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In This Issue

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SPECIAL Features

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Relief Valve Sizing for Two Phase Flow

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Process upsets are inevitable in every process facility built, and two-phase flow can develop in pressure vessels requiring over pressure protection. Over pressure situations can result on an account of a host of events such as inadvertent fires, control valve failures, runaway reactions, etc. This is can result in not only single phase flow, but also two phase flow where due to liquids flashing, the total volumetric flow rate can be higher than if the liquid in the vessel did not flash.

To perform two phase calculations, API 520, Part 1, provides distinct methods to estimate the mass flux and the required orifice area to choose the pressure relieving device (PRD). They are chiefly classified as,

1. Homogenous Equilibrium Model (HEM)

2. Two Point Omega [ω] method

3. Omega Method for subcooled liquids

The HEM method is a direct integration method whereby a series of isentropic flashes are made to arrive at the maximum mass flux emerging from a PRD. As the name suggests, omega [ω] method utilizes an omega [ω] factor to capture the relative expansion rate, as the fluid depressurizes across the PRD.

The following article covers sizing of two phase relief valves (RV) based on the HEM method, and Two point Omega method for flashing liquids with a boiling range > 150^0F.

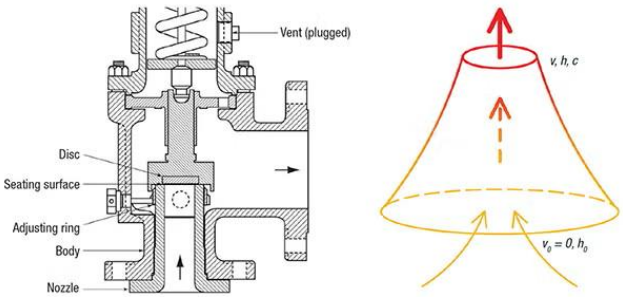


Figure 1. Two Phase Relief Flow [1]

General Notes

1. Two phase flow is often caused due to liquid entrainment, foamy or viscous fluids which hinder separation of vapour from liquids, liquid swelling due to dissolved gases that can generate bubbles, fluid expansion due to heating, high superficial vapour velocity, and runaway reactions.

Additionally when the process piping is not routed to avoid excessive pressure losses, then any excessive loss in pressure can cause liquid flashing prior to entering the pressure relief valve (PRV).

2. The HEM method assumes the two phase mixture consisting of a vapour and liquid, is sufficiently mixed until homogeneity due to high fluid velocities. Any phase change which occurs, proceeds until equilibrium at the saturation pressure, and the fluid continues to flow at the same velocity with no slip between the phases.

3. The Omega [ω] method correlates fluid expansion during depressurization with the critical pressure ratio [P_c]. The higher the ω value, that much more the fluid expands, and more likely to result in choked / critical flow. Depending on the fluid category, the value of ω changes [2],

   - For subcooled liquids, $\omega = 0$

   - Non flashing two phase flow, $0 < \omega < 1$

   - Vapour/Gas flow, $\omega \sim 1$

   - Flashing two phase flow, $\omega > 1$

4. A non-condensable gas (e.g., air, oxygen, nitrogen, CO_2, etc) does not easily condense under normal process conditions

5. A highly subcooled liquid can be defined as a fluid which does not flash when passing through the PRV.

6. Compared to ω method, the HEM method is more versatile [3] for,

   - Two phase liquid-vapour mixtures (including saturated liquids) where upon entering the pressure relief valve (PRV), the liquid flashes. Also applicable for fluids which are above and below the thermodynamic critical point in condensing flow. For e.g., Saturated liquid-vapour propane with no non-condensables enters the PRV, and the liquid propane flashes.

   - Saturated two phase mixture with non-condensable mixture, and the liquid flashes, e.g., Nitrogen and propane enter the PRV with liquid propane flashing.

   - Subcooled liquid without non-condensables enters the PRV and the liquid flashes, e.g., subcooled propane enters the PRV and flashes

   - Highly subcooled liquid with non-condensable gases and condensable vapour. For e.g., highly subcooled propane and nitrogen enters the PRV, but the propane component does not flash.

7. The Omega [ω] though reported in API 520, Part 1, Annexure C, has its limitations when applying to multicomponent systems with wide boiling point differences, or where composition changes are significant due to chemical reactions or large pressure changes [4]. Therefore its

use is limited to pure component systems. The omega [ω] method is also covered in Part 10 - ISO 4126, but cautions against its use in cases involving dissolved gases and immiscible liquids.

8. It is to be noted that irrespective of the method used, flow slip can affect the mass flux estimated, especially for viscous systems. Neglecting the slip can underestimate the mass flux, which results in an oversized two phase relief valve.

Methodology - HEM Method

The HEM method is a direct integration method of an isentropic nozzle flow. To determine the mass flux through a converging nozzle (which goes into estimating the PRV size), the nozzle is assumed to be adiabatic and reversible, of which both conditions must be satisfied for flow to be treated as isentropic. To estimate the mass flux, API 520 Part 1, Annexure C provides as follows,

$$G = \rho_t \times \left[\sqrt{-2 \times \left[\int_{P_{relief}}^{P} \frac{dP}{\rho}\right]}\right]_{max} \qquad (1)$$

Where,

G = Mass Flux [kg/s.m^2]

ρ_t = Mass density at nozzle throat [kg/m^3]

P = Stagnation pressure of the fluid [Pa]

P_{relief} = Pressure at nozzle inlet [Pa]

The subscript 'max' represents the point of potential choking when the mass flux passing through the nozzle is maximum. The PRV is also sized based on the maximum mass flux.

A key point when performing HEM direct integration is that the above energy balance holds true irrespective of the non-ideality or compressibility of the fluid. Solving the energy balance involves, estimating the mass density at various stagnation pressures from the nozzle inlet to the throat of the nozzle, at constant entropy. This can be arrived at by using a suitable physical property database and a thermodynamic model.

It is also important to note, that when generating successive pressure and mass density data points at constant entropy, the data points are generated until either the mass flux correlation reaches a maxima (representing choked conditions), or the actual back pressure on the nozzle is reached, whichever is first.

One can also employ an isenthalpic model across the nozzle, but this is under the condition that the mixture is of low mass quality, and far from the critical point in the phase envelope.

To solve for the mass flux integral in the energy balance, the integral can be evaluated numerically for any fluid by direct summation over small pressure intervals as follows,

$$\int \frac{dP}{\rho} \approx \sum_{i=i}^{i=i+1} \left[2 \times \left(\frac{P_{i+1}-P_i}{\rho_{i+1}+\rho_i}\right)\right] \qquad (2)$$

Where,

ρ_i = Overall mass density at stagnation pressure [P_i] [Pa]

Based on the maximum mass flux estimated over various pressures across the nozzle, the required orifice area can be estimated as,

$$A = \frac{277.8 \times W}{K_d \times K_b \times K_c \times K_v \times G} \qquad (3)$$

Where,

W = Mass flow rate, [kg/h]

A = Required effective discharge area [mm^2]

K_d = Rated coefficient of discharge that should be obtained from the valve manufacturer. For preliminary sizing, a discharge coefficient of 0.85 can be used for a two phase mixture or saturated liquid entering the PRV inlet [-]

K_b = Capacity correction factor due to back pressure. This can be obtained from the manufacturer's literature or estimated for preliminary sizing from Fig 30 of Ref [3]. The back pressure correction factor applies to balanced bellow PRV only. For conventional and pilot operated valves, K_b is equal to 1.0 [-]

K_c = Combination correction factor for installations with a rupture disk (RD) upstream of PRV. K_c = 1.0 when an RD is not installed, and 0.9 when RD+PRV combination is installed with no certified value [-]

K_v = Correction factor due to viscosity [-]. For conventional or pilot operated relief valve, K_v can be taken as 1.0. For balanced bellow as determined from Figure 37 of Ref [3] or from the following equation,

$$K_V = \left(0.9935 + \frac{2.878}{Re^{0.5}} + \frac{342.75}{Re^{1.5}}\right)^{-1} \qquad (4)$$

$$Re = \frac{Q \times 18800 \times SG}{\mu \times \sqrt{A_r}} \qquad (5)$$

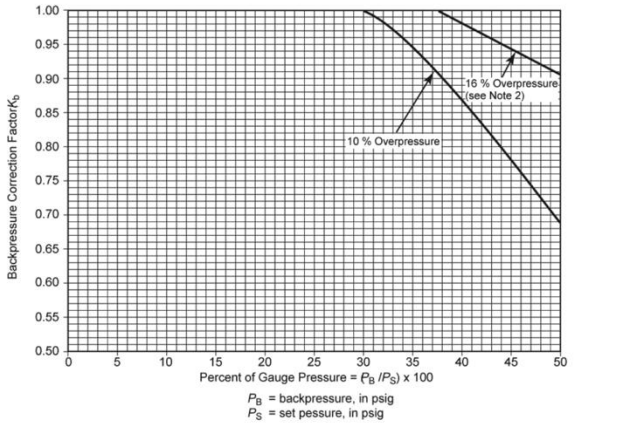
Q = Volumetric flow rate [lit/min]

Re = Reynolds number [-]

SG = Liquid specific gravity at flowing temperature referred to water at standard conditions [-]

μ = Viscosity at flowing temperature [cP]

A_r = Effective discharge area [mm^2] [API 526]



NOTE 1 The curves above represent a compromise of the values recommended by a number of relief valve manufacturers and may be used when the make of the valve or the critical flow pressure point for the vapor or gas is unknown. When the make of the valve is known, the manufacturer should be consulted for the correction factor. These curves are for set pressures of 50 psig and above. They are limited to back pressure below critical flow pressure for a given set pressure. For set pressures below 50 psig or for subcritical flow, the manufacturer must be consulted for values of K_b.

NOTE 2 See 5.3.3.

NOTE 3 For 21 % overpressure, K_b equals 1.0 up to P_B/P_S = 50 %.

Figure 30—Backpressure Correction Factor, K_b, for Balance Spring-loaded PRV (Vapors and Gases)

Figure 2. Fig 30, API 520 Part 1, 9th Edition [3]

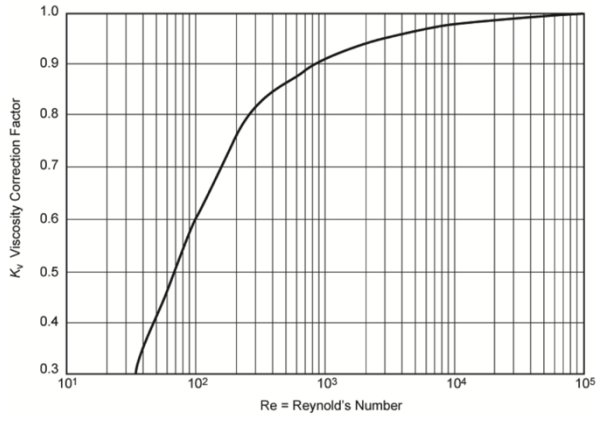


Figure 37—Capacity Correction Factor, K_v, Due to Viscosity

Figure 3. Fig 37, API 520 Part 1, 9th Edition [3]

Methodology – Two Point ω Method

In the two point ω method, (can be used for both flashing and non-flashing fluids without non-condensables in condensing two phase flow, both above and below the thermodynamic critical point), the omega [ω] parameter is determined using the specific volume data for a mixture at the stagnation conditions and one additional pressure point. The steps involved include,

1. Estimate the Omega [ω] parameter, using two pressure specific volume data points

2. Determine if flow is critical or subcritical

3. Calculate the Mass flux

4. Calculate required Orifice Area

The omega parameter can be calculated as,

$$\omega = 9 \times \left[\frac{\upsilon_9}{\upsilon_0} - 1\right] \qquad (6)$$

Where,

υ_9 = Specific volume evaluated at 90% of the PRV inlet pressure [P_{relief}] [ft^3/lb]

υ_0 = Specific volume of the two phase system at PRV inlet pressure [P_{relief}] [ft^3/lb]

It is to be noted that when estimating υ_9, the flash calculation must be carried out isentropically. Any isenthalpic calculation

must be performed only for low mass quality mixtures, far from the critical point.

To determine if flow is critical or subcritical,

$$P_c = \eta_c \times P_{relief} \qquad (7)$$

$$P_c \geq P_b \rightarrow Critical\ Flow \qquad (8)$$

$$P_c < P_b \rightarrow Subcritical\ Flow \qquad (9)$$

Where,

P_c = Critical Pressure [psia]

P_b = PRV Total backpressure [psia]

P_{relief} = PRV relieving pressure [PRV Set pressure, psig + Allowable overpressure, psi + Atmospheric Pressure] [psia]

η_c = Critical Pressure ratio[-] [Fig C1, API 520]

To simplify the critical pressure ratio estimation from using iterative methods, an approximation can be taken as follows,

$$\eta_c = \left[1 + \left[\left(1.0446 - \left(0.0093431\sqrt{\omega}\right)\right) \times \omega^{-0.56261}\right]\right]^{\left[-0.70356 + \left(0.014685 \times ln(\omega)\right)\right]} \qquad (10)$$

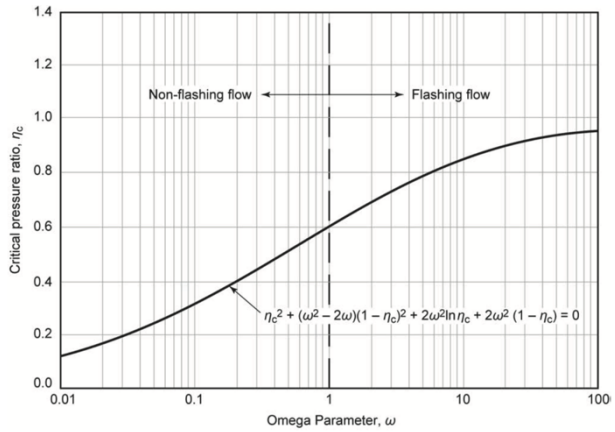


Figure C.1—Correlation for Nozzle Critical Flow of Flashing and Nonflashing Systems

Figure 4. Fig C.1, API 520 Part 1, 9th Edition [3]

The mass flux is calculated as follows,

For Critical Flow,

$$G = 68.09 \times \eta_c \sqrt{\frac{P_{relief}}{v_0 \times \omega}} \qquad (11)$$

For Subcritical flow,

$$G = \frac{68.09 \times \sqrt{\left[-2 \times \left[\omega ln(\eta_a) + [(\omega - 1) \times (1 - \eta_a)]\right]\right]}}{\left[\omega \times \left(\frac{1}{\eta_a} - 1\right)\right] + 1} \times \sqrt{\frac{P_{relief}}{v_0}} \qquad (12)$$

Where,

$\boldsymbol{\eta}_a$ = Backpressure ratio [P_b / P_{relief}] [-]

G = Mass flux [lb/s.ft^2]

P_{relief} = PRV relieving pressure [PRV Set pressure, psig + Allowable overpressure, psi + Atmospheric Pressure] [psia]

v_0 = Specific volume of the two phase system at PRV inlet pressure [P_{relief}] [ft^3/lb]

The required area of the PRV is estimated as,

$$A = \frac{0.04 \times W}{K_d \times K_b \times K_c \times K_v \times G} \qquad (13)$$

Where,

W = Mass flow rate, [lb/h]

A = Required effective discharge area [in^2]

K_d = Rated coefficient of discharge that should be obtained from the valve manufacturer. For preliminary sizing, a discharge coefficient of 0.85 can be used for a two phase mixture or saturated liquid entering the PRV inlet [-]

K_b = Capacity correction factor due to back pressure. This can be obtained from the manufacturer's literature or estimated for preliminary sizing from Fig 30 of Ref [3]. The back pressure correction factor applies to balanced bellow PRV only. For conventional and pilot operated valves, K_b is equal to 1.0 [-]

K_c = Combination correction factor for installations with a rupture disk (RD) upstream of PRV. K_c = 1.0 when an RD is not installed, and 0.9 when RD+PRV combination is installed with no certified value [-]

K_v = Correction factor due to viscosity [-]. For conventional or pilot operated relief valve, K_v can be taken as 1.0. For balanced bellow as determined from Figure 37 of Ref [3].

Case Study

A two phase pressure relief valve (PRV) is to be selected during the front end engineering stage (FEED) for hydrocarbon service. The process parameters are as follows,

Table 1. Vapour Composition

Components	mol%
Methane	40.0
Ethane	20.0
Propane	10.0
i-Butane	10.0
n-Butane	10.0
i-Pentane	4.0
n-Pentane	4.0
n-Hexane	2.0
Total	100

The relief valve settings are as follows,

Table 2. Two Phase PRV Inputs

PRV parameters	Value	Units
RV Inlet Temperature [T_{relief}]	27	^{0}C
RV Set Pressure [P_{set}] [=DP]	13.64	barg
Allowable Over Pressure [R_o]	10	%
Total backpressure [P_b]	1.364	barg
Relief rate [W]	35,961	kg/h
Mass Entropy	145	kJ/kmol.K
Liquid Density [ρ_l]	555.1	kg/m^3
Liquid Dynamic Viscosity	0.1483	cP
Molecular Weight	35.83	kg/kmol
Inlet 2 Phase Sp. volume [v_0]	0.4516	ft^3/lb
Inlet 2 Phase Sp. volume [v_9]	0.5070	ft^3/lb
Rupture Disc Present	Yes	

HEM Direct Integration Method

The relieving pressure is calculated as,

$$P_{relief} = 1.1 \times 13.64 = 15\ barg \qquad (14)$$

$$\%\ Gauge\ Pressure = \frac{1.364}{1364} \times 100 = 10\% \qquad (15)$$

With a % gauge pressure of 10%, either a conventional or bellow type PRV can be chosen. Plotting the two phase mass density for various pressures (from P_{relief} = 15 barg to P_b of 1.364 barg) at a constant entropy of 145 kJ/kmol.K, the plot is as follows,

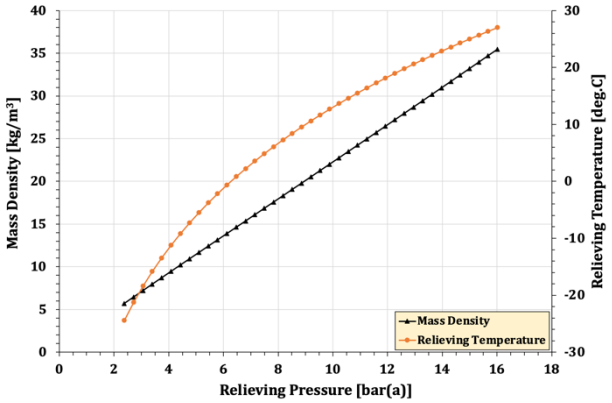


Figure 5. Two phase Mass Density and Temperature vs Relieving Pressure

The mass flux is calculated at each relieving pressure point. At the beginning of PRV relief, the relieving pressure is 15 barg (16.01325 bara) and 27^0C, with a mass density of 35.47 kg/m^3. Since no mass exits at 15 barg, the mass flux is zero.

