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SPECIAL FEATURES

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Letter from the Editor: New Year

Welcome to 2022. Everyone hopes we will soon recover from the Pandemic. It has changed how we will view life in the near future – at some point we will forget these lessons learned like many other generations. Mass Transit will need to be revised, the cities with large mass transit were the hardest hit. Air Travel will need to be revised, many things will need to undergo a review and develop new strategies. In 1918 the Spanish Flu resulted in deaths of 50 million people. Currently we have about 5.5 million deaths, which is more a credit to our advanced medical abilities, than our ability to control the virus- many flus are transmitted by birds and other animals.

How can we make 2022 better for ourselves and others? Most people think they are linked. We strive to make ourselves better, and then began to help others. We make ourselves better by learning from our mistakes and the mistakes of others – we do not need to make all our learning mistakes ourselves, in fact it is better if we learn from the mistakes of others.
Mistakes of Others – most people might agree that there are several mistakes that many people make.

Not gaining enough knowledge before starting a project. Many people learn expensive lessons as they go through a project. Better to study and gain knowledge before starting a project.
Not having practical knowledge. Many people study and gain some knowledge and then they struggle to apply the knowledge they have gained. Wisdom is the soundness of an action or decision regarding the application of experience, knowledge, and good judgment. We can be knowledgeable but not wise.

Not building a solid network of friends. Life is often hard, unfair, and deadly. One of the few real treasures in life are good friends. There are good people out there – go find them and befriend them.

We believe that IACPE can assist in overcoming some of these common mistakes of others. IACPE Level One builds a foundation of knowledge. Level Two builds how to apply the engineering fundamentals learned in Level One. IACPE also can assist in helping build a great network of good people. One way to tell if a person is good, is that then go beyond what is required. IAPCE member go beyond what is required to gain extra knowledge to help them in their live.
One of our members Kelson Gando, wrote this great comment about IACPE. Thanks for his kind words.

“By attending the IACPE program and later becoming a CPE, I have realized that the program, to a greater extent, helped enhance my professional skills - I can confidently say that I feel more comfortable now taking on big roles and/or responsibilities in my profession than I was before attending the program. Thus, I would highly recommend anyone interested in the program to enroll and become a Certified Practicing Engineer (CPE).”

All the best in your Career and Life,
Karl

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Improving the Reliability & Availability of the Refining Hardware | Management of Naphthenic Corrosion

Marcio Wagner da Silva

INTRODUCTION AND CONTEXT

The highly competitive environment of the refining industry requires high availability and reliability of the refining hardware in the sense to maximize the operational lifecycle of the process units avoiding unnecessary shutdowns and production losses. One of the great threats to the availability and integrity of equipment in the refining industry is the corrosion phenomena that can lead to a reduction in the operation lifecycle of process equipment and, in extreme cases, serious accidents.

Among the corrosion mechanisms find the crude oil refining industry, the naphthenic corrosion. This corrosion mechanism occurs in hot sections of the process units like crude oil distillation and delayed coking units, naphthenic corrosion leads to quickly material loss, representing a dangerous threat to the integrity of these process units.

CRUDE OIL DISTILLATION UNIT GENERAL OVERVIEW

The crude oil distillation unit defines the processing capacity of the refinery and, normally the others process units are sized on the basis of his yields. Figure 1 shows a basic process flow diagram for a typical atmospheric crude distillation unit.

The crude oil is pumped from the storage tanks and preheated by hot products that leaving the unit in heat exchangers battery, then the crude oil stream receives an injection of water aim to assist the desalting process, this process is necessary to remove the salts dissolved in the petroleum to avoid severe corrosion problems in the process equipment. The desalting process involves the application of an electrical field to the mixture

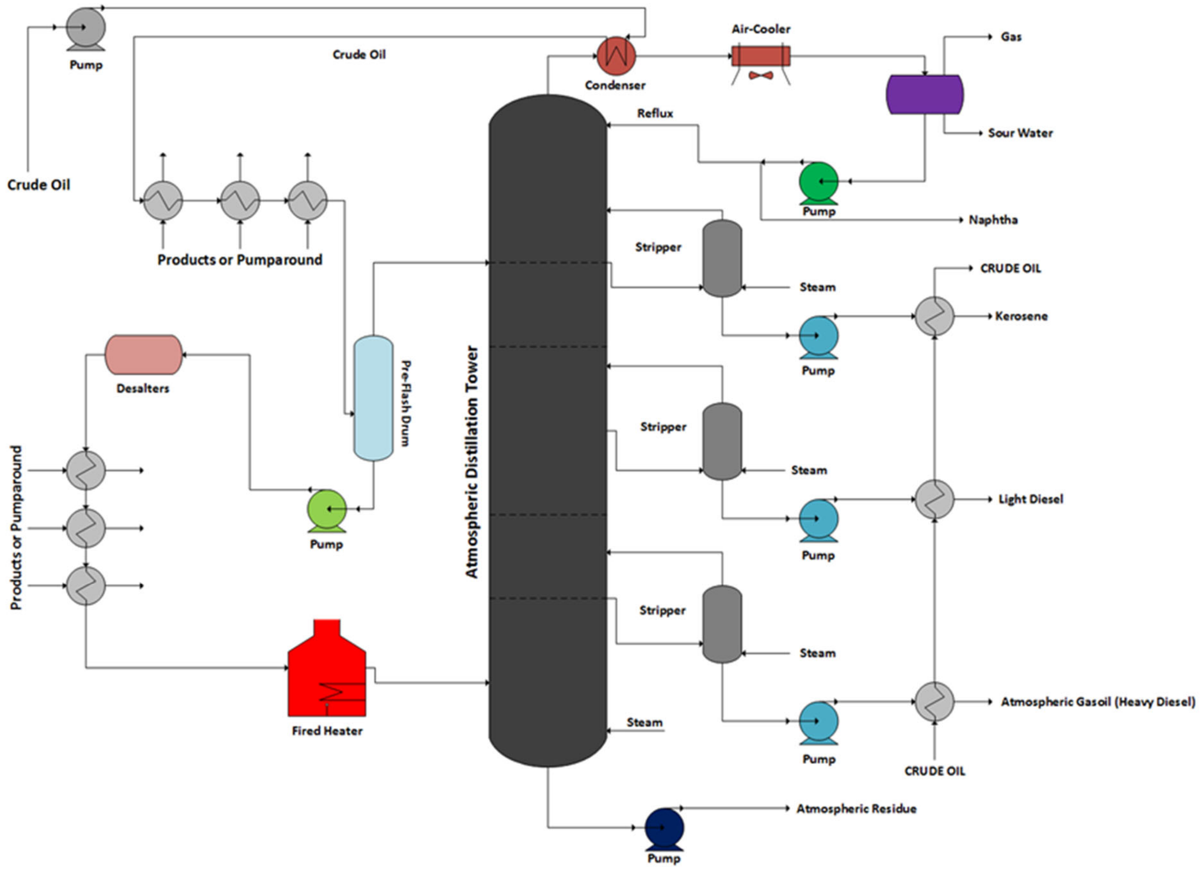

Figure 1 – Process Flow Diagram for a Typical Atmospheric Crude Oil Distillation Unit

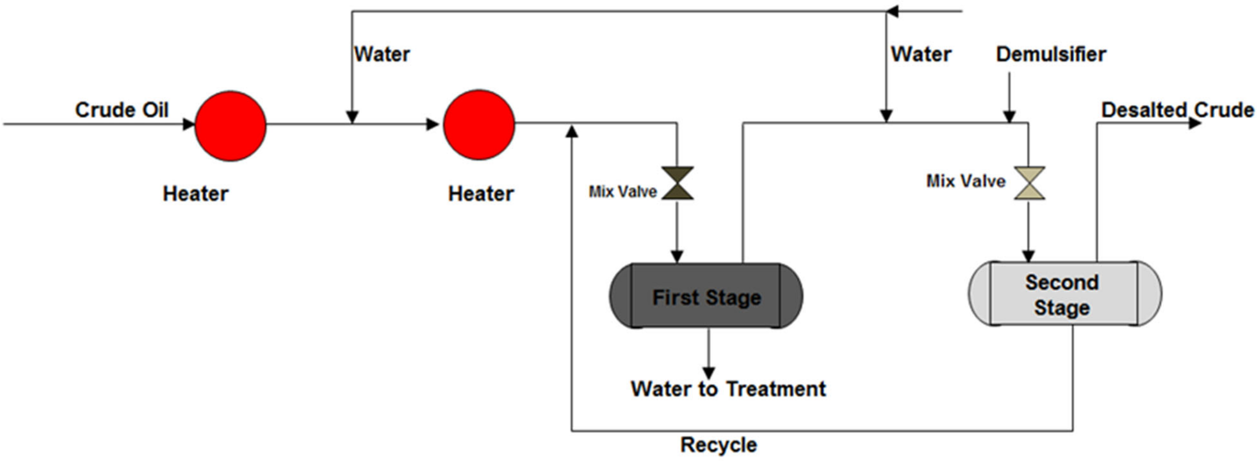

Figure 2 – Crude Oil Desalting Process with Two Separation Stage

crude oil-water aim to raise the water droplets dispersed in the oil phase and accelerate the decanting, as the salts solubility is higher in the aqueous phase the major part of the salts is removed in the aqueous phase effluent from the desalter, called brine. Normally the petroleum desalting process is carried out at temperatures among 120 and 160 °C, higher temperatures raise the conductivity of oil phase and prejudice the phase separation, and this can lead to drag oil to the brine and result in process inefficiency.

The desalting process involves the mixture of crude oil with water aiming the dissolution of the salts considering the higher solubility of these compounds in the aqueous phase. Figure 2 depicts a typical desalting process with two separation stages, the salt content in de desalted crude is normally controlled bellow of 5,0 ppm (as NaCl).

In the desalter exit, the desalted oil is heated again by hot products or pump around and fed into a flash drum, in this equipment the lighter fractions are separated and sent directly to the atmospheric tower, the main role of this vessel is reducing the thermal duty needed in the furnace. Following, the stream from the bottom of the flash vessel is heated in the fired heater to temperatures close to 350 to 400 °C (depending on the crude oil to be processed) and is fed the atmospheric tower where the crude oil is fractionated according to the distillation range, like example presented in Table 1.

At the exit of the atmospheric tower, the products are rectified with steam aim to remove the lighter components.

The gaseous fraction is normally directed to the LPG (C3-C4) pool of the refinery and the fuel gas system (C1-C2) where will feed the furnaces and boilers. The light naphtha is normally commercialized as petrochemical intermediate or is directed to the gasoline pool of the refining complex, the heavy naphtha can be sent to the gasoline pool and in some cases, this stream can be added to the diesel pool since not compromise the specification requirements of this product (Cetane number, density and flash point). Kerosene is normally commercialized as jet-fuel while the atmospheric residue is sent to the vacuum distillation tower, in some refining schemes it's possible sent this stream directly to the residue fluid catalytic process unit (RFCC), in this case, the contaminants content (mainly metals) of the residue needs to be very low to protect the catalyst of the cracking unit.

Nowadays, face to the necessity to reduce the environmental impact of the fossil fuels associated with the restrictive legislations,

Table 1 – Example of Crude Oil Distillation Cuts

Fraction	**Distillation Range (°C)**
Gases (C_1 – C_4)	≤ 30
Light Naphtha (C_5 – C_7)	30 - 100
Heavy Naphtha (C_8 – C_{11})	80 - 200
Kerosene (C_{11} – C_{12})	170 - 280
Light Diesel (C_{13} – C_{17})	220 - 320
Heavy Diesel (C_{18} – C_{25})	290 - 350
Atmospheric Residue (C_{25+})	350 - 390

difficultly the straight run products can be commercialized directly. The streams are normally directed to the hydrotreating units aim to reduce the contaminants content (sulfur, nitrogen, etc.) before being marketed.

In distillation units with higher processing capacity, normally the flash drum upstream of the atmospheric tower is substituted by a prefractionation tower. In these cases, the main advantage is the possibility of reduction of the atmospheric tower dimensions that implies in cost reductions associated with the unit implementation and improve the hydraulic behavior in the distillation tower, consequently with better fractionation. This arrangement is shown in Figure 3.

VACUUM DISTILLATION SECTION

The bottom stream of the atmospheric column (Atmospheric Residue) still contains recoverable products capable to be converted into high added value derivatives, however, under the process conditions of the atmospheric unit, the additional heating led to thermal cracking and coke deposition.

Aiming to minimize this effect, the atmospheric residue is pumped to the vacuum distillation column where the pressure reduction leads to a reduction in the boiling point of the heavy fractions allowing the recovery while minimizing the thermal cracking process.

The vacuum generated in the column can be humid, semi-humid and dry. Humid vacuum occurs when is applied steam injection in the fired heater and in the column aiming to reduce the partial pressure of the hydrocarbons improving the recovery while in the semi-humid vacuum the steam is injected only in the fired heater minimizing the residence time reducing the coke deposition. The dry vacuum does not involve the steam injection, in this case, is possible to achieve pressures between 20 to 8 mmHg while in the humid vacuum the column operates under pressures varying between 40 to 80 mmHg, however, it’s possible to achieve comparable yields through the injection of stripping steam. Figure 4 presents a process arrangement for a typical vacuum generation system in a vacuum crude oil distillation unit.

