 5 presents a block diagram showing a case study demonstrating how the petrochemical FCC unit, in this case the INDMAX™ technology by Lummus Company, can maximize the yield of petrochemicals in the refining hardware.

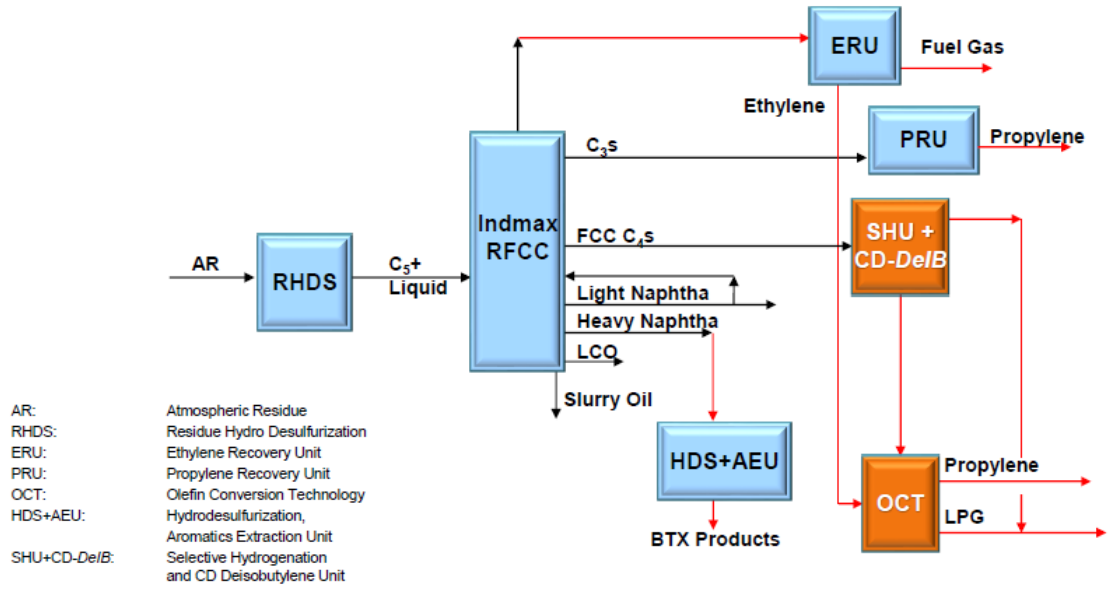

Figure 5 – Olefins Maximization in the Refining Hardware with INDMAX™ FCC Technology by Lummus Company (SANIN, A.K., 2017)

It's interesting to note that, in the case presented in Figure 5 the refiner can maximize olefins production from atmospheric residue, as aforementioned, due to the feedstock characteristic is necessary to apply a residue hydrotreating unit upstream to the FCC unit (the RHDS unit) to control the contaminants content to the FCC catalyst.

The FCC Catalyst – Converting Residues to Added Value Derivatives

A key factor in the FCC operation is the catalyst applied in the process. The catalyst normally employed in fluid catalytic cracking units is a solid constituted by small particles of alumina (Al_2O_3) and silica (SiO_2) (zeolite). By the catalyst characteristics and the operational conditions in the catalytic cracking process (temperature higher than 500 ^oC), the process is inefficient to cracking aromatic compounds, therefore, how much more paraffinic is the feed stream, higher is the unit conversion.

The active phase in the FCC catalyst is composed by the zeolite that is responsible by the catalytic activity and selectivity of the catalyst and by the alumina that is responsible by the cracking of heavier molecules allowing these molecules to reach the access to the zeolitic phase. The other

components of the FCC catalysts are the inert (kaolim) and synthetic matrixes that are responsible to the mechanical resistance, hardness, and act as binder agent between the active phases and the matrix.

According to the process conditions, can be added some compounds to the catalyst with specific purpose. In refineries that processing feed streams with high amount of nickel it's common to add antimony as that act as passivator agent, another deleterious metal is the vanadium, in this case is applied some trap agent to minimize his effects. The Figure 6 presents an arrangement of a typical design of FCC catalyst.

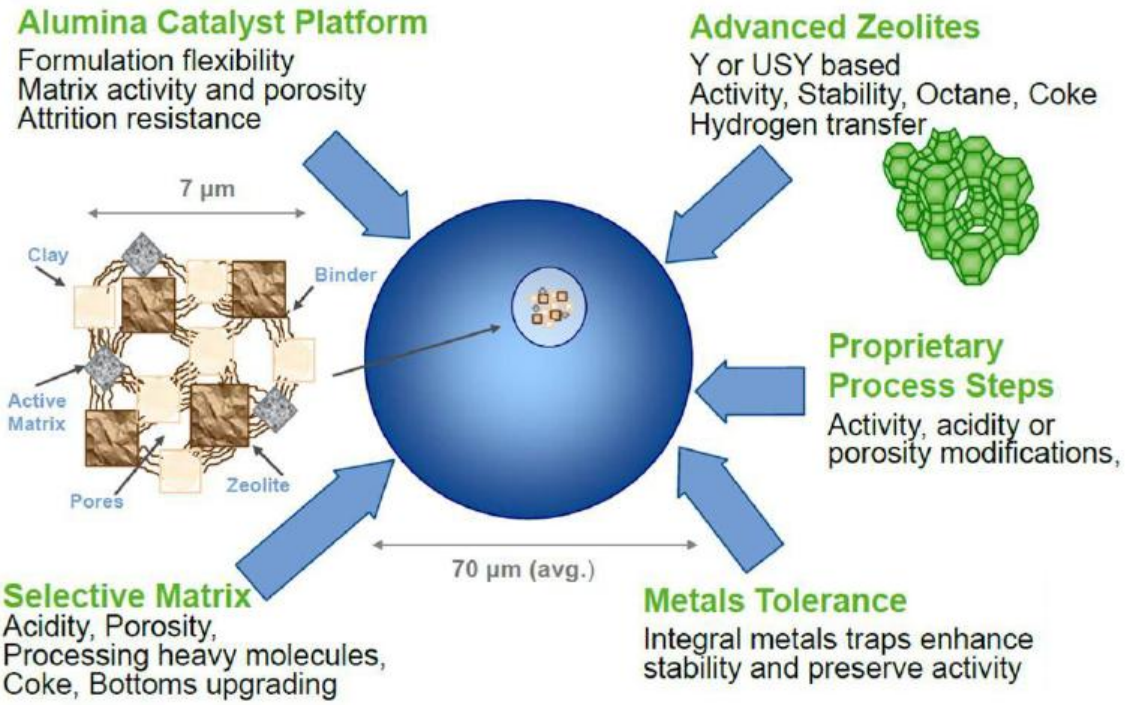

Figure 6 – Typical Design of FCC Catalyst (Grace Company, 2015)

As aforementioned, the processing of heavier crude oils leads to a more challenging feedstocks to FCC units due to the higher concentration of residual carbon and mainly contaminants as Nickel and Vanadium. The nickel acts as dehydrogenation agent leading to the coke deposition over the catalyst and raises the hydrogen production, normally the refiners used to process heavier feeds apply metals passivators as boron to keep under control the deleterious effect of metals, the most common form to control the nickel effects is to inject Antimony in the FCC feed.

The vanadium effect over the FCC catalyst involves the degradation of the zeolite matrix leading to the reduction in the catalytic activity and his action is keep under control through vanadium traps. In the last years some catalyst developers are focusing his research to study the effects of iron in the FCC catalyst, the high concentration of iron is a characteristic of the shale oils produced in the North America and the availability of these crudes raises significantly in the last years, especially after 2015, when the United States starts to export his internal production. The iron is not catalytically active, but this compound can accumulate over the catalyst surface reducing the porosity reducing the activity and leading to dehydrogenation reactions as well as carbon monoxide (CO) promoter, furthermore the high concentration of Iron can raise the SOx emissions in the catalyst regenerator.

Another dangerous contaminant of FCC catalyst is sodium, this compound promotes a irreversible deactivation of the catalyst through the chemical degradation of the zeolitic matrix. By this reason, an adequate control of the crude oil desalting process is fundamental to control the sodium content in the FCC feeds, preserving the catalyst lifecycle. Nowadays, some refiners are injected caustic soda in the crude to improve the desalting characteristics, and a stricter control is required in these cases. A less common contaminant founded in some crude oils is the copper, his effect is the promotion of dehydrogenation reactions, raising the yield of hydrogen and coke. The copper is present in some NOx reducing agents.

Aiming to improve the catalytic activity, some developers apply rare earth compounds to the FCC catalyst as Lanthanum and Cerium. These compounds raises significantly the activity and selectivity of the final catalyst, but his high cost made the refiners avoid his application, furthermore, the presence of rare earth in the catalyst improve yield of gasoline and reduces the light olefins production in FCC units, in the current scenario this is exactly the inverse that the refiners are looking for.

The trend of reduction in transportation fuels demand is making refiners to optimize his FCC units to maximize petrochemical intermediates against transportation fuels. To achieve this goal, normally the refiners are applying most severe conditions as higher catalyst/oil ratios, higher reaction temperature (TRX), and the use of ZSM-5 as additive to the catalyst.

The presence of ZSM-5 in the catalyst is capable to improve the yield of light olefins in the FCC unit by up to 8,0 %. One of the most important roles of the refineries optimization teams is to analyze the FCC equilibrium catalysts to find the improvement alternatives based on the contaminants content and the reached conversion of the unit as well as the degradation observed on the equilibrium catalysts. The volumetric conversion of an FCC unit is defined as Equation (1).

(1) Volumetric Conversion (%) = [Feed – (LCO + Decanted Oil)]/ Feed x 100

The fraction LCO and Decanted Oil (DO) is considered non converted fractions.

The main FCC catalyst developers present in the market nowadays are, BASF Catalysts Company, Albermarle, and W. R. Grace Company.

Meeting the Market Demand through FCC Optimization

According to the market demand, the FCC units can be optimized to produce the most demanded derivatives. In traditional FCC units there are normally four operating campaigns:

1 – Maximum Gasoline – In Maximum gasoline campaigns the process unit operates under medium or high severity. The severity is limited by the octane number achieved in the cracked naphtha, in refining configurations where the refiners rely on octane boosting units like alkylation, catalytic reforming, and isomerization there are more flexibility to maximize the gasoline yield in the FCC operating in maximum gasoline mode.

The catalyst formulation to maximum gasoline campaigns involves high zeolite and active matrix once the presence of rare earth compounds tends to raise hydrogen transfer reactions, reducing the olefins content and consequently the octane number in the cracked naphtha. Another alternative to improve the yield of gasoline in the is operation mode is change the final boiling point of cracked naphtha to higher values, in this case the limitation is the quality requirements,

mainly the sulfur content in the final derivative, according to the operating capacity in cracked naphtha hydrodesulfurization unit.