Taking the next pressure point at P_{relief} = 15.67 bara, the mass density is 34.71 kg/m^3, and the integrand becomes,

$$Integrand\ 1 = -2 \times \left[\frac{(15.67-16.01325)\times 10^5}{34.71+35.47}\right] \quad (16)$$

$$Integrand\ 1 \approx 971.6\ m^2/s^2 \quad (17)$$

$$Cumulative = 0 + 971.6 = 971.6\ m^2/s^2 \quad (18)$$

$$G_1 = 34.71 \times \sqrt{2 \times 971.6} \approx 1530\, kg/s.m^2 \quad (19)$$

Taking the next pressure point at P_{relief} = 15.33 bara, the mass density is 33.95 kg/m^3, and the integrand becomes,

$$Integrand\ 2 = -2 \times \left[\frac{(15.33-15.67)\times 10^5}{33.95+34.71}\right] \quad (20)$$

$$Integrand\ 2 \approx 993\ m^2/s^2 \quad (21)$$

$$Cumulative = 971.6 + 993 \approx 1964.5\ m^2/s^2 \quad (22)$$

$$G_2 = 33.95 \times \sqrt{2 \times 1964.5} \approx 2128\, kg/s.m^2 \quad (23)$$

Taking the next pressure point at P_{relief} = 14.99 bara, the mass density is 33.2 kg/m^3, and the integrand becomes,

$$Integrand\ 3 = -2 \times \left[\frac{(14.99-15.33)\times 10^5}{33.2+33.95}\right] \quad (24)$$

$$Integrand\ 3 \approx 1015.3\ m^2/s^2 \quad (25)$$

$$Cumulative = 1015.3 + 1964.5 \approx 2979.8\ m^2/s^2 \quad (26)$$

$$G_3 = 33.2 \times \sqrt{2 \times 2979.8} \approx 2563\ kg/s.m^2 \quad (27)$$

Repeating for all other pressure points until 1.364 barg, a mass flux [G] vs. relieving pressure is plotted to find the maxima [G_{max}]

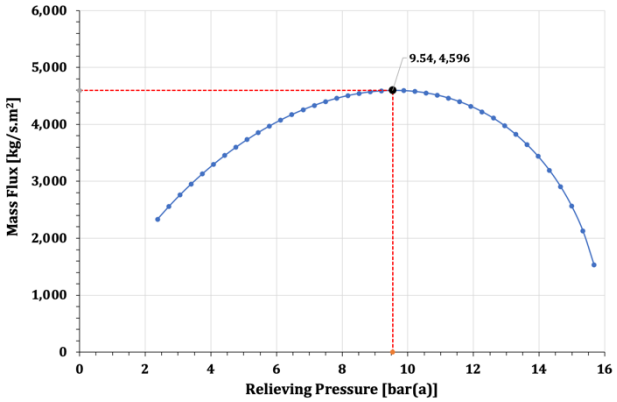


Figure 6. Mass Flux [G] vs Relieving Pressure

From the above plot, it is seen that the maximum mass flux is 4,596 kg/s.m^2. This occurs at a relieving pressure of 9.54 bara (8.52 barg) which is greater than the total backpressure of 1.364 barg, indicating that the flow through the pressure relief valve (PRV) is critical / choked flow.

Based on the estimated maximum mass flux of 4,596 kg/s.m^2, the two phase PRV is now sized. Since the PRV can be chosen between a conventional type and balanced bellow, a balanced below is chosen initially to check.

For the initial orifice area, with no viscosity correction factor, K_v is taken as 1.0. the value of K_d for preliminary sizing is 0.85; and K_c = 0.9 since a rupture disc (RD) is present in combination with the PRV.

The value of K_b is estimated based on Fig 30 of API 520, Part 1 as K_b = 1.0 for a % gauge backpressure of 10%.

The initial orifice area is therefore,

$$A = \frac{277.8 \times 35{,}961}{0.85 \times 1.0 \times 0.9 \times 1.0 \times 4596} \quad (28)$$

$$A = 2841.6\ mm^2 = 4.404\ in^2 \quad (29)$$

Calculating the corrected K_v with the initial orifice area size,

$$Re = \frac{\left(\frac{35{,}961}{555.1} \times \frac{1000}{60}\right) \times 18800 \times \frac{555.1}{1000}}{0.1483 \times \sqrt{2841.6}} = 1{,}424{,}942 \quad (30)$$

$$K_V = \left(0.9935 + \frac{2.878}{1424942^{0.5}} + \frac{342.75}{1424942^{1.5}}\right)^{-1} \quad (31)$$

$$K_V = 1.0 \quad (32)$$

Recalculating the required orifice area with the corrected K_v,

$$A = \frac{277.8 \times 35{,}961 \times 0.00155}{0.85 \times 1.0 \times 0.9 \times 1.0 \times 4596} = 4.404\ in^2 \quad (33)$$

From the above calculation, it can be seen that between a conventional type PRV and a balanced bellow type, it makes no difference since the % gauge backpressure is low [10%]. From an operations and cost perspective, for the same orifice area, balanced bellow PRVs are more expensive to purchase and maintain, leaving conventional type PRVs as a viable choice for this case study.

Based on the orifice area of 4.404 in^2, from API 526, the next higher size of PRV available is designation 'P' with an orifice area of 6.38 in^2. The rated capacity [W_{rated}] and % capacity utilization of the P designated PRV is,

$$\%\ Capacity = \frac{4.404}{6.38} \times 100 = 69\% \quad (34)$$

$$W_{rated} = \frac{35{,}961}{0.69} = 52{,}092\ kg/h \quad (35)$$

Omega [ω] Method

Performing the same case study using the omega method in imperials units, the PRV set pressure [P_{set}] is,

$$P_{set} = 13.64\ barg = 197.8\ psig \quad (36)$$

$$W = 35{,}961\ kg/h = 79{,}281 lb/h \quad (37)$$

$$P_b = 1.364\ barg = 19.78\ psig \quad (38)$$

$$P_{relief} = 1.1 \times 197.83 = 217.62\ psig \quad (39)$$

$$\%\ Gauge\ Pressure = \frac{19.78}{197.8} \times 100 = 10\% \quad (40)$$

To calculate the Omega [ω] parameter, the specific volume [v_9] is calculated at 90% of relieving pressure and flashed isentropically.

$$90\%\ of\ P_{relief} = 0.9 \times (217.62 + 14.7) = 209.1\ psia \quad (41)$$

$$v_9 = 0.507\ ft^3/lb \quad (42)$$

$$v_0\ at\ P_{relief}[217.62\ psig] = 0.4516\ ft^3/lb \quad (43)$$

$$\omega = 9 \times \left[\frac{0.507}{0.4516} - 1\right] = 1.105 \qquad (44)$$

With an omega [ω] factor of 1.105, i.e., $\omega > 1$, the fluid flow state is flashing two phase flow.

The next step is to check if the flow is critical or subcritical, and is calculated as,

$$\eta_c = \left[1 + \left[\left(1.0446 - \left(0.0093431\sqrt{1.105}\right)\right) \times 1.105^{-0.56261}\right]\right]^{\left[-0.70356 + \left(0.014685 \times ln(1.105)\right)\right]} \qquad (45)$$

$$\eta_c = 0.6194 \qquad (46)$$

The critical pressure is calculated as,

$$P_c = 0.6194 \times [217.62 + 14.7] = 143.9\ psia \qquad (47)$$

Since P_c = 143.9 psia > P_b = 19.78 psia, the flow is critical / choked flow.

Estimating the mass flux for critical flow, we have,

$$G = 68.09 \times 0.6194 \times \sqrt{\frac{232.32}{0.4516 \times 1.105}} \qquad (48)$$

$$G = 910.1\ lb/s.ft^2 \qquad (49)$$

Based on the estimated maximum mass flux of 910.1 lb/s.ft^2, and taking a balanced below type PRV, the initial orifice area, with no viscosity correction factor, K_v is taken as 1.0; K_d for preliminary sizing is 0.85; and K_c = 0.9 since an RD is present in combination with the PRV. The value of K_b is estimated based on Fig 30 of API 520, Part 1 as K_b = 1.0 for a % gauge backpressure of 10%.

The initial orifice area is therefore,

$$A = \frac{0.04 \times 79{,}281}{0.85 \times 1.0 \times 0.9 \times 1.0 \times 910.1} = 4.55\ in^2 \qquad (50)$$

Calculating the corrected K_v with the initial orifice area size,

$$Re = \frac{\left(\frac{35{,}961}{555.1} \times \frac{1000}{60}\right) \times 18800 \times \frac{555.1}{1000}}{0.1483 \times \sqrt{(4.55 \times 645.16)}} = 1{,}401{,}234 \qquad (51)$$

$$K_V = \left(0.9935 + \frac{2.878}{1401234^{0.5}} + \frac{342.75}{1401234^{1.5}}\right)^{-1} \qquad (52)$$

$$K_V = 1.0 \qquad (53)$$

Recalculating the required orifice area with the corrected K_v = 1, the orifice area remains the same at 4.55 in^2.

Based on the orifice area of 4.55 in^2, from API 526, the next higher size of PRV available is designation 'P' with an orifice area of 6.38 in^2. The rated capacity and % capacity utilization of the P designated PRV is,

$$\% \, Capacity = \frac{4.55}{6.38} \times 100 = 71.4\% \qquad (54)$$

$$W_{rated} = \frac{79{,}281}{0.714} = 111{,}052 \ \frac{lb}{h} \left[50{,}373 \ \frac{kg}{h}\right] \quad (55)$$

References

1. *https://www.chemengonline.com/sizing-pressure-relief-valves-two-phases/*

2. "Two Phase Flow", Robert A. Sadowski, E^2G Industry Insights, Vol 7, 2019

3. "Sizing, Selection, and Installation of Pressure relieving devices", Part 1-Sizing and Selection, API Standard 520, 9^{th} Ed, July 2014

4. "Forget the Omega Method", An ioMosaic White Paper, G.A. Melhem

Appendix A – HEM Direct Integration Method

Two Phase Relief Flow [HEM] - API 520 9th Ed [Sec C 2.1] [Blocked Outlet Case]		
Gas Composition	**Value**	**Unit**
Methane	40.00	mol%
Ethane	20.00	mol%
Propane	10.00	mol%
i-butane	10.00	mol%
n-butane	10.00	mol%
i-Pentane	4.00	mol%
n-Pentane	4.00	mol%
n-Hexane	2.00	mol%
Total	100.00	mol%
PSV Operating Parameters		
RV Inlet Temperature [T_{relief}]	27.00	^{0}C
RV Set Pressure [P_{set}] [Same as Design Pressure]	13.64	barg
	14.65	bara
Allowable Over Pressure [R_o]	10	%
RV Total Back Pressure [P_b] [10% x P_{set}]	1.364	barg
	2.377	bara
Relief Rate [W]	35,961	kg/h
Molar Entropy	145.00	kJ/kgmol.K
PRV Relieving Pressure [P_{relief}] [1.1 x P_{set}]	15.00	barg
% Gauge Back Pressure	10.00	%
Maximum Mass Flux [G_{max}]	**4,596**	**kg/s.m^2**
Relieving Pressure at G_{max}	**9.54**	**bara**
PSV Orifice Area [A]		
PSV Type	Balanced Bellow	
Liquid Density [For Specific Gravity Calc] [ρ_l]	555.1	kg/m^3
Specific Gravity of Liquid [SG]	0.5551	-
Dynamic viscosity of liquid [μ]	0.1483	cP
Flow Rate at flowing temperature [Q]	1079.8	lit/min
Coefficient of Discharge [K_d]	0.85	-
Backpressure correction factor for vapour [K_b]	1.0000	-
Rupture Disc present	Yes	-
RD Correction Factor [K_c]	0.9	-
Initial Orifice Area (no μ correction) [K_v = 1.0] [A_R]	2841.6	mm^2
Reynolds Number	1,424,942	-
Viscosity correction factor [K_v]	1.0000	-
Orifice Area [A] with viscosity correction [K_v]	2841.6	mm^2
	4.404	in^2
Selected PSV Orifice Size [API 526]		
Selected Orifice Size	**6.380**	**in^2**
Selected Orifice Designation	**P**	**-**
% Utilization Capacity	**69.0**	**%**
Rated Capacity [W_{rated}]	**52,092**	**kg/h**

Fluid Properties from Aspen HYSYS V11 [Property Table - Utilities]						
Relieving Pressure [P_{relief}]	**Phase**	**Temperature**	**Mass Density [ρ_l]**	**Integrand**	**Cumulative**	**Mass Flux [G]**
bara	**-**	**[^{0}C]**	**kg/m^3**	**m^2/s^2**	**m^2/s^2**	**kg/s.m^2**
16.01	L-V	27.00	35.47	0.0	0.0	0
15.67	L-V	26.34	34.71	971.6	971.6	1,530
15.33	L-V	25.68	33.95	993.0	1964.5	2,128
14.99	L-V	24.99	33.20	1015.3	2979.8	2,563
14.65	L-V	24.30	32.45	1038.6	4018.4	2,909
14.31	L-V	23.59	31.69	1063.0	5081.4	3,195
13.97	L-V	22.86	30.94	1088.5	6169.8	3,437
13.63	L-V	22.12	30.19	1115.2	7285.1	3,645
13.29	L-V	21.36	29.44	1143.2	8428.3	3,823
12.95	L-V	20.59	28.70	1172.7	9601.0	3,977
12.60	L-V	19.80	27.95	1203.6	10804.6	4,109
12.26	L-V	18.98	27.20	1236.2	12040.8	4,221
11.92	L-V	18.15	26.46	1270.6	13311.4	4,317
11.58	L-V	17.30	25.71	1306.9	14618.3	4,397
11.24	L-V	16.42	24.97	1345.2	15963.5	4,462
10.90	L-V	15.52	24.23	1385.8	17349.3	4,513
10.56	L-V	14.60	23.49	1429.0	18778.3	4,551
10.22	L-V	13.65	22.75	1474.8	20253.1	4,578
9.88	L-V	12.67	22.00	1523.6	21776.7	4,592
9.54	L-V	11.66	21.27	1575.7	23352.4	4,596
9.20	L-V	10.63	20.53	1631.5	24983.9	4,588
8.85	L-V	9.55	19.79	1691.3	26675.1	4,571
8.51	L-V	8.44	19.05	1755.6	28430.7	4,543
8.17	L-V	7.29	18.31	1824.9	30255.6	4,505
7.83	L-V	6.10	17.58	1899.9	32155.4	4,457
7.49	L-V	4.87	16.84	1981.2	34136.7	4,400
7.15	L-V	3.58	16.10	2069.9	36206.5	4,333
6.81	L-V	2.24	15.37	2166.8	38373.3	4,257
6.47	L-V	0.84	14.63	2273.2	40646.5	4,171
6.13	L-V	-0.62	13.89	2390.7	43037.2	4,076
5.79	L-V	-2.16	13.15	2520.9	45558.1	3,971
5.45	L-V	-3.77	12.42	2666.3	48224.4	3,856
5.10	L-V	-5.47	11.68	2829.6	51054.1	3,732
4.76	L-V	-7.28	10.94	3014.5	54068.5	3,597
4.42	L-V	-9.19	10.20	3225.5	57294.0	3,452
4.08	L-V	-11.24	9.46	3468.7	60762.7	3,297
3.74	L-V	-13.44	8.71	3752.3	64515.0	3,130
3.40	L-V	-15.82	7.97	4087.4	68602.4	2,951
3.06	L-V	-18.40	7.22	4489.8	73092.2	2,760
2.72	L-V	-21.25	6.47	4982.4	78074.6	2,555
2.38	L-V	-24.41	5.71	5600.1	83674.7	2,335
					Max G	**4,596**

Appendix B – Omega [ω] Method for Flashing Fluids

Two Phase RV Sizing [Omega Method] - API 520 9th Ed [Sec C 2.2] - Flashing flow [Blocked Outlet Case]		
Gas Composition	**Value**	**Unit**
Methane	40.00	mol%
Ethane	20.00	mol%
Propane	10.00	mol%
i-butane	10.00	mol%
n-butane	10.00	mol%
i-Pentane	4.00	mol%
n-Pentane	4.00	mol%
n-Hexane	2.00	mol%
Total	100.00	mol%
PSV Operating Parameters		
Molecular Weight [MW]	35.96	kg/kmol
Liquid Density [For Specific Gravity Calc] [ρ_l]	555.07	kg/m^3
Liquid Density [For Specific Gravity Calc] [ρ_l]	0.5551	-
Absolute viscosity of Liquid [μ]	0.1483	cP
RV Inlet Temperature [T_{relief}]	80.6	^{0}F
RV Set Pressure [P_{set}] [DP of Vessel]	197.83	psig
	212.53	psia
Allowable Over Pressure [R_o]	10	%
RV Total Back Pressure [P_b] a.k.a [P_a] [10% x P_{set}]	19.783	psig
Relief Rate [W] [a.k.a. Q in Reynolds number]	79,281	lb/h
	1079.8	lit/min
Two Phase Specific Volume at PSV Inlet [$\mathcal{V}_0$]	0.4516	ft^3/lb
PRV Relieving Pressure [P_{relief}] [1.1 x P_{set}] [P_0]	217.62	psig
	232.32	psia
% Gauge Back Pressure	10.0	%