As shown in Figure 5, the traditional arrangement of vacuum units presents two side drawn, heavy and light gasoil. These streams are normally directed to conversion units like hydrocracking or fluid catalytic cracking (FCC), according to the adopted refining scheme. The fractionating quality achieved in the crude oil vacuum distillation column has a direct impact upon the reliability and conversion units’ operation lifecycle, once which in

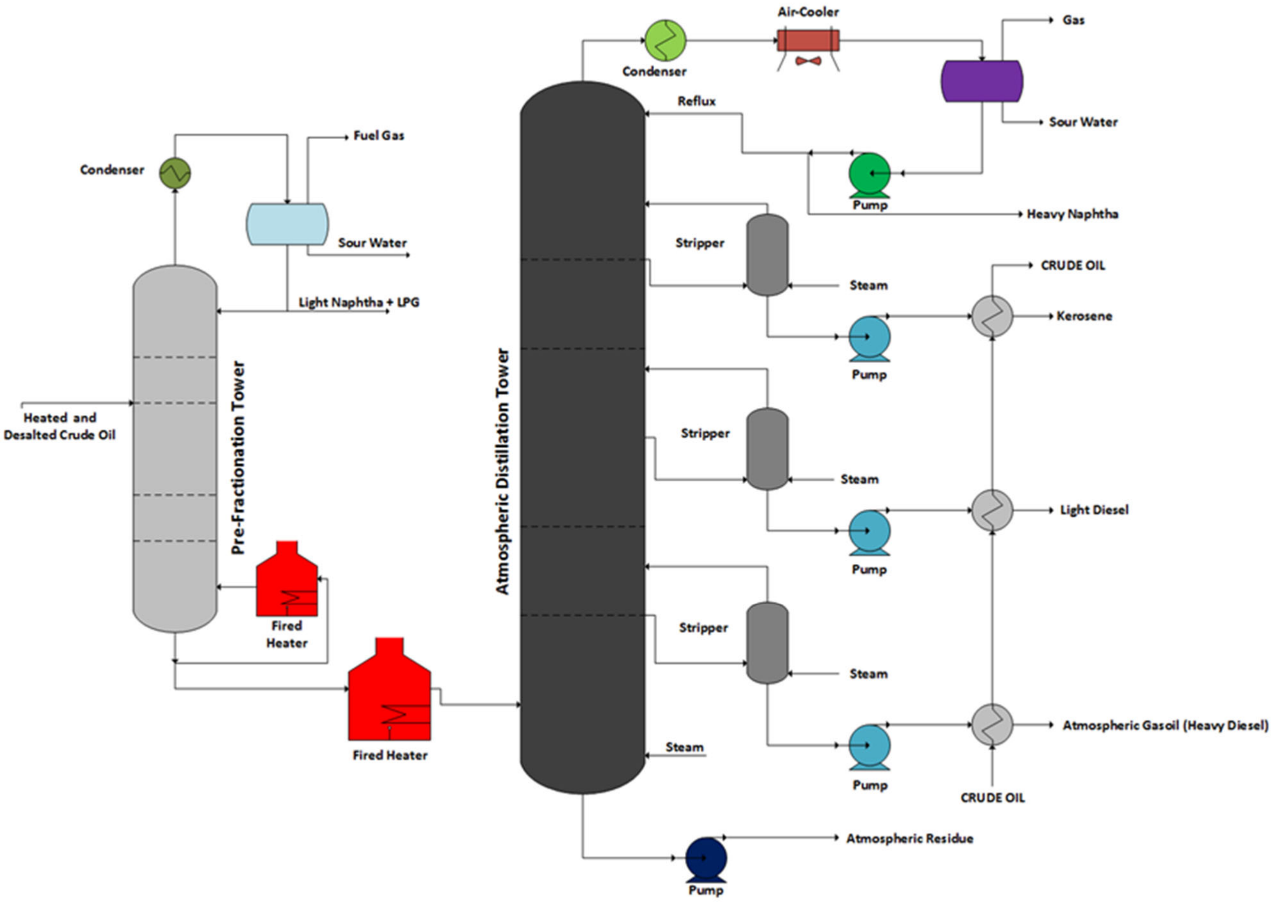

Figure 3 – Typical arrangement to Atmospheric Distillation with Pre-Fractionation Tower.

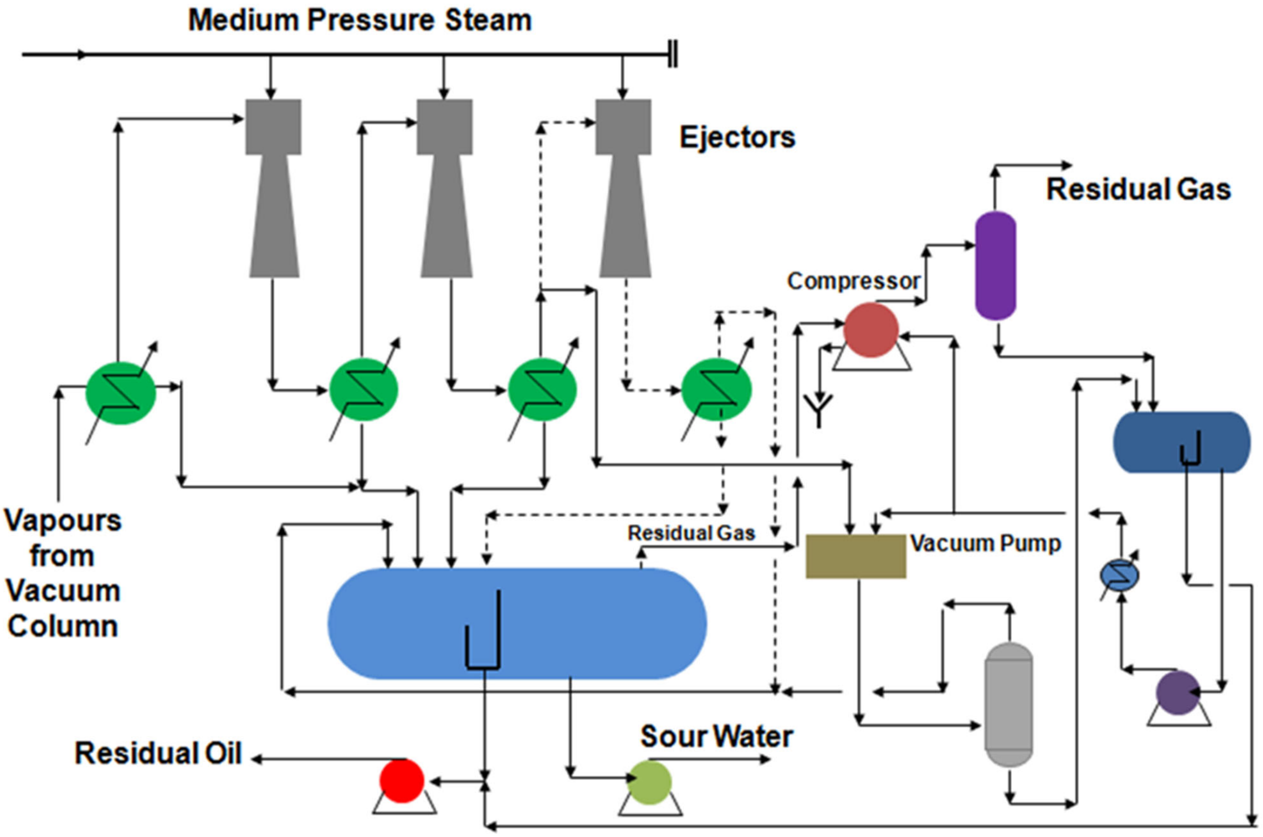

Figure 4 – Process Arrangement for a Typical Vacuum Generation System for a Vacuum Crude Oil Distillation

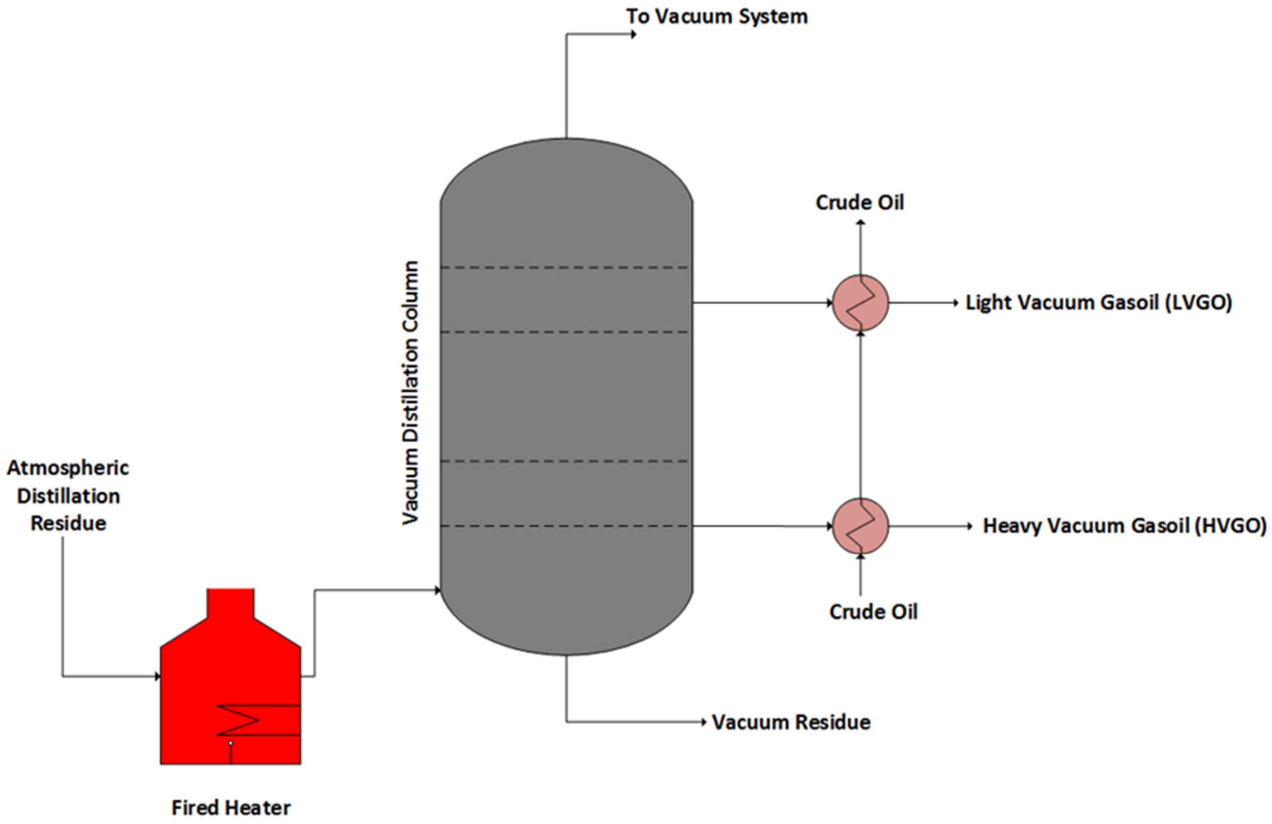

Figure 5 – Schematic Process Flow Diagram for Vacuum Distillation

this step is controlled the metals content and the residual carbon (CCR) concentration in the feedstock to these processes, high values of these parameters lead to a quickly catalyst deactivation raising operational costs and reducing profitability.

Some refiners include additional side withdraws in the vacuum distillation column. When the objective is to maximize the diesel production, it's possible to add a withdraw of a stream lighter than light vacuum gasoil that can be directly added to the diesel pool or after hydrotreating, according to the sulfur content in the processed crude oil. When the crude oil presents high metals content, it's possible to include a withdraw of fraction heavier than the heavy gasoil called residual gasoil or slop cut, this additional cut concentrates the metals in this stream and reduce the residual carbon in the heavy gasoil, minimizing the deactivation process of the conversion processes catalysts as aforementioned. The vacuum residue is normally directed to produce asphalt and fuel oils, however, in most modern refineries this stream is sent to bottom barrel units as delayed coking and solvent deasphalting to produce higher-value products.

According to the refining scheme, the installation of vacuum distillation units can be dispensed. Refiners that rely on residue fluid catalytic cracking units (RFCC) can sent the atmospheric residue directly to feed stream of these units, however, it's necessary to control the contaminants content (metals, sulfur, nitrogen, etc.) and residual carbon (CCR) aiming to protect the catalyst, this fact restricts the crude oil slate that can be processed, reducing the refiner operational flexibility. On the other hand, in refineries that process extra heavy crudes, normally the crude oil distillation unit is restricted to the vacuum unit once the yields of the atmospheric column would be very low and the coking risk very high.

In refineries optimized to produce lubricants, the distillation process is modified face to the paraffinic characteristics of the crude oil processed, mainly the vacuum distillation step. The necessity to separate the lubricants fractions requires higher fractionation quality in the column and some configurations rely on two columns, as presented in Figure 6.

The distillation unit design is strongly dependent by the characteristics of crude oil that will be processed by the refinery, for extra-heavy oils normally the crude is fed directly to the vacuum column. The design is generally defined based on a limited crude oil range that can be processed in the hardware (Contaminant content, API grade, etc.).

DELAYED COKING TECHNOLOGIES GENERAL OVERVIEW

Delayed coking employs the thermal cracking concept under controlled conditions to produce light and middle streams (LPG, naphtha and gas oils) from residual streams which would normally be used as diluents in fuel oils production.

The typical feed stream for delayed coking units is the residue from vacuum distillation process that contains the heavier fractions of processed crude oil, however, streams like decanted oil from FCC unit and asphaltic residue produced in solvent deasphalting can compose the feed stream to the delayed coking unit, depending upon the refining scheme

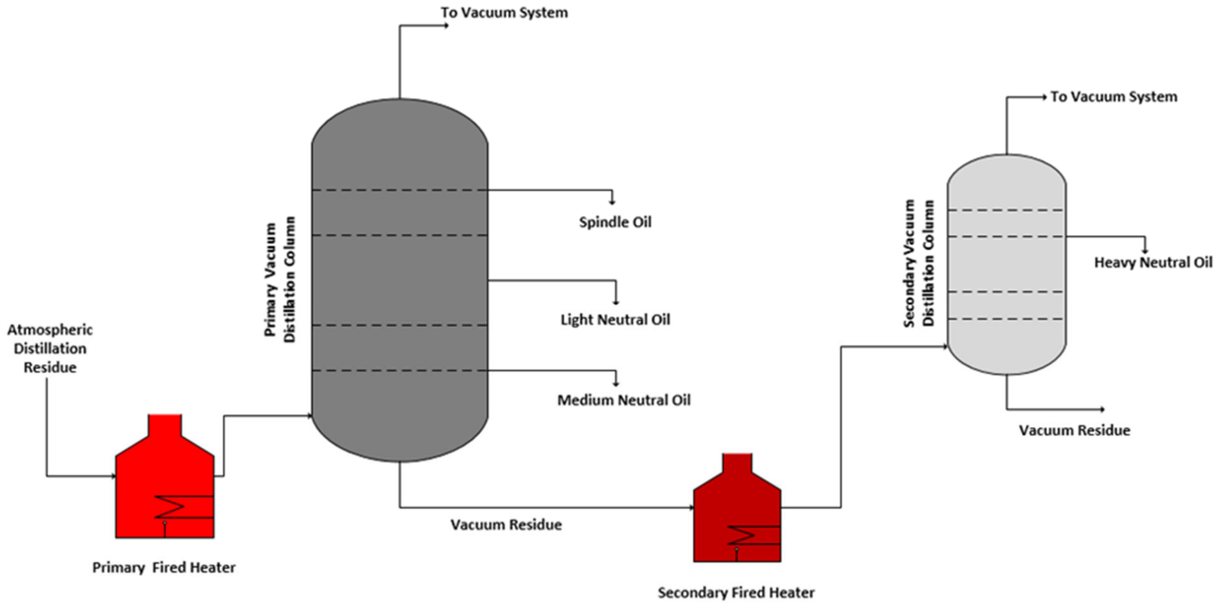

Figure 6 – Vacuum Distillation Process to Produce Lubricants

adopted by the refiner. Another possibility is sent the residue from atmospheric distillation directly to the delayed coking unit, in this case, the unit design is quite modified demanding greater robustness of the fractionating and gas compression section.

Due to the thermal cracking characteristics (low availability of hydrogen during the reactions), the streams produced by delayed coking unit have a high concentration in olefinic compounds, which are chemically unstable. Furthermore, due to the processing of residual streams that have high contaminants content like nitrogen, sulfur, and metals, therefore the refiners that apply delayed coking units need high hydrotreating capacity to convert these streams into added value products and that meets the contaminants level according to the environmental regulation. Figure 7 presents the process flow scheme for a typical delayed coking unit.