The main restrictions in the process unit in maximum gasoline mode are the gas separation section capacity, especially related to the cold area compressors as well as the debutanizer columns.

2 – Maximum LPG – In this operation mode the FCC unit operates under high severity translated to high operation temperature (TRX), high catalyst/oil ratio. The catalyst formulation taking into account higher catalyst activity through addition of ZSM-5 zeolite. There is the possibility to a reduction in the total processing capacity due to the limitations in blowers and cold area capacity.

It's observed an improvement in the octane number of cracked naphtha despite a lower yield, due to the higher aromatics concentration in the cracked naphtha. In some cases, the refiner can use the cracked naphtha recycle to improve even more the LPG yield.

In the maximum LPG operation mode, the main restrictions are the cold area processing capacity, metallurgic limits in the hot section of the unit, treating section processing capacity as well as the top systems of main fractionating column.

3 – Maximum LCO – The maximum LCO (Light Cycle Oil) operation mode is applied by refiners with great demand by middle distillates, especially diesel, and adequate hydrotreating capacity to convert the LCO in high quality diesel. In this case, the FCC unit operates under relative low severity conditions with low TRX, low catalyst/oil ratios, and the catalyst formulation tends to minimize the catalyst activity.

In process units where is observed a restriction related to the cold area and blowers processing capacity, the maximum LCO operation mode can allow the raise in the total processing capacity of the unit. This fact can be positive, once allow lower time contact between catalyst and feed, improving even more the LCO and decanted oil (DO) yield.

It's important to taking into account the effect of the feed stream quality over the produced LCO. Paraffinic feeds tends to produce higher quality LCO, refiners operating FCC units in maximum LCO mode tends to minimize the final boiling point of cracked naphtha and maximize this parameter in the LCO aiming to improve the LCO yield, but this action is limited by the quality of the final diesel.

4 – Maximum Aromatic Residue – This is the less common operation mode in FCC units, where the main objective is to maximize the yield of decanted oil and achieve the quality requirements of aromatic residue. The aromatic residue is normally applied to produce carbon black, these derivative presents great demand in some markets.

The main difficult to comply with the aromatic residue specification is regarding to ash content in the decanted oil. This parameter is strictly related to the cyclones efficiency in the catalyst regeneration section, to achieve this objective some refiners apply additives to promote de ash decantation in the final tanks or specific filtration systems that requires more capital spending.

Another key quality parameter to meet aromatic residue specification is the BMCI (Bureau of Mines Correlation Index) that is related to the aromaticity of the decanted oil, to achieve the current specifications of black carbon it's necessary to achieve a minimum BMCI higher than 120. The BMCI is calculated based on viscosity of the decanted oil at the temperature of 210 oF. The

metals content in the decanted oil needs to be also controlled, especially, sodium, aluminum, and silicon.

The operating severity in maximum aromatic residue mode tends to be high with high TRX, high catalyst/Oil ratio, and high catalyst activity. As side effect is observed the raise of octane number in cracked naphtha due to the incorporation of aromatics compounds in this intermediate.

In maximum aromatic residue operation mode, the main restrictions are the temperature of the bottom section in the main fractionators that can lead to coke formation, metallurgic limitations in the hot sections as well as the capacity of blowers and cold area compressors.

Synergy between Hydrocracking and FCC Units – A High Bottom Barrel Conversion Refining Hardware

Considering the changes in the downstream market where the gasoline demand faces a severe downturn, some refiners are concerned in promote changes in the FCC operations aiming to minimize the cracked naphtha yield while maximize petrochemical intermediates like propylene and streams able to feed hydroprocessing units to be converted in diesel, mainly light cycle oil (LCO). The great carbon emissions from FCC processing units are another concern which tends to reduce the participation of FCC technologies in modern refining configurations and reduce the feed flow rate processed in the next years in brown route investment.

Despite these concerns, FCC processing units are still relevant to the refining sector, and the solution is promoting synergies with other residue upgrading technologies aiming to maximize added value to the processed feed and minimize carbon intensity. Traditional refining mind set indicates hydrocracking and FCC technologies are faced by competitors technologies in the refining hardware due to the similarities of feed streams that are processed in these units. In some refining schemes, the mild hydrocracking units can be applied as pretreatment step to FCC units, especially to bottom barrel streams with high metals content that are severe poison to FCC catalysts, furthermore the mild hydrocracking process can reduce the residual carbon to FCC feed, raising the performance of FCC unit and improving the yield of light products like naphtha, LPG, and olefins. Figure 7 presents an example of high integrated refining configuration relying on hydrocracking and FCC technologies.

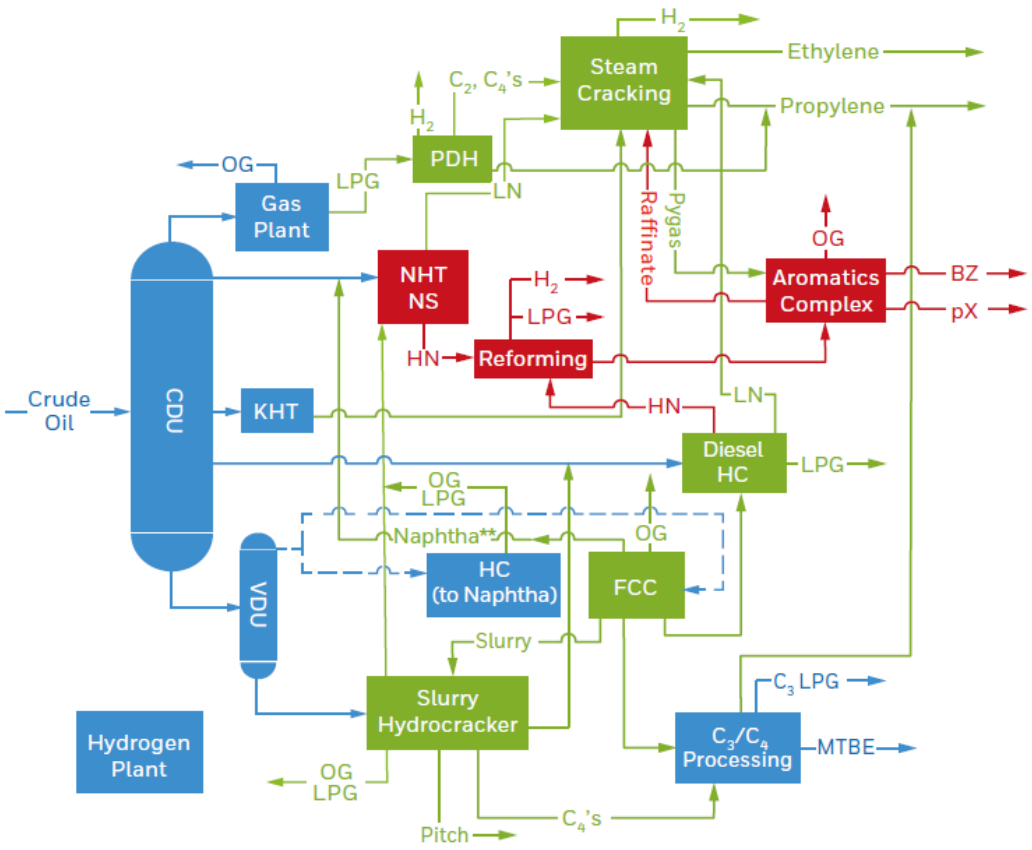

Figure 7 – Example of an Integrated Refining Focusing on Petrochemicals Scheme by UOP Company

As presented in Figure 7, the integrated refining scheme relies on deep residue upgrading technologies as hydrocracking and fluid catalytic cracking (FCC) is capable to reach the production of high-quality petrochemicals, according to the market trends. In some refining configurations, mild hydrocracking technologies are applied upstream of FCC units aiming to improve the quality of the FCC products, mainly the sulfur content.

Examples of mild hydrocracking technologies are the IsoTherming™ process developed by the Dupont Company, and the SPC™ Process licensed by Haldor Topsoe Company, a simplified process flow diagram for SPC™ process is presented in Figure 8.

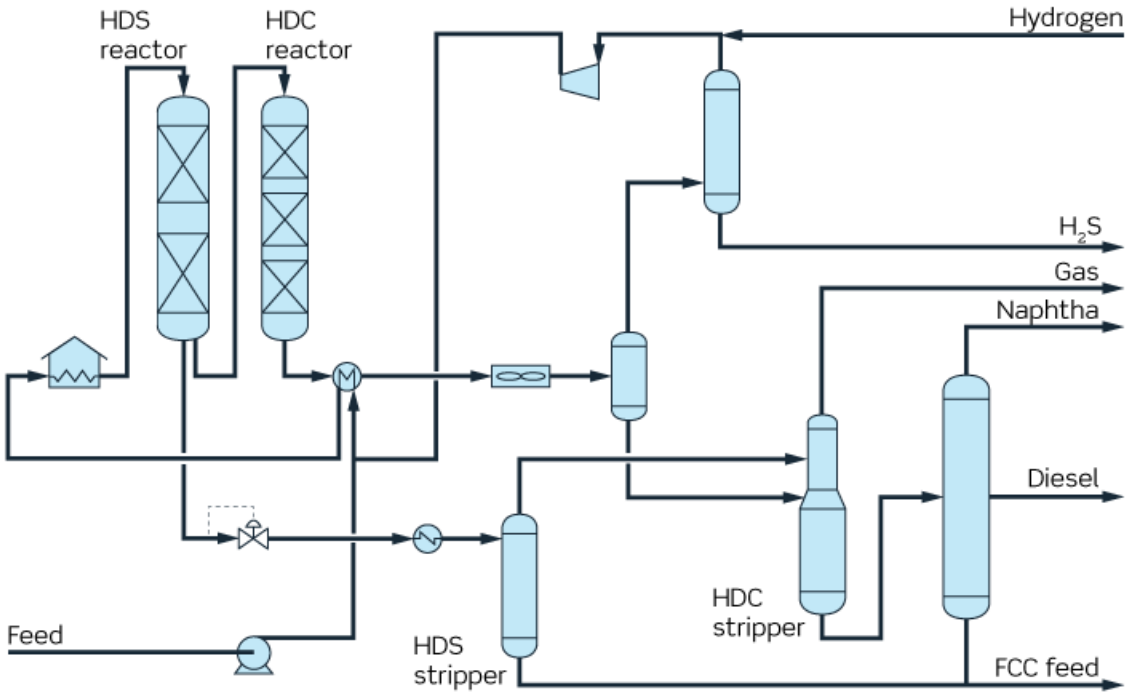

Figure 8 – Process Flow Diagram for the SPC™ by Haldor Topsoe Company

Considering the great flexibility of deep hydrocracking technologies that are capable to convert feed stream varying from gas oils to residue, an attractive alternative to improve the bottom barrel conversion capacity is to process in the hydrocracking units the uncracked residue in FCC unit aiming to improve the yield of high added value derivatives in the refining hardware, mainly middle distillates like diesel and kerosene. This configuration can be especially attractive for gasoline Tier 3 production (with maximum 10 ppm sulfur) for refiners processing high sulfur crudes.