Step 1: Calculate Omega Parameter [ω]		
90% of Relieving Pressure [0.9 x P_{relief}]	209.08	psia
Specific Volume at 209.08351368 psia [$\mathcal{V}_9$]	0.5070	ft^3/lb
Omega [ω] [Flashing liquids with boiling range > 150 ^{0}F]	**1.105**	-
Step 2: Determine is Flow is Critical or Subcritical		
Critical Pressure Ratio [η_C] [Eq C.15, API 520, 9th Ed]	0.6194	-
Critical Pressure [P_c] [$P_c = \eta_C$ x P_{relief}]	143.89	psia
Flow Condition	**Critical flow**	
Step 3: Calculate Mass Flux [G]		
Back Presure Ratio [η_a]	0.1484	-
Mass Flux [G] [Eq. 16, API 520, 9th Ed] [Critical Flow]	910.1	lb/s.ft^2
Mass Flux [G] [Eq. 16, API 520, 9th Ed] [Subcritical Flow]		lb/s.ft^2
Mass Flux [G]	910.1	lb/s.ft^2
Step 4: Calculate Area of PSV Orifice [A]		
PSV Type	Balanced Bellow	
Coefficient of Discharge [K_d]	0.85	-
Backpressure correction factor for vapour [K_b]	1.0000	-
Rupture Disc present	Yes	-
RD Correction Factor [K_c]	0.9	-
Initial Orifice Area (no μ correction) [K_v = 1.0] [A_R] [Eq. C.20, API 520, 9th Ed	4.55	in^2
Reynolds Number [Eq. 33 Sec 5.8.1.4, API 520, 9th Ed]	1,401,234	-
Viscosity correction factor [K_v] [Eq. 30, Sec 5.8.1.3, API 520, 9th Ed]	1.0000	-
Calculated Orifice Area [Eq. C20, API 520, 9th Ed]	4.55	in^2
Selected PSV Orifice Size [API 526]		
Selected Orifice Size	**6.380**	**in^2**
Orifice Designation	**P**	**-**
% Utilization Capacity	**71.4**	**%**
Rated Capacity [W_{rated}]	**111,052**	**lb/h**

Author

Jayanthi Vijay Sarathy holds a Master's Degree in Chemical Engineering from Birla Institute of Technology & Science (BITS), Pilani, India and is a Chartered Engineer from the Institution of Chemical Engineers, UK. His expertise over 18 years of professional experience covers Front End Engineering, Process Dynamic Simulation and Subsea/Onshore pipeline flow assurance in the Oil and Gas industry. Vijay has worked as an Upstream Process Engineer with major conglomerates of General Electric, ENI Saipem and Shell

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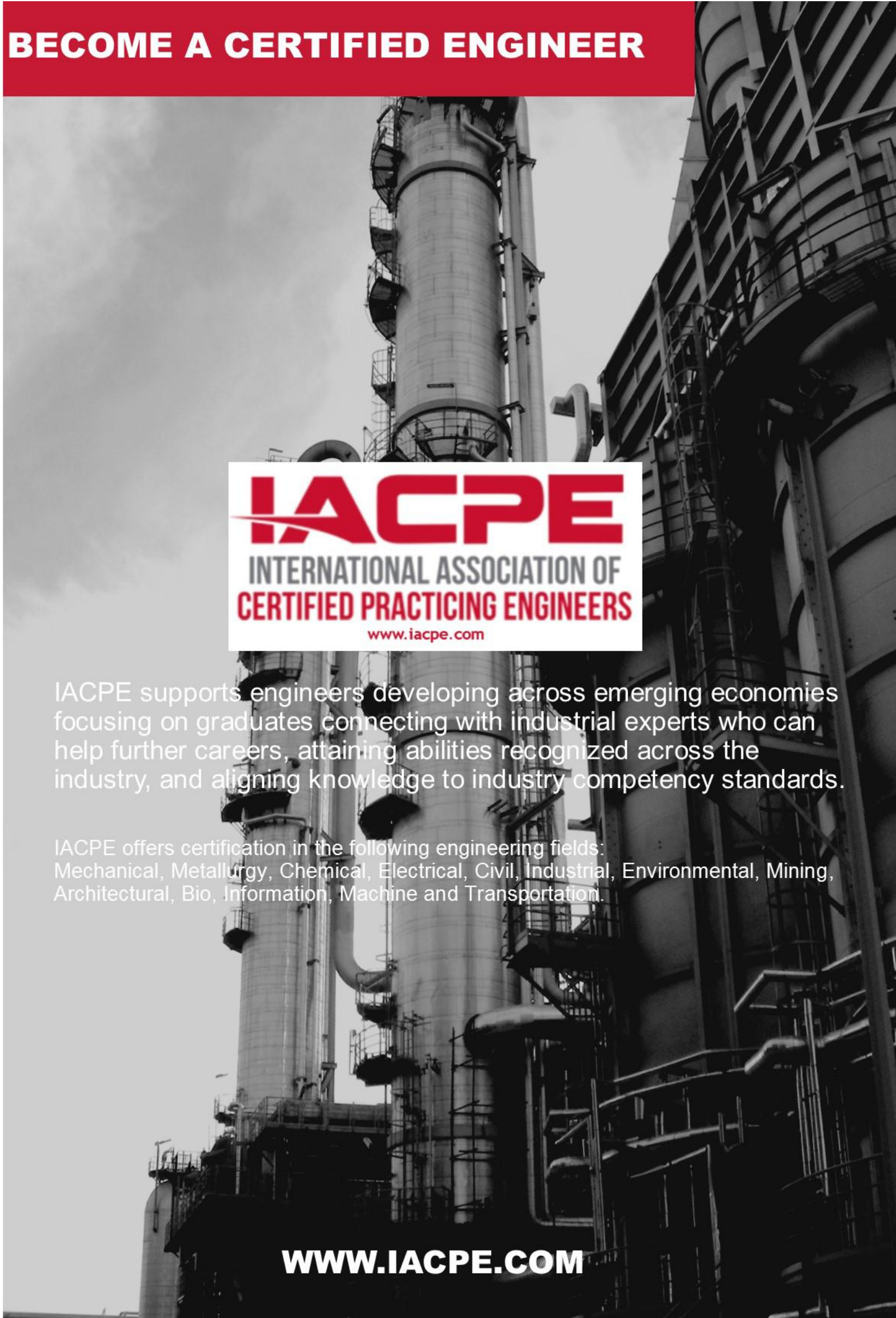

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Introduction and Context

According to some recent forecasts, the petrochemical market tends to rise in the next years and, in middle term, will be responsible by a major part of the crude oil consumption over passing the transportation fuels this fact have been made the refiners to looking for closer integration with petrochemical assets through the maximization of petrochemical intermediates in their refining hardware as a strategy to ensure better refining margins and higher value addition to the crude oil. Figure 1 presents an overview of the trend of growth in the petrochemical market in middle term.

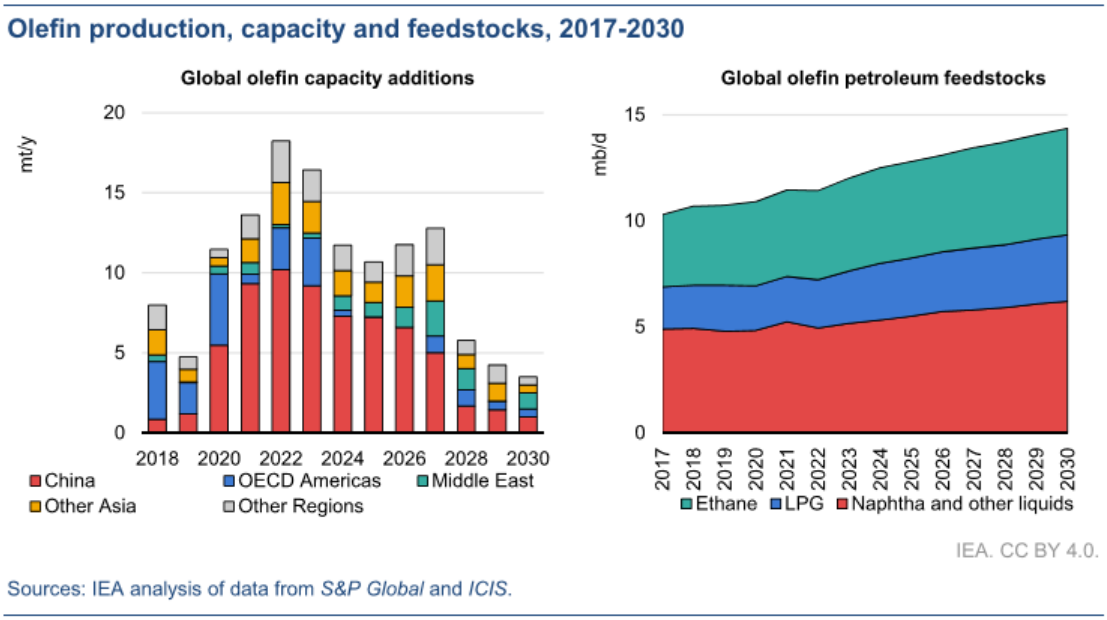

Figure 1 – Growing Trend in the Demand by Petrochemical Intermediates (IEA, 2025)

Some of the most promising petrochemical intermediates are the aromatics benzene and p-xylene. The maximization of aromatics in the refining hardware is possible through the installation of catalytic reforming technologies associated with aromatics separation unit. The catalyst applied to catalytic reforming units have a fundamental role in the aromatics yield and consequently to allow the achievement of profitable and reliable operation.

In the current scenario of the downstream industry, the refiners are facing important trends, the petrochemicals maximization as a strategy to ensure added value to the processed crude oil and the hydrogen question, which the refiners are facing a growing demand and environmental restrictions related to the CO2 emissions of the traditional steam reforming generation route. In this sense, the catalytic reforming units can develop a fundamental role in the strategy of some refiners.

Beyond the aromatics production, in markets with surplus of gasoline, some alternatives like blending the heavier fraction of naphtha with diesel and jet fuel can be an interesting strategy, but this alternative presents limitations due to the middle distillates specifications like volatility and Reid Vapor Pressure (RVP). In this case, technologic routes capable of managing naphtha molecules aiming to direct these streams to petrochemical intermediates can ensure closer integration with petrochemical assets as well as higher added value to refiners.

Again, being a high demand and most profitable market, the alternative to convert naphtha to petrochemicals should be a trend for refiners inserted in markets with gasoline surplus in the next years. According to data from Wood Mackenzie Company (2021), the highly integrated refiners can increase from US$ 0,68 to US$ 2,02/ bbl. Still according to Wood Mackenzie, the Asian Market presents the major concentration of integrated refining plants.

It's interesting to quote the potential competitive imbalance of the downstream industry in short term due to the growing demand for petrochemicals. Based on data from 2019 the total capital investments in crude to chemicals refineries is 300 billion US dollars and 64 % of this investment was made by Asian players, to reinforce this trend Figure 4 present a comparison between the relation of crude oil distillation capacity and the integrated refinery capacity for each continent.

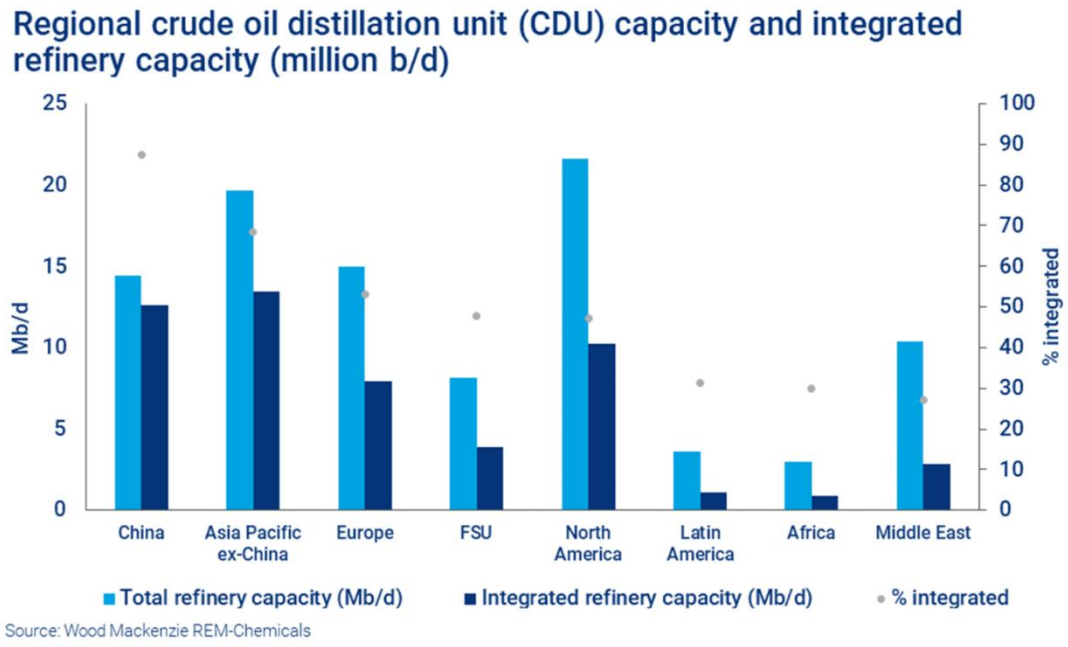

Figure 2 – Crude Oil Distillation Capacity and Integrated Refinery Capacity for Each Continent (Wood Mackenzie, 2023)

Figure 2 shows that the Asian players have a superior integration capacity of their refining assets in comparison with another continents, as mentioned above, this can be translated in a significant competitive advantage to the Asian players and a great potential o competitive imbalance of the downstream market considering the recent forecasts which indicates growing demand for petrochemicals. Furthermore, it's possible to see the power of the China in the Asian and global downstream market.

As presented in Figure 3, the petrochemicals demand tends to drive crude oil demand for the next years.

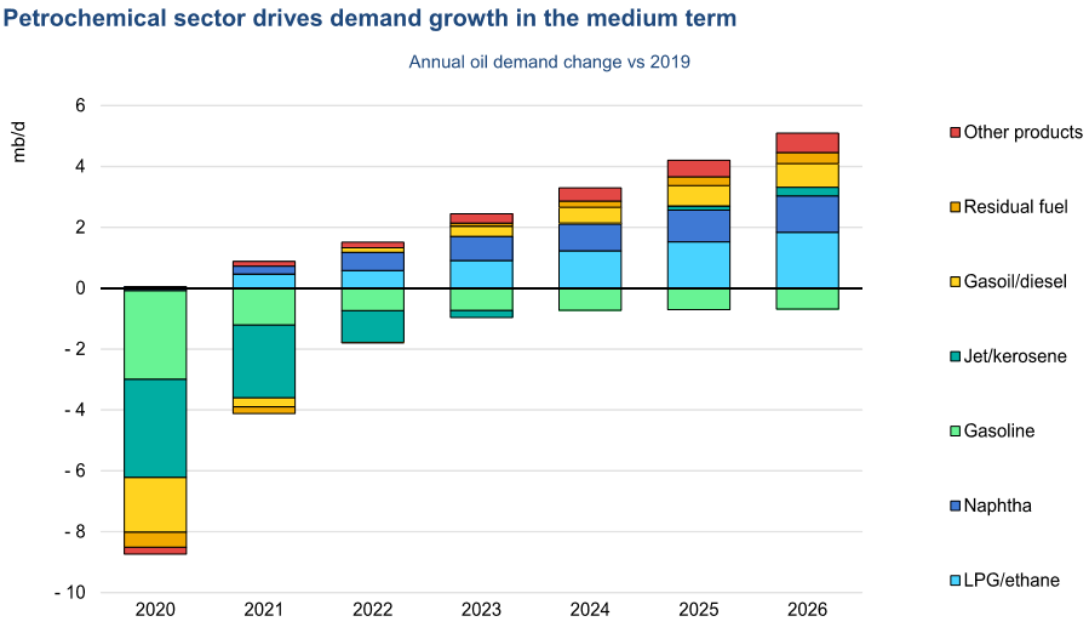

Figure 3 – Growth of Petrochemicals as Driver for Crude Oil Consumption (IEA, 2021)

Additionally, it's important to quote that the gasoline demand will be sustained by the in developing economies, as presented in Figure 4.

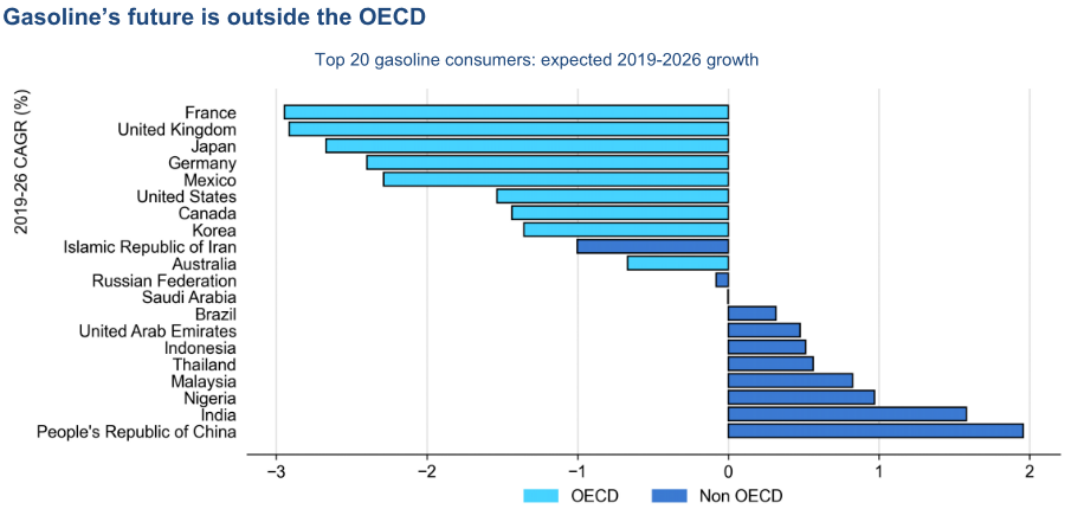

Figure 4 – Growth of Gasoline Demand for the Next Years (IEA, 2021)

This fact tends to restrict the consumer market which tends to offer lower refining margins, another great advantage to refiners capable of converting naphtha to petrochemicals against gasoline.

Based on description above it's possible to apply the article published by W. Chan Kim and Renée Mauborge called "Blue Ocean Strategy" in Harvard Business Review, to classify the competitive markets in the downstream industry. In this article the authors define the conventional market as

a red ocean where the players tend to compete in the existing market focusing on defeating competitors through the exploration of existing demand, leading to low differentiation and low profitability. The blue ocean is characterized by look for space in non-explored (or few explored markets), creating and developing new demands and reaching differentiation, this model can be applied (with some specificities once is a commodity market) to the downstream industry, considering the traditional transportation fuels refineries and the petrochemical sector.