The feed stream is fed into the bottom of the main fractionating tower where is mixed with the heavier fraction of the thermal cracking products and then sent to the fired heater where thermal cracking reactions are initiated, the reaction conditions are controlled so that the reactions are completed in the coke drums, the residence time in the fired heater must be the lowest possible to minimize the coke precipitation in the fired heater tubes. A manner of minimizing the coke formation in the walls of tubes is the steam injection that raises the velocity and consequently reduces the residence time.

After the fired heater the feed stream is sent to the coke drum or reactor, where the thermal reactions are completed, and the coke is deposited. The thermal cracking products are removed from the top of the reactor and receive an injection of quench with a cold process stream (normally heavy or middle gas oil) and directed to the main fractionators where the products are separated. The coke deposited in the reactor is removed through a cut with water under high pressure (about 250 bars).

Delayed coking is a process that occurs in batch, in order to make a semi-continuous process are always employed pairs numbers of reactors and each two reactors is applied one fired heater when one reactor is under reaction the other is in decoking step and so on. The delayed coking process occurs in cycles that can vary from 14 to 24 hours.

The main operational variables of the delayed coking unit are recycle ratio which is the quantity of the total feed stream which corresponds to the heavier fraction of the reaction products that are mixed with the fresh feed, reactor temperature, normally considered in the top of the coke drum, pressure in the top of reactor and the time of the reactor cycle.

The recycle ratio vary normally between 5 to 10% (to units dedicated to producing fuels) and the refiners seeks to operate the unit with de lower recycle ratio possible in order to

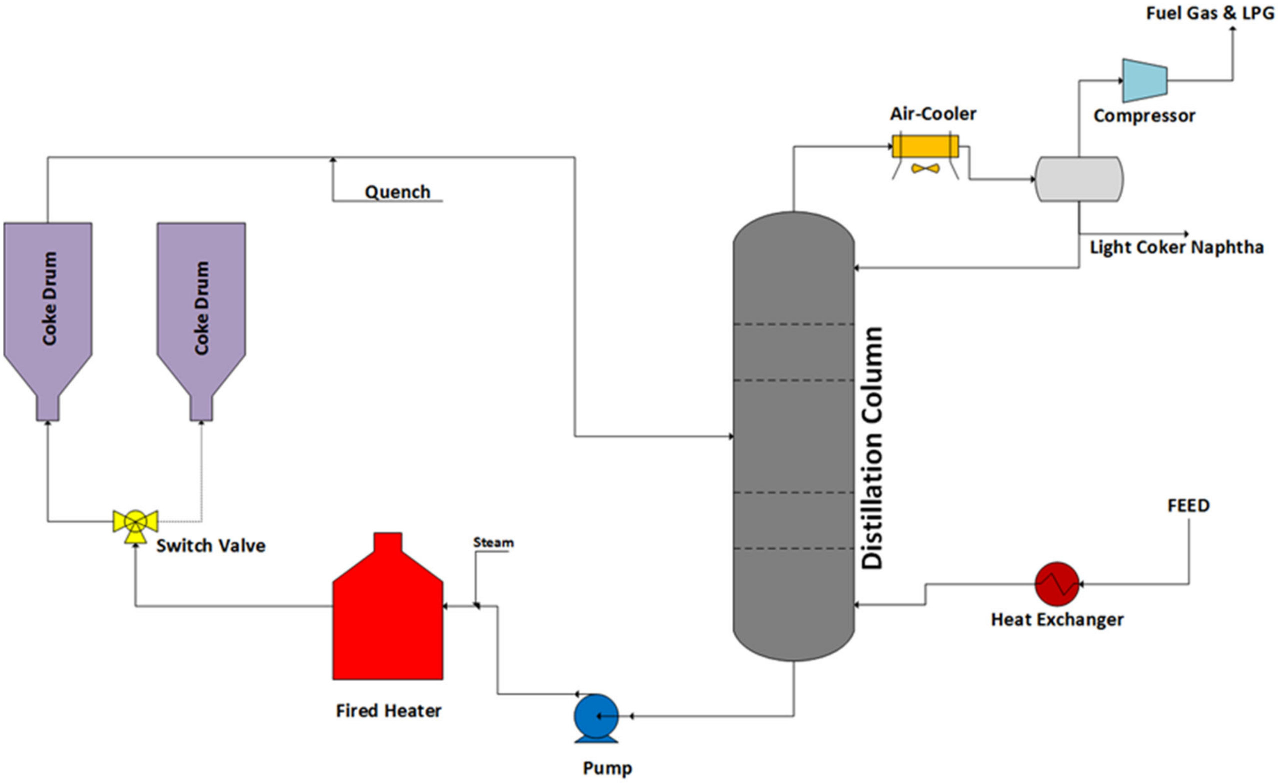

Figure 7 – Typical Arrangement for Delayed Coking Unit

maximize the capacity of the plant in processing residual streams. The reactor temperature is close to 430 oC and is linked with the fired heater temperature, throughout the thermal cracking reactions the temperature fall due to the endothermic characteristics of the reactions.

The pressure in the reactor can vary between 1 to 3,5 bars, in units optimized to producing fuels the variable is maintained at lower levels, on the other hand, when the unit is dedicated to producing high-quality coke, the unit is operated under higher pressures.

Reactor cycle time is linked to the function performed by the delayed coking in the refining scheme. Units dedicated to producing fuels operate at shorter cycles and units optimized to producing high-quality coke operate under longer cycles.

The coke produced normally is seen as a by-product of the delayed coking unit, however, in some cases, the delayed coking process is optimized to producing high-quality coke and the coke becomes to the principal product of the process.

Depending on the feedstock quality that will be processed, three types of the coke can be produced:

- Shot coke – Poor quality coke produced from feedstock with high asphaltenes and contaminants (sulfur, nitrogen and metals) content, normally this type of coke is commercialized as fuel;
- Sponge coke – In this case, the feedstock have a lower asphaltenes and contaminants content and the coke can be directed to raw material to anodes production process to the aluminum industry;
- Needle coke – The production of this type of coke require the processing of feedstock with high aromatics content (decanted oil from FCC, for example) and these products are sent as raw material to producing anodes to the steel industry;

As mentioned above, production of high-quality coke requires a quality control of the feed stream that will be processed, in the most of the cases the refiners choose to install delayed coking units focusing in the production of middle and light distillates. Therefore, the unit optimization to produce needle coke occurs only in specific cases.

The heavy gas oil stream is normally directed to the fluid catalytic cracking unit or can be utilized as fuel oil, in refining schemes that have deep hydrocracking units this stream can be used like feedstock to the unit. The sending of this stream to the fluid catalytic cracking unit needs be controlled to avoid the premature deactivation of catalyst, face of the high level of contaminants, mainly nitrogen and metals.

Middle and light gas oils are normally sent to severe hydrotreating units to compose the diesel pool of the refinery. The heavy coker naphtha can be directed like feed stream to FCC units. When the flash point specification of diesel is not restricted this stream can be sent to the diesel pool, after deep hydrotreating process.

The lighter fraction of naphtha can be sent to the gasoline pool of the refinery after hydrotreatment or directed to FCC units, in this case this stream contributes to raise de LPG production in the FCC unit. In some cases the light coker naphtha can be sent to catalytic reforming units aiming to produce high octane gasoline or petrochemical precursors (benzene, toluene and xylenes).

The overhead products from main fractionator are still in gaseous phase and are sent to the gas separation section. The fuel gas is sent to the refinery fuel gas ring, after treatment to remove H2S, where will be burned in fired heaters while the LPG is directed to treatment and further commercialization.

On the other hand, the delayed coking technology obliges the refiners the necessity of high hydrotreatment capacity once the streams produced by the unity needs severe treating process before being sent to the commercialization, this fact can raise the operational and installations costs.

NAPHTHENIC CORROSION

The naphthenic corrosion occurs in processing units that process bottom barrel streams that tends to present high concentration of naphthenic acids and operates under high temperatures. Due to these characteristics, naphthenic corrosion phenomenon is observed in crude oil distillation units and residue upgrading units like delayed coking.

The characteristics of the processed crude oil slate are a determinant factor in the naphthenic corrosion. A very relevant characteristic of oils for refining hardware is naphthenic acidity. Naphthenic acidity is determined based on the amount of KOH required to neutralize 1.0 gram of crude oil, normally a mixture of crude oils is sought in the refinery load so that it does not exceed 0,5 mg KOH/g, above this reference, the bottom sections of the distillation units can undergo a severe corrosive process, leading to shorter periods of operational campaign and higher operating costs in addition to problems associated with integrity and safety. Naphthenic acidity is directly linked to the concentration of oxygenated compounds in the crude oil that tend to be concentrated in the heavier fractions, giving instability and odor to the intermediate currents.

The sulfur content in the crude oil is another key factor in the naphthenic corrosion phenomena. In crude oils with sulfur content higher than 2,0 %, a protective layer of iron sulfide (FeS) is formed in the metal surface that is insoluble, avoiding then the attack by naphthenic acids as presented in Figure 8. In this sense, refiners processing very low sulfur crude oils with high Total Acid Number (TAN) can face severe issues with naphthenic corrosion in their refining hardware.

The equation 1, presents the chemical representation of naphthenic corrosion.

$\mathrm{Fe} + \mathbf{2RCOOH} \rightarrow \mathrm{Fe(RCOO)2} + \mathrm{H2}$ (1)

According to the literature, above of 400 oC the iron naphthenate (corrosion product) is that is soluble in hydrocarbons is attacked by hydrogen sulfide (H2S) leading to the regeneration of the naphthenic acid, as represented in equation 2.

$\mathrm{Fe(RCOO)2} + \mathrm{H2S} \rightarrow \mathrm{FeS} + \mathbf{2RCOOH}$ (2)

In crude oil distillation units, the bottom section of atmospheric column and the vacuum distillation column are the most common regions where is observed naphthenic corrosion while in delayed coking units the phenomenon is observed in bottom section of main fractionators column.

The carbon steel, series 300 and 400 stainless steel, and nickel alloys tend to suffer naphthenic corrosion.

Among the actions to control the naphthenic corrosion in the refining hardware is the blending of crude oils aiming to keep the Total Acid Number (TAN) and sulfur content inside the adequate limits. Other alternatives are the injection of neutralizers or corrosion inhibitors in the processing streams and the selection of materials with higher resistance to naphthenic acid attack.

Refiners processing crudes with high acidity and low sulfur normally applies chromium and molybdenum alloys in the bottom sections of crude oil distillation columns and transfer pipes as well as in residue upgrading units. In extreme conditions, it's possible to consider apply stainless steel 317 L that presents high resistance to naphthenic corrosion, this decision needs to consider the higher capital investment due to the high cost of this material.

CONCLUSION

The availability of the refining hardware is a key parameter to ensure the economic sustainability of the refiners, especially those inserted in highly competitive markets.

Like described above the naphthenic corrosion can compromise the reliability and availability of the key units to the refining hardware, by this reason, the naphthenic corrosion issues can need to take into account in the crude oil selection aiming to minimize integrity risks and shorter operational campaigns. In this sense, adequate monitoring, and control of corrosion process in these units is fundamental to ensure the competitiveness of the players in the downstream sector.

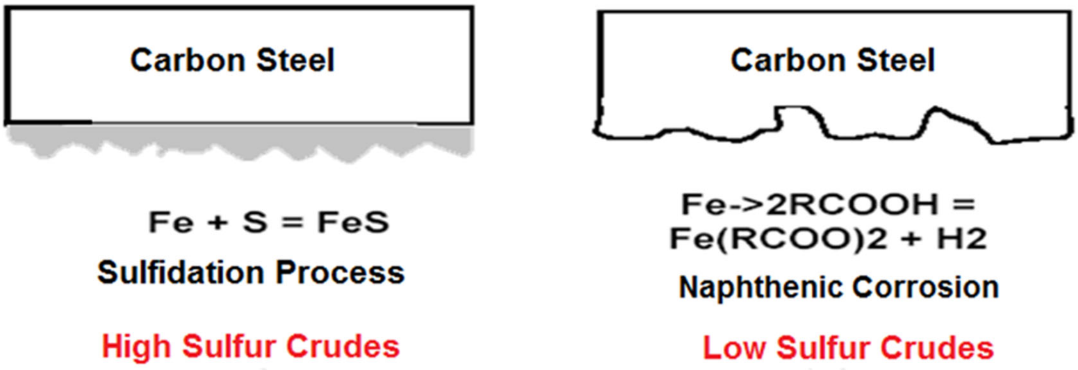

Figure 8 – Naphthenic Corrosion Process

REFERENCES

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

RAMANATHAN, L.V. Corrosion and his Control. 1st ed. Hemus Press, 1978.

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

SILVA, M. W.; CLARK, J. – Flexible Upgrading of Heavy Feedstocks. PTQ Magazine, 2021.

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Considering Feedforward Strategy for Process Control

Alex Michinel

The use of feedback control strategies is one of the most common ways to control a process, but it is important to consider that feedback control is an error-driven strategy; correction to any upset in the process we want to control depends on the error or difference between the desired value of the “controlled” variable and its current value.

By the other hand, the feedforward control strategies use the knowledge of the process behavior to take action before a perturbation to the process has any effect on the controlled variable.

Several methods has been discussed to implement a feedforward control strategy, one of them is explained by Soundar Ramchamdram (1). He uses steady-state models of the process to determinate the value that the manipulated variable should has to compensate the effect of perturbations on the controlled variable. He successfully demonstrated that Process-Model-Based control was appropriate to control the trays temperature in a typical wastewater column.

A SIMPLE EXAMPLE

To give an example of the advantage that a feedforward strategy can have over a pure feedback control, let’s have a look on a very simple system; consider a tank were we want to control the level, as shown in figure (No 1). This is a typical first order system.

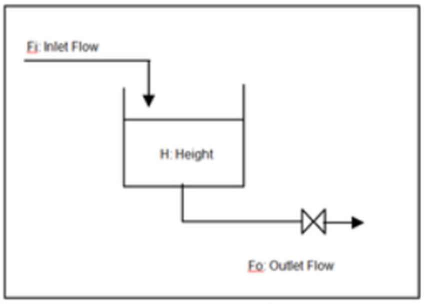

Figure (N° 1): Typical Tank System

We would like to keep the level of the tank under control. The Feed flow can vary and the Outlet flow will be manipulated through a control valve.

MODELING THE PROCESS

To simulate the process, MATLAB 5.1 can be used. This requires the process equations and the transfer function of the process representation in a block diagram.

The equations that describe this process are as follows:

$$Fi = Fo + A(dH/dt) \qquad (1)$$

Where, Fi is the inlet flow, Fo is the outlet flow, A is the transversal area of the thank and H is the height of liquid in the tank.

It is assumed that at the initial stead-state the process parameters are: H=2m, A=5m2 , Fi and Fo are 0.2m3/seg.