The Side Effect of Cracked Feeds – A Special Challenge to Hydrocracking Units

The most common cracked feeds directed to hydrocracking units are residual streams from FCC like Light Cycle (LCO) and Decanted Oil (DO) and Heavy Coker Gasoil (HCGO) from Delayed Coking units. Another less common feed is residue from Visbreaking units.

The main characteristics that influence the hydrocracking performance for each feedstock is presented below:

- FCC Cycle Oils – Present high aromaticity that are normally refractory to cracking reactions as well as refractory sulfur components, raising the sulfur content in the final products and reduction in diesel cetane number, on the other side, normally presents low basic nitrogen content that is a poison to the hydrocracking catalysts.
- Thermal Cracking Feeds – Normally presents low aromatics content but concentrate refractory sulfur components.

The Heavy Coker Gasoil (HCGO) is an interesting case study as a feed to hydrocracking unit. Refiners with high complexity refining hardware can rely on the synergy between delayed coking and hydrocracking technologies to ensure added value to bottom barrel streams.

The quality of the HCGO relies on the quality of the feed to the delayed coking unit as well as the operating mode of the unit, mainly the recycling ratio. Higher recycling ratios produces better quality HCGO once reduces Conradson Carbon Residue (CCR), reducing the contaminants content like metals, sulfur, and nitrogen.

Despite this advantage, the delayed coking operators normally minimize the recycle ratio to minimum as possible aiming to raise the fresh feed processing capacity and the quality of HCGO is not an optimization focus of the refinery. For this reason, normally HCGO is a hard feed to hydrocracking units due to the high content of refractory sulfur components, high CCR, high nitrogen content, and aromatics concentration.

The sulfur and nitrogen content raises the heat release in the first bed (Higher exothermal profile) that can produce damage to the catalysts, the nitrogen tends to inhibit the cracking reaction leading to lower conversion in the unit. Hydrocrackers processing feeds with high nitrogen content tend to apply processing configuration with intermediate gas separation to control the catalyst activity. The higher production of H2S and NH3 due to the higher concentration of sulfur and nitrogen reduces the hydrogen partial pressure, raises the necessity of washing water to the units, and can raise the corrosion rate in the processing unit.

Aromatics compounds tend to raise the hydrogen consumption, the heat release in the catalyst bed, and are precursors of coking deposition that deactivate the catalyst. Other side effects of the cracked feeds to hydrocracking units are the impact over the quality of the final products like lower cetane number of diesel, higher smoke point of kerosene, lower viscosity index in the lubricating oils and higher sulfur content.

As described above, processing cracked feeds in hydrocracking units present some additional challenges to refiners related to hydrogen consumption, better quench design of the catalyst bed due to the higher exothermic profile of the reactions, and lower global activity of the catalyst due to the higher poison content, like basic nitrogen. These characteristics lead the refiners processing cracked feeds in hydrocracking units to invest more capital in feed treating systems like filtering and guarding beds. Despite this apparent disadvantage, refiners able to add value to bottom barrel streams can enjoy highly competitive advantage considering the downstream market post IMO 2020. For refiners processing extra-heavy bottom barrel streams, the deep hydrocracking technologies like slurry phase hydrocracking can be an interesting option, despite the high capital and operating costs.

The Petrochemical FCC Alternative – Raising Competitive Advantage in Changing Market Consumer

Considering the current scenario of the downstream industry and the last forecasts, it's observed trend of reduction in transportation fuels demand followed of a growing market of petrochemicals, leading the refiners to optimize his FCC units to maximum LPG yield aiming to improve the capacity to produce light olefins and promote closer integration with petrochemical assets. The major part of the catalytic cracking units is optimized to maximize transportation fuels, especially gasoline, however, face to the current scenario some units have been optimized to maximize the production of light olefins (ethylene, propylene, and butenes). As aforementioned, units focused on this goal have these operational conditions severely changed, raising the cracking rate.

The reaction temperature reaches 600 oC and higher catalyst circulation rate raises the gases production, which requires a scaling up of gas separation section. The higher thermal demand makes advantageous operates the catalyst regenerator in total combustion mode leading to the necessity of installation a catalyst cooler system.

Installation of catalyst cooler system raises the process unit profitability through the total conversion enhancement and selectivity to noblest products as propylene and naphtha against gases and coke production. The catalyst cooler necessary when the unit is designed to operate under total combustion mode due to the higher heat release rate as presented below.

$C + ½\ O_2 \rightarrow CO$ (Partial Combustion) $\Delta H = -27$ kcal/mol

$C + O_2 \rightarrow CO_2$ (Total Combustion) $\Delta H = -94$ kcal/mol

In this case, the temperature of the regeneration vessel can reach values close to 760 oC, leading to higher risks of catalyst damage which is minimized through catalyst cooler installation. The option by the total combustion mode needs to consider the refinery thermal balance, once, in this case, will not the possibility to produce steam in the CO boiler, furthermore, the higher temperatures in the regenerator requires materials with noblest metallurgy, this raises significantly the installation costs of these units which can be prohibitive to some refiners with restricted capital access.

Conclusion

As discussed above, the FCC units offer great operation flexibility to refiners and can raise significantly the refining margins and, according to the local market demand, the process unit can be optimized to produce different kind of intermediates. Following recent trends, the synergy between refining and petrochemical processes raises the availability of raw material to petrochemical plants and makes the supply of energy to these processes more reliable at the same time ensures better refining margin to refiners due to the high added value of petrochemical intermediates when compared with transportation fuels.

Another advantage is the risks reduction of transportation fuels oversupply, facing the current scenario of demand reduction and restriction of fossil fuels. It's important to consider that integrated processes lead to a higher operational complexity, however, given current and middle term scenarios to refining industry, a better integration between refining and petrochemical processes is fundamental to the economic sustainability of the downstream industry. In this scenario, the FCC technologies can ensure higher added value to processed crude oils through the maximization of petrochemical intermediates, like propylene, in the refining hardware. Another important variation of the FCC technologies is the unit focused to processing residues, the Residue Fluid Catalytic Cracking (RFCC) that can allow even more added value to bottom barrel streams, especially for refiners processing heavier crudes.

Like a flexible refining technology the Fluid Catalytic Cracking (FCC) units have a highlighted role in the future of the downstream industry, especially considering the transitive period, where the FCC units can help to supply the demand by petrochemicals without shortage of transportation fuels.

References

GARY, J. H.; HANDWERK, G. E. ***Petroleum Refining – Technology and Economics***.4th ed. Marcel Dekker., 2001.

MYERS, R.A. Handbook of Petroleum Refining Processes. 3a ed. McGraw-Hill, 2004.

ROBINSON, P.R.; HSU, C.S. ***Handbook of Petroleum Technology***. 1st ed. Springer, 2017.

SALGADO, H. Controlling the metals content of FCC equilibrium catalyst. PTQ Magazine, Q3 2018.

SARIN, A.K. – Integrating Refinery with Petrochemicals: Advanced Technological Solutions for Synergy and Improved Profitability – Presented at Global Refining & Petrochemicals Congress (Mumbai, India), 2017.

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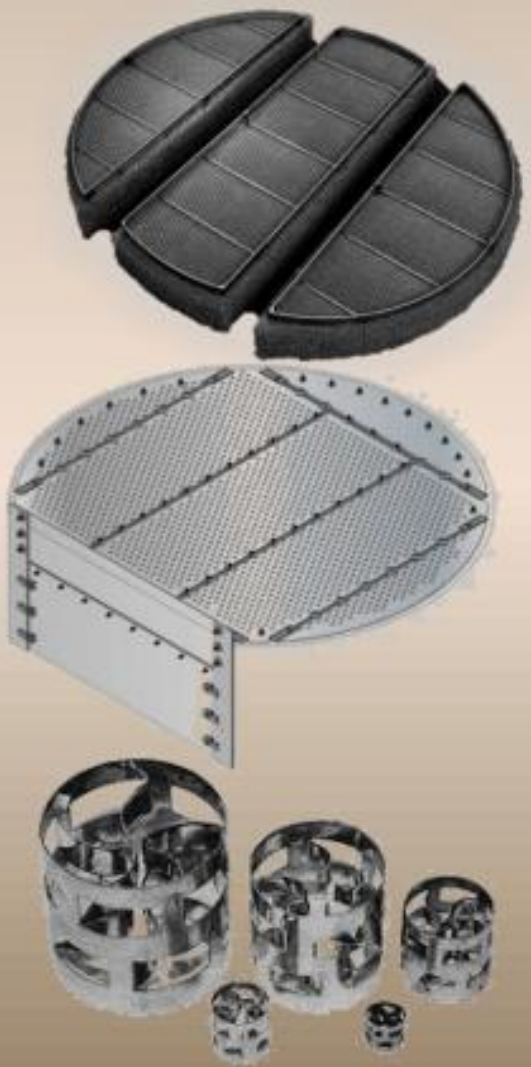

All About Downcomers and Weirs

Dr. -Ing. Volker Engel

Tower trays and internals are the heart of all distillation columns. Their design is an essential part of a process engineer's task and determines the process reliability and economy.

This article is the part of a series on different kinds of trays and internals.

In almost all tray towers, the liquid flows horizontally from the inlet, gets in contact with the vertical streaming gas, generates a two-phase layer on the active area, leaves the tray at the outlet weir and degases in the downcomer while passing to the next tray below.

The main focus in tray towers is often only the active area (type of tray, pressure drop, froth height, efficiency, ...), where the mass transfer takes place. To understand all the complexity of trays, it is necessary to get an overview of the various shapes and layouts of downcomers.

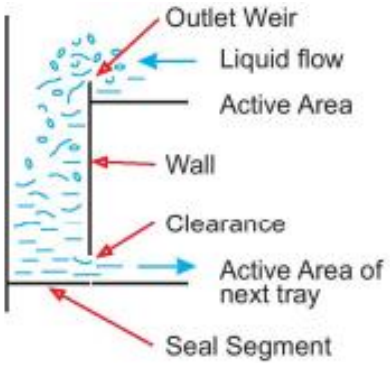

Figure 1: Parts of a Downcomer

DOWNCOMERS (DC)

The main function of the downcomer is to collect all liquid from the active area, degas the liquid, lead the liquid to the next tray and seal the downcomer against gas bypass (Fig. 1).