Due his characteristics, the transportation fuels market can be imagined like the red ocean, where the margins tend to be low and under high competition between the players with low differentiation capacity. On the other side the petrochemicals sector can be faced like the blue ocean where few players are able to meet the market in competitive conditions, higher refining margins, and significant differentiation in relation to refiners dedicated to transportation fuels market. Figure 5 presents the basic concept of blue ocean strategy in comparison with the traditional red ocean where the players fight to market share with low margins.

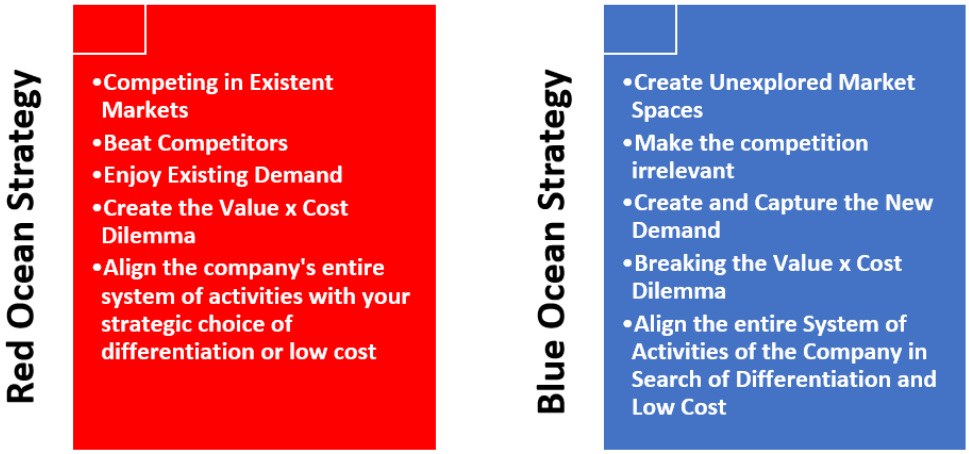

Figure 5 – Differences between Blue and Red Ocean Strategies (KIM & MAUBORGNE, 2004)

As presented above, the market forecasts indicate that the refiners able to maximize petrochemicals against transportation fuels can achieve highlighted economic performance in short term, in this sense, the crude oil to chemicals technologies can offer even more competitive advantage to the refiners with capacity of capital investment.

Can be difficult to some people to understand the term “differentiation” in the downstream industry once this is a market that deal with commodities, but the differentiation here is related to the capacity to reach more added value to the processed crude oil and as presented above, nowadays this is translated in the capacity to maximize the petrochemicals yield, creating differentiation between integrated and non-integrated players.

Considering 2024 as the base year, the petrochemical market size reached a total value of USD 659,22 billion with an expected compound annual growth rate (CAGR) of 6,11 % between 2024 and 2034 as presented in Figure 6.

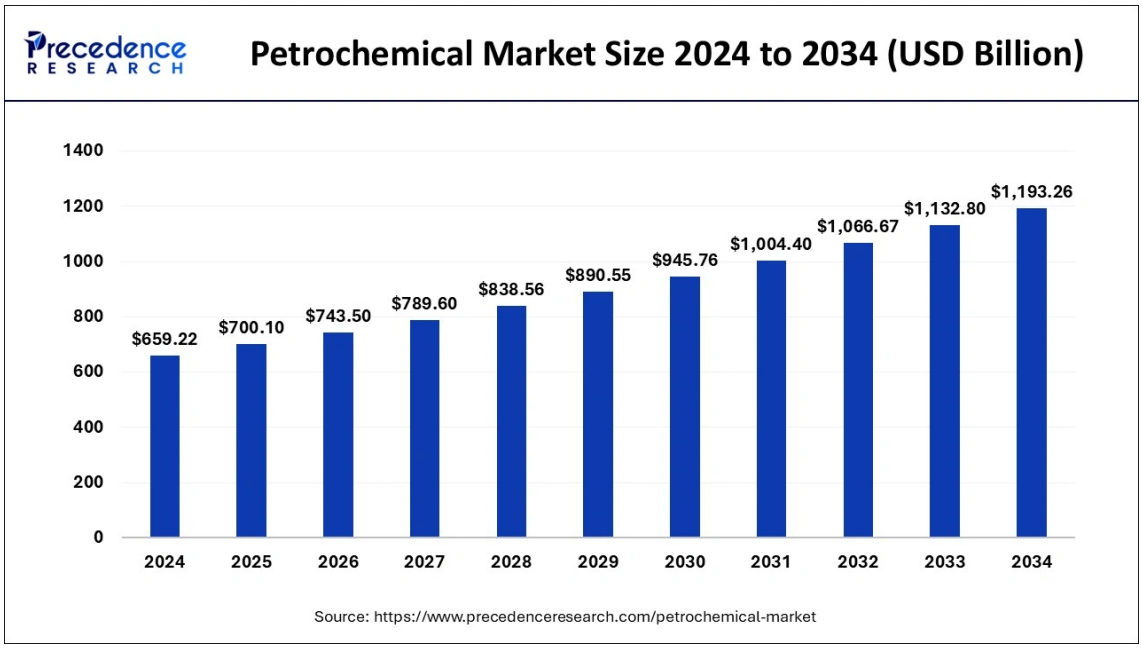

Figure 6 – Petrochemical Market Size Forecast 2023-2033 (Precedence Research, 2025)

Based on these data, the petrochemical market size can reach a total value of close USD 1.193,26 billion in 2034, reinforcing the attractiveness of the petrochemical market for the refiners under a scenario where the transportation fuels show in contraction demand and hostile scenario due to the necessity to reduce the carbon intensity of the energetic matrix.

Considering just the aromatics solvent market (Benzene, Toluene, and Xylenes) the CAGR expected between 2025 and 2034 is 4,80 % leading the aromatics solvent market size to reach USD 12,66 billion in 2030 still according to Precedence Research data.

Maximizing Added Value to the Processed Crude – Petrochemical Integration

The focus of the closer integration between refining and petrochemical industries is to promote and seize the synergies existing opportunities between both downstream sectors to generate value to the whole crude oil production chain. Table 1 presents the main characteristics of the refining and petrochemical industry and the synergies potential.

Table 1 – Refining and Petrochemical Industry Characteristics

Refining Industry	Petrochemical Industry
Large Feedstock Flexibility	Raw Material from Naphtha/NGL
High Capacities	Higher Operation Margins
Self Sufficient in Power/Steam	High Electricity Consumption
High Hydrogen Consumption	High Availability of Hydrogen
Streams with low added Value (Unsaturated Gases & C2)	Streams with Low Added Value (Heavy Aromatics, Pyrolysis Gasoline, C4's)
Strict Regulations (Benzene in Gasoline, etc.)	Strict Specifications (Hard Separation Processes)
Transportation Fuels Demand in Declining at Global Level	High Demand Products

As aforementioned, the petrochemical industry has been growing at considerably higher rates when compared with the transportation fuels market in the last years, additionally, represents a noblest destiny and less environmentally aggressive to crude oil derivatives. The technological bases of the refining and petrochemical industries are similar which lead to possibilities of synergies capable of reducing operational costs and adding value to derivatives produced in the refineries.

Figure 7 presents a block diagram that shows some integration possibilities between refining processes and the petrochemical industry.

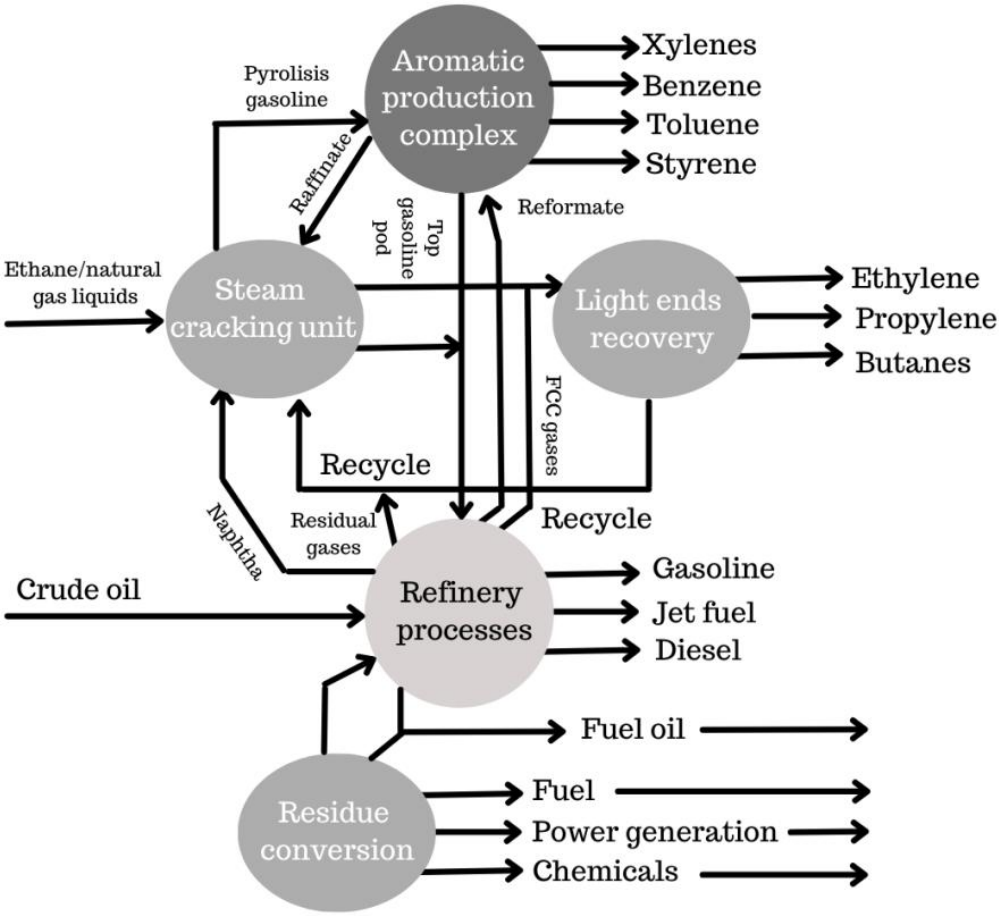

Figure 7 – Synergies between Refining and Petrochemical Processes

Process streams considered with low added value to refiners like fuel gas (C2) are attractive raw materials to the petrochemical industry, as well as streams considered residual to petrochemical industries (butanes, pyrolysis gasoline, and heavy aromatics) can be applied to refiners to produce high quality transportation fuels, this can help the refining industry meet the environmental and quality regulations to derivatives.

The integration potential and the synergy among the processes rely on the refining scheme adopted by the refinery and the consumer market, process units such as Fluid Catalytic Cracking (FCC) and Catalytic Reforming can be optimized to produce petrochemical intermediates to the detriment of streams that will be incorporated to fuels pool. In the case of FCC, installation of units dedicated to producing petrochemical intermediates, called petrochemical FCC, aims to reduce to the minimum the generation of streams to produce transportation fuels, however, the capital investment is high once the severity of the process requires the use of material with noblest metallurgical characteristics.

The IHS Markit Company proposed a classification of the petrochemical integration grades, as presented in Figure 8.

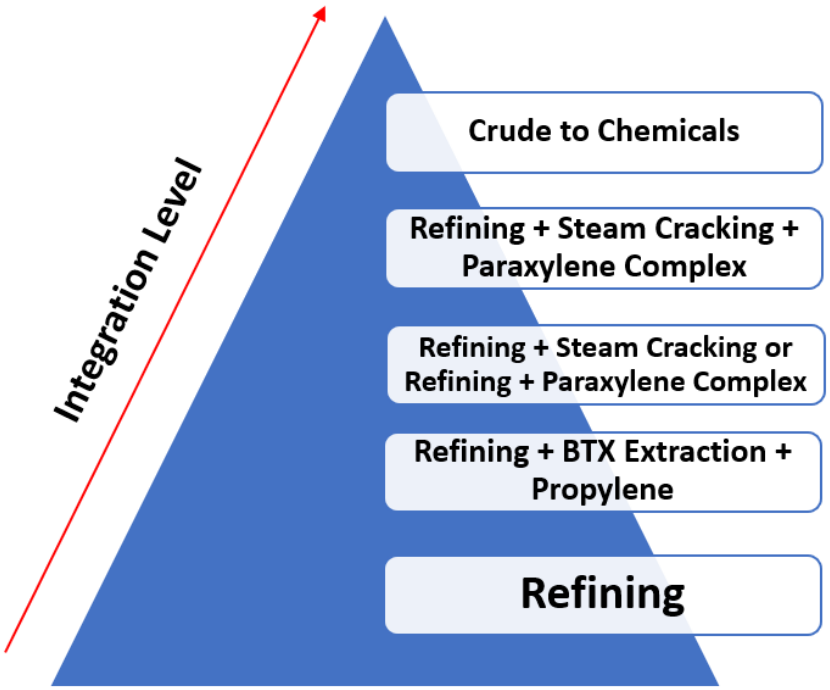

Figure 8 – Petrochemical Integration Levels (IHS Markit, 2018)

According to the classification proposed, the crude to chemicals refineries is considered the maximum level of petrochemical integration, where the processed crude oil is totally converted into petrochemical intermediates.

Catalytic Reforming Technologies – Naphtha to BTX

The main objective of the Catalytic Reforming unit is to produce a stream with high aromatics hydrocarbons content that can be directed to the gasoline pool or to produce petrochemical intermediates (benzene, toluene, and xylenes) according to the market served by the refiner, due the high content of aromatics compounds the reformate can significantly raise the octane number in the gasoline, in the current scenario this a less attractive route.

A typical feedstock to the catalytic reforming unit is the straight run naphtha, however, in the last decades due to the necessity to increasing the refining margin through installation of bottom

barrel units, hydrotreated coke naphtha stream has been consumed like feedstock in the catalytic reforming unit.

The catalyst generally employed in the catalytic reforming process is based on platinum (Pt) supported on alumina treated with chlorinated compounds to raise the support acidity. This catalyst has bifunctional characteristics once the alumina acid sites are active to molecular restructuring and the metals sites are responsible for hydrogenation and dehydrogenation reactions.

The main chemical reactions involved in the catalytic reforming process are:

- Naphthene Compounds dehydrogenation;
- Parafinns Isomerization;
- Isomerization of Naphthene Compounds;
- Paraffins Dehydrocyclization;

Among the undesired reactions can be cited hydrocracking reactions and dealkylation of aromatics compounds.

Figure 9 presents a basic process flow diagram for a typical semi-regenerative catalytic reforming unit.

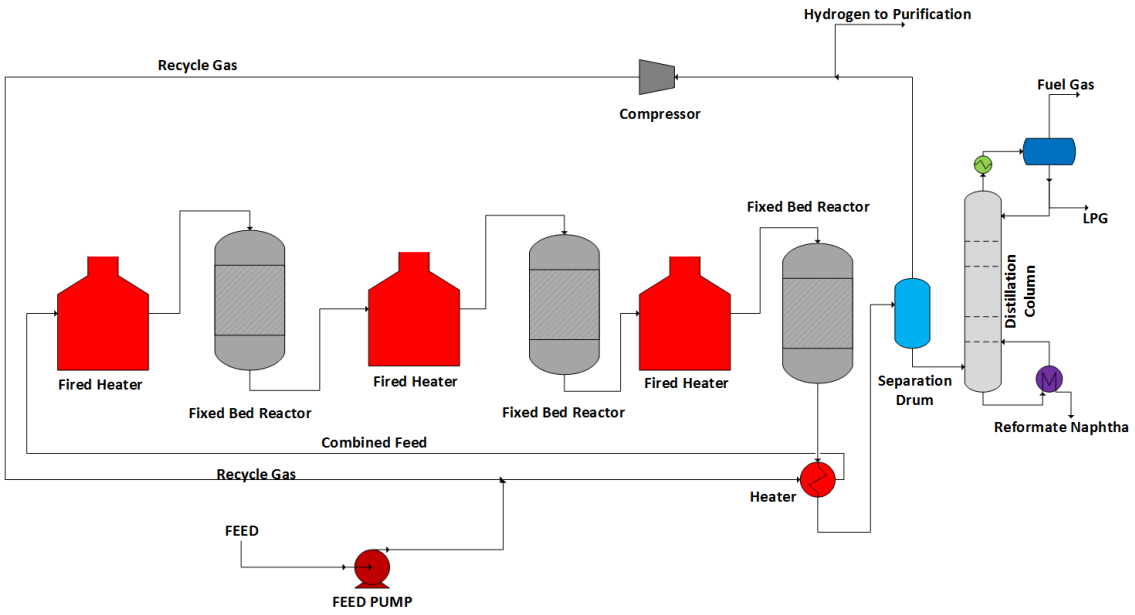

Figure 9 – Typical arrangement to Semi-regenerative Catalytic Reforming Process Unit

The naphtha feed stream is blended with recycle hydrogen and heated at a temperature varying from 500 to 550 oC before entering the first reactor, as the reactions are strongly endothermic the temperature fall quickly, so the mixture is heated and sent to the second reactor and so on. The

effluent from the last reactor is sent to a separation drum where the phases liquid and gaseous are separated.

The gaseous stream with high hydrogen content is shared in two process streams, a part is recycled to the process to keep the ratio H2/Feed stream the other part is sent to a gas purification process plant (normally a Pressure Swing Adsorption unit) to raise the purity of the hydrogen that will be exported to others process plants in the refinery.

The liquid fraction obtained in the separation drum is pumped to a distillation column wherein the bottom is produced the reformate and in the top drum of the column is produced LPG and fuel gas.

The reformate has a high aromatics content and, according to the market supplied by the refinery, can be directed to the gasoline pool like a booster of octane number or, when the refinery has aromatics extraction plants is possible to produce benzene, toluene, and xylenes in segregated streams, which can be directed to petrochemical process plants. The gas rich in hydrogen produced in the catalytic reforming unit is an important utility for the refinery, mainly when there is a deficit between the hydrogen production capacity and the hydrotreating installed capacity in the refinery, in some cases the catalytic reforming unit is operated with the principal objective to produce hydrogen.

The main process variables in the catalytic reforming process unit are pressure (3,5 – 30 bar), which normally is defined in the design step, in other words, the pressure normally is not an operational variable. The temperature can vary from 500 to 550 oC, the space velocity can be varied through feed stream flow rate control and the ratio H2/Feed stream that have a direct relation with the quantity of coke deposited on the catalyst during the process. To semi-regenerative units, the ratio H2/Feed stream can vary from 8 to 10, in units with continuous catalyst regeneration this variable can be significantly reduced.

