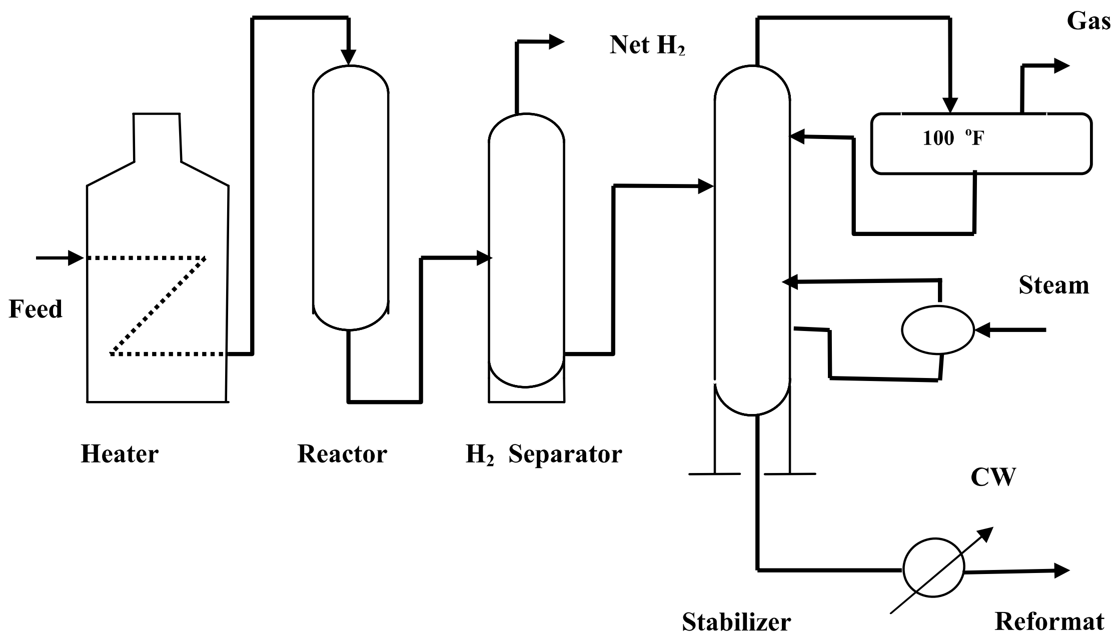
Catalytic Reforming

It is mainly used for the production of high- octane number hydrocarbons. In catalytic reforming the change in the boiling point of the feed stock passed through the unit is relatively small as the HC molecules are not cracked but their structures are rearranged to form higher octane aromatics. Thus catalytic reforming primarily increases the octane number of motor gasoline rather than increasing its yield. Typical feed stock is HSR gasoline and naphtha (180 to 375 ^{o}F).

<u>(PONA)</u> {Paraffin, Olefin, N**aphthene, Aromatic}**
1) P → isomerizes to some extent converted to **N**, and **N** subsequently converted to **aromatics**.
2) O → saturated to form **P** which then react as in **(1) {hydro-cracking}**.
3) N → converted to **aromatics. {dehydrogenation}.**
4) A→ unchanged.
<u>Reforming are classified as :</u>
1) <u>Continuous :</u> Catalyst can be regenerated continuously and maintained at a high activity . ***(Higher capital cost)***.
2) <u>Semi- regeneration:</u> Regeneration is required at intervals 3 to 24 months depending on the severity of operation. ***(Low capital cost)***.
High H_2 recycle rates and operating pressure is utilized to minimize coke.
3) <u>Cyclic:</u> compromise between the two extremes having a swing reactor for regeneration.



Catalytic Reforming, semi- regeneration

<u>Reforming Catalyst :</u> All of the reforming catalyst contains ***platinum supported on a silica alumina base***. In most cases ***rhenium is combined with platinum*** to form a more stable catalyst which permits operation at lower pressure.
Platinum serve as a catalytic site for hydrogenation and dehydrogenation reactions. Chlorinated alumina provides an acids site for isomerization and hydro- cracking reactions and cyclization.