Note: The downcomer belongs to the tray, where the liquid leaves the tray. Its numbering is therefore the same as the tray. Any downcomer above the top tray (e.g. for inlet) is called "False Downcomer" (refer Figure 33)

The size and shape of the downcomer is specified by the liquid flow rate, whereas the size of the active area is determined by the gas load.

One of the main tasks at designing a proper tray is to choose suitable values for these two areas. This sets the tower diameter!

FLOW PASSES

One of the very first steps in the design of a tray is to deal with the liquid load: It shows, how many downcomers per tray are required and how the downcomer(s) have to be designed. To handle the liquid load, you have to supply the appropriate downcomer area as well as enough weir length.

To achieve this goal, there are different layouts with one or more downcomers and special down-comer shapes. (Some other designs like Dual-flow trays, baffle trays, shower decks, ... have no down-comers.)

In the easiest case, there is only one downcomer (Fig. 2). The liquid is streaming from one side of the tower to the other. There is one active area and one flow path. This (common) design is called 1-pass or Singlepass tray. The tray design is turned by 180° at each stage. It is used for tower diameters up to 3m. (For some special applica-tions you will even find 1-pass trays up to 8m.)

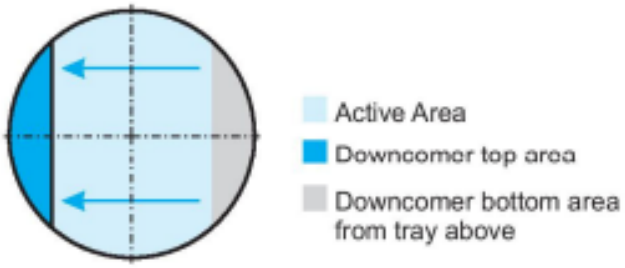

Figure 2: 1-Pass Tray Layout

For higher liquid loads and larger diameters, two liquid paths are needed (Fig. 3). This leads to two different tray designs, which are used alternately for the stages (therefore the odd tray numbers belong to one design, the inboard design has a center downcomer, the other (outboard design) has two side downcomers. The liquid is strea-ming from the center down-comer to the side downcomers and on the next stage from the sides to the center downcomer. The active area is symmetrical on each tray. 2-pass designs are often used in practice.

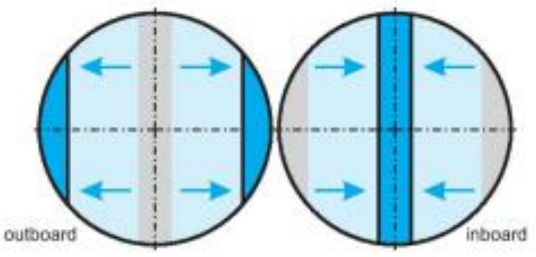

Figure 3: 2-Pass Tray Layout

The next logical step is the 3-pass tray design (Fig. 4). There is a side downcomer and an off -center downcomer. The design is rotated by 180° per stage. There is only one design for all trays (as for the 1-pass design). This is an ad-vantage in terms of the investment costs compared to the 2- and 4-pass tray design. On the other hand there are three different shaped active areas per stage and a very different geometry of the downcomers. Therefore it is not easy to design this 3-pass tray for a wide operation range and you will not find this type very often in practice (but you will!).

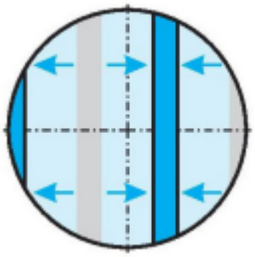

Figure 4: 3-Pass Tray Layout

The 4-pass tray has two tray layouts (Fig. 5): One stage is equipped with two side downcomers and one center downcomer (outboard), the other with two off-center downcomers (inboard). As each layout is symmetrical to the tower center line, there are only two different active area shapes. Therefore it is easier to calculate than the 3-pass-design – but as the liquid splits up in two different active areas (with different froth heights, pressure drop, ...), the calculation of 4-pass trays needs a complex iteration for the individual load of each active area for each design case (min / design / max).

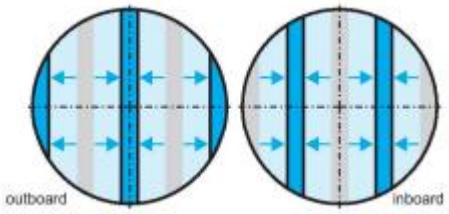

Figure 5: 4-Pass Tray Layout

The next logical number of downcomers would be the 5-pass tray. You will find it in literature, but rarely in practice. For large tower diameters the 6-pass tray is used (Fig. 6). Since it is symmetrical to the tower center line, there are “only” three different active areas per tray and four different down-comer shapes per tray. Therefore the calculation is similar to that of the 3-pass tray. The outboard layout has two side downcomers and two off-center downcomers, the inboard layout has one center downcomer and two off-center downcomers.

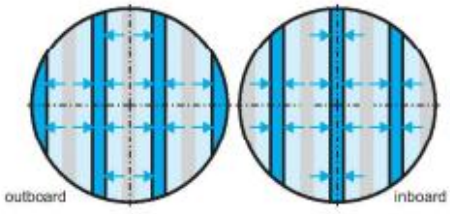

Figure 6: 6-Pass Tray Layout

The choice of a certain design (number of passes) and setting the width of the downcomer results in a certain length of the flow paths. This flow path length is relevant in terms of hydrau-lics (contact time of the gasliquid mix-ture) as well as in terms of practice (the flow path length is the maximum width of one dimen-sion of the manway). Therefore you have to check these hydraulic and security aspects, too.

MULTI-DOWNCOMERS

The designs described above are called conventional multi-pass trays and they are “bound” to the circle geometry of towers. Instead of using the tower shell as part of the downcomer wall, you can place downcomer 38 boxes in the active area. These designs are called Multi-Downcomer designs (”MD trays”).

There are different principles:

You can place the downcomer boxes over the entire diameter of the tower and rotate the design by 90° per stage (Fig. 7). The crossing points of the boxes are blocked to have no liquid shortcuts.

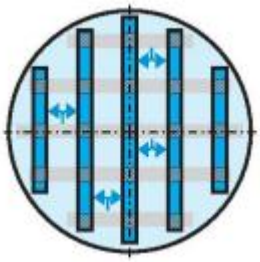

Figure 7: MD Layout (UOP Design)

Another option is to place the downcomer boxes only on one half of the cross section area and rotate the designs by 180° per stage (Fig. 8). These designs are also known as Calming Section trays.

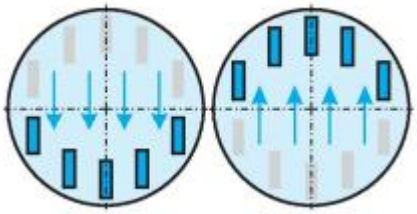

Figure 8: MD Layout (“Calming Section”)

Another type is shown in Fig. 9. It is also known as HiFi-tray (Shell).

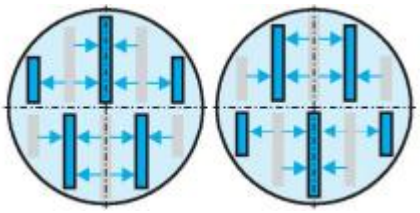

Figure 9: MD Layout (“HiFi-Tray”)

These MD trays are normally used for large liquid load when volumetric ratio between vapor and liquid rate is low (medium to highpressure systems).

REVERSE-FLOW

Another design for the flow path is the socalled Reverse-Flow tray (Fig. 10). It is used to achieve long flow paths and long contact times between gas and liquid.

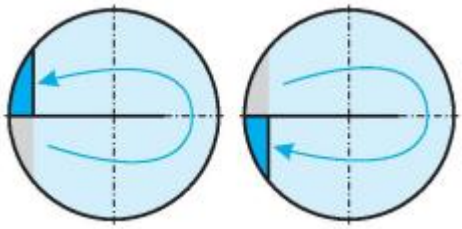

Figure 10: Reverse-Flow Layout

DOWNCOMER SHAPES

The side downcomer in a classic design is nor -mally chordal (Fig. 11). This design is comparably easy in construction and fabrication for tower attachments and tray parts.

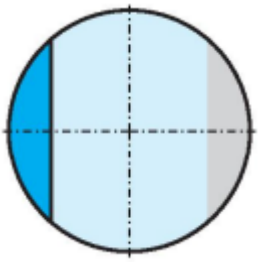

Figure 11: Chordal Downcomer

For high liquid loads you can have multichordal designs (Fig. 12). Especially at the side down-comers you are struggling with the geometry of round towers: changing the width of the side downcomer has great effect on the area, but small on the weir length. With a multi-chordal design you can achieve a long(er) weir length without increasing the corresponding chordal downcomer area.

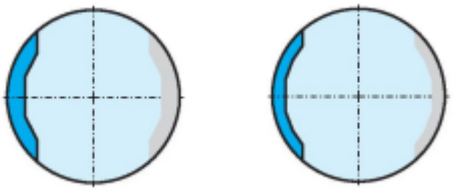

Figure 12: Multi-Chordal Downcomers

At the design of multi-chordal downcomers, you have some degrees of freedom. Normally you will find 5 segments as multi-chordal, some-times 7 segments. The construction and fabrication of a multi-chordal tray is more expensive than a chordal design. SLOPED DOWNCOMER The gas-liquid mixture entering the downcomer has the average density of the phases. During the degassing process, the density increases and ideally becomes the liquid density at the bottom of the downcomer. You can take benefit from this change in density (i.e. change in volume) by reducing the cross sectional area of the down-comer over its height. This adjustment is called sloped or stepped downcomer (Fig. 13).

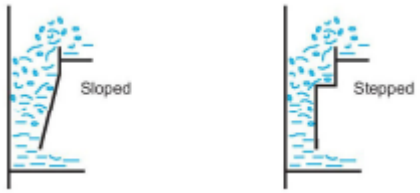

Figure 13: Sloped and Stepped Downcomers

It can be applied on all types of downcomers (side, center, off-center, multi-downcomers, ...). The benefit of this design is the gain of active area on the next tray.

TRUNCATED DOWNCOMER

Another possibility to maximize the active area is to use truncated downcomers (Fig. 14). In this design the floor of the next tray is not the bottom of the downcomer – the downcomer ends above the active area. For Multi-Down --comers this design is standard. The area below the downcomer can be used (partly) as active area.