Due to the process severity, the high coke deposition rate on the catalyst and consequently the quick deactivation leaves to short operational campaign periods to semi-regenerative units that employ fixed bed reactors. To solve this problem some technology licensors developed catalytic reforming process with continuous catalyst regeneration steps.

The process Aromizing™ developed by Axens company apply side by side configurations to the reactors while the CCR Platforming™ developed by UOP apply the configuration of stacked reactors to catalytic reforming process with continuous catalyst regeneration. Figure 10 presents a flow diagram to Aromazing™ catalytic reforming unit.

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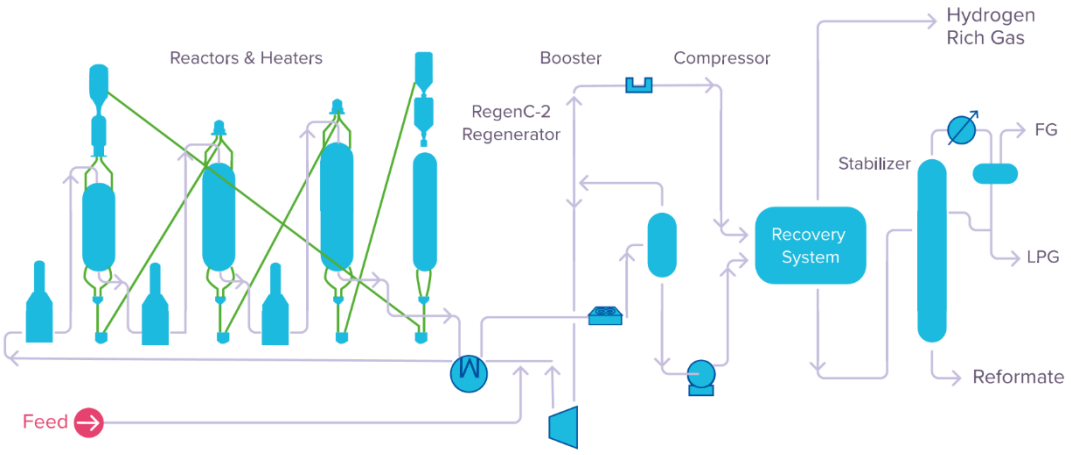

Figure 10 – Aromizing™ Reforming Technology by Axens Company

Both technologies are commercial and some process plants with these technologies are in operation around the world. Figure 11 presents a basic process flow diagram to CCR Platforming™ developed by UOP Company.

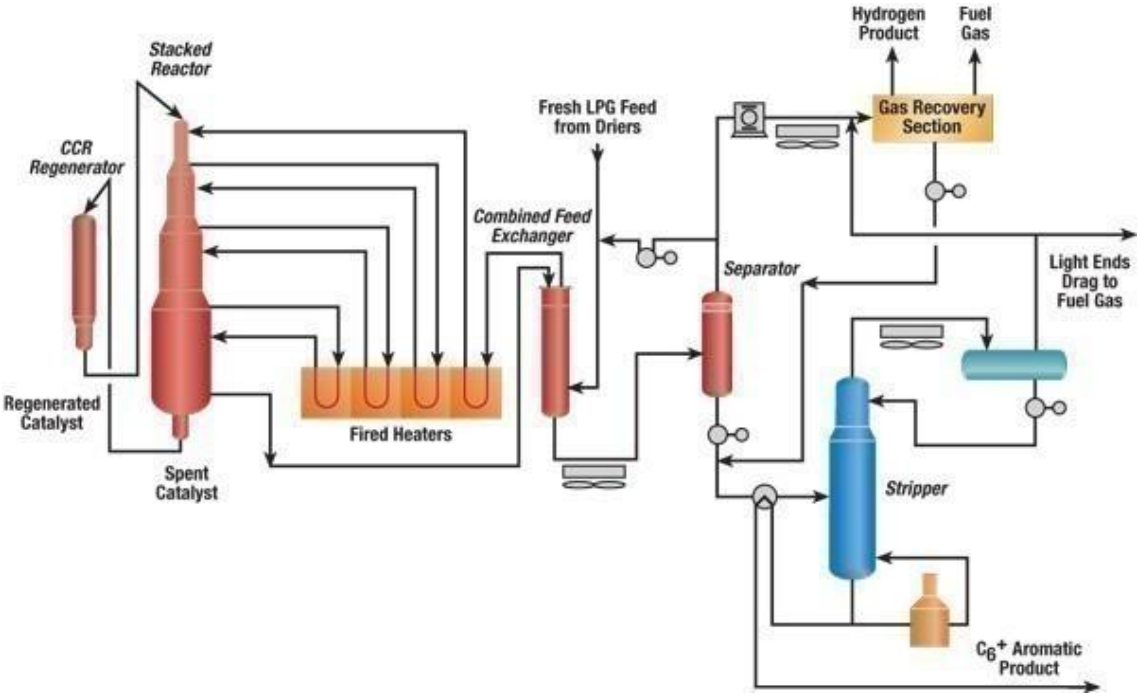

Figure 11 – CCR Platforming™ Reforming Technology by UOP Company

In the regeneration section the catalyst is submitted to processes to burn the coke deposited during the reactions and treated with chlorinated compounds to reactivate the acid function of the catalyst. Despite the higher capital investment, the rise in the operational campaign and higher flexibility in relation to the feedstock to be processed in the processing unit can compensate for the higher investment in relation to the semi-regenerative process.

Catalytic reforming technology gives great flexibility to the refiners in the gasoline production process, however, in the last decades there has been a strong restriction on the use of reformate in gasoline due to the control of benzene content in this derivate (due to the carcinogenic characteristics of this compound). This fact has been reduced the application of reformate in the gasoline formulation in some countries. Furthermore, the operational costs are high, mainly due to the catalyst replacement and additional security requirements linked to minimizing leaks of aromatics compounds.

Catalytic Reforming Catalysts

The catalysts applied to naphtha reforming are based on platinum carried on high purity alumina, in many cases is applied ruthenium or germanium as promoters to the catalyst activity. The catalyst has dual function. The metal site is responsible by hydrogenation and dehydrogenation reactions while the acid function, determined by the chlorine content, is responsible by molecular arrangement reactions like paraffin cyclization, isomerization, and hydrocracking.

In some formulations, Tin (Sn) can be added to the catalyst, especially in most severe operating conditions. The tin promotes a better dispersion of platinum leading to a more selective catalyst to aromatics like xylenes, reducing the coke deposition and gases production.

As aforementioned, the main disadvantage of the semi regenerative catalytic reforming units is the relatively short operating cycles due to the catalyst deactivation. The main deactivation mechanisms to catalytic reforming catalysts are the poisoning due to the contaminants in the feed, pore plugging, chemical attack to the structure, sintering, and leaching due to the catalyst cracking leading to fines production.

Regarding contaminations, the sulfur and nitrogen are temporary poison to catalytic reforming catalysts and normally the content of these contaminants is controlled through the feed hydrotreating, it's important to quote that in some cases sulfur and nitrogen can be purposefully added to the feed to keep under control the acid function of the catalyst. As permanent poisons are quoted metals like Lead (Pb), Silicon (Si), Mercury (Hg), Copper, (Cu), Vanadium (V), etc. The poisoning involves the selective adsorption in the active sites in detriment of reactants. Metal contamination involves the chemical bond of the contaminant with platinum leading to metal alloy with activity loss, especially to dehydrogenation reactions.

Sintering is normally caused by high temperatures as well as excessive water concentration in the feed and is related to the agglomeration of metal particles reducing the active surface area.

The most common deactivation mechanism in catalytic reforming units is the coke deposition that leads to pore plugging with drastic reduction of the catalyst activity. In catalytic reforming units that rely on catalyst regeneration sections like the CCR Platforming™ by UOP Company and Aromizing™ by Axens Company, the catalyst is subjected to a sequence of process aiming to restore the catalyst activity. The first step is a controlled burning process to burn the coke deposited over the catalyst; in the sequence the catalyst crosses the oxychlorination section where is added chlorine to restore the acid function and normally is applied perchlorethylene as chlorine

source. Following the catalyst regeneration process, the catalyst is dried and cooled before to back to the process.

Due to their formulations, the catalytic reforming catalyst presents a high cost and adequate management actions are fundamental to maximizing the catalyst lifecycle.

Aromatics Separation Section – Ensuring Maximum Added Value to the Naphtha

As aforementioned, in markets where there is demand, the production of petrochemical intermediates is economically more advantageous than the production of transportation fuels, especially in countries with easy access to lighter oils. The production and separation of aromatics are processes with great capacity of adding value to crude oil.

The aromatics production complex is a set of processes intended to produce petrochemical intermediates from naphtha produced in the catalytic reforming process or by pyrolysis process. An aromatics production complex can take on different process configurations, according to the petrochemical market to be served, an example is shown in Figure 12.

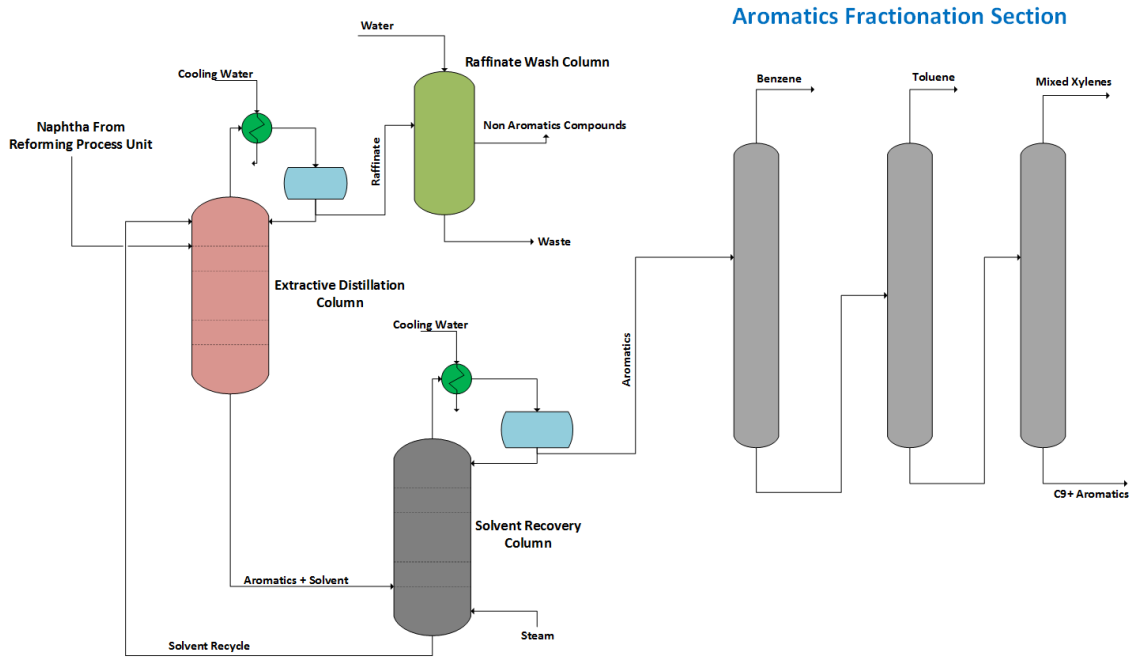

Figure 12 – Basic Process Configuration for a Typical Aromatics Separation Unit

The naphtha rich in aromatics, produced in catalytic reforming or pyrolysis units (in some cases from both), is fed to an extractive distillation column where the separation of aromatic compounds is conducted, which are withdrawn in the extract phase, are recovered at the bottom of the column while the non-aromatic compounds are withdrawn from the top in the raffinate phase. The aromatics are separated from the solvent in the solvent recovery column and directed to the fractionation section of aromatics where the essentially pure benzene and toluene streams and xylenes blend are obtained. The raffinate is sent to a wash column, and the non-aromatic hydrocarbons are usually sent to the refinery's gasoline pool.

The process shown in Figure 12 involves only physical separation steps, that is, the process yields in each stream depends on the concentration of this compound in the feed stream.

The growing demand for high-quality petrochemical intermediates and the higher added value of these products have made it necessary to develop conversion processes capable of converting lower interest aromatics (Toluene) into more economically attractive compounds (Xylenes).

Aromatics separation, mainly xylenes, is a great challenge to modern engineering. The similarities between the molecules make the separation through simple distillation extremely hard, for this reason, several researchers, and technology licensors dedicate their efforts to develop new processes which can lead to pure compounds with lower costs. A basic scheme for a xylene separation process is shown in Figure 13.

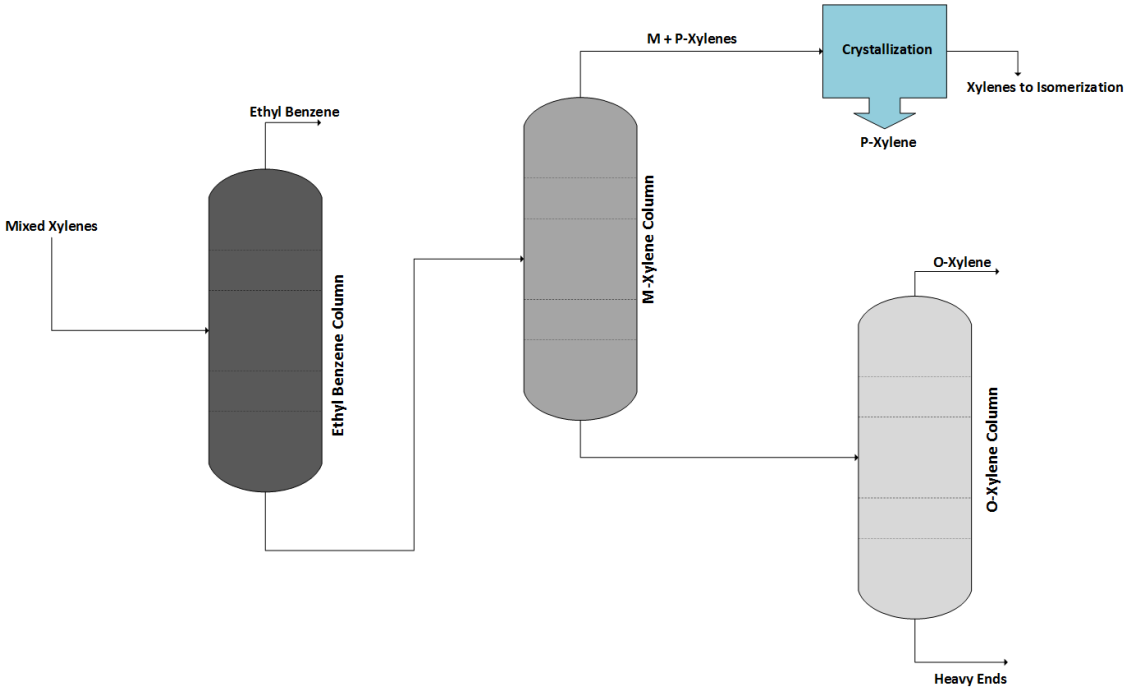

Figure 13 – Basic Process for Xylenes Separation

The xylenes blend is fed to a distillation column where the ethylbenzene is separated in the top and sent to styrene production market while the bottom stream is pumped to another column where the mixture of Meta and Para-xylenes is withdrawn in the top and the Ortho-xylene and heavier compounds are removed in the bottom.

Ortho-xylene is separated from heavy aromatics in another distillation column while the Meta and Para-xylene are fed to a crystallization process, where a stream is obtained with a high concentration in Meta-xylene and the residual stream is directed to an isomerization unit, aiming to promote the conversion of residual Meta and Orto-xylenes in Para-xylene. The aromatics production units are normally optimized to maximize the Para-xylene production because this is a petrochemical intermediate with higher interest, this compound is raw material to produce terephthalic acid that is used to produce PET (Polyethylene terephthalate). Figure 14 present the chemical arrangement of the xylenes isomers.

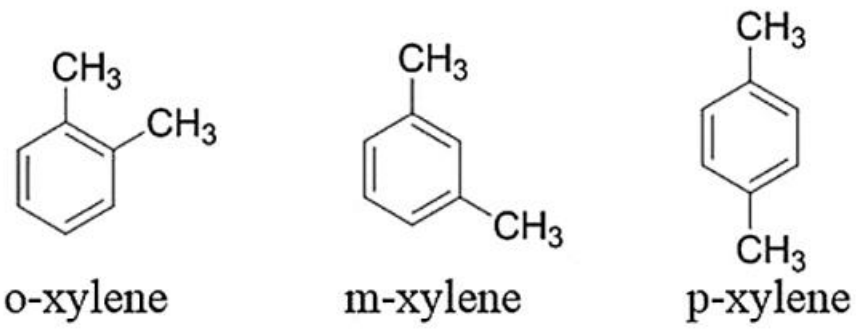

Figure 14 – Chemical Arrangement of the Xylene Isomers

To raise the production of higher commercial and economic interest compounds (P-Xylene and Benzene), technology licensors developed several processes to convert streams with low added value in these compounds. One of the main developers of this technology is the UOP Company, the PAREX™ process apply the separation through adsorption to obtain high purity P-xylene from xylenes blend.

Another UOP technology is the ISOMAR™ process, which promotes the xylenes isomerization to Para-xylene raising the recovery of this compound in the aromatic complex. TATORAY ™ process was developed to convert toluene and heavy aromatics (C9+) in benzene and xylenes through transalkylation reaction. Another economically attractive technology is the SULPHOLANE ™ process that applies liquid-liquid extraction operations and extractive distillation to reach high purity aromatics separation from hydrocarbon mixture.

The UOP Company developed an integrated aromatics complex aiming to maximize the production of benzene and P-xylene, which lead to a higher profitability to the refiner. A UOP Aromatics Complex scheme is presented in figure 15.