Laplace can be applied to find the transfer functions of the process.

$$Fi(S) = Fo(S) + ASH(S) \qquad (2)$$

$$H(S) = (Fi(S) – Fo(S)) / (A\ S). \qquad (3)$$

The figure (No 2) shows the resulting transfer function for the tank system. Using SIMULINK of MATLAB 5.1 this process can be dynamically simulated to study the effect of perturbations in the inlet flow on the tank level.

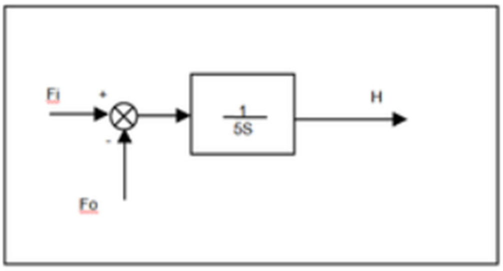

Figure (N° 2): Transfer function of the process

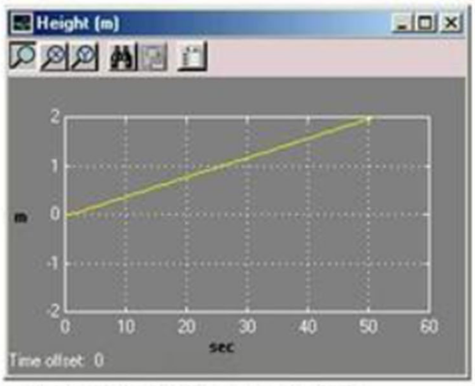

Figure (No 3): Response to Step

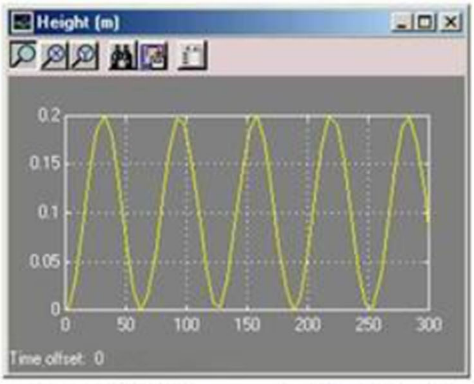

Figure (No 4): Response to 0.05sin(0.1 t)

The figures (No 3) and (No 4) show the response to a step and to a sinusoidal function as perturbation to the inlet flow respectably. Note that graphics show variation from initial steady-state in meters.

The sinusoidal function applied on the inlet flow had amplitude of 0.05m3/sec and a frequency of 0.1.

CONTROLLING THE SYSTEM WITH A FEEDBACK STRATEGY

As discussed above, feedback control is the most common strategy used to control a desired variable. So, this kind of strategy was tested on the modeled system. The controlled variable is the tank level and the manipulated variable is the outlet flow handled through the control valve.

The control strategy used can be observed in the figure (No 5).

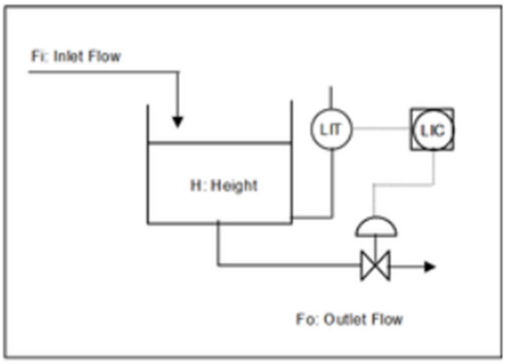

Figure (N° 5): Feedback control strategy

This control strategy was tested with two different kind of perturbation on the inlet flow. The figure (No 6) shows the block diagram used to simulate the control strategy. Note that the control valve has a transfer function like a first order process. This is to have a better approximation to what really happens with the valve positioning in response to a signal from the level controller.

The strategy was exposed to a change in inlet flow of 0.05m3/seg. The response to this perturbation can be seen in the figure (No 7).

The maximum overshot obtained was about 2% in the level. The gain and the integral time used for the PID were 0.5 and 0.2 respectively. This set of tuning parameters can be improved, but they will not be adequate for the next perturbation that this strategy was tested against lately.

The strategy was then tested with a sinusoidal inlet flow. It can be observed in the figure (No 8) that the strategy can not control properly this perturbation.

The level kept fluctuating around the set-point; the amount of the fluctuation was about ±5%.

Other tests were carried out such as changes in the level set-point and the combination of changes in the set-point and the introduction of perturbations in the inlet flow.

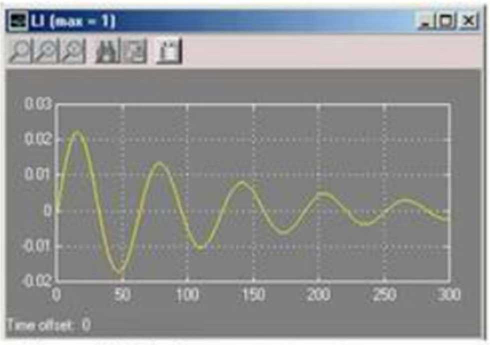

Figure (N° 7): Response to a step

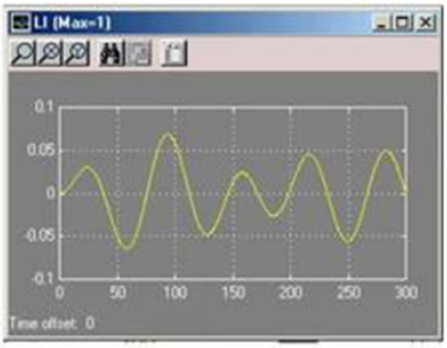

Figure (N° 8): Response to sinusoidal

IMPROVING WITH A FEEDFORWARD STRATEGY

A feedforward strategy was designed to evaluate its performance under the same perturbations that the feedback strategy was exposed.

The process diagram of the strategy can be seen in the figure (No 9).

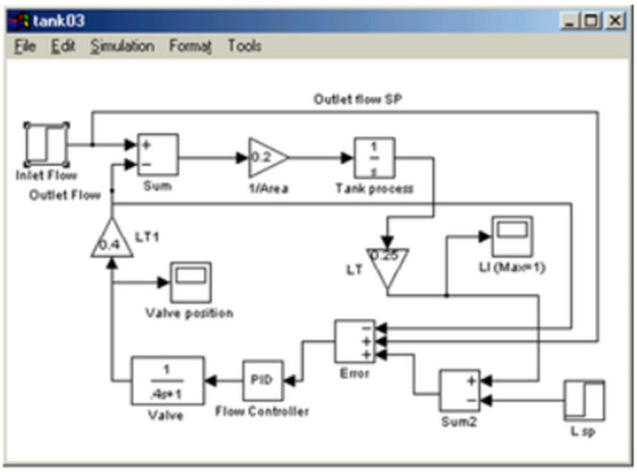

Figure (No 9): Feedforward strategy

The results obtained with the feed-forward control strategy are better than the obtained with the feedback strategy. In the figures (No 10) and (No 11), can be observed the response of the system to two kinds of perturbation; one step and one sinusoidal function respectively.

Basically in this strategy two errors are controlled at once; the error between the measured outlet flow and its set-point, and the error between the current level and its set-point.

The steady-state of this system tells that the inlet flow to the tank must be equal to the outlet flow from the tank. For this reason the inlet flow is send as set-point to the outlet flow controller to take action in advance, before the level is actually perturbed by the inlet flow upset.

Any error in the level will be corrected by the same outlet flow controlled since the error between the current level and its set-point is added to the error discussed above.

In the figure (No 10), can be observed that the maximum overshot obtained is less than 0.3% significantly less than the 2% obtained with the feedback strategy.

In the figure (No 11) it is shown that the level will be oscillating around the set-point about ±0.5% which is less than the 5% resulting in the feedback strategy.

WHAT IS AGAINST

“Nothing is better for nothing”; Having a better strategy should involve some extra cost. In this case the installation of flow instruments at the inlet and outlet of the tank involve some extra costs. How important is to keep the controlled variable with a permissible error is the key question to evaluate if the extra cost involved in a better strategy is profitable.

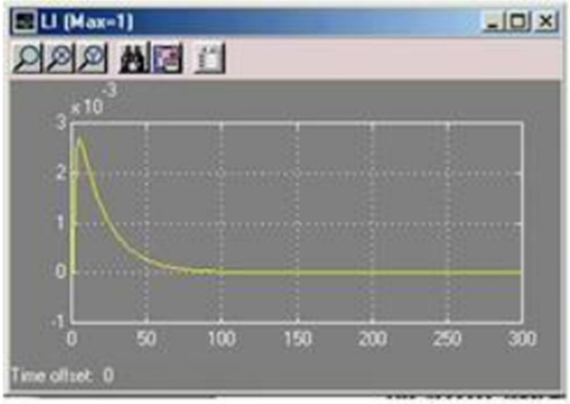

Figure (No 10): Response to step

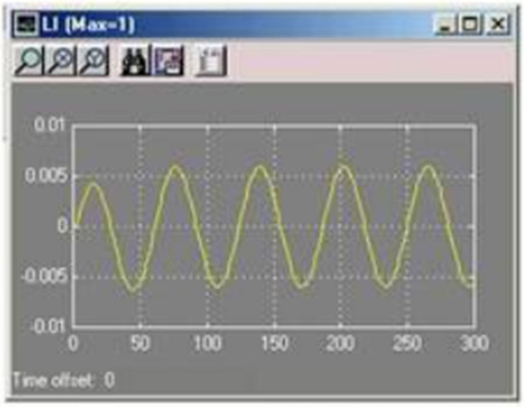

Figure (No 11): Response to sinusoidal

IN THE REAL LIFE

This designed strategy can be implemented in a real application such as bottom of a distillation column, where the inlet flow will be proportional to the column feed flow. In most of the cases of distillation columns the feed and bottom flows are measured, so this strategy will not represent an extra cost.

Another implementation could be in boilers and evaporators, where a steady level of condensate is very important.

It will be very interesting to evaluate the performance of the discussed strategy in processes with delays or a higher order.

CONCLUSION

Feedforward control strategies have a better performance than pure feedback control strategies when implemented on processes of first order, but this improvement can involve some extra costs. The balance between the profitability of a better strategy and the importance of having a very steady controlled variable, is the key point to choose or design the control strategy.

REFERENCES

Soundar Ramchandran, “Consider Steady-State Models for Process Control”, Chemical Engineering Progress, Feb-1998, pp 75 – 81.

Katsuhiko Ogata, “Ingenieria de Control Moderna”. 3rd Edition, 1997, pp 9 – 40

S. Majhi; D.P. Atherton, “Online tuning of controllers for and unstable FOPDT process”, IEE Proc.-Control Theory Appl., Vol 147 No 4, July 2000, pp. 421 – 427.

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Since 1987 until now I have accumulated a vast experience in Process Engineering, Project Management, plant pre-commissioning and commissioning activities, Turnaround maintenance planning, process simulation and troubleshooting, Feasibility Studies, Technology Evaluation, in the petrochemical and refining business.

Rock Bottom View

Into a Clearing, But Not Out of the Woods Just Yet…

Ronald J. Cormier, *Engineering Practice* Contributing Author

Hello again from the screened porch on this old central Texas ranch, above the Pedernales River. As 2022 opens, the weather is starting to cool, which makes my trusty old propane heater feel mighty welcome as I pen these thoughts. A quick look at our business as we emerge from Covid Round 1, (pre-Omicron) reveals that the US Chemical Industry is on track for a strong recovery, having witnessed a solid rebound since the beginning of 2021. Demand increases from the major end markets such as construction and health and safety, lead the growth trend.

This growth was driven partly by a rebound in US GDP, which will likely grow between 6.0% and 6.5% during 2021 after declining by 3.5% in 2020. Industrial production in the United States is expected to grow by 5.5% during 2021 and by 4.3% during 2022. US chemical volumes are expected to grow around 1.5% in 2021 and 3.0% in 2022, while shipments will likely increase by 8.0% in 2021 and 2022, following a 13.5% decline in 2020.

During the first half of 2021, the industry experienced supply chain disruption caused by extreme weather events when significant chemical capacity along the US Gulf Coast went idle during summer hurricanes. With supply chain challenges easing, idled capacity should come back online and support inventory buildup. In fact, some commodity chemicals have already achieved pre-pandemic sales levels in Q3 2021 on a year-over-year basis. 2022 could mark the full recovery for the US chemical industry post-pandemic.

US chemical exports are also expected to grow significantly as major economies reopen and import demand in partner economies improves, especially with trade between the United States and China returning to business as usual. The rapid recovery in the automotive sector in China is expected to keep chemical imports high in 2022. After declining 7.6% in 2020, chemical exports are expected to grow draftbetween 5.5% and 6.0% in 2021 and reach pre–COVID-19 levels by the end of 2022.

As always there are possible structural risks at hand. One of the risks for a strong recovery is inflation. For example, Brent crude oil spot prices rebounded strongly and remained at an average of $74 per barrel in September 2021. In a recent survey by Deloitte, 60% of the chemical industry respondents saw volatile costs as the biggest risks facing their respective companies in 2022.

The September 2021 inflation figures provide further evidence that the supply of raw materials and labor is struggling to keep up with resurgent demand. Central banks in the United States and Europe expect supply bottlenecks and inflation to ease in 2022 once the global recovery finds firmer footing and government quantitative easing programs are unwound.

As well, it is still prudent to monitor possible pandemic variants that may change/slow the scope and timing of improvements, should such occur. However, if there are any silver linings to be had from the pandemic so far, demand for personal protective equipment such as masks, gowns, and gloves should remain high, driving ethylene and propylene sales. The industry could still pivot some of its production capacity toward the products and materials needed to combat the pandemic, such as isopropyl alcohol and ethanol. Industry players are likely to ensure a strong supply of chemicals required to produce antibacterial wipes, disinfectants, and surfactants for soaps and hand sanitizers.

As the industry moves into 2022, strong demand for both commodity and specialty chemicals should keep prices robust throughout the year. The industry should also experience increased capital expenditure as leading industry players focus on building capacity and expanding into growing end markets through both organic and inorganic routes. However, as mentioned above, keep an eye on cost inflation. The industry could face margin pressures amid higher raw material and fabrication material costs too, which will likely remain elevated through the first half of 2022. Additionally, historic inventories for LPG going into winter, appear low and hence higher priced. Similar dynamics will apply for naphtha as historic summer driving season patterns approach. Also, industry margins could come under increased pressure toward the second half of 2022 as pent-up demand starts to clear out.

A key initiative on the minds of many chemical leaders in 2022 will be returning employees to work. While the industry quickly implemented the required safety standards to incorporate an element of remote work, the transformed talent landscape likely requires chemical companies to adapt further. Baby boomers who entered the workforce in the '80's and in many cases rose to senior leadership and thought/change management positions, are retiring in record numbers (though thankfully heavy implementation of experienced consultant expertise remains available to bridge bench strength and talent requirements).

Changing demographics and skills requirements should draw a more diverse workforce to chemical companies than ever before. Consider International Association of Certified Practicing Engineers (IACPE) certification for young operations/maintenance/engineering professional personnel. Besides being a value-added CV credential, IACPE training ensures a professional level of competency and ethics, developing and strengthening principles in three main areas: personal, professional, and networking—many times perspectives not covered in undergraduate degree programs.