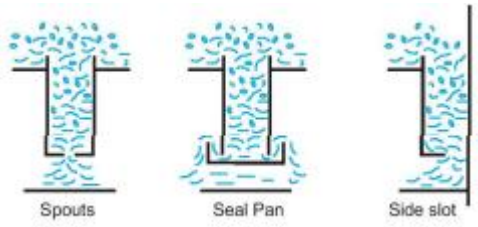

Figure 14: Truncated Downcomers

Truncated downcomers bring us to some fundamental hydraulic considerations.

DOWNCOMER FLOODING

There are two flooding effects related to down -comers. One deals with the maximum through-put, the other with the liquid level in the down-comer.

The first one is called Choke Flood and is related to the cross-sectional area of the downcomer and the physical properties of the liquid and the gas. If there is not enough area for the degassing (upflowing gas and downflowing liquid) the downcomer “chokes”. In result, no liquid will pass trough the down-comer and the tower starts flooding.

The standard model for calculating the max. volume flow rate through a downcomer was published by Glitsch in 1993 (here in SI-units):

$$\dot{V}_{DC,lmax,1} = 0.1698 \cdot SF \cdot A_{DC,top}$$

$$\dot{V}_{DC,lmax,2} = 0.006955 \cdot \sqrt{\rho_L - \rho_G} \cdot SF \cdot A_{DC,top}$$

$$\dot{V}_{DC,lmax,3} = 0.0079824 \cdot \sqrt{TS} \cdot \sqrt{\rho_L - \rho_G} \cdot SF \cdot A_{DC,top}$$

$$\dot{V}_{DC,lmax} = \min(V_{DC,lmax,1}, V_{DC,lmax,2}, V_{DC,lmax,3})$$

The equations are based on gas density rG, liquid density rL, tray spacing TS and system factor SF. This last parameter describes the difficulty of degassing. You can find lists of values for different appli-cations in literature (ref. to References at the end of article).

The second flooding effect is the so-called Aerated Downcomer Flood (also called Downcomer Backup Flood): In this case the liquid level in the downcomer exceeds the total down-comer height. The level in the downcomer is the result of "border effects" (Fig. 15):

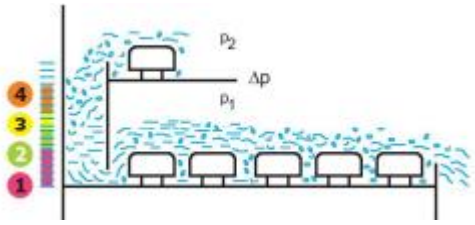

Figure 15: Liquid Level in Downcomer

1. The two-phase layer on the next tray seals the downcomer. It is calculated by the physical height of the weir plus the weir crest height.
2. The hydraulic gradient of a tray produces an additional liquid head in the downcomer. (Depends on contact elements and flow path length.)
3. The liquid is accelerated by leaving through the clearance. This rise in kinetic energy is taken from potential energy and results in an additional liquid head.
4. The pressure p1 of the gas is higher than the pressure p2 at the tray above. This difference in pressure results in liquid head in the down-comer.

All these values are calculated as "clear liquid". In praxis the liquid in the downcomer contains gas (calculated as so-called Aeration Factor) and leads to a higher level in the downcomer than the clear liquid height. If this level exceeds the downcomer height plus weir height, the down-comer will not be able to handle the load and floods.

Note: The effect of Aerated Downcomer Flood is not correlated with the downcomer area. Therefore the change of downcomer area will not affect this issue!

There are some calculation models for the Aeration Factor in literature (ref. to References at the end of article).

A short glance back: We started the discussion with truncated downcomers. Since the height of the downcomer is relevant for the Aerated Downcomer Flood (and a truncated downcomer is less in height), truncated downcomers are only applicable, if there is no problem with Aerated Downcomer Flood.

OUTLET WEIRS

The liquid-gas mixture enters the downcomer at the outlet weir. The standard weir is a plain bar of about 50 mm height. In liquid limited systems or on trays running in the froth regime, the weir height might be more. The specific weir load should be more than 4.5 m3/m/ h (or 5 mm weir crest height). Why?

If the weir crest height is very low, any leveling issues of the tower (due to tower attachments, tower installation or even wind) can cause problems: If a part of the weir is not in use, this part of the active area is stagnant – no mass transfer will happen (Fig. 16).

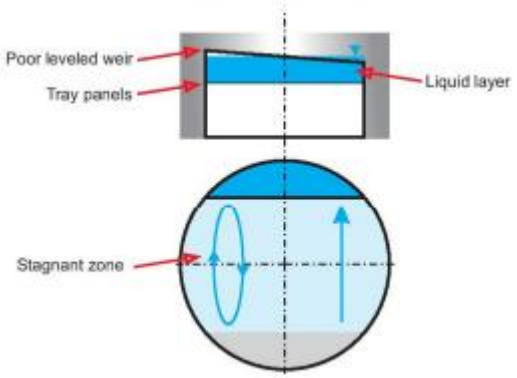

Figure 16: Malfunction of Weir

To achieve a uniform overflow at the entire length of the downcomer, you can use Notched Weirs (Fig. 17). At low loads the liquid uses the bottom parts of the notches. At higher loads the entire weir is used by the liquid.

To achieve a uniform overflow at the entire length of the downcomer, you can use Notched Weirs (Fig. 17). At low loads the liquid uses the bottom parts of the notches. At higher loads the entire weir is used by the liquid.

Figure 17: Notched Weir

For safety reasons (e.g. during installation or inspection) it is strongly recommended to have plain tops at the notches.

Another possibility to adapt long weirs for small liquid loads is to block the weir (Fig. 18). The weir length is reduced to the openings. An additional effect of a blocked weir: it keeps the two-phase layer on the active area.

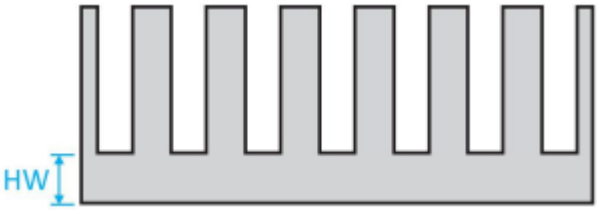

Figure 18: Blocked Weir

The number and width of the blocks depend on the weir crest height you have to achieve. The method of blocking is not only relevant for small liquid loads, but also for balancing the different weir lengths of 2-and-more-pass layouts. The block should be at least as high as the two- phase layer of the active area.

Blocked weirs prevent workers/inspectors from climbing over the center/off-center downcomers to the other pass! If there is no manhole in each pass, there should be manways in the blocked weir.

Sometimes you will find weirs with adjustable height. These are built in when it is known that the load will change in the future.

ANTI-JUMP BAFFLES

At center, off-center and multi-downcomers, the liquid enters the downcomer from both sides. At high liquid loads and small downcomer widths, you have to install so-called Anti-Jump Baffles (Fig. 19). These baffles ensure, that only half of the area is used by each side.

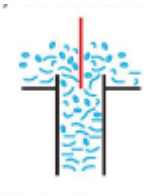

Figure 19: Anti-Jump Baffle

Just like for the weir blocks, Anti-Jump Baffles have to be as high as the two-phase layer on the active area. Therefore they block the passage during inspections. It is good practice to have bolted manways in the baffles.

SWEPT-BACK WEIR

To achieve a long weir, you can use a socalled Swept-back Weir (Fig. 20). The downcomer stays chordal, but the weir is – similar to the shape of multi-chordal downcomers – longer than the chordal length.

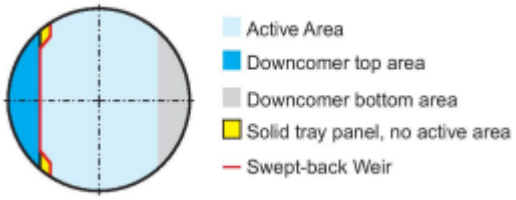

Figure 20: Swept-Back Weir

WEIRS FOR STARTUP

For high capacity trays you will find so-called Weir Spouts (Fig. 21): At tray floor level, there are rectangular openings.

These holes help to fill the downcomer initially at startup of the tower. During operation they help to reduce the weir crest height. They are only used, if the tower is always running at high liquid loads and there is no minimum load issue.

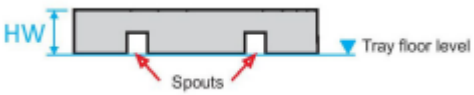

Figure 21: Weir Spouts

CLEARANCE

The last aspect of downcomers is the design of the liquid outlet (= inlet to the next tray).

The liquid is leaving the downcomer through the so-called Clearance. The outlet velocity of the liquid should be less than 0.45 m/s. If it is higher, the liquid will splash and overrun the first rows of contact elements (see later).

The height of the clearance HCL is normally less than the outlet weir height HW to have a Static Seal

$$Sealstatic = HW - HCL.$$

For liquid-tight trays (e.g. bubble cap or tunnel trays) this seal is present, whenever liquid is on the tray. For all other tray types (e.g. sieve, float valves, fixed valves) the seal is only working,

if there is enough gas for no weeping. If there is no Static Seal (due to design, poor hydraulics or at startup of the tower) there is the risk of gas bypassing the active area through the downcomer.

During operation the weir crest generates additional liquid height (How) (Fig. 22) and generates a Dynamic Seal

$$Sealdynamic = How + HW - HCL.$$

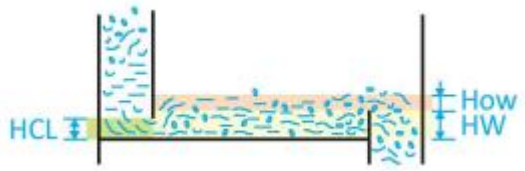

Figure 22: Static and Dynamic Seal

So-called high capacity designs are often running with no static but only dynamic sealing.

RADIUS LIP

The acceleration of the liquid in the clearance results in liquid head (ref. to Fig. 15, Aerated Downcomer Flood). To reduce this effect, you can add a Radius Lip (Fig. 23). This shape lowers the outlet orifice coefficient.

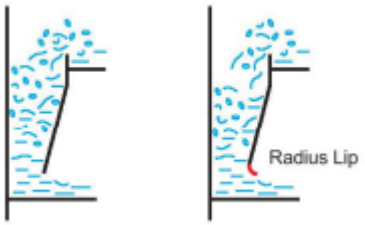

Figure 23: Radius Lip

INLET WEIR

To ensure sealing of downcomers you can place an Inlet Weir in front of the clearance (Fig. 24). Even at a very small liquid load, there will be no gas bypass through the downcomer. The inlet weir can only function if it is higher than the clearance. It can also be notched (for low liquid loads).