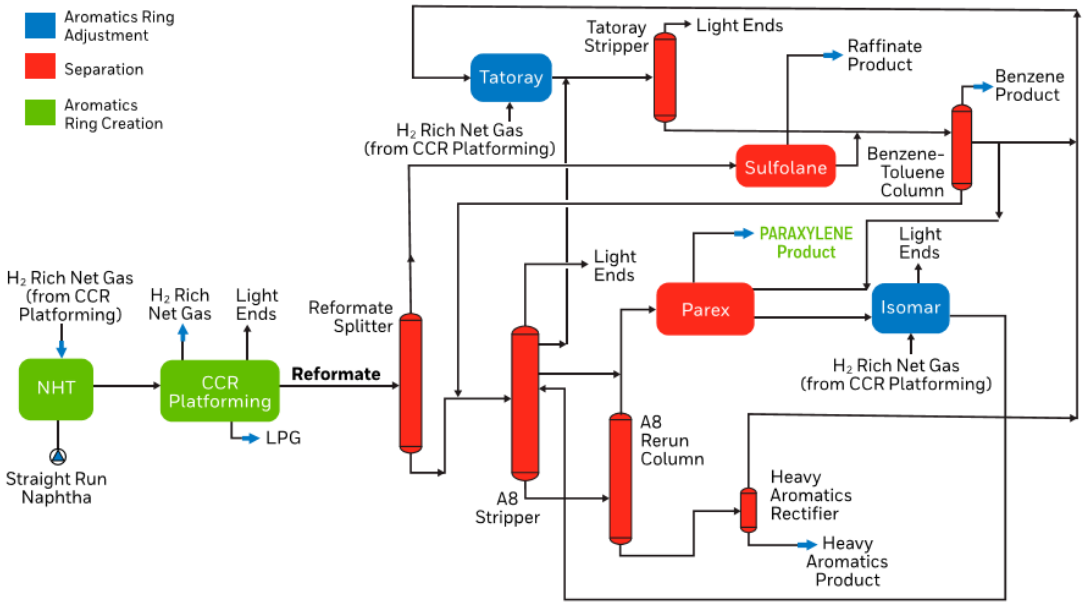

Figure 15 – Aromatics Complex by UOP Company

Other companies have attractive and efficient technologies to produce high purity aromatics, the Axens Company license an aromatics production complex also based on separation and conversion processes, called ParamaX™ that can be optimized to produce P-xylene. This process is presented in Figure 16.

The ParamaX™ technology offers the possibility of Cyclohexane production (Raw material to synthetic fibers) through benzene hydrogenation beyond raise the production of this component through toluene HydroDealkylation (HDA).

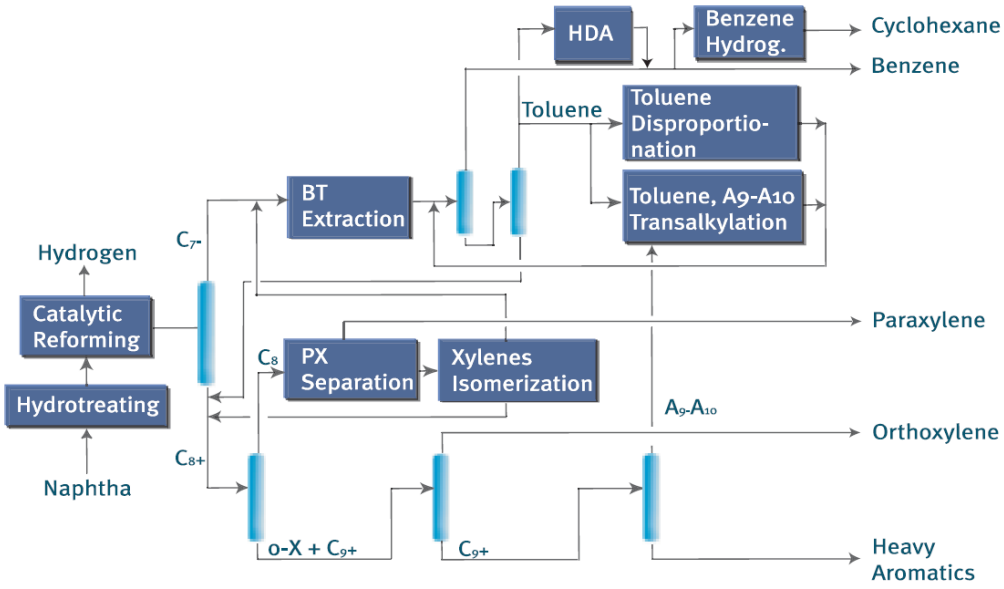

Figure 16 – Schematic Process Flow Diagram for ParamaX™ technology, by Axens Company.

As aforementioned, the capital investment to installation of aromatics production complexes is high, however, the obtained products have high added value and rely on great demand, and even the compounds with low interest can be commercialized with high margin. In countries with easy access to light oil reserves as Saudi Arabia and United States (Tight Oil) the installation of these process plants is even more economically attractive. As presented in Figure 15, the main reactions carried out in the aromatics production process aiming to improve the yield of benzene and xylenes are the toluene transalkylation presented in Figure 16 and the toluene disproportionation, presented in Figure 17.

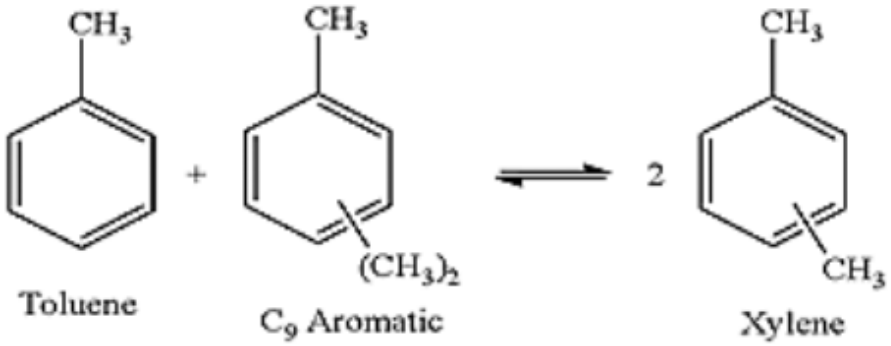

Figure 17 – Toluene Transalkylation Reaction

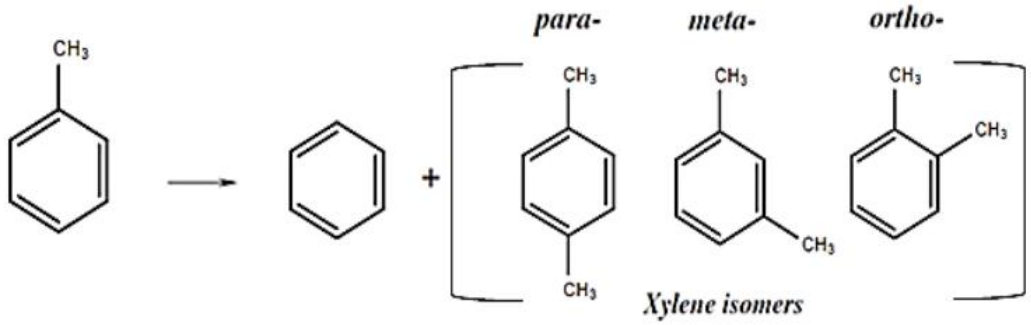

Figure 18 – Toluene Disproportionation

It's important to quote that all technologies have molecular management processes in order to improve the yield of p-Xylene, the most added value aromatic. Recent forecasts indicate great potential growth to the BTX market in the next years, as presented in Figure 19.

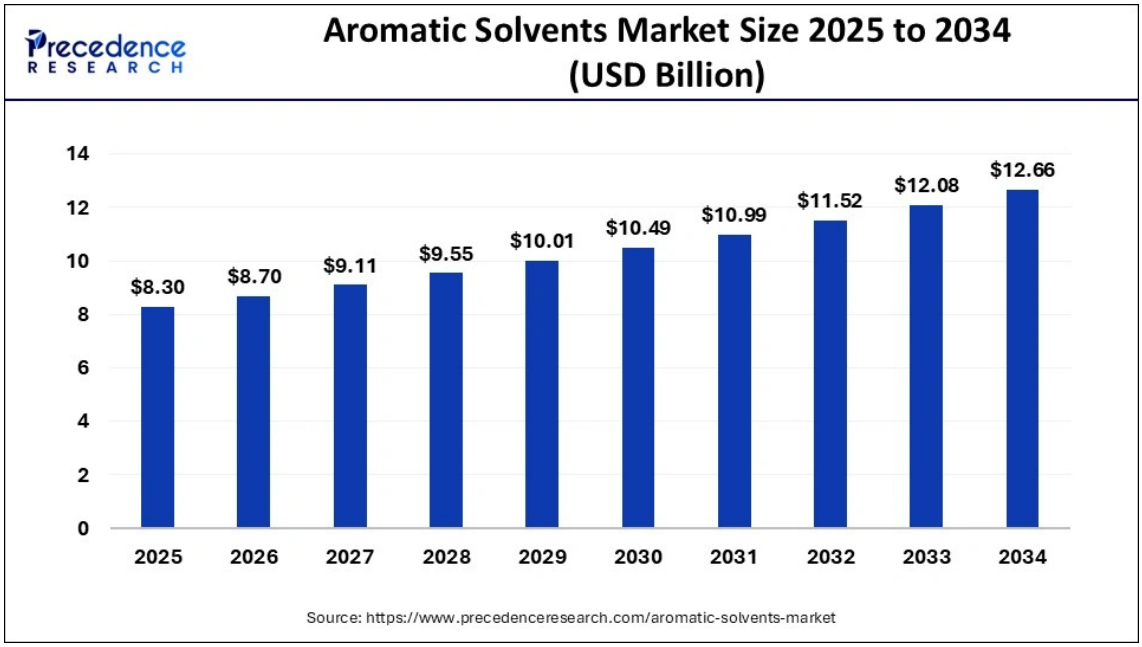

Figure 19 – Evolution of Aromatics Market Size 2025 to 2034 (Precedence Research, 2025)

Considering the data from Figure 19 the market size of the BTX market can reach a total value of USD 12,66 billion in 2034 under a compound annual growth rate (CAGR) of 4,80 % between 2025 to 2034. This data reinforces the relevance of the BTX market, especially considering the hostile scenario imposed on fossil fuels like gasoline which is the most conventional destiny of naphtha in non-integrated refineries.

The Synergy between Aromatics Production Complex and Steam Cracking Units

As presented in Figure 1, light aromatics and olefins presents growing demand and high added value when compared with gasoline, in this sense, maximize the yield of these petrochemical intermediates in the refining hardware can ensure high economic result to refiners, despite the high capital spending and operation costs related to a more complex refining hardware.

Among the synergy possibilities between steam cracking and aromatics production complexes is the use of pyrolysis gasoline produced in the steam cracking units as feed stream to aromatics production complex, improving the refinery capacity to produce aromatics against gasoline. By his turn, the raffinate stream from aromatics complex can be used to improve the olefins yield in steam cracking units, mainly ethylene and propylene.

An example of refining configuration relying on the synergy between aromatics production complex and steam cracking units is presented in Figure 20.

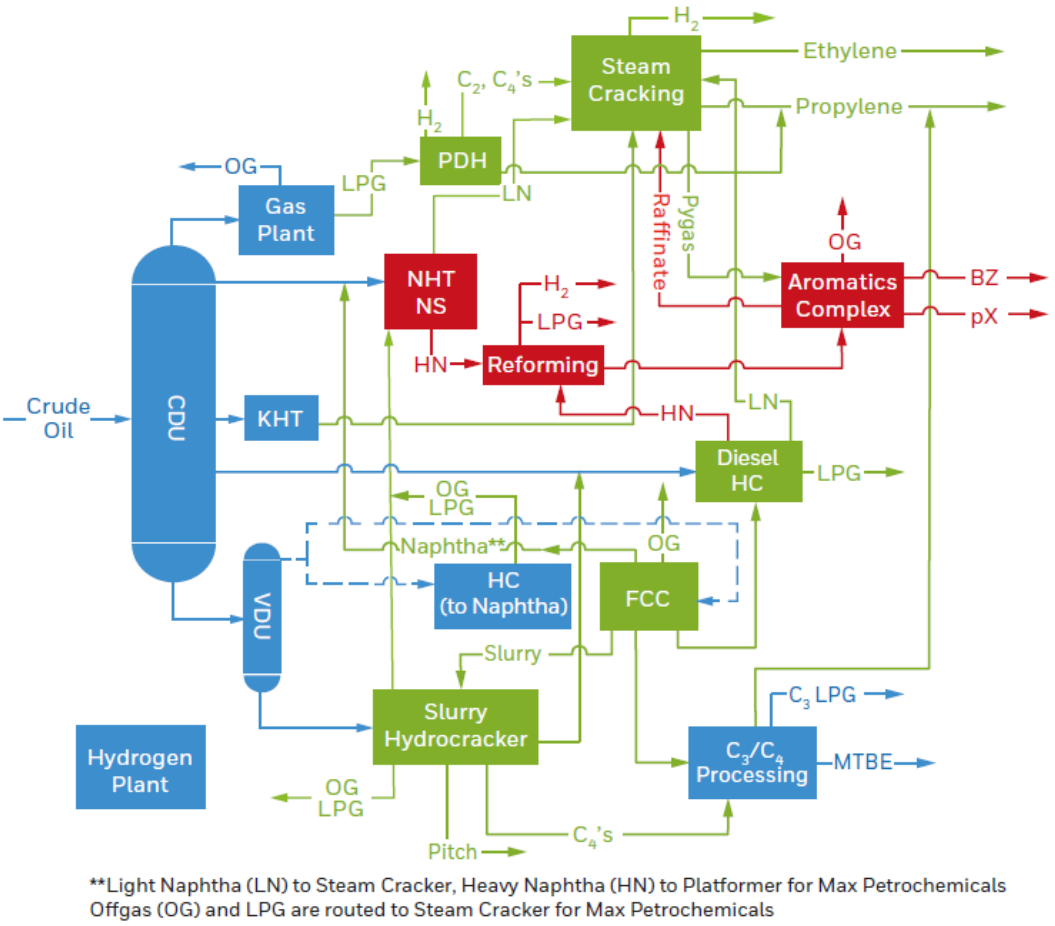

Figure 20 – Integrated Refining Scheme Base on Aromatics Complex and Steam Cracking Units (UOP, 2019)

Considering the recent trend of reduction in transportation fuels demand followed by the growth of petrochemicals market makes the synergy between aromatics production complex and steam cracking units an attractive way to maximize the petrochemicals production in the refining hardware and achieve closer integration between refining and petrochemical assets, a growing trend in the downstream industry.

As aforementioned, face the current trend of reduction in transportation fuels demand at the global level, the capacity of maximum adding value to crude oil can be a competitive differential to refiners. Due to the high capital investment needed for the implementation that allows the conventional refinery to achieve the maximization of chemicals, capital efficiency becomes also an extremely important factor in the current competitive scenario as well as the operational

flexibility related to the processed crude oil slate. Considering just the petrochemical complexes focused on PX (Para Xylene), we have total capital investments around 87 US billion dollars presented in Figure 21.

Project	Refinery Capacity (MMt)	P-Xylene Capacity (MMt)	Ethylene Capacity (MMt)	Propylene Capacity (MMt)	Est. Chemical conversion/ bbl. of oil (%)	Investment ($bn)	Full line Operation
Hengli Petrochemical	20	4.3	1.5	1.0	42	11.4 (Excl. SC)	May 17, 2019
Zhejiang Petroleum and Chemical (ZPC) Phase 1	20	4.0	1.4	0.65	45	12	Dec 31. 2020
Hengyi (Brunei) PMB Refinery-Petrochem Phase 1	8	1.5	0.5	0.2	>40	3.45	Nov 3, 2019
Zhejiang Petroleum and Chemical (ZPC) Phase 2	20	4.8[a]	1.5	0.7	50*	12	Jan 12, 2022
Shenghong refinery and Integrated Petrochem	16	4.0[c]	1.4[c]	0.5	60[b]	9.6[c]	2022
Hengyi (Brunei) PMB Refinery-Petrochem Phase 2	14	2.0	1.5	0.7	>40	10	2022
Tangshan Xuyang (Risun)[d]	15	3.5	1.5	0.6	>50	8.5	**On Hold**
Shandong Yulong (Phase 1) [e]	20	4.0	3.0	1.2	> 50	20 (1st phase)	2024 (1st phase)
Total	**133**	**28.1**	**12.3**	**5.6**	**--**	**87**	**--**

a. ZPC/UOP press release Jan. 17, 2019 announced that Phase 2 configuration and technology will be changed from Phase 1.
b. Based on information obtained by IHSM from a visit to Shenghong in November 2018
c. Reduced investment by 12.6% from the original announcement by reducing capacity or 10 process units and eliminating 8 product units. However, refinery capacity remained unchanged, and PX capacity in fact increased from original 2.8 to 4.0 MMt/y. Ethylene capacity will also increase from 1.2 to 1.4 MMt/y.
d. A new project which is under environmental impact Assessment.
e. A new project in three phases in Shandong Province. The first phase with investment of $20bn has been approved and under environmental evaluation. The projects are focusing on petrochemical production. With each barrel of fuel production, 1.25 barrel of Teapot refinery capacity will be closed to reduce the refinery over capacity.

Figure 21 – PX focused Crude to Chemicals Capital Investments (S&P Global Commodity Insights, 2024)

The Role of Catalytic Reforming Units in the Refineries Hydrogen Balance

Demand for hydrogen rose strongly in the last decades following the necessity of hydrotreatment units installations in refineries to comply with the pressure to reduce the content of contaminants like sulfur and nitrogen in the petroleum derivates and consequently minimizing the environmental impact caused by fuels burn. This scenario became the hydrogen one of the most important production inputs in modern refineries and adequate hydrogen management actions reach strategic character to keep under control the operating costs and refining margins, contributing to economic sustainability in the downstream industry.

The hydrogen matter is one of the most important questions to the future of downstream, the growing participation of renewable raw material in the refining hardware as a decarbonization strategy tends to raise even more the hydrogen consumption. The renewable streams have a great number of unsaturations and oxygen in their molecules which lead to high heat release rates and high hydrogen consumption, this fact leads to the necessity of higher capacity of heat removal from hydrotreating reactors aiming to avoid damage to the catalysts. The main chemical reactions associated with the renewable streams hydrotreating process can be represented as below:

$R\text{-}CH{=}CH_2 + H_2 \rightarrow R\text{-}CH_2\text{-}CH_3$ (Olefins Saturation)

$R\text{-}OH + H_2 \rightarrow R\text{-}H + H_2O$ (Hydrodeoxigenation)

Where R represents a hydrocarbon.

These characteristics lead to the necessity of higher hydrogen production capacity by the refiners as well as quenching systems of hydrotreating reactors more robust or, in some cases, the reduction of processing capacity to absorb the renewable streams. In this point it's important to consider a viability analysis related to the use of renewables in the crude oil refineries once the higher necessity of hydrogen generation implies in higher CO2 emissions through the natural gas reforming process that is the most applied process to produce hydrogen in commercial scale.