One of the critical areas of focus for most US chemical companies in 2022 will likely be sustainability and decarbonization. Mark Nicolich, US CEO of Braskem spoke on this very topic at December's fourth quarter 2021 meeting of The Northeastern Chemical Association (NECA) in New York City. Similar strategic Board decisions and deployment of capital towards sustainability have become foundational at most large firms such as Dow and LyondellBasell. This upcoming paradigm shift toward process optimization which emphasize emissions reductions/elimination designs will be factored at the beginning of the design equation, vs. handling at the end as has been the case in the past. Effectively a new field of expertise in its own right, sustainability will present challenging new opportunities for new and young talent entering petrochemicals.

In parallel, ever more demanding environmental requirements will be required to produce the same desired end-use petrochemicals via alternate or optimized stoichiometric routes. In lockstep (hopefully), such optimization will also provide for mandatory process and talent upgrade opportunities on the fuels refining side of the industry as well. After WW2, demand for modern performance products exploded; these previously-unoptimized streams derived added value vs. being flared or reacted away in the fuel refinery. In other words, time-honored accepted petrochemical synthesis methods, which pollute excessively, will start to pare away from their original roots as refining co-products, then becoming "on-purpose" routes to these desired products.

Do not forget that the refining sector will simultaneously be coping with changing carbon balance and overall demand volume reductions required as passenger and goods transport platforms move to electrification. Many chemical companies are expected to increase investment in research and development (R&D) capabilities and leverage advances in decarbonization and recycling technologies to lower their and their customers' carbon footprint, as well as reduce plastic waste. Another issue is whether demand for many conventional plastics and chemicals could wane as the public becomes more educated about the environmental impacts of end products and ready to accept eco-friendly substitutes. As in 2021, the market could show that people are willing to switch to more environmentally friendly substitutes, even if they cost slightly more or function less effectively.

In closing then, 2022 should see more industry leaders create their firms' goals and plans around abatement of emissions and monetization of waste. Toward this end, some will most likely study divestment from processes which currently cause undue cost associated with emissions conversion and/or distribution risk in handling. Such could pare back investments in traditional products and services to free up capital for more future-oriented projects in 2022. Some companies are shifting investments from gas to liquids and refining projects to developing differentiated applications for performance chemicals. Many other companies will likely face similar decisions in the near future, though the range of opportunities will continue to change as technologies and markets evolve.

For chemical companies, the past two years' events may be a warning that better intelligence is needed to navigate disruption. Increased visibility (providing increased lead time for corrective action) is likely to become the most critical capability for the industry in the coming year. Increasing visibility (including costs and prices) depends on how a company is experiencing disruption; digital technologies could be essential enablers for increasing chain visibility. For example, companies experiencing a surge in demand should ensure visibility across their supply network as they ramp up production. As supply shortages could derail production flow, chemical producers in this scenario could consider multi-sourcing strategies. Moreover, companies experiencing shifts in demand may want to increase visibility into operations to help them focus on cost-cutting opportunities. They can use the visibility gained to create flexibility across their battery limits, better decreasing costs and suppressed demand in order to more quickly respond to an eventual uptick in demand.

Until the March issue of Engineering Practice, I will leave you with these possibilities to ponder. Hopefully our industry's ingenuity, creativity, and technical expertise will continue to provide efficient solutions demanded by our clients and suppliers, as we have faithfully demonstrated for well over a century.

REFERENCES

American Chemistry Council, "Mid-Year Outlook: U.S. Chemical Industry Rebounds as Global Recovery Continues," accessed June 24, 2021; US Energy Information Administration (EIA), "Short-Term Energy Outlook," accessed September 8, 2021.

Bloomberg Intelligence, "Methanol Contract U.S. Gulf Coast Non-Dis- counted Ref. Price," accessed September 8, 2021.

Bloomberg Intelligence, "Fed Minutes Flag Chance of Earlier Hikes, Balance Sheet Rundown," January 5, 2022.

Deloitte analysis based on data from American Chemistry Council, "Mid-Year Outlook."

Deloitte analysis based on data from DuPont de Nemours, Inc., Q2 2021 Earnings Call, 2021, p.6.; Eastman Chemical Co, Q2 2021 Earnings Call, 2021, p.3; Dow Chemical Company, Q2 2021 Earnings Call, 2021, p.3; and LyondellBasell, Q2 2021 Earnings Release, 2021, p.1.

Deloitte, 2022 Industrial Chemical Outlook, Chronis, Hardin, and Yankovitz

Deloitte, The future of work in chemicals.
EIA, US Natural Gas Production Will Establish a new record High in 2022, December 16, 2021

Mark Nicolich, CEO, Braskem America, Inc., Featured Speaker, December 2021 Meeting, Northeastern Chemical Association, New York City, NY

Chemical Engineering for the Energy Transition

Rupsha Bhattacharyya

INTRODUCTION AND CONTEXT

The energy infrastructure that supports human life and ensures its well-being today reflects the efforts, developments, and initiatives of all domains of science, engineering and technology, carried out over centuries, gaining particular momentum after the Industrial Revolution in the nineteenth century. Abundance of fossil fuels and the development of processes to recover and use them cheaply have been the bed rock of this phase of growth and advances in human well-being. When it comes to primary energy forms and their conversion to secondary energy sources, an undeniably important role has always been that of the chemical sciences, embodied by the work and tireless efforts of chemists and chemical engineers. Be it the identification of the grade of coal and its calorific value through proximate and ultimate analysis, extraction of crude oil, natural gas and its refining and processing to yield various kinds of hydrocarbon gases, liquids and solids for use as fuel and industrial feed stock, the bulk synthesis of enormous variety of organic and petrochemicals for both energy and non-energy applications – chemical engineering has been deeply involved with the energy sector every step of the way. This also means that the chemical industry has contributed directly or indirectly to huge amounts of carbon emissions and other negative environmental externalities.

But this fossil fueled growth and prosperity has exacted a heavy price. The perils of today's energy sources and utilization patterns are unknown to nobody. Today's energy system is heavily dependent on fossil fuels. Use of fossil fuels and the resultant emissions of greenhouse gases (GHGs) such as CO2, oxides of sulphur and nitrogen, particulate matter are largely responsible for human induced global warming and climate change. Increasing levels of atmospheric carbon are causing acidification of the oceans and loss of marine biodiversity. This in turn has begun to wreak havoc on the planet, causing rapid temperature rise, severe weather events, and abnormal changes in the water cycle – all of which has had devastating effects on the lives of millions, if not billions. No wonder climate change has been described as the defining crisis of our time. This crisis also puts in serious jeopardy the attainment of sustainable and inclusive development for mankind, as embodied in the United Nations Sustainable Development Goals. The clamour for deep decarbonization of industry and widespread adoption of non-fossil energy has never been louder and more widespread than now.

Does it mean then, that the enormous body of work and technical knowledge developed via the chemical sciences and the industrial infrastructure and assets built up around fossil fuels in course of a century or more will become redundant overnight? Does it mean that chemical engineering discipline will become irrelevant? No, absolutely not. In fact, the chemical engineers will continue to play an increasingly important role in decarbonizing much of its existing operations and in building up an environmentally friendly new energy system, one that would be sustainable and inclusive and would ensure a just and holistic transition that benefits everyone on our planet Earth. Some insights are briefly provided in the following paragraphs.

CHEMICAL ENGINEERING PRINCIPLES FOR NEW ENERGY MATERIALS AND PROCESS SYSTEMS

The academic basis of chemical engineering is built on fundamental themes such as thermodynamics, momentum, heat and mass transport, chemical reactions and separation sciences, materials sciences and material processing technologies, process dynamics and control. Energy technologies of the future may look very different from what it is like today, but to go from natural resources to final infrastructure and delivery of energy services, the importance of understanding and applying these fundamental lessons will not diminish.

They will have to be augmented with new lessons and learning. Crucial components of the energy transition are vastly enhanced use of renewable energy and the materials and technologies to harness them, grid scale energy storage systems especially electrochemical systems such as grid scale batteries, new low carbon energy vectors and industrial feed stock such as hydrogen and its derivatives (ammonia, methanol, liquid organic hydrogen carriers, etc). Overall, much greater dependence on extractive industries is easily foreseen in course of the energy transition.

Thus, instead of applying the fundamental chemical engineering principles to extraction, transportation and purification of crude oil and natural gas, we would apply them to extraction and recovery of various kinds of mineral resources such as lithium from brines and hard rock deposits, rare earth elements from beach sands, noble metals, cobalt, nickel, vanadium, titanium, phosphorus, manganese and others from their ores and so on. The use of these materials will increase by orders of magnitude for the energy transition. Much greater amounts of aluminum, copper will also be required to support widespread electrification across sectors. Novel separation technologies (based on the vast amount of knowledge of conventional unit operations like solvent extraction, distillation, membrane separations, adsorptive separations) including new solvents (e.g., supercritical CO2, ionic liquids, green solvents) and contactors would have to be developed and made available for extracting and concentration of the useful minerals that will support energy transition. New materials of construction including composite materials such as fiber reinforced polymers and cermets would be required which would be compatible with these new service fluids and operating conditions of the process equipment. New materials with widespread applications in the new energy systems, such as graphite, carbon nanotubes and graphene and various kinds of materials for heat energy related applications will have to be produced in bulk quantities for applications in electrochemistry, electrochemical energy storage and thermal energy storage.

Solar photovoltaics will require scaling up and cost reduction of facilities which convert silica in sand to very high purity silicon (>99.999 % pure) through a number of unit operations performed in fluidized bed reactors, chemical vapor deposition systems and high purity distillation operations. The solar grade silicon is used to produce the solar cells. Alongside silicon, huge quantities of high-quality glasses are required to be manufactured and glass making is a process that's not only energy intensive but currently almost entirely fossil energy based. Thus, renewable energy will have to be used to scale up renewables deployment to have a really low life cycle carbon emission associated with the process.

Green hydrogen, i.e., hydrogen produced by water splitting using low carbon energy sources such as renewable and nuclear energy is an extremely promising route to decarbonizing sectors such as heavy transport, chemicals and green fertilizers synthesis (via ammonia production). These will require massive increase in the electrolyser development industry along with ancillary components such as hydrogen compressors, pipelines and storage vessels, instruments, detectors and safety systems. The accrued experience with the hydrocarbon industry can be harnessed very fruitfully for transitioning to the hydrogen economy.

While new process technology development is clearly required, many of the current industries will have to keep working but with changes in their material and energy inputs. For example, ammonia can be produced using green hydrogen instead of current practice of using hydrogen from steam methane reformers. Process optimization and intensification will be the norm for the existing industries; it is in fact an embodiment of sustainable design where the maximal utilization of an input resource (material, energy) to get to a value-added product must be carried out, while ensuring safety of the operating personnel and producing minimal wastes and hence negative environmental impacts while running the process. This process engineering philosophy will be well supported by advances in digital tools and technologies for the process industry. Computational material modelling and material design tools, process simulation software, advanced process control, energy integration and conservation measures for the process industries, the power of data analytics and artificial intelligence for monitoring the operation of process plants, fault detection and diagnosis, reliability analysis and condition monitoring of plants, development of digital twins of process plants for operator training and predictive maintenance of these facilities only broadly represent the possibilities. This also means that chemical engineers will continuously need to upskill themselves to be able to fully utilize these advanced technologies, which are progressively becoming ubiquitous in this sector.

CARBON CAPTURE TECHNOLOGIES

Carbon dioxide removal (CDR) technologies are expected to have a big role in the energy transition, even though the techno-economic case for it is not very well established today. Technologies for the removal of CO2 already present in the atmosphere via direct air capture (DAC) are typically based on CO2 scrubbing (by alkaline solutions such as amines) principles well known in the petroleum industries, for example in the removal of CO2 and recovery of hydrogen in steam methane reformers. Solid state adsorbers such as hydrotalcites are also possible which form solid carbonate like species after CO2 uptake. While these technologies are transferable to thermal power plants or cement industries for flue gas CO2 capture, the real challenge is in working with very low CO2 levels in atmospheric air (currently at about 420 ppm) rather than in flue gases (~40-50% by volume). This is what makes these DAC processes extremely capital and energy intensive and developing cost is an open challenge for chemical engineers. Another possibility of integrating carbon capture into processes is by way of chemical looping technologies, which may work very well in industries such as cement where carbon emissions occur from the fundamental process chemistry, alongside those from the energy use in these sectors.

Apart from carbon capture, innovation is truly the need of the hour for ensuring that the captured carbon can be used to develop value added products. This is one of the important ways that will make carbon capture processes techno-commercially viable, and it is something that needs to be done on a very large scale.

WASTE MANAGEMENT TECHNOLOGIES AND CIRCULAR ECONOMY

The current nature of consumption of resources, products and services is once-through or open-loop – extract, use and discard are the operating words along which most industries including the chemical process industries work. This unsustainable and wasteful attitude needs to shift a circular resource utilization pattern, giving rise to what is broadly known as the ‘circular economy’. This philosophy strongly advocates for reducing consumption, repairing, re-using and recycling materials and energy as far as is techno-economically feasible. As newer technologies are developed using a different set of raw materials that what we are currently used to, material processing technologies, process plants making use of these materials and end-of-life management protocols will all need to evolve alongside it.

In the context of the energy transition, crucially important is the development of process technologies for battery, degraded solar photovoltaic panels and used wind turbine blades recycling, regenerating spent solvents and solids for carbon capture, plastics recycling technology, cement/clinker recycling and domestic and industrial water purification and recycling technologies. Entire infrastructure planning, development and deployment have to be based on sustainable and circular principles embodying the waste-to-wealth or waste valorisation philosophy.

Integration of nature based solutions into chemical process infrastructure (i.e., the development of grey-green infrastructure) needs more attention. For example, the use of specific plant species for phyto-remediation of contaminated soils and water could complement traditional and new waste water purification technologies at small and large scales.

CHALLENGES AND THE WAY FORWARD

It is obvious that the entire chemical engineering profession has to rapidly evolve and reinvent itself to meet these upcoming challenges and phase of disruptive progress. It has to begin from the education and training phase itself. The undergraduate and post graduate university curricula do not change as fast as industrial practices, norms and legislations get altered, thus, as students one hardly has the opportunity to prepare for actual work place roles. More case study and internship-based education should complement acquiring traditional knowledge and fundamental principles of the chemical engineering discipline.

Energy system modeling, life cycle analysis, green house gas accounting, carbon budgeting, energy economics and finance as well as elements of project management must all be taught at universities to prepare the work force of the future. Including use of digital tools for the process industry in academic curricula is essential for chemical engineers to play their crucial roles in the future energy sector and industry. It is also important to highlight the interdisciplinary nature of the energy transition challenge, thus an understanding of related engineering disciplines such as metallurgy, materials sciences, mechanical engineering, manufacturing technologies and mining engineering must also be built into chemical engineering education. University and other academic research programs need to be better tuned to industrial demands and they should be executed in mission mode. Industry too has a role in providing support for such programs which would ultimately solve their challenging problems. Synergy and cooperation are the needs of the hour.