The inlet weir is helpful for the startup of the tower and at low liquid load – but the downcomer tends to blocking in fouling systems.

Figure 24: Inlet Weir

As discussed at the Aerated Downcomer Flood, one part of the liquid level in the downcomer depends on the liquid level on the next tray. For part 1 (Fig. 15) you have to take into account the inlet weir height plus the inlet weir crest height. Therefore an inlet weir "costs" down-comer height in respect to downcomer flooding (especially at low tray spacings).

RECESSED SEAL PAN

The effect of sealing can also be achieved by the so-called Recessed Seal Pan (also called Inlet Pot, Fig. 25). The main advantage over the inlet weir is, that it "costs" no downcomer height. But of course it is more complex in construction and fabrication. It is normally combined with a sloped downcomer.

Figure 25: Recessed Sealed Pan

INTERRUPTER BAR

At float valve trays you may sometimes find elements, that are looking similar to inlet weirs. They are not for sealing the downcomer but to keep the first valve row functional (Fig. 26). The bar is about 13mm high and is called Interrupter Bar.

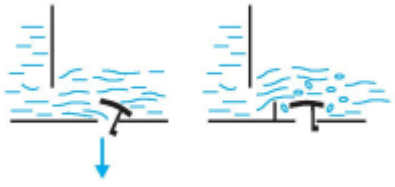

Figure 26: Interrupter Bar

BUBBLING INITIATORS

At high liquid outlet velocity, the liquid overruns the first rows (of fixed valves, sieve holes, ...). To break the impulse at high capacity trays there are Bubbling Initiators (also called Bubbling Promoters, Fig. 27). They are placed instead of the first row of "normal" contact elements. The gas outlet openings of the Bubbling Initiators are not oriented towards the downcomer (to prevent gas entry through the clearance).

Figure 27: Bubbling Initiator

INLET PUSHER

A special design ("NYE-tray") for the inlet is to push gas near the clearance through holes to the active area (Fig. 28). It also helps to initiate the two-phase layer. It is not easy to design this layout for a large operating region.

Figure 28: Inlet Pusher ("NYE-Tray")

SEAL PAN

The last downcomer of a tower/section is sealed by a special Seal Pan (Fig. 29) or submerged in the bottom liquid (Fig. 30).

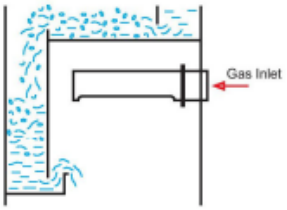

Figure 29: Last Downcomer with Seal Pan

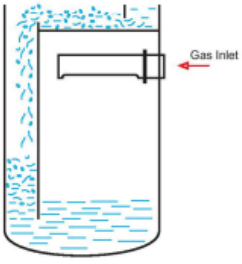

Figure 30: Submerged Downcomer

In any case, the gas inlet should not affect the outlet of the liquid. It is good practice to make the last downcomer long enough to bypass the gas inlet.

DRAW-OFF (DRAW)

You will find seal pans not only at the very last downcomer of a tower. Whenever there is a change in flow path number and/or tray orientation ("transitions"), the liquid is transferred to the next tray (or to a liquid distributor) by pipes. Fig. 31 shows some examples for Draw -Offs from a seal pan.

In all cases you have to ensure that the seal is not affected by the Draw-Off.

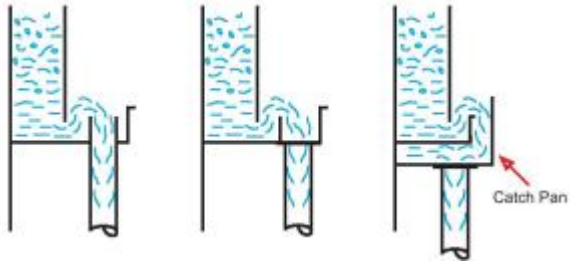

Figure 31: Draw-Offs at Seal Pan

One of the advantages of tray towers is the possibility to draw off liquid at each stage. These intermediate Draw-Offs are normally done with the help of Recessed Seal Pans (Fig. 32).

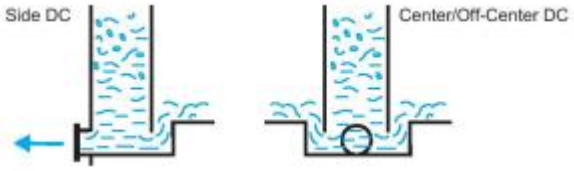

Figure 32: Draw-Offs at Recessed Seal Pan

FEED TO DOWNCOMER

As mentioned before, you will install so-called False Downcomers (FDC) to feed liquid on the top tray of a section (Fig. 33). They are designed accordingly to the downcomers of the following trays.

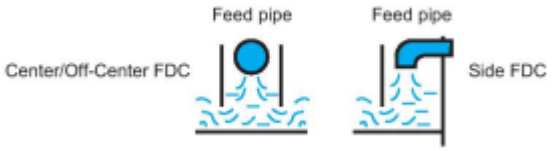

Figure 33: False Downcomer

To feed liquid to a tray within a section, it is good practice to feed near the downcomer clearance (feed pipe with holes directed to the area above the clearance, Fig. 34). If there is the danger of two-phase or super-heated liquid, you will shield the downcomer wall by a so-called Impingement Baffle.

Figure 34: Feed to Tray

MECHANICAL ASPECTS

The active area is carried by the support ring, the downcomer (at the outlet weir side) and the downcomer seal segment (at the inlet). There-fore these downcomer elements have to be designed to withstand the dead load plus the liquid load.

For large tower diameters, the upper part of the downcomers (so-called Downcomer Truss) will be fabricated in a higher material thickness (Fig. 35).

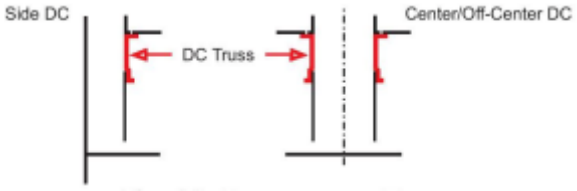

Figure 35: Downcomer Truss

Whenever the downcomer elements would not fit through the manhole, the elements have to be divided, too. (A benefit of a downcomer wall consisting of two parts is the opportunity to adjust the clearance height during installation.)

In downcomers there are used so-called Down-comer Brackets (Fig. 36). They support the downcomer seal segment and prevent the walls from vibration. In center/off-center downcomers the brackets additionally connect the walls and keep them in place – in best case even at a pressure surge. (A well-known failure pattern is when the downcomer walls are pushed inwards after a pressure surge and the panels of the active area therefore slip and fall.)

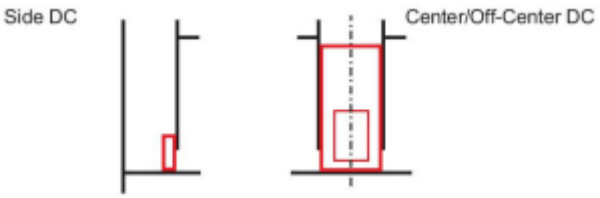

Figure 36: Downcomer Brackets

CALCULATED PARAMETERS

The following parameters have to be calculated and checked for all downcomers at each load. This can hardly be done by hand, but with suitable software. Any software that does not show all of

these parameters carries the risk of overlooking relevant values. Since the supplier's free software does not do this, commercial software (ref. TrayHeart) should be used.

Parameters of the downcomer:

- Residence Time
- Clear liquid level in downcomer
- Aeration Factor of downcomer
- Aerated liquid level in downcomer
- Liquid head caused by clearance
- Liquid outlet velocity
- Max. liquid volume flow through downcomer

For the outlet weir the following parameters have to be calculated:

- Specific liquid load of weir
- Weir crest height
- Throw width over weir
- Liquid ratio through spouts

If there is an inlet weir:

- Specific liquid load of weir
- Weir crest height

CONCLUSION

There are many layouts and designs for downcomers and weirs. The effort for the engineer is to combine the best options to achieve an optimum for the very different aspects as startup, statics, hydraulics and interaction with the active area. The hydraulic design of downcomers has to be as accurate as the design of the active area.

AUTHOR

Volker Engel studied process engineering at the Technical University of Munich and did his Ph.D. thesis on packed columns with Prof. Johann G. Stichlmair. Since 1998 he has been the managing director of WelChem Process Technology GmbH and head of the TrayHeart software. TrayHeart has developed into a state-of-the-art design tool for trays and internals in process technology.

REFERENCES

Glitsch Ballast Tray Design Manual. Bulletin 4900, 6th edition, Dallas (1993)

Hoppe, K.; Mittelstrass M.: Grundlagen der Dimensionierung von Kolonnenböden, Steinkopff Verlag, Dresden (1967)

Kister, H. Z.: Distillation Operation, Mc Graw Hill (1989)

Liebermann, N.: Process Equipment Malfunctions. Techniques to Identify and Correct Plant Problems. McGrawHill (2011)

Liebermann, N.; Liebermann, E.: A working guide to Process Equipment. McGrawHill Education (2014)

Lockett, M. J.: Distillation tray fundamentals, Cambridge University Press, New York (1986)

Norton Valve Tray Design Manual: Company publication, 7/96 VTDM-1-E (1996)

Nutter Engineering: Float Valve Design Manual, (1976, April, rev1)

Stichlmair, J.: Grundlagen der Dimensionierung des Gas/Flüssigkeit-Kontaktapparates Bodenkolonne, Verlag Chemie, Weinheim (1978)

Stichlmair, J.; Fair, J.: Distillation – Principle and Practice, Wiley-VCH New York (1998)

WelChem Process Technology: TrayHeart Software. Tower

Internals Calculation Software. Internet: www.welchem.com; Info: service@welchem.com