$CH_4 + H_2O = CO + 3H_2$ (Steam Reforming Reaction - Endothermic)

$CO + H_2O = \mathbf{CO_2} + H_2$ (Shift Reaction - Exothermic)

This fact leads some technology licensors to dedicating their efforts to looking for alternative routes for hydrogen production in large scale in a more sustainable manner. Some alternatives pointed out can offer promising advantages:

· Natural Gas Steam Reforming with Carbon Capture – The carbon capture technology and cost can be limiting factor among refiners;

· Natural Gas Steam Reforming applying biogas – The main difficult in this alternative is a reliable source of biogas as well as their cost.;

· Reverse water gas shift reaction ($CO_2 = H_2 + CO$) – One of the most attractive technologies, mainly to produce renewable syngas;

· Electrolysis – The technology is one of the more promising to the near future.

Refiners and technology developers are looking for alternatives to produce hydrogen on an industrial scale with lower CO2 emissions, and some attractive routes have been considered as competitive in the future.

Despite the advantages of the green production routes of hydrogen, they are still in development and poor attractive to the most part of the refiners, in the current scenario the refiners to look for more efficient operations aiming to optimize the hydrogen balance the refining hardware as well as apply CO2 capture technologies (the blue route), in this sense an attractive alternative is to apply technologies capable to recovery hydrogen from refinery off-gases and apply control strategies capable to minimize the hydrogen losses to flare system.

As exposed above the hydrogen generation is a key matter to refiners, and refineries that rely on Catalytic Reforming units apply the hydrogen produced in this processing unit to compose a relevant part of the hydrogen network becoming an important internal source of hydrogen. In some markets, where the demand by petrochemicals is lower, the main relevance of the catalytic reforming to the refining hardware is the hydrogen generation against the production of light aromatics. Figure 22 presents an example of hydrogen network in a crude oil refinery with high hydroprocessing capacity.

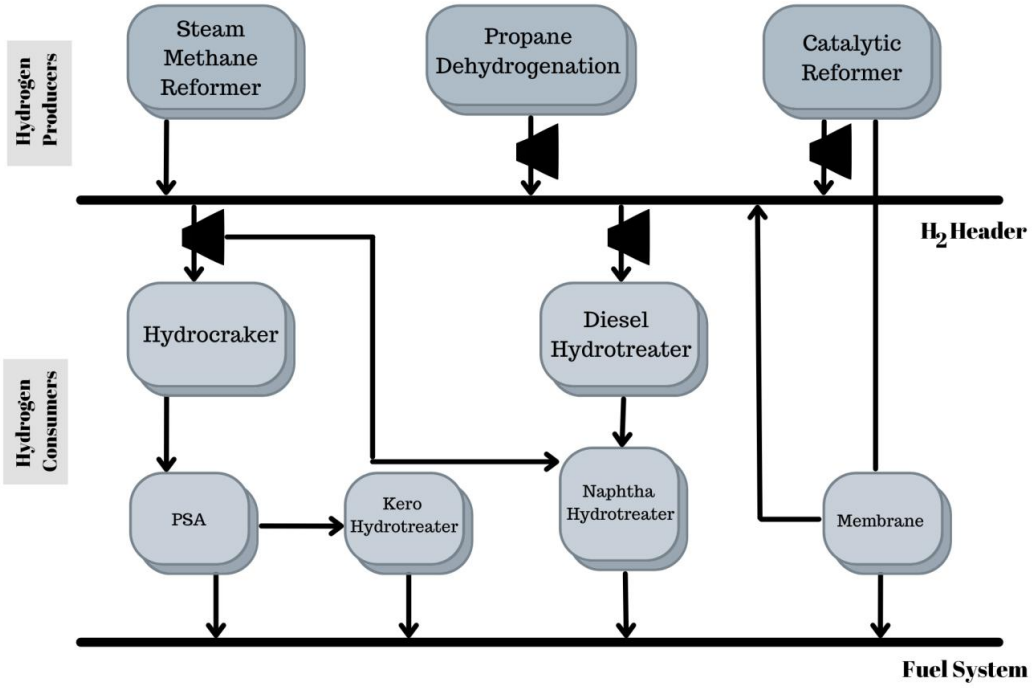

Figure 22 – Example of Hydrogen Network to a Crude Oil Refinery

In refineries with bottlenecked hydrogen generation units, the hydrogen from catalytic reforming units is fundamental to ensure compliance with the current quality and environmental regulations, becoming a fundamental enabler to profitable and reliable operations of the refining hardware. Nowadays, it's not uncommon to find refiners operating catalytic reforming units with the main objective to hydrogen generation, especially to refiners that operate with octane giveaway in the gasoline pool.

Closing the Sustainability Cycle – Plastics Recycling Technologies

As described above, we are facing a continuous growing of petrochemicals demand, and a great number of these crude oil derivatives have been applied to produce common use plastics. Despite the higher added value and significant economic advantages in comparison with transportation fuels, the main side effect of the growth of plastics consumption is the growth of plastic waste.

Despite the efforts related to the mechanic recycling of plastics, the increasing volumes of plastics waste demand most effective recycling routes to ensure the sustainability of the petrochemical industry through the regeneration of the raw material, in this sense, some technology developers have been dedicated investments and efforts to develop competitive and efficient chemical recycling technologies of plastics.

One of the most applied technologies for plastics recycling is in the catalytic pyrolysis where the long chain polymeric are converted into smaller hydrocarbon molecules which can be fed to steam cracking units to reach a real circular petrochemical industry. Another route is the thermal

pyrolysis of plastics, is this case, it's possible to quote the Rewind™ Mix technology developed by Axens Company.

Another promising chemical recycling route for plastics in the hydrocracking of plastics waste, in this case the chemical principle involves the cracking of carbon-carbon bonds of the polymer under high hydrogen pressure which lead to the production of stable low boiling point hydrocarbons. The hydrocracking route present some advantages in comparison with thermal or catalytic pyrolysis, once the amount of aromatics or unsaturated molecules is lower than the achieved in the pyrolysis processes, leading to a most stable feedstock to steam cracking or another downstream processes as well as is more selective, producing gasoline range hydrocarbons which can be easily applied in the highly integrated refining hardware.

The chemical recycling of plastics is a great opportunity to technology developers and scientists, especially related to the development of effective catalysts to promote depolymerization reactions which can ensure the recovery of high added value molecules like BTX. More than that, the chemical recycling of plastics is an urgent necessity to close the sustainability cycle of an essential industry to our society.

Conclusion

The search to add maximum value to processed crude oil is a constant among the refiners, especially considering the competitive scenario faced by the downstream market, in this sense, the flexible refining technologies like Catalytic Reforming and aromatics recovery section can offer a significant competitive advantage.

Installation of aromatics production units can significantly raise the profitability to refiners inserted in markets with high demand for petrochemical intermediates and surplus in gasoline, this fact is especially true in the current scenario where the transportation fuels consumption suffered drastic reduction due to the economic crisis caused by the COVID 19. The catalytic reforming technologies can develop a fundamental role in the downstream industry to allow profitable and reliable operations to refiners both to maximize petrochemicals and allow closer integration with petrochemical assets and ensure a positive contribution to the hydrogen balance, reducing the necessity to higher capacity of traditional steam methane reformers with consequent lower CO2 emissions. These advantages can be even more relevant in market with great gasoline surplus aiming to ensure higher added value to the processed crude.

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. The development of crude to chemicals technologies reinforces the necessity of closer integration of refining and petrochemical assets by the brownfield refineries aiming to face the new market that tends to be focused on petrochemicals against transportation fuels, it's important to note the competitive advantage of the refiners from Middle East that have easy access to light crude oils which can be easily applied in crude to chemicals refineries. As presented above, crude oil to chemicals refineries is based on

deep conversion processes that require high capital spending. This fact can put under pressure the refiners with restrict access of capital, again reinforcing the necessity to look for close integration with petrochemical sector aiming to achieve competitiveness.

Despite the benefits of petrochemical integration, it's fundamental taking in mind the necessity to reach a circular economy in the downstream industry, to achieve this goal, the chemical recycling of plastics is essential. As presented above, there are promising technologies which can ensure the closing of the sustainability cycle of the petrochemical industry.

References

Advances in Catalysis for Plastic Conversion to Hydrocarbons – The Catalyst Group (TCGR), 2021.

CHANG, R.J. – ***Crude Oil to Chemicals – Industry Developments and Strategic Implications*** – Presented at Global Refining & Petrochemicals Congress (Houston, USA), 2018.

CHARLESWORTH, R. ***Crude oil to chemicals: Developments and Trends –*** S&P Global Commodities Insights Company Presentation at CERAWEEK 2024, 2024.

Deloitte Company. ***The Future of Petrochemicals: Growth Surrounded by Uncertainties***, 2019.

DOMERGUE, B.; LE GOFF, P. Y.; ROSS, J. ***Octanizing Reformer Options.*** PTQ Magazine, 2006.

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

GELDER, A.; BAILEY, G. ***The Future of Petrochemicals: A Tale of Two Transitions,*** Wood Mackenzie, 2020.

International Energy Agency (IEA). ***Oil 2021: Analysis and Forecast to 2026***, 2021.

KIM, W.C.; MAUBORGE, R. - ***Blue Ocean Strategy***. Harvard Business Review, 2004.

LAFLEUR, A. ***Use and Optimization of Hydrogen at Oil Refineries***. Shell Company, Presented at DOE H2@Scale Workshop – University of Houston, 2017.

OYEKAN, S.O. ***Catalytic Naphtha Reforming Process***. 1st ed. CRC Press, 2019.

SILVA, M. W. – ***More Petrochemicals with Less Capital Spending***. PTQ Magazine, 2020.

VU, T.; RITCHIE, J. ***Naphtha Complex Optimization for Petrochemical Production***, UOP Company, 2019.

ZHOU, T.; BAARS, F. ***Catalytic Reforming Options and Practices***. PTQ Magazine, 2010.

International Energy Agency (IEA) - ***OIL 2025 : Analysis and Forecast to 2030.***

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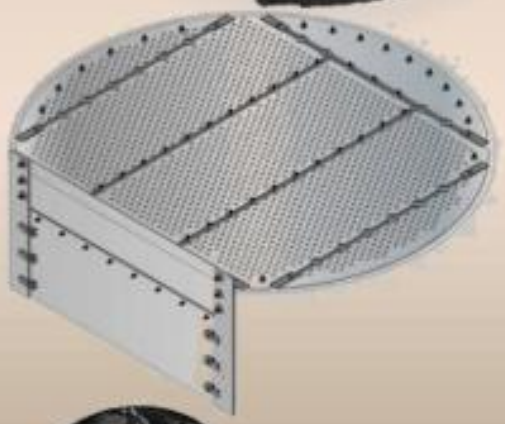

Looking at Current Society from a 70-year-Old Perspective

Karl Kolmetz

Last October I was 70 years old. Sometimes I feel every day, other times I still feel like the teenager riding my horse through the old neighborhood swamps. I struggle to understand the current society and culture. I know every generation thinks the next generation is bad, but our current statics are unbelievably bad, whether I like the nuances or not.

Loss of Morality

We have lost our way morally. We have substituted thousands of years of moral fundamentals for doing what you feel as an individual. In 1955, the year I was born, Walter Litman said, If what is right and wrong is what an individual feels, we are outside of the boundaries of civilization.

We have these thousands of years of moral fundamentals because they are what is best for society, and they are simple. Do not kill, do not steal, do not lie, keep your promises.

Today we kill others if we do not like what they say. I am still surprised by the depth of hate it takes to kill someone for their words. I did not agree with everything Charlie Kirk stated, but I do not believe he deserved to die for his statements. There are good reasons to kill some people, but not for their words. Most people's words will not be remembered very long, as Lincoln said at Gettysburg, yet we immaturely kill people for short term vocalizations.

Honesty is very expensive, do not expect it from cheap people. Unfortunately, most people will give away their honor for 30 pieces of silver, the price Judas received for Jesus. Given the choice to accept a bribe, or reject the bribe and do the right thing, a very high percentage of the people in the world take the money.

There is a direct correlation between corruption and the living standard of the population. If you visit a poor country there is a good chance corruption is high.

Do not steal, do not commit adultery. As a young man I thought adultery was a sexual sin, as I get older, I now believe it is a breaking of a commitment.

First World? There is a divorce rate of over 50% in most of the 1st world nations – hard to call any country with a divorce rate of over 50% 1st in any measure. This is due to a lack of morals and forgiveness. The family is the building block of society. The family is the unit that ensures children are fed, loved, protected, nurtured and raised in the virtues they need to become the

responsible citizens of tomorrow, but family breakdown has become epidemic with nearly half of children experiencing the dissolution of their parents' relationship.

The collapse of marriage rates particularly among low-income groups has exacerbated poverty and disadvantage the impact of family breakdown on children is profound it is the single biggest predictor of poor teen mental health and correlated with worse outcomes in every aspect of adult life. The support of extended family has been weakened and loneliness increased as young people have moved away from their communities.

Christianity is based on a sacrifice. Humbling yourself, giving to others and to the community. We have the voice of Gensis – we are all made in the image of God – that is the foundation of democracy. The Voice of Exodus – that we wonder thought the wilderness. The Voice of Jesus – blessed are the meek, blessed are the poor in spirit.

George Marsden a historian, said what gave Martin Luther King rhetoric its power was the sense that there is a moral order built into the universe. If slavery is not wrong, then nothing is wrong and if segregation is not wrong, then nothing is wrong.

Mental Health

Without a strong moral order it is hard to have trust, it is hard to find your meaning in life. This leads to a sadder society. Higher rates of mental health issues and higher rates of suicides. Fifteen percent of people have an addictive personality with drugs and alcohol dependency.

It is higher than 15% in my family. Every time I fly into Singapore the airline is required to announce that Singapore has a policy of death to drug traffickers. I do not really like that statement, but the thing I like less than that statement is that 107,081 people died in USA of drug overdose in 2022.

Forty-five percent of high school students say they are hopeless and despondent. The number of people who say they do not have close friends is increasing. Since 2000, people that say they are in the lowest happiness category have increased by fifty percent.

Ten percent of the people have mental issues, if you travel to any major city in the USA, you will think it is also higher than 10% - so 25% are not stable if you add in the addictive personalities.

In the first world countries - one in seven adults (14%) now takes anti-depressants and suicide is the most common cause of death for young men. Our families are in crisis and the social fabric of our neighborhoods is also unraveling. Shrinking membership organizations and religious attendants have eroded a sense of common purpose.

Loss of Opportunity

In the USA and other countries, we have created a chaste system. In the USA 54% of the leading positions in commerce went to the same 34 elite universities graduates. My sister worked in the

stockbroker industry, and if you did not have an elite degree, you did not get hired to senior positions.

People with high school degrees die nine years younger than people with college degrees. People with high school degrees are five times more likely to have children out of wedlock. People with high school degrees are twenty-four times more likely to say they have no friends.

How People and Nations Renewed

So how are people and nations repaired and renewed? Normally we get to a point where we are broken. Some nations and people go to the bottom and never recover. Many broken people commit suicide but going to the bottom is the only way to get better for some people. I have never understood why some people need to go all the way to the bottom to see the obvious, but many do not see they are on the wrong path till they hit bottom.

Olives can only make olive oil when they are crushed. We start to recover when we are broken open. People and nations must hit a spiritual and cultural crisis and then be revived. We do not grow through good times – we grow through a process of rupture and repair

1770s – America Revolution - Colonial Order was broken

1830 – US East Coast Elites dominated – Andrew Jackson brought in an area of populism

1860 – Slavery order was broken – Abraham Lincoln brought in a new order

1890 – Industrializing and Robber Barrons – Teddy Roosevelt brought a new order

1960 – Flower Power – brought a new order – 4000 bombings on American College Campuses – maybe not real love

The temptation of those who do not read history is to think this time is different.

The way culture is renewed is when a small group of people find a better way to live. That is the story of the early Christian Church. Good culture is brought about by good people, not by laws. I am always surprised at laws that require you to be a certain religion, which is like a law requiring you to love your wife, they will never truly work.

The way societies improve is that you have a cultural shift, a civil shift and then a political shift. The shift in ideas have to proceed before the political shifts. Milton Freeman wrote about conservative valves in 1945, but Ronald Regan was not elected until 1980.

You could be that Person

“There may not be as much humanity in the world as one would like to see. But there is some. There's more than one would think. … Walk down the street of any city, any afternoon, and look around you. What you've got to remember is what you're looking at is also you. Everyone you're

looking at is also you. You could be that person. You could be that person, you could be that cop. And you have to decide in yourself not to be." James Baldwin

You will choose the person you will become. You can choose a path to make a better society, or you can be the person that makes society worst.

How can We Improve?

I have written several times, we must have a war on drugs and drug traffickers – 25% of the population has mental instabilities (not my statistics, go search) – this is like leaving the sheep out for the wolves, we must have some shepherds. We need to go after the wolves hard and there may only be 35% capable shepherds. In Malayia and Singapore, they put high taxes on alcohol, I would be in favor of this given the 25% instabilities.

We must get some values back. We must understand that taking bribes and being dishonest actually hurts ourselves. This is also an IQ thing; many people are not smart enough to see the correlation. They view it as my turn at the hog trough, but remember hogs live in very poor conditions. It is very easy to review the average salary in a country and compare that to the level of corruption – almost the same. If we want to raise the standards of living in the world we need to go after corruption. We must train our voters to vote wisely – which is subjective – but must be better than we are doing today.

Empathy – 65% of the population have challenges. As a young man I was very hard on others because I tried to live a very good life. I am fortunate to have very high IQ and I saw what drugs and alcohol did to my family members. I may have a small additive personality, but I was smart enough to be scared of the temptation.

As a young teenager I saw a man who I respected make a moral error, I asked my dad how people could fail so easily, he replied, you have the same blood in your veins. So how can we help the majority of our population – it is an up hill task. We need to move slowly and improve in these areas.

Freedom, prosperity, and happiness are not values, they're not a map, they're not even principles. They may be the fruits of a successful society, but they're not its roots. No good tree bears bad fruit and to restore the fruit we must first attend to the roots. The true roots, the foundation stones of a civilization are not freedom, prosperity, and happiness, but the pursuit of the good deeds, hard work, and truth in action.