Selectivity: The difference between the RON and MON of a given gasoline. Alkylate is an excellent low sensitivity and reformat a high sensitivity gasoline component.
Severity: The degree of intensity of the operating conditions of a process unit.

$$\text{Severity} = \frac{(\text{Cat. / oil})\ \text{ratio (lb/lb)}}{\text{Space velocity}}$$

$$\text{Space velocity} = \frac{\text{lb oil /hr}}{\text{lb cat.}}$$

Space velocity: The volume or weight of gas and / or liquid passing through a given catalyst or reactor space per unit time, divided by the volume or (weight) of catalyst through which the fluid passes. High space velocity corresponds to short reaction time.
WHSV: Weight hour space velocity = weight of feed per hour per weight of catalyst.
LHSV: Liquid hour space velocity = volume of feed per hour per volume of catalyst.
Example (2):
Calculate the length of time between regeneration of catalyst in a reformer operation at the following conditions:
Liquid hourly space velocity (LHSV) = 3.0 v/hr/v
Feed rate = 5000 BPSD
Feed gravity = 55 API
Catalyst bulk density = 50 lb / ft^3
H_2 to feed ratio = 8000 scf /bbl (standard cubic feet, 14.7 psia, 60 ^oF)
No. of reactors = 3
Catalyst deactivates after processing 90 bbl of feed per pound of catalyst. If the catalyst bed is 6 ft deep in each reactor. What are the reactor inside diameters? Assume an equal volume of catalyst in each reactor.
Solution :
5000 BPSD = 5000 (bbl/day) * (day/ 24 hr) * (42 gal/bbl) * (0.13 ft^3/ 1 gal)
= 1170 ft^3/ hr

$$\text{LHSV} = \frac{\text{Oil volume /hr}}{\text{Catalyst volume}}$$

Total catalyst = 1170 / 3 =390 ft^3
Weight of catalyst = volume * density of catalyst
= 390 * 50 = 19500 lb catalyst

$$\text{Time between regeneration} = \frac{19500\ \text{lb}_{\text{catalyst}} * 90\ \text{bbl/ lb}_{\text{catalyst}}}{5000\ \text{bbl/day}} = 351\ \text{days}$$

Volume of catalyst per reactor = 390 / 3 =130 = 130 ft^3
Inside area = 130 ft^3 / 6 ft
Inside diameter = 5.25 ft.

Example (3): On processing 1200 ton / day of 27 API catalyst crackers feed stock at a temperature of 450 ^oC, pressure =1050 mm Hg the following products were obtained:

Products	wt %	API	Mw
Gases	15	--	32
C_5^+ gasoline	55	63	110
TCGO	26	5	260
Coke	4	--	12

Given that: WHSV = 0.7 hr^{-1}, Linear velocity of vapor (U) = 0.3 m/s,
$\rho_{catalyst}$ = 420 Kg /m^3
Calculate: a) diameter of the cracker, b) weight of catalyst needed, c) conversion, and d) efficiency.

Solution :

$$m_{feed} = \frac{1200 \text{ ton/day} * 1000 \text{ Kg/ton}}{24 \text{ hr/day}} = 50000 \text{ Kg /hr}$$

m_{gases} = 0.15 * 50000 / 3600 =2.08 Kg/s
$m_{C5^+ \, gasoline}$ = 0.55 * 50000/ 3600 =7.64 Kg/s
m_{TCGO} = 0.26 * 50000/ 3600 = 3.61 Kg/s
Total moles of vapor = (2.08/ 32) + (7.64/ 110) + (3.61/ 260)
n = 0.1479 Kg mole / s

$$R = \frac{22.4 \, (m^3/\text{ kg mole}) * 760 \text{ mm Hg}}{1 \text{ Kg mole} * 273} = 62.359$$

$$V = \frac{n R T}{P} = \frac{0.1479 * 62.359 * (450+273)}{1050} = 6.35 \, m^3/s$$

$$A = \frac{\Pi}{4} D^2 = \frac{V}{U} = \frac{6.35 \, m^3/s}{0.3 \, m/s} = 21.17 \, m^2$$

$$D = \sqrt{\frac{4 * 21.17}{\Pi}} = 5.19 \, m$$

$m_{catalyst}$ = 50000/0.7 = 71428 Kg
$V_{catalyst}$ = m/ ρ = 170 m^3

$$H = \frac{4 * 170}{3.14 * (5.19)^2} = 8 \, m$$