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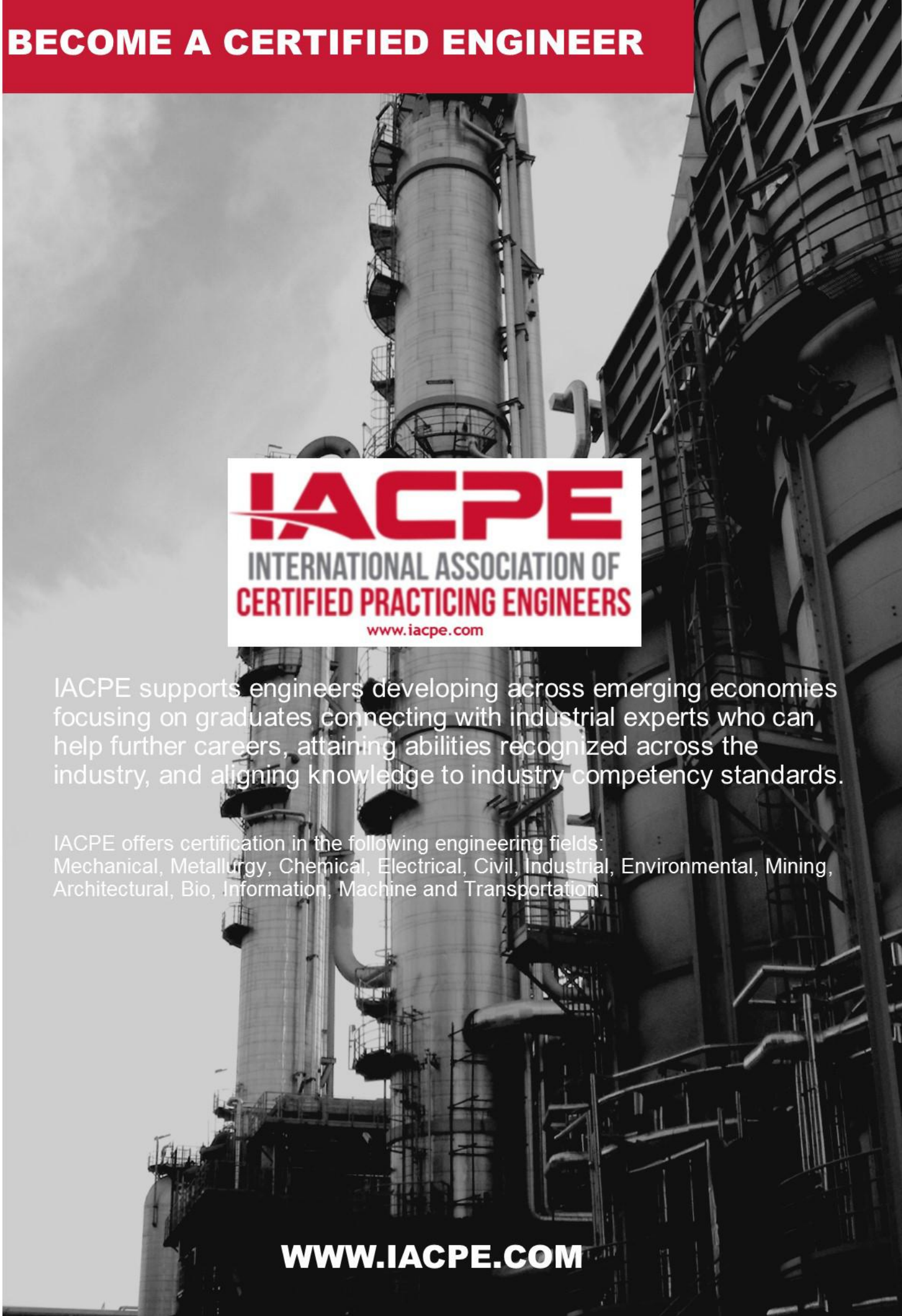

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Design and Revamp Guidelines for Palm Oil Stripper Columns

Karl Kolmetz

Introduction

The palm oil industry has emerged as one of the vital manufacturing sectors in the world. Further, it has gradually become the most abundant traded vegetable oil in the world, owing to the growing demand.

Palm oil processing refers to the extraction and refining of palm oil from the fruit of the oil palm tree (Elaeis guineensis). Palm oil is widely used in various industries, including food, cosmetics, and biofuel production.

The crude palm oil is gathered from the mesocarp (seed) of the palm oil. However, the crude palm oil that has been extracted contains unwanted impurities and requires a purification process to partially or eliminate them to produce edible oils, mainly cooking oil.

In the current scenario, palm oil is widely used in tropical countries in Southeast Asia, Africa, and parts of Brazil for cooking purposes. Further, due to its low costs and oxidative stability, palm oil has grown popular in these regions.

As mentioned before, crude palm oil that is extracted contains impurities that require a purification process to improve the quality, physical appearance, oxidative stability, and other properties. The impurities are removed at a processing plant, and thus, it is essential to ensure that the plant is modern and equipped with advanced technologies to make the process efficient.

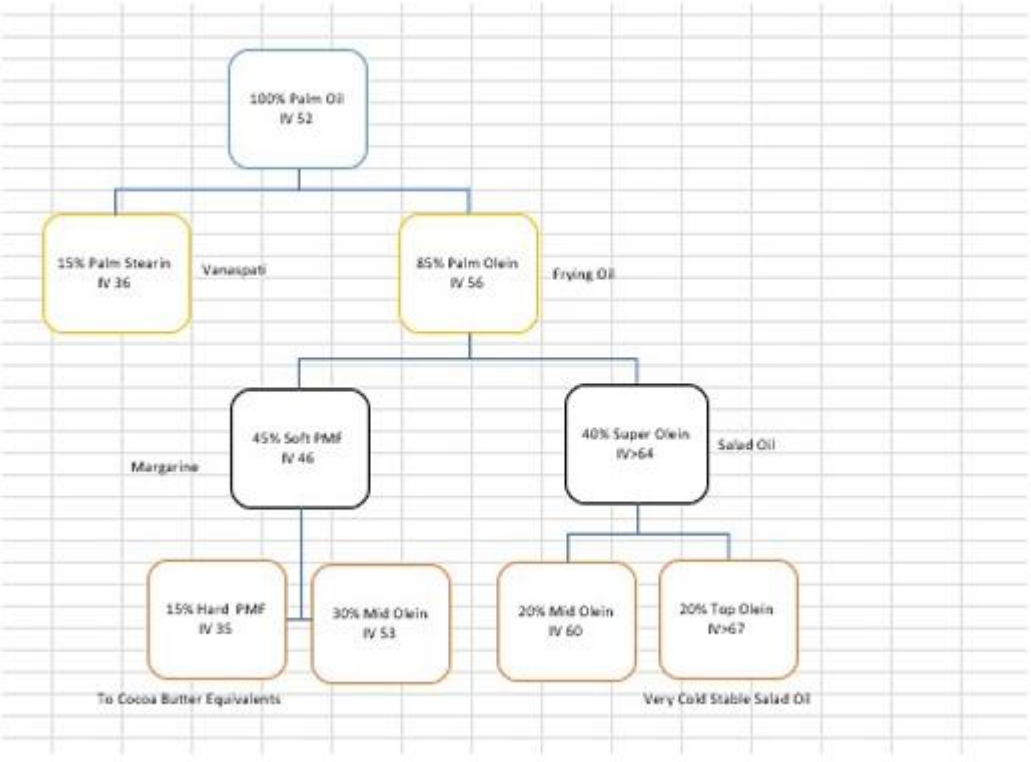
The palm oil processing process typically involves the following steps:

- Harvesting
- Sterilization
- Threshing
- Digestion
- Extraction of oil
- Clarification
- Refining – Physical and Chemical
- Fractionation

Fractionation

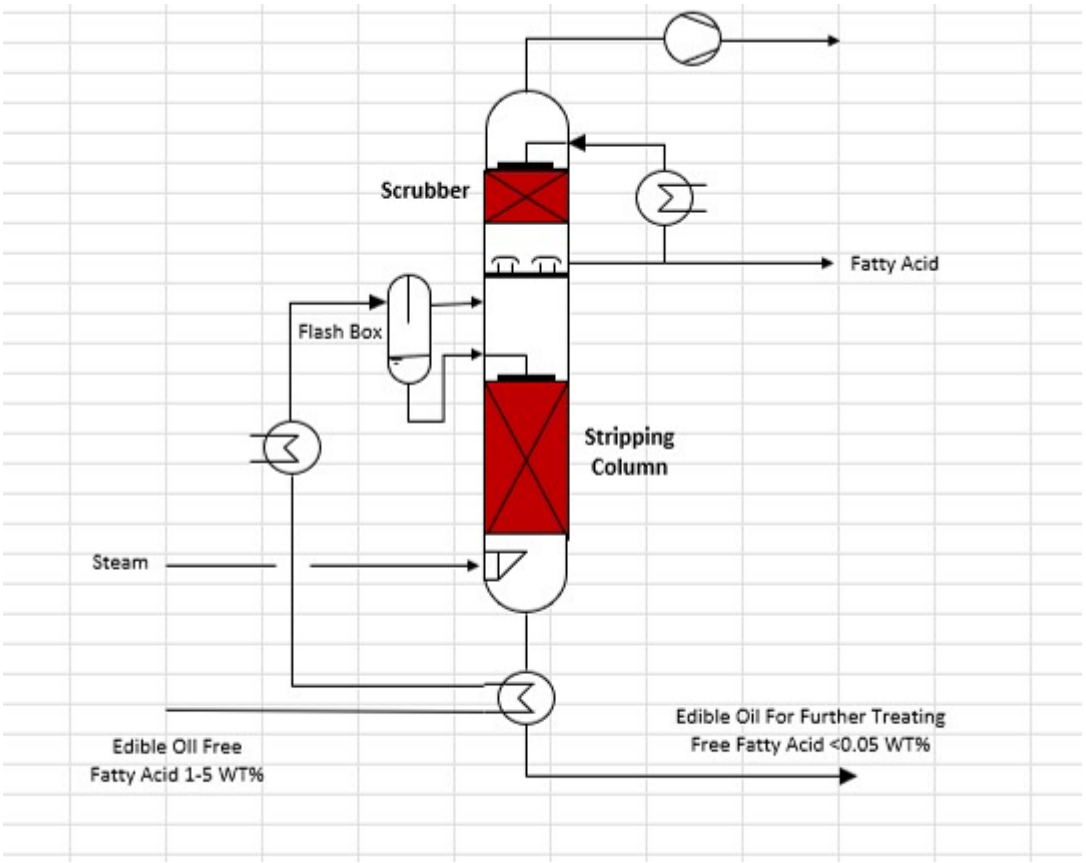
Fractionation is a process that can be employed to separate the palm oil into different fractions, such as palm olein (liquid fraction) and palm stearin (solid fraction). This process is used to obtain oils with specific melting points and different applications.

After processing, the palm oil is usually stored and transported in bulk for various uses in the food industry, such as cooking oil, margarine, and processed foods. It is also used in the production of personal care products, soaps, and biodiesel.



Palm Oil Stripper

The Palm Oil Stripper is one of the first fractionation columns in the palm oil process. The Palm Oil Stripper is typically one or two columns. For a single column, the bottoms section is called the stripper and the top section is called the scrubbing section. Preheated, hot crude palm oil is entering the stripper column, which is equipped with structured packing and operated at very low pressure, where the FFA are stripped off by introducing stripping steam in the bottom. Some processes have the palm oil stripper as part of the deodorization system.