Reference

1. David Brooks Video at ARC

2. K Kolmetz, Generations, Engineering Practice Magazine, Volume 10 Number 42, January 2024

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How to... DIRECT CONTACT TRAYS

How to design all kinds of Direct Contact Trays

Dr.-Ing. Volker Engel

Direct Contact Trays (short DCT; as umbrella term for shed decks, shower decks, splash decks, cascade trays, disc-donut trays, baffle trays and iron angles) are open-path designs and used in quenching and degassing applications. Even if their area of application is limited and their function/design is comparatively simple, there is a variety of designs and some hydraulic aspects must be taken into account when designing them.

There are several types of trays used to facilitate direct contact between gas and liquid for heat exchange (quenching). The same types of trays are also used for degassing liquids (e.g., after pressure reduction). The link between these very different applications lies in the phase flow: The gas should not penetrate the liquid in the form of a two-phase layer, but rather the liquid should form a large surface area. Thus, these trays are similar in their application to grid packings.

All these tray types are based on an open path for gas: There are no downcomer plates (as for classical trays). Therefore, gas can take the easy way through the column. Liquid is going almost the same way through the column.

Figure 1 shows an overview of the various types of designs and their namings. The different types are described in the next chapters.

1. Baffle Trays

A Baffle Tray (sometimes called Splash Tray) is the easiest variant of those Direct Contact Trays. The trays consist of horizontal decks onto which the liquid splashes. As at classical trays, there are single-pass as well as multi-pass designs.

As long as there is no overlap of the decks, each deck covers half of the cross-sectional area of the column. Therefore, the type is also called 50%-cut. To ensure a good re-mixing on the next tray, there may be an overlap of the tray decks. But there are also layouts with a gap: Standing on the top tray you are able to look through all trays because of the central gap. Those designs are used at high loadings (gas and/or liquid). The liquid is carried along by the rising gas and thus does not follow gravity alone.

One feature is the easy construction. It is easy to have a strong and robust design – even for heavy duty service. The main disadvantage is the poor efficiency of the tray. There is only a short contact of gas and liquid, since the liquid has a small surface area by splashing.

Figure 2 shows a single pass of a Baffle Tray as well as a two-pass design. Such single-pass designs are also named 50%-cut (when covering half of the cross sectional area) or side-to-side design.

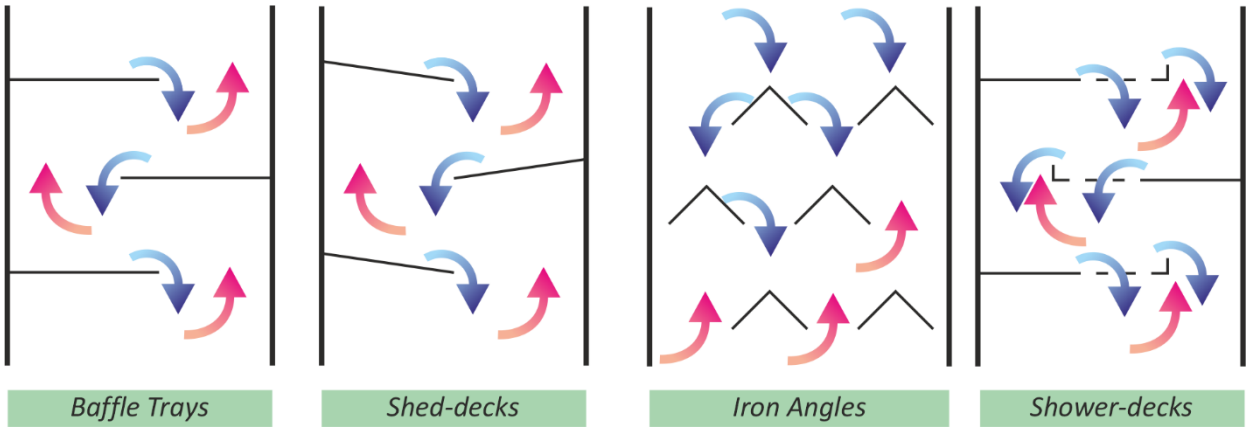

Figure 1: Types of Direct Contact Trays

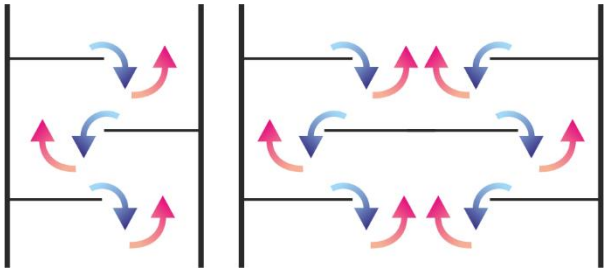

Figure 2: Baffle Tray Design: Single-Pass (Left) and Two-Pass Design (Right)

2. Shed-Decks

Whenever the decks of a Baffle Tray are sloped, the naming of the trays become Shed-deck (see Figure 3). Such an inclination of the panels is done to ensure that all solids will pass the column and do not agglomerate on the decks.

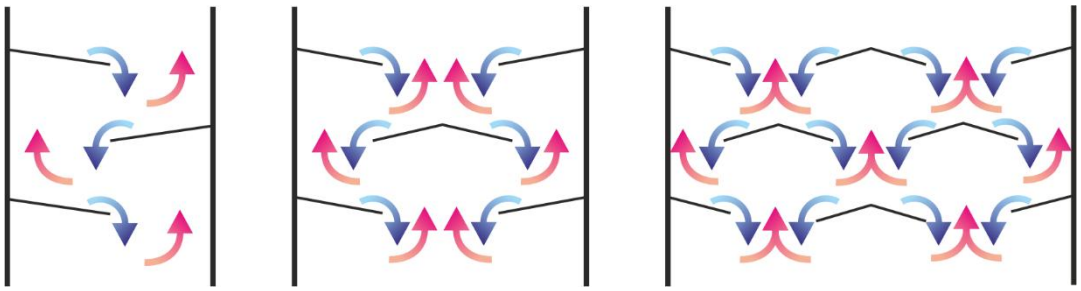

Figure 3: Shed-deck design: Various number of sheds per stage

As for the Baffle Tray type there might be single-pass designs as well as multiple sheds per stage. Shed-decks are also used for large tower dia-meters, since the shape of the sheds has good structural properties and it is quite easy to add major beams.

Each of the top sheds has to be fed the right quantity of liquid. For this task liquid distributors (perforated pipes as well as trough distributors) are used. Since there is only little transversal liquid transport on the sheds (no weirs), any liquid maldistribution will stay.

For these Shed-decks, the inclination of the panels is about 5°. For heavy duty designs, it may be more. Number, dimension, inclination angle, material thickness – all these parameters can be varied by the designer. This point should be emphasized here, as there is a special type of design in which these degrees of freedom do not exist due to the use of standard components. These are discussed in the next chapter.

3. Iron Angle Tray or Cascade Tray

Instead of fabricating the shed panels from sheet metal material, profile elements (L-bars) are used. Because these standard elements are 90°-bended profiles, the inclination of the sheds is 45°. Figure 4 shows such an arrangement of the angles. The direction of the angles is the same at each tray or it is rotated by 90° at each tray.

Those "Iron Angle" (also called „Cascade Tray") designs are closely related to grid packings in design and function: Both variants ensure that the falling liquid is deflected and mixed by small sheet metal surfaces. That is why both designs are used in quenching, and the liquid distribu-tion for both designs is similar: Liquid is initially applied via spray nozzle distributors or trough distributors.

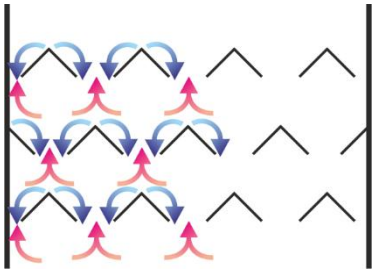

Figure 4: Iron Angles (Cascade Tray)

4. Shower Decks

Compared to the Baffle Trays, there are two additional features at Shower-decks: The panels are partly perforated and there are weirs. These features have a very positive impact on the function of the tray:

The perforation on part of the panel area generates droplets falling to the next tray. By this, there is not only the falling liquid from the edge of the tray but also liquid droplets with a much higher surface area. To ensure that liquid will pass all openings, there is a weir to generate a certain backup on the tray. Figure 5 shows the weirs and the liquid through the perforations. For a stable operation (constant level) the weir should be used by the liquid, too!

In most cases, this overlapping area is also perforated. This perforation (open area) and the weir should be designed so that at design load about 40% of liquid runs through the per-foration and 60% across the weir. In addition the weirs should be notched for transversal equali-zation of liquid.

Those Shower-deck configurations are often used in quench applications.

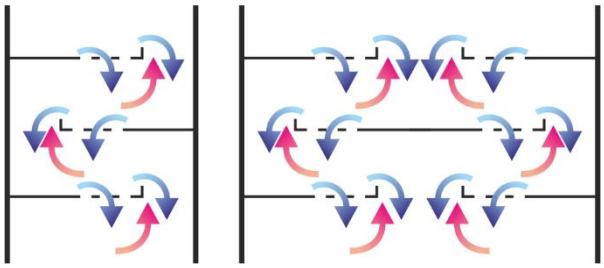

Figure 5: Shower-Deck Designs

When they are used in degassing applications (e.g. for feed water of steam boilers), the open area for gas has not to be that large. In this case, all deck area is perforated and gas will escape and bypass the trays (see Figure 6).

At this design all liquid is supposed to drain by the perforation and there should be no liquid over the weir (This liquid would reach the column sump without further treatment).

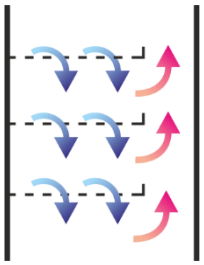

Figure 6: Shower-Decks for Degassing (Single-Pass Design)

Instead of using chordal openings for the gas flow, there can be one or multiple risers to let the gas escape (see Figure 7). Those designs are very likely to deck distributors of packings. But unlike deck distributors, shower-decks are stacked one on top of the other.

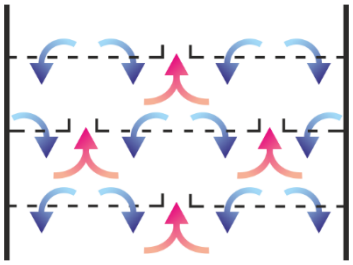

Figure 7: Shower-Decks for Degassing (Multi-Pass / Multi-Riser Design)

5. Disc-Donut Trays (DD, D+D , D&D)

This variant is not actually a separate function type; it is merely a special geometric variant. However, since it is used in many applications, it is described separately here.

The variants discussed so far are based on chordal designs or profile materials. The Disc-Donut Trays consist of radial components: alternating donuts and discs. The donut leads the liquid to the center opening, the liquid falls to the central disc and is led back to the circumferential opening (Figure 8).

This geometry also allows for the incorporation of all design aspects of the configurations described so far. Therefore, a Disc-Donut tray can be constructed either as a simple Baffle Tray or as a more complex Shower-deck.

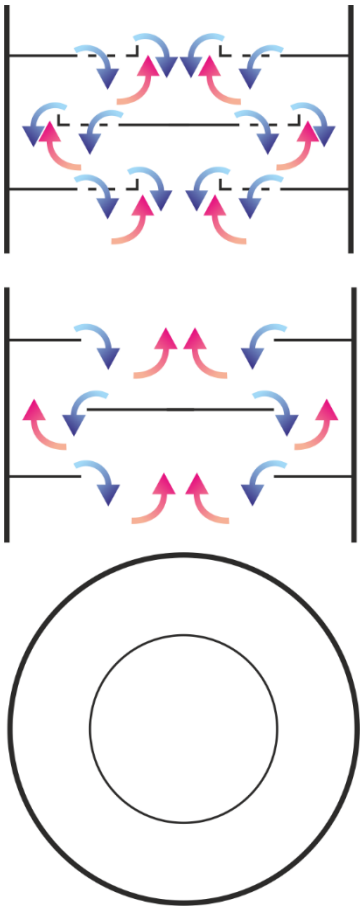

Figure 8: Disc-Donut Tray Design

Hydraulic limits

For all trays and internals that are operating at counter current there are limiting effects. Figure 9 shows the operation diagram for Direct Contact Trays. Compared to classical trays, there are few limits.

1 System Flood FFSF

When the vapor velocity exceeds the settling velocity of liquid droplets („Stokes Law Criterion"), vapor lifts and takes much of the liquid with it. If these conditions apply to the cross-sectional column area (lowest gas velocity), then operation of the column is hardly possible. This System Flood condition is considered as a hard limit (flooding factor).

5 Capacity

The gas velocity changes as it passes through the column: Figure 10 shows the three basic cross-sectional areas of a simple baffle-tray column.

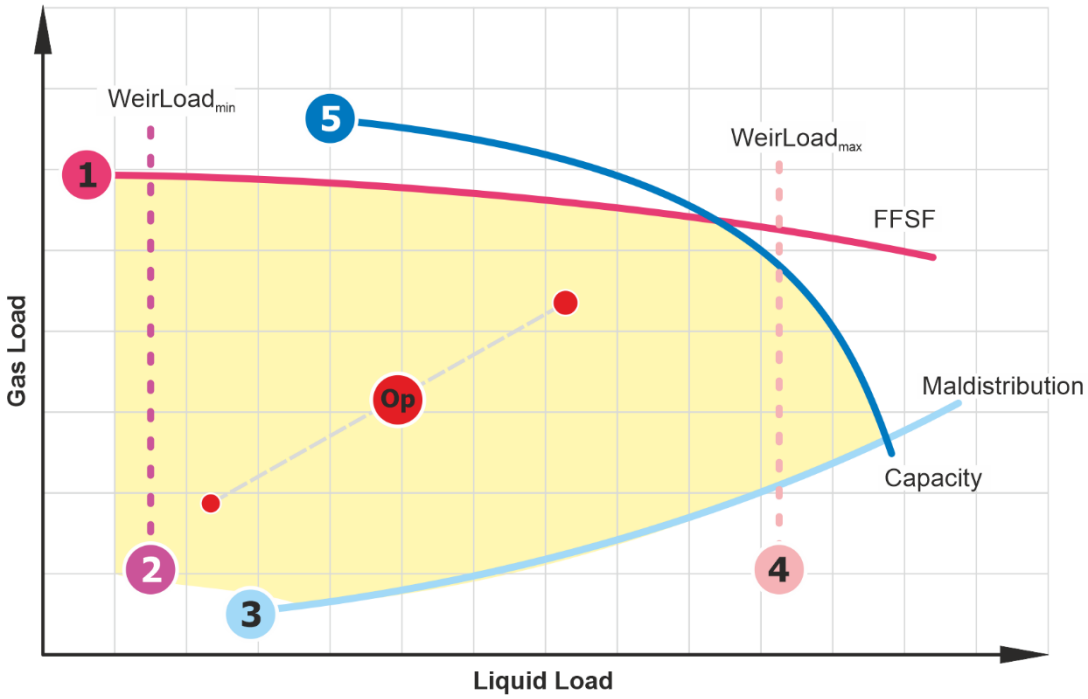

Figure 9: Qualitative Operation Diagram for Direct Contact Trays

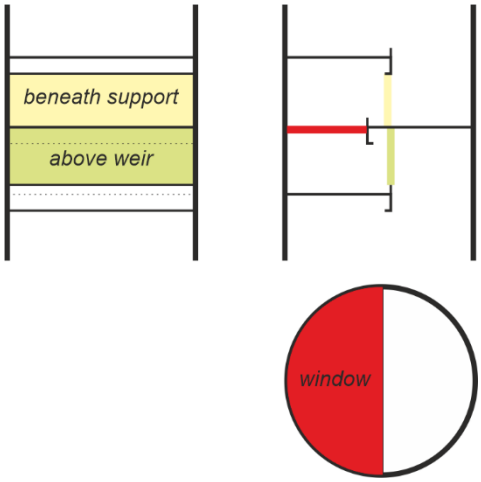

Figure 10: Areas of Phase Contact

In each of these cross-sections, the gas is in contact with liquid. The entrainment criteria for system flooding can also be calculated for the conditions prevailing here. However, values exceeding 100% of this "local system flooding" do not necessarily indicate a flooding limit in this context, but rather signify liquid agglomeration and liquid slugging – and in result a loss in efficiency. All these effects are summarized as Capacity in curve 5.

The curves 2 and 4 are recommendations for the min and max load of weirs. Both lines are not an operational limit and – of course – only present, if a weir is implemented (Shower-decks). Line 2 indicates a potential influence of out-of-level tolerances at very low weir loadings. Line 4 shows up at high load and suggests to enlarge the weir lengths (more flow paths).

3 Maldistribution

Gas as well as liquid maldistribution for quench applications is an effect limiting the efficiency. At degassing applications this effect is not a limit since there is no need for phase contact.

Since the pressure drop across all Direct Contact Tray types is quite low, there is no balancing effect on the gas flow. This means that any gas maldistribution at the inlet is not evened out by the trays. Therefore, it is very important to ensure good gas distribution (e.g. using a vane inlet device).

The aspect of liquid maldistribution is relevant for all kinds of Direct Contact Trays except Shower-decks. Since there is no lateral distribution of the liquid on the trays, the liquid distribution remains virtually the same as it was when set on the top tray. At Shower-decks there is a weir, that enables a liquid holdup and equalization on each tray. Therefore – even if the tray design is simple – it is important to ensure proper liquid distribution on the top tray.

Summary

Figure 11 shows an overview of the features of the variants of Direct Contact Trays. Depending on the application, different types are used. Even though their basic design is simple, it is still important to use the specific features of the different variants correctly.

	Baffle Trays	Shed-decks	Iron Angles	Shower-decks
Handling of solid particles / fouling	o	++	++	-
Suitable for large tower diameters	o	+	++	+
Compensation of liquid maldistribution	o	-	-	++
Compensation of gas maldistribution	o	-	-	o
High interfacial area	o	o	o	++
Free open area (Suitable for high gas load)	o	+	++	o

Figure 11: Specific Features of Direct Contact Trays

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

Fair, J. R.: How to design baffle tray columns, Hydrocarbon Processing (1993)

Górak A., Schoenmakers, H.: Distillation: Equipment and Processes. Academic Press (2014)

Kister, H. Z.: Distillation Design, McGraw Hill (1992)

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

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

WelChem Process Technology: TrayHeart Software. Tower Internals Calculation Software.

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