The chemical process industry has made enormous contributions to human welfare so far, and without a doubt, this will continue through the energy transition. All of us involved in this profession have to rise up to this challenge to make meaningful contributions to this journey.

REFERENCES

The 2030 Decarbonization Challenge – The Path to the Future of Energy: Chemicals. Deloitte, 2020.

The big choices for oil and gas in navigating the energy transition. McKinsey, 2021.

How to build a more climate-friendly chemical industry. World Economic Forum, 2020.

Chemical Engineering role in the use of renewable energy and alternative carbon sources in chemical production. G Centi, G Iaquaniello, S Perathoner, BMC Chemical Engineering, 1:5, 2019.

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An Overview of Pipeline Costing & Economics

Jayanthi Vijay Sarathy

In any pipeline project, an economic analysis has to be performed to ensure the project is a viable investment. The major capital components of a pipeline system consists of the pipeline, Booster station, ancillary machinery such as mainline valve stations, meter stations, pressure regulation stations, SCADA & Telecommunications. The project costs would additionally consist of environmental costs & permits, Right of Way (ROW) acquisitions, Engineering & Construction management to name a few.

The following article explains some of the key factors in performing a pipeline economic analysis using the method of Weight Average Capital Cost (WACC) to estimate gas tariffs, project worth in terms of Net Present Value (NPV), Internal Rate of Return (IRR), Profit to Investment Ratio (PIR) and payback period. The cost of equity is estimated using the Capital Asset Pricing Model (CAPM).

COST ESTIMATION ASPECTS & ASSUMPTIONS

1. The weight average capital cost (WACC) comprises of a firm's cost of capital in which each category of capital is proportionately weighed. All sources of capital, including common stock & preferred stock, bonds & any other long-term debt, are included in a WACC calculation. For the following module, the cost of equity [ke] & cost of debt [kd], (Sum of which is the cost of capital), expressed as a percentage for the pipeline proposal are included to estimate the WACC.
2. The revenue generated comes from the gas tariffs delivered to the power station. Capital is necessary to initiate a project for which some of the capital can come from private equity (where investors lend for the project) and debt, where the money is borrowed from an institution (e.g., bank).
3. In order to set the minimum gas tariff & pay for the equity, lenders/investors or financial institutions impose a hurdle rate or minimum acceptable rate of return (MARR) (i.e., rate at which NPV = 0 at IRR = WACC) to offset the cost of the investment (i.e., break-even).
4. Generally the hurdle rate is equal to the company's cost of capital. In the current module cost of capital is taken as a weighted average, the WACC becomes the hurdle rate i.e., at IRR = WACC.
5. The internal rate of return (IRR) is the expected annual cash flows, (expressed as a percentage) that the investment can be expected to produce over and above the hurdle rate. Therefore for a project's acceptability, IRR > MARR.
6. The prime lending rate [r] is the rate imposed by the bank to service the project. The average risk free interest rate [rf] represents the rate of return where there is no risk of defaulting/loss. This can be the interest offered as government/treasury bonds. In actual practice, inflation is deducted from the interest rate to get a real risk free interest rate.
7. The expected market portfolio return [rm] is a measure of the expected returns on the investments made. This is however correlated to market risks that is quantified by the beta [b] ratio and is useful in determining the volatility when arriving at the equity costs using CAPM. b<1 indicates that the security issued by the lender providing the equity, is less volatile (i.e., less risky) than the systematic market risks. Whereas a b>1 indicates that the security issued by the lender is more volatile (i.e., more risky) than the systematic market risks. For this module, b of 0.9 is taken assuming the venture is by an established firm with a history of low volatility. In actual practice, low b values need not necessarily show lower volatility if the security's value changes minimally in the short run but indicates a downward trend in the long run. Therefore the estimation of b requires historical data to arrive at a suitable value.
8. The term [rm – rf] is expressed as the equity risk premium which indicates the additional compensation for the risk taken by the investors/institutions towards providing equity. When multiplied by the b value, it accounts for the responsiveness of

the venture to market changes. With higher b value, the gains/return on investment risks is higher. Similarly, with lower b value, the return on the investment risks is lower.

8. Assets over time depreciate in value due to wear and tear. Two methods used to depreciate an asset are the straight line depreciation method and declining balance depreciation method. In the straight line method, the depreciation rate is considered to be annually uniform, where the total life of the asset is divided by the operational years. Whereas in declining balance method, the asset is considered to lose more value in the early years and less in later years of operation. For this module, a straight line depreciation method is applied for the stipulated period of 25 years. It must be noted that the salvage value at the end of the operational period becomes zero. However the asset is considered to have a salvage value which is estimated at a depreciation rate of 10% for this exercise.
9. Depreciation also relates to taxable income, since the annual depreciation cost is neglected from the annual gross tax on revenue. Annual Cash flow is estimated by addition of the annual tax benefit from depreciation [ATBD] to the net annual revenue [NAR] before subtracting from the annual gross tax on revenue [AGTR] (i.e., Net Annual revenue – Annual Gross Tax on Revenue + Annual Tax Benefit from Depreciation). It may be noted that when companies tend to depreciate the asset value more, the tax payable decreases and increases the annual savings from depreciation.
10. The profit to Investment ratio (PIR) is a measure of investment efficiency. It is estimated as the ratio of NPV to the discounted capital costs (Disc. CAPEX). For a project to be viable, the PIR > 1.
11. The key differences between using NPV, IRR & PIR is that NPV and PIR measure a project value, while PIR allows screening a project against a company benchmark. For project to be viable, NPV>0, PIR>1 and IRR>WACC. The IRR value indicates the returns possible above which break even cannot be achieved and NPV<0.

CAPEX ESTIMATION

To estimate CAPEX, the pipeline and booster station tonnage and installed power is required to be estimated. The pipeline tonnage is estimated as,

$$Pipe\ Tonnage\ [tonnes] = \frac{\pi \times L \times \rho}{4 \times 1000}[OD^2 - ID^2]$$

(1)

$$PMC\ [\$] = Cost_{PMC} \times Pipe\ Tonnage$$

(2)

$$PCW\ [\$] = Cost_{PCW} \times Pipe\ Tonnage$$

(3)

$$LC\ [\$] = Cost_{LC} \times Pipe\ Tonnage$$

(4)

$$PIC\ [\$] = Cost_{PIC} \times Pipe\ Tonnage$$

(5)

The total pipeline [PC] cost becomes,

$$PC\ [\$] = PMC + PCW + LC + PIC$$

(6)

The Station Compressor cost is estimated as,

$$SC[\$] = SMC + SLC + SMiC + SLaC$$

(7)

Total Cost of Booster Station based on Comparison factor [C_f] for various GT/CC,

$$Total\ SC[\$] = SC \times C_f$$

(8)

The other machinery costs [OMC] including mainline valve stations [MVS], meter stations [MS] & pressure regulator stations [PRS] can be computed by accounting them as a percentage of the pipeline costs as,

$$MVS[\$] = PC \times \%factor$$

(9)

$$MS[\$] = PC \times \%factor$$

(10)

$$PRS[\$] = PC \times \%factor$$

(11)

For SCADA & Telecommunications, the cost is taken as a percentage of pipeline & station,

$$S\&T[\$] = [PC + SC] \times \%factor$$

(12)

The total machinery costs are calculated as,

$$OMC[\$] = MVS + MS + PRS + S\&T$$

(13)

The project costs involving environmental costs & permits [ECP], Right of Way Acquisition [ROWA], Engineering & Construction Management [ECM], Contingency [Con] is estimated similarly as,

$$\mathrm{ECP}[\$] = [PC + SC] \times \%factor$$

(14)

$$\mathrm{ROWA}[\$] = [PC + SC] \times \%factor$$

(15)

$$\mathrm{ECM}[\$] = [PC + SC] \times \%factor$$

(16)

$$\mathrm{Con}[\$] = [PC + SC] \times \%factor$$

(17)

The total project costs are calculated as,

$$Project[\$] = ECP + ROWA + ECM + Con$$

(18)

In addition to the pipeline costs, total station costs, other machinery costs & project costs, a working capital [WC] & AFUDC is included. AFUDC which stands for "*Allowance for funds used during construction*" represents costs associated with financing the project during various stages of construction. The total CAPEX is now calculated as,

$$CAPEX[\$] = PC + Total\ SC + OMC + Project + WC + AFUDC$$

(19)

OPEX ESTIMATION

To estimate annual OPEX, the basis is made on the annual salaries payable [S] and multiplying the annual salary by a %factor in this module.

$$\mathrm{Salary}[\$] = S$$

(20)

$$\mathrm{Payroll}[\$] = S \times \%factor$$

(21)

$$\mathrm{Administrative\ Expenses}[\$] = S \times \%factor$$

(22)

$$\mathrm{Vehicle\ Expenses}[\$] = S \times \%factor$$

(23)

$$\mathrm{Office\ Expenses}[\$] = S \times \%factor$$

(24)

$$\mathrm{Misc\ Materials\ \&\ Tools}[\$] = S \times \%factor$$

(25)

Similarly the related costs for compressor station maintenance is estimated as,

$$\mathrm{Consumables}[\$] = S \times \%factor$$

(26)

$$\mathrm{Periodic\ Maintenance}[\$] = S \times \%factor$$

(27)

$$\mathrm{ROW\ Payments}[\$] = S \times \%factor$$

(28)

$$\mathrm{Utilities}[\$] = S \times \%factor$$

(29)

$$\mathrm{Gas\ Control}[\$] = S \times \%factor$$

(30)

$$\mathrm{SCADA\ Maintenance}[\$] = S \times \%factor$$

(31)

$$\mathrm{Corrosion\ Inspection}[\$] = S \times \%factor$$

(32)

$$\mathrm{Cathodic\ Protect.\ Survey}[\$] = S \times \%factor$$

(33)

The Total O&M [OPEX] is computed as,

$$OPEX[\$] = S + Eq.23 + \cdots + Eq.36$$

(34)

CAPEX & OPEX COST FACTORS

Mainline Valve Stations

Block Valves on mainlines are installed to isolate pipeline sections for safety & maintenance such as blow down. In case of any pipeline rupture, block valves close to isolate the pipeline section and shut-off flow. For mainline valve stations a lump sum figure may be obtained from a construction contractor based on the size.

Meter & Pressure Regulating Stations

Meter Stations are installed for the purpose of measuring gas flow rates. Similar to mainline valve stations, a lump sum can be quoted for a given size that includes meters, valves, fittings, instrumentation & controls. Pressure regulation stations are installed for the purpose of reducing gas delivery pressure prior to delivering to the buyer. The price can be quoted as a lump sum figure.

SCADA & Telecommunications

SCADA allows pipeline process conditions data to be transmitted by electronic signals from remote control units installed on valves & meters. The signal transmission can be either by telephone lines, microwave or satellite communications. The pricing of these provisions can be expressed as a percentage of the total project cost, typically 2% to 5%.

Environmental Costs & Permits

Costs associated with environment and permits pertain to the modification of pipeline and booster station to prevent pollution. The costs also include compensation for acquisition of land to compensate for the areas disturbed due to pipeline construction.

Permitting costs also include environmental study, environmental impact report and permits for railroad, stream & river crossings. These costs for a gas pipeline project can vary from 10% to 15% of the total project costs.

Right of Way (ROW) Acquisitions

Right of Way Acquisitions can be applied from private and government entities. The fee may be lump sum based at the time of acquisition, with additional fees paid on an annual basis. The initial cost for acquiring the ROW will be included in the capital cost, whereas annual ROW lease would be included as an O&M cost. For most gas pipelines, the initial ROW would be typically in the range of 6% to 10%.

Engineering & Construction Management

E&C management costs are associated with preparing design documents & drawings at both front end & detailed design phases. Details would also include specifications, operating manuals, purchase documents & equipment acquisition. Construction management costs would also include costs for field personal, transportation, rentals & all associated costs to manage the pipeline construction efforts. Engineering & Construction costs can typically vary from 15% to 20% of total project costs.

Other Project Costs

As the pipeline project progresses, it is expected that there could always be some unforeseen issues during project execution. To account for these uncertainties, contingencies & allowances for funds used during construction (AFDU) are also allotted. These costs can vary from 15% to 20% of the total project costs.

Salvage & Depreciation Costs

Pipeline infrastructures experience wear and tear thereby rendering depreciation to the project value. The salvage value [SV] based on the depreciation rate over the operational life of the project is computed as,

$$SV[\$] = [PC + SC + OMC + ROWA] \times [1 - d_r]^n$$

(35)

Where, n = Project Operational Life [Years]

The pipeline salvage deduction [*SD*] is computed as the difference between the Salvage value [SV] & the initial investment cost i.e., [PC+SC+OMC+ROWA]. Therefore the salvage costs is as follows,

$$\mathrm{SD}[\$] = [PC + SC + OMC + ROWA] - SV$$

(36)

Annual depreciation of the Asset [*ADA*]

$$\mathrm{ADA}[\$] = SD/n$$

(37)

Annual Tax Benefit from Depreciation [*ATBD*]

$$\mathrm{ATBD}[\$] = ADA \times T_{sale}$$

(38)

Cost of Capital

Based on the cost of capital structure of the project (that involves both debt & equity), the appraisal is made using the Weight average Cost of Capital [WACC] from which all the cash flows generated will be discounted to arrive at the Net Present Value [NPV]. WACC can be expressed as,

$$\mathrm{WACC}[\%] = \frac{E}{C}k_e + \frac{D}{C}k_d$$

(39)

Where, $C = E + D$

The cost of debt [k_d] is taken as after-tax cost of debt and the cost of equity [k_e] is based on CAPM. The cost of debt is estimated as

$$k_d[\%] = r \times [1 - T_C]$$

(40)

The cost of equity is estimated as,

$$k_e[\%] = r_f + \left[\beta \times \left[r_m - r_f\right]\right]$$

(41)

The loan annual amortization cost [AAC] is computed as,

$$\mathrm{AAC}[\$] = \frac{[D \times CAPEX] \times r}{1 - \left[\frac{1}{1+r}\right]^n}$$

(42)

where, n=Loan period during project life [Yrs]

The Annual Revenue on Equity Capital [AREC] adjusted to effective corporate tax rate [T_C] is,

$$\mathrm{AREC}[\$] = \frac{E \times k_e}{1 - T_c}$$

(43)

Gas Tariff & Economic Viability

The minimum gas tariff required for economic viability is estimated based on the total cost of capital for an $NPV = 0$ at $IRR = WACC$ and yearly gas production. The pipeline operational availability is 8,040 hours/year. Therefore, the yearly gas production is,

$$\mathrm{Gas\ Rate}[Mscf] = \frac{1000 \times Q_{MMscfd}}{24} \times Available[hr]$$

(44)

The Annual cost of service [ACOS] is computed as,

$$\mathrm{ACOS}[\$] = OPEX + AAC + AREC + ADA$$

(45)

The Gas Transportation Tariff is estimated as,

$$\mathrm{Tariff}_{Trans}[\$/Mscf] = \frac{ACOS[\$]}{Gas\ Rate[MScf]}$$

(46)

Setting an initial guess sales tariff [Min_{Tariff}], the Net Annual Cash Flow [CF] is calculated for the condition NPV=0 and IRR=WACC. The Annual Revenue [AR] without tax is found as,

$$\mathrm{AR}[\$] = Min_{Tariff} \times Gas\ Rate[Mscf]$$

(47)

The Net Annual Revenue [NAR] becomes,

$$\mathrm{NAR}[\$] = AR - ACOS$$

(48)

The Annual Gross Tax on Revenue [AGTR] is,

$$\mathrm{AGTR}[\$] = NAR \times T_{sales}$$

(49)

The Tax Payable is calculated as,

$$\mathrm{Tax\ Payable}[\$] = \mathrm{AGTR} - \mathrm{ADA}$$

(50)

The Annual Cash Flow [ACF] is calculated as,

$$\mathrm{CF}[\$] = \mathrm{NAR} - \mathrm{AGTR} + \mathrm{ATBD}$$

(51)

The Net Present Value (NPV) is calculated as,

$$\mathrm{NPV}[\$] = \left\{\sum_{n=1}^{n=n}[CF]_n[1+i]^{-n}\right\} - CAPEX$$

(52)

Where,
n = Operational Life [years]
i = Internal Rate of Return (IRR) at WACC

The Profit to Investment Ratio (PIR) is then calculated based on the discounted cash flow [DCF] as follows,

$$\mathrm{PIR}[-] = \frac{Cash\ Outflow}{Cash\ Inflow} = \frac{[CF]_n[1+i]^{-n}}{NPV}$$

(53)

In the above PIR expression, after the discount factor $[1+i]^{-n}$ is applied to each year, the negative cash flow, i.e., cash outflow of each year is summed up and divided by the NPV value. At NPV=0 and WACC=IRR, PIR=0.