For a single column the top section is called the scrubbing section. The scrubbing section can be one to three beds. The scrubbing section is typically structured packing of KLM 250X EC style. The EC (Enhanced Capacity) is 2nd Generation Structured Packing.

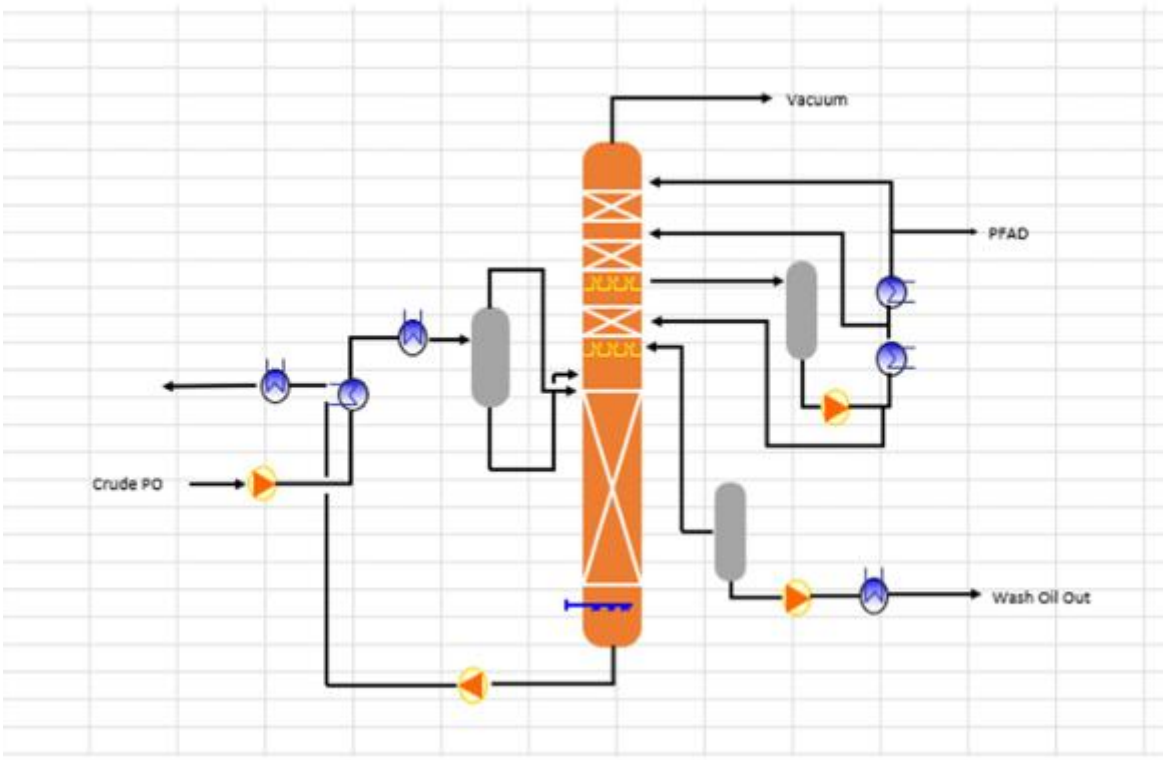
If the column has two stripping section beds, the top bed height of 1.3 meters with 1.3 mbar pressure drop. The higher-pressure drop is due to the pump around bed which has higher flow parameters. The middle bed height is approximately 1.6 meters with 0.6 mbar pressure drop. Total tower pressure drop may be in the range of 4 to 6 mbar. A typical column diameter is one to two meters.

The stripping section is structured packing of KLM 250Y EC style with a bed height of four to 6 meters. The 6-meter bed might have approximately 3.92 mbar pressure drop.

Typical metallurgy is SS 304, SS 316 and SS 316L. Typical stripping steam ratio is 0.20 kg/hr for each MTPD. Free Fatty Acid (FFA) in the feed is about 5% wt and FFA out is about 0.05% wt.

Typically, about one stage per meter and packing pressure drop about 1 mbar per meter except for the pump around sections. Typically, design with about 20% safety factors.

Some Palm Oil Strippers may be more complex with an extra draw from the column.



Steps of a Column Revamp

1. Determine Current Capacity and Limitations

The first step of a revamp is to understand your current capacity and limitations. One way to determine your current capacity and limitations is a high load test. You develop a test and measure all of the relevant data. Determine what is the current capacity and limitations. Limitations might include packing, heat exchangers, or pumps.

2. Develop a simulation matching current data

From the high load test data, a simulation is developed for the complete system including heat exchangers and pumps. The simulation should match the field data. KLM is happy to assist in your simulation.

3. Review the equipment.

Heat Exchangers

Heat exchangers are rarely optimized or designed well because they are awarded to the low bidder. Then the operating unit pays for the low unit performance in heat exchanged and run length. Equipment should be bid on a cost and performance matrix.

When a performance matrix is utilized, the lowest bidder rarely wins the heat exchanger bidding. For many units heat exchangers set the run time between maintenance outages. With

today's high energy cost heat exchanger is a larger focus. KLM can assist with a better design and supply of heat exchangers.

Pumps

Pumps again are awarded to the lowest bidder. Typical low-cost pump efficiency might be 80% where newer pumps are above 90% - this is a large energy saving. If you need to replace a pump do not replace it with a low efficiency pump. KLM can assist with a high efficiency pump.

Packing

Structured Packing has performed well in Palm Oil Strippers. Distillation of Oleochemicals requires attention to the following key features:

- High Vacuum
- Low pressure drop
- Low bottom temperature
- Minimum holdup
- Short residence time

There is a trade off in structured packing between capacity and fractionation ability. Early columns were designed with KLM 450Y which has higher fractionation capacity, but lower capacity ability. Most newer columns are designed with KLM 250Y EC (Enhanced Capacity - 2nd Generation Structured Packing) which has less fractionation ability but higher capacity.

From the High load test, we can develop vapor and liquid flows. We can then calculate tower hydraulic loads. KLM is happy to assist in developing hydraulic profiles and hydraulic capacity.

To increase fractional ability

If your tower diameter is not the limiting case from the hydraulic profile, we can increase your style of packing to increase your fractioning ability. If you had KLM 250Y before, we could consider KLM 300Y, or KLM 350Y for better fractionation. If you had KLM 350Y before, we could consider KLM 400Y or KLM 450Y for better fractionation.

To increase capacity

If your tower diameter is the limiting case from the hydraulic profile and your fractionation ability is good, we could slightly decrease your style of packing to increase your capacity. If you had KLM 250X before we could consider KLM 200X. If you had KLM 350Y before we could consider KLM 300Y or KLM 250Y.

If you had KLM 250Y before we could consider KLM 250Y EC for increased capacity. The KLM 250Y and KLM 250Y EC has roughly the same efficiency but the KLM 250Y EC has about 25% to 40% more capacity.

Here are some average Number of Theoretical Stage Per Meter (NTSM) and Height of Equivalent Theoretical Plate (HETP). These numbers are based on low pressure of about 1 bar, low relative volatility, and good vapor / liquid distribution – see next section on distributors.

Style	NTSM	HETP
KLM 200 Y	2	20 Inches (508 mm)
KLM 250 Y	2.5	16 inches (406 mm)
KLM 350 Y	3.33	11 inches (279 mm)
KLM 500 Y	4	9.75 Inches (248 mm)

Distributors

For Packing systems, every designer will mention that the packing only works as well as the distributor, and then they normally provide a not optimized distributor. Normal industry standard distributors average about 90 to 92% distribution quality. Below 90% is considered poor and 94% is achievable.

For structured packing, the top two layers distribute the liquid. If each layer is 300mm and you have a 3-meter bed, you have lost efficiency in the top two layers (600mm) which is 20% of the bed height. Remember, higher efficiency will save energy and reduce energy costs.

For random packing, the packing will mix the liquid some, but not like the structured packing. It is more important to have an optimized distributor for random packing to increase efficiency and reduce energy.

KLM optimizes distributor design to gain as much distribution as possible, to improve the packing efficiency, reduce energy costs and save money. With energy in USA at USD $4.00 MMBtu and in Asia at 15.00 MMBtu, it does not take much energy saving to be real money.

Palm Oil has a viscosity challenge – make sure you have an optimized distributor for your Palm Oil Application.

Cost of Quality

There are hundreds of Random Packing Suppliers, some at very low cost. KLM only buys from Quality Suppliers that we have vetted.

Suppose you are buying random packing and thickness is 0.30 mm. Low-cost vendor provides you thickness of 0.28, which is difficult to measure. His raw material cost is 6.66% percent lower. If you have a large bed height, the crush strength of the thinner packing is lower and the bottom of the bed may be crushed, leading to a higher pressure drop.

Suppose you are buying random packing and the size is 25 mm. Low-cost vendor provides you packing of 27mm, which is difficult to measure. Packing is sold in 1-meter bags. The bags should be full and weighted, make sure the bags are weighted. Again, His raw material cost is 3 to 5% percent lower by using less metal.

We know that larger packing size gives more capacity, but less efficiency. If your energy cost is low this should not have a large effect, but most companies do not have low energy cost.

With energy in US at USD $4.00 MMBtu and in Asia at 15.00 MMBtu, it does not take much energy saving to be real money. There may be a good reason someone's packing is lower cost – and it may not actually save you money.

Conclusions

Palm Oil Stripper Columns are typically the first fractionation column in palm oil processing. There are many types and styles of Palm Oil Strippers. There are opportunities to revamp these towers. Revamps, when possible, should increase capacity and reduce energy. This can be accomplished with the proper design of heat exchangers, pumps, and distillation equipment. KLM would be happy to work with your team.

References

1. https://lipidlibrary.aocs.org/edible-oil-processing/palm-oil
2. https://centpro.com/news/everything-you-need-to-know-about-palm-oil refinery-process/
3. Peter Faessler, Karl Kolmetz, Kek Wan Seang, Siang Hua Lee, Advanced fractionation technology for the oleochemical industry, July 2007, Asia Pacific Journal of Chemical Engineering
4. Peter Faessler, Karl Kolmetz, Kek Wan Seang, Revamp Strategies for Fatty Acid Distillation Section In Oleo-chemical Plant, Sept 2004
5. K Kolmetz et al, Kolmetz Handbook of Process Equipment Design, Distillation Column Selection Sizing and Troubleshooting, Engineering Design Guidelines, May 2021
6. K Kolmetz et al, Kolmetz Handbook of Process Equipment Design, Distillation Column Packing Hydraulics Selection Sizing and Troubleshooting, Engineering Design Guidelines, March 2011

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