Payback Period

With detailed NPV calculations, the payback period for \$/Mscf can be computed as,

$$\text{Payback Year} = [\mathit{Cumulative\ Annual\ CF} = \mathit{Total\ CAPEX}] \quad (54)$$

REFERENCES

"Gas Pipeline Hydraulics", E. Sashi Menon, 2005, Taylor & Francis

"Pipeline Compressor Station Construction Cost Analysis", Y. Zhao, Z. Rui, Int J. Oil, Gas and Technology, Vol. 8No. 1, 2014 (https://www.researchgate.net/publication/275590240)

"Economic Analysis of Gas Pipeline Project in Nigeria", Ahmed Adamu, Journal of Economics and Sustainable Development, Vol.8, No.2, 2017

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Zero Sulfur in Refineries

Joseph C. Gentry

INTRODUCTION

The refining industry is pressured. Environmental regulations dominate investment budgets, and often take the top priority in operating expenditures. Mandates on compliance rule over all else and must be adhered as a license to operate regardless of the logic or cost/ benefit analysis. The Gas Processing Industry faces some of the same challenges and has found different ways to manage H2S that may be useful in the downstream refinery industry.

REDUNDANCY IN SRU

Refiners typically have a sulfur recovery unit (SRU) as catch-all for sulfur from the various processing units. Organic sulfur is removed from the oil in hydrodesulfurization units, where H2S is generated; then removed in an off gas. This gas is treated in an amine contacting unit to remove the H2S. The acid gas is then stripped and routed to the SRU, which uses the Claus technology to produce elemental sulfur. The SRU has long been a mainstay of the refinery operation; but now it is even more so required. New regulations require 100% redundancy of critical emissions control systems, including the sulfur recovery and tail gas treating (TGT) units. The goal is to completely eliminate flaring of the H2S gas during outages of the SRU and TGT, even during start up and shutdown1.

In times past, it was sufficient to have surplus capacity in the sulfur plant or multiple SRUs, in case one unit had to be taken offline. Now, there is need to coordinate the start up and shutdown sequences of the sulfur plant and the tail gas treating unit, which involve different types of unit operations and process dynamics. This requires a cumbersome coordination of process units just to make a normal startup or shutdown, which would otherwise require to temporarily flare a small amount of H2S gas.

One interesting option is to use a solid scavenger system, such as SweetTreatTM to remove the H2S during plant start up/shutdown, or during process upsets. This de-couples the main process units from emergency operation. The SweetTreatTM process is a passive, always-ready system to ensure that the acid gas is treated in every situation. The solid adsorbent converts 100% of the H2S into an iron-sulfide, which is suitable for disposal as non-hazardous waste. SweetTreatTM is a fixed vessel operation that is activated simply by opening valves to the unit. When the adsorbent is spent, it is dumped and re-loaded for the next incident. This is far better than keeping a full-size redundant amine and incinerator within the TGT unit ready on hot standby, or a caustic scrubber that requires disposal of spent caustic.

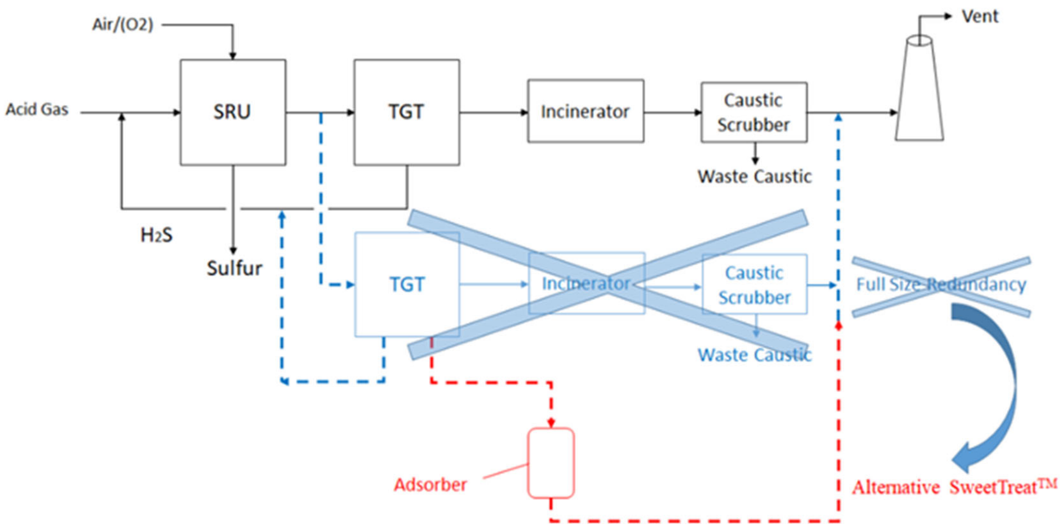


Fig. 1: Redundant Tail Gas Treatment

ENERGY EFFICIENCY BY BURNING SULFUR-FREE GAS

The SRU with TGT units are designed for certain sulfur recovery, to meet an overall site allowance of SO_2 emissions. The refinery-generated fuel gas contains some amount of H_2S that converts mainly to SO_2; but a small portion is further oxidized to SO_3, which condenses with water to form H_2SO_4. To avoid the corrosive effects of sulfuric acid on the process equipment, the furnace is operated to keep the flue gas at a temperature level above the acid-gas dew point. This limits the furnace efficiency, which would otherwise maximize the extraction of useful heat and result in a lower flue gas temperature. If the refinery fuel gas had zero H_2S, then the acid-gas dew point condition would be substantially lower, giving flexibility to optimize the furnace operation.

Let’s look at an example:

- Refinery fuel gas H2S content: 100 wppm (as set by recovery in the amine absorbers)
- Resultant flue gas, Sox: 180 wppm
- Resultant flue gas, SO3: 5.6 wppm (assumes 3% is further oxidized)
- Acid-Gas Dew Point: 130 DegC

- To guard against sulfuric acid being formed upon condensing the SO3 with water, the stack outlet temperature would normally be kept at 150 -160 DegC or higher. This precludes use of low-level heat recovery into combustion air or boiler feed water2,3. However, if the acid-gas dew point is not governing, the stack gases can be further cooled to the H2O dew point or lower (subject to reasonable temperature approach), thereby increasing the furnace efficiency4.

- Refinery fuel gas H2S content: 0.1 ppm (with H2S removal by SweetTreatTM)
- Resultant flue gas, Sox: 0.2 ppm
- Resultant flue gas, SO3: 0
- Acid-Gas Dew Point: < 100 DegC

By availing of the full air preheating potential, there may be energy savings of 10-15% in the furnace. Higher combustion air temperature also makes it easier to control the level of excess O2, to reduce this to the minimum level. The energy savings comes with reduction in CO2 emissions from the refinery process heating, which can be useful to add to the refinery ESG scorecard7.

Even if the normal furnace operation maintains the stack temperature above the acid gas dew point, there can be times when the temperature may drop due to sudden weather change or turndown operation, bringing the system into the severe corrosion regime. Thus, it is important to avoid having H2S in the fuel system in any case. Sulfur is not our friend.

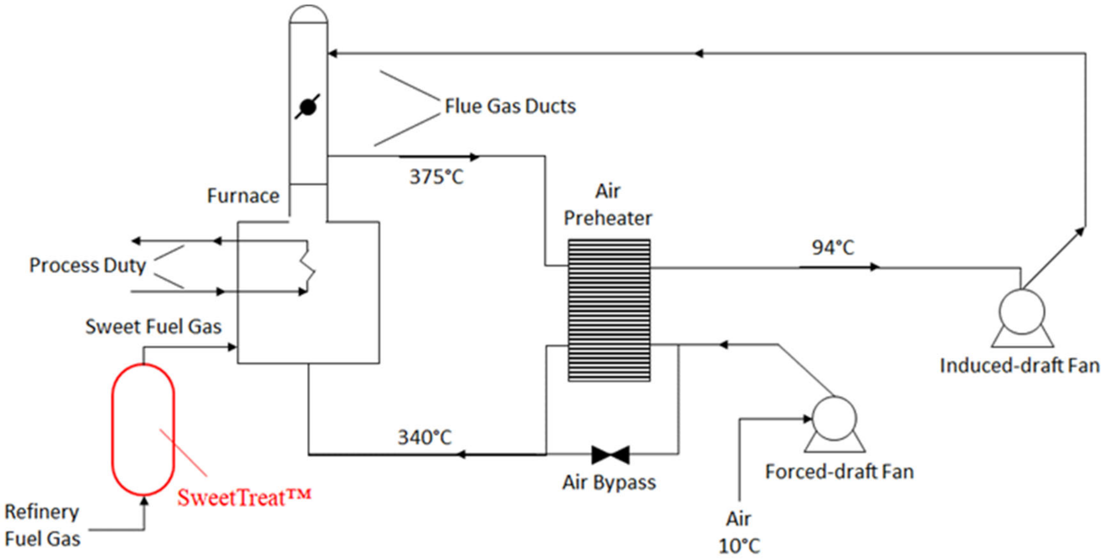


Fig. 2: Air Preheating System[5]

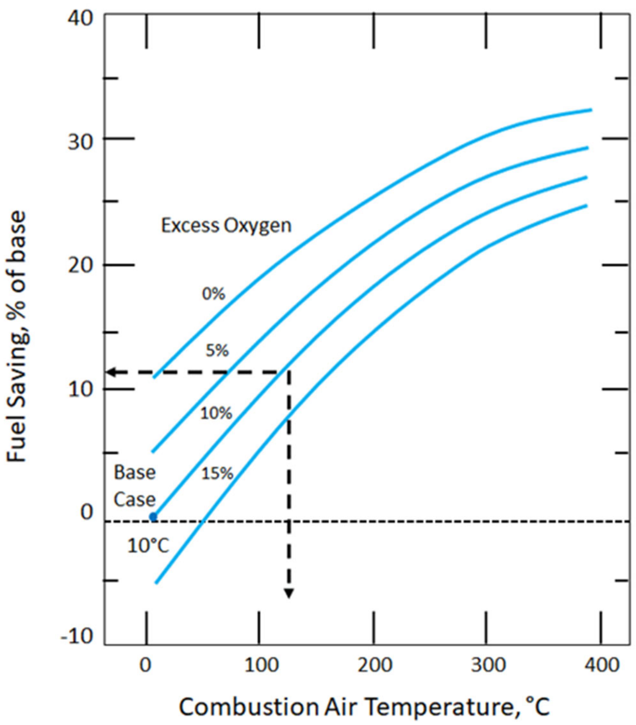


Fig. 3. Effect of Combustion Air Temperature on Furnace Fuel Saving[6]

Another example:

Many refinery gas systems operate at low pressure. In order to get a satisfactory removal of H2S, they may use a strong, unhindered amine such as MEA. There are two issues with MEA which make it less efficient than hindered amines such as MDEA: 1) MEA holds the H2S more strongly and requires more energy for regeneration. The MEA also absorbs most of the CO2, which requires regeneration cost. 2) CO2 in the acid gas would be recycled back to the Claus unit and TGT units, which incurs operating costs that would not be present if CO2 were excluded.

If the SweetTreatTM system were applied to the fuel gas, the refiner could relax the requirement for H2S removal in the amine system and use a lower-cost system for bulk H2S removal. SweetTreatTM removes the final quantity of H2S in the fuel gas from the relaxed operation; nearly all the CO2 can be rejected into the fuel gas away from the Claus and TGT units; and energy required to operate these units is reduced.

SULFUR SCAVENGING IN HYDROGEN SYSTEMS

The majority source of H2S in the refinery is from hydrodesulfurization units, which convert the organic sulfur into hydrogen sulfide. These units typically operate with hydrogen recycle gas to maintain the hydrogen partial pressure in the reactor. Some of the gas is purged to the fuel system, to maintain purity of the hydrogen and remove H2S. Depending on the quantity of sulfur in the liquid feed, there may be need to operate an amine absorber on the hydrogen recycle gas stream to remove the bulk of H2S. Even so, the amine will not capture all the hydrogen sulfide, which returns to the reactor via the recycle hydrogen. H2S is generally a catalyst poison and should be removed as low as possible. The SweetTreatTM system can remove all the H2S from the recycle hydrogen to improve the cycle length and operability of the hydrotreater.

One example is in the area of FCC gasoline selective hydrotreating. Here the objective is to remove the organic sulfur while avoiding saturating the olefins. The Δ RON between an olefin and its corresponding paraffin is about negative 50! If H2S remains in the recycle hydrogen, it will recombine with olefins to create new mercaptan sulfur species, defeating the purpose of the hydrotreatment and requiring higher severity. Additionally, H2S poisons the catalyst. Thus, it is important to eliminate all H2S from the hydrogen gas.

We can apply the SweetTreatTM process within the hydrogen recycle loop to greatly improve the functional efficiency of the selective gasoline hydrotreatment system. This fixed bed adsorption system is installed on the recycle loop itself, or after the amine absorption unit if the sulfur content is extremely high. In either case the octane retention is improved, hydrogen consumption lowered, and operating costs outside the HDS unit are reduced.

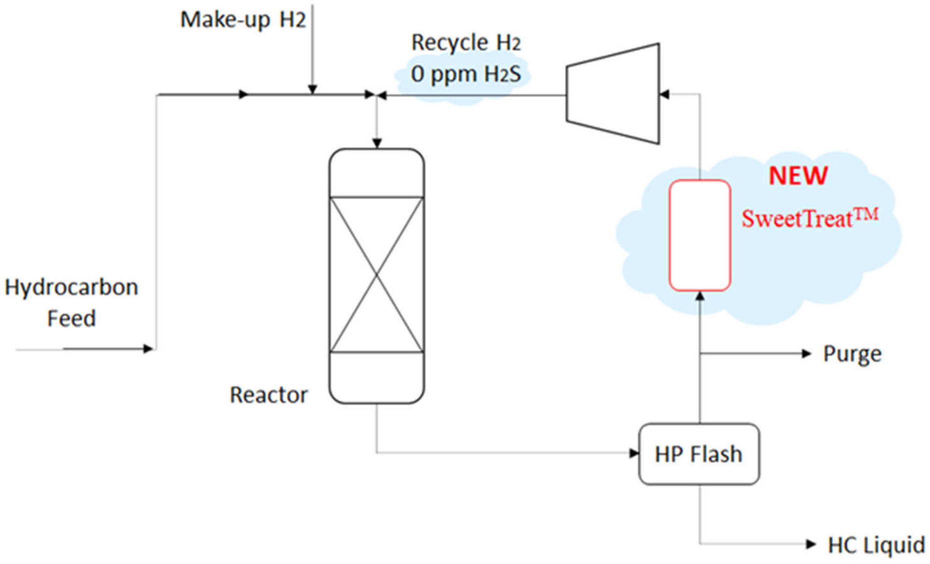


Fig. 4: Improvement in Selective Gasoline Hydrogenation Unit

CLAUS AND TGT IMPROVEMENT

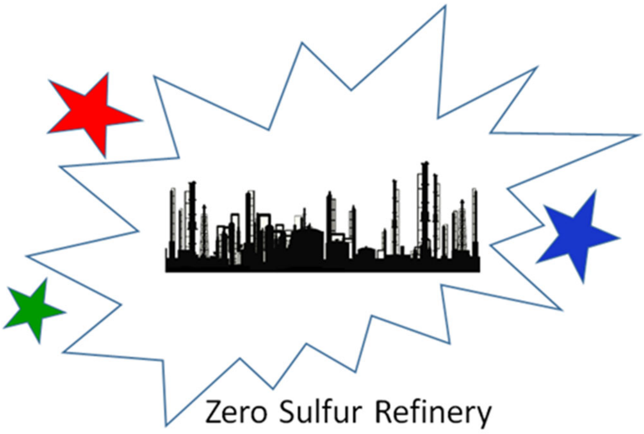
The first area to get attention for sulfur reduction was the Claus unit off gas, to add tail gas treating (TGT). One of the well-known options is the SCOT8 unit offered by Shell. The SCOT consists of a hydrogenation step to convert all the sulfur species into H2S, which is then reabsorbed into an amine for recycle to the sulfur recovery unit. In some cases, the remaining off-gas from the amine absorber may be sent to thermal oxidizer (TO) to convert all residual H2S into SO2, then absorbed with caustic wash to yield a final product suitable for venting. This is a complex arrangement to get the last traces of hydrogen sulfide removed from the vent.

One improvement that can be made to this arrangement is to use the SweetTreatTM process as replacement of the thermal oxidizer and caustic wash. This will eliminate the energy required in the TO, and the waste caustic to be handled. Alternatively, the SweetTreatTM process can be operated as a back-up to the amine and caustic wash / T.O. part of the TGT unit, if these needs to be shut down for a short duration.

A further enhancement would be to use a weaker amine in the TGT absorber. This permits easier and less costly regeneration of the H2S and could further reduce the CO2 that is captured and recycled to the Claus plant. Diisopropyl amine (DIPA) for example, may take 0.10-0.15 moles of CO2 with each mole of H2S that is recycled to the Claus reaction section. Optimizing the amine choice and operating conditions may drop the CO2 capture to 5% or less. Any elevated content of H2S in the resulting off-gas would be removed in the SweetTreatTM adsorber, similar to the redundancy application of Fig. 1.

ZERO SULFUR REFINER

The logical progression of environmental activism is to have zero emission of whatever is offending. ESG initiatives provide the impetus to act on reducing noxious waste to the atmosphere. In the case of sulfur, there is a pathway to get to zero sulfur leaving the refinery in the atmospheric emissions by removing all H2S from the flare gas, TGT off-gas, and fuel gas streams. Traditional unit operations leave some small quantity of H2S, which may require further treatment in an incinerator or caustic scrubbing system. SweetTreatSM is a solid adsorbent system that eliminates all the H2S in these streams.



Conclusion

Sulfur is objectionable on account of equipment corrosion and acid rain, and general toxicity of H2S and SO2. Environmental regulations are squeezing every aspect of refinery emissions, including sulfur. Thus, refiners should avail of the best options to reduce sulfur emissions. The Gas Processing Industry extensively uses solid adsorption systems as scavengers for H2S. Similar systems can also apply to its cousins in the downstream refining industry.

1. Solid adsorption scavenger systems are excellent means to provide redundancy to the TGT and eliminate flaring. They give continuous and complete removal of sulfur.
2. Removing all H2S from the fuel gas allows the furnaces to operate with a lower stack temperature, extracting more heat from the fuel gas. Overall, there could be an increase of 3-4% in refinery energy efficiency by recovery of the low-level heat from the fired heaters.
3. Scavenging sulfur in the hydroprocessing units reduces hydrogen consumption and improves operating performance such as gasoline octane value.
4. Claus/TGT and amine optimization can be made by judicious use of fixed-bed scavenger system at key points in the process. This will reduce emissions going to the flare and directionally debottleneck the sulfur processing units.

A ‘Zero Sulfur Refiner’ will improve its ESG score and be recognized as a leader that is sensitive to environmental issues.

REFERENCES

Crowther, David, presentation at 2021 GPA Midstream Conference

Pim Van Keep, Robert Sakko, PTQ Q2, 2019.

Robert Torgerson, Gayesco publication, 11/10/2020

Byron Eslinger, Tulsa Heaters, Inc., consultant note, Dec. 8, 2021

SRI Intl. Report 29C; original ref Oil & Gas Journal, Sept. 2, 1974, p. 59-67

SRI Intl. Report 29C; original reference Dittman H., et. al; Linde Ber Tech. Wiss., 51, 1982

S. Singh and R. Mukherjee, Hydrocarbon Processing, October 2021.

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Quench Oil Tower Operation Problems & Troubleshooting With Solutions

Abhishek Sharma

INTRODUCTION

Quench tower is the mother of the ethylene plant and every product comes from this tower. Liquid steam cracker units have a quench oil tower or gasoline fractionator or pyrolysis fractionator, which is used to separate quench oil from the mixture of cracked hydrocarbon. Gas steam crackers don't produce much amount of quench oil so they don't require this tower. Cracked gas with dilution steam, from the furnace first, cooled down in transfer line exchanger to reduce the secondary reactions.

Secondly, cracked gas is cooled down by direct contact with quench oil via injection which is called quench fitting or quench injection. Dual-phase, quench oil (liquid) and cracked gas diverted towards quench oil tower where quench oil and heavier separated from lighter gas and removed as a top product. All steam condenses in the quench water tower after the quench oil tower and water is recovered for reuse as dilution steam (Figure-1).

Quench oil Tower is very sensitive to steam condensation if column top temperature is lower side than higher chances of steam condensation. Steam condensate (water) may form emulsion or cavitation of the bottom pump. It also creates pressure surges due to the sudden vaporization of water at the bottom and can dislodge packing.

This article will elaborate on the quench oil tower steam condensation and calculation to avoid condensation during normal operation or start-up. This also explains the major issues during the operation of the tower and troubleshooting with the solution of the problems.

KEY WORDS

QO (Quench oil), Troubleshooting, Steam condensation, Dew point, HSS (High steam standby), FCCU (Fluidized catalytic cracking unit), DCU (Delayed cracking unit)

PROBLEM-1

Steam condensation in the quench oil tower is a very perilous event that can cavitate the bottom quench oil pump and may shut down the whole plant and production loss. Steam condensate may also form an emulsion with quench oil and we are unable to demulsify the solution. The major impact of water in quench oil tower is pressure surge which is described in many case studies1 and can break the packing of the tower.

To avoid this condensation problem in quench oil tower, all designers or licensers set the top temperature of the quench oil tower higher than the dew point of water. Let's see the calculation of the dew point and top temperature setpoint by an example. This problem is similar to the main fractionators in FCCU and DCU in refinery.

The normal operating pressure of the Tower is 0.5bar (g) and 1.5bar (absolute)

Composition of the water and hydrocarbon in the gas stream

Mole% of water (steam) =50% and hydrocarbon Mole% =50%

By Raoult's law of partial pressure of the gas for ideal gases.

The partial pressure of the steam in the total gas in the quench oil tower

PW = X*PT

Where PW is the partial pressure of the steam.

PT is total pressure of the column in absolute.
X is mole fraction of the steam.

PW = 0.5*1.5= 0.75bar (a)

Water dew point at the 0.75 bar (a) pressure is 920C

For safety margin of 5-100C is considered based on the designer.

Quench oil top temperature should be 100^0C to avoid steam condensation. It can

be calculated based on operating conditions and gas compositions.

All the set points of the top and bottom temperature of the Tower, provided by designers or licensers are for normal operation but during the startup, they deviate and may create the problem.

During the startup of the steam cracker plant, Cracking furnaces run without hydrocarbon only with steam in the tube side which is called Heat steam standby (HSS).When the furnace's effluents is diverted towards the quench oil tower from HSS mode. This effluent only contains steam, not any hydrocarbon. At this time if we want to maintain the parameters of normal operation like top temperature and pressure of the tower then definitely steam will condense.

Let's see the calculation of partial pressure.

$$P_T = 1.5\text{bar (a)}$$

X= 1 (mole fraction of the steam because it is free from hydrocarbon vapor)

$$P_W = 1*1.5 = 1.5 \text{ bar (a)}$$

The dew point of water at this pressure is 111^0C and we are trying to maintain the top temperature of the quench oil tower is 100^0C.

During the startup of the plant, we should maintain the top temperature of the quench oil tower more than 115^0C for the above example or reduce the pressure according to if possible.

The problem with steam condensation is hiccups in the quench oil tower because the top temperature is lower than the water dew point and bottom temperature is higher than the dew point of water so steam condensates (Water) may trap in the middle of the column. That's creating the hiccups in the tower[2].

Quench oil circulates from the bottom of the tower to exchangers for removal of heat and after cooled down injected in furnace effluent. Reflux for quench oil tower comes from Oil-Water separator and total draw-off from quench oil tower to Product.

PROBLEM-2

Water in the quench oil tower other than steam can come from the steam line which is provided from the steam-out line during the

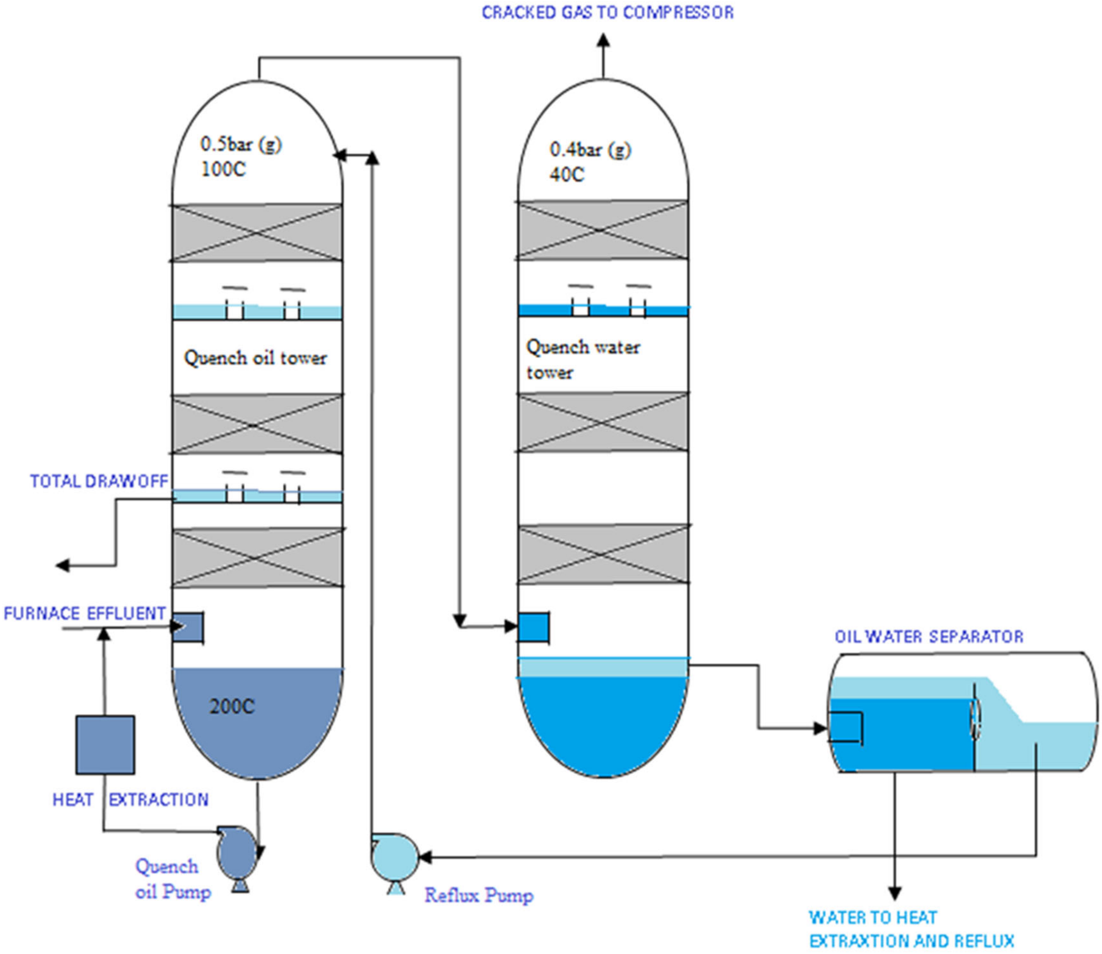

Figure 1 Overview of the system. (Quench oil tower and Quench water tower)

shutdown. If the valve is passing or condensate accumulated over there.

Another source of water in the quench oil tower is reflux. Reflux in the tower comes from the water-oil separator. Any design error in the oil-water separator or any leakage in the separating baffle can cause the water into the oil compartment and reach the tower. Every oil-water separator has a separating baffle with a manway for inspection which may leakage due to gasket issue or loose bolting. (Settling baffle should be tight sealed)

When we send water as reflux then the top temperature of the reflux reduces drastically due to the heat capacity and latent heat of the water. This creates the same problem as steam condensation.

Some plants have an extra line of gasoline from caustic wash gasoline to quench oil tower as reflux. From where also any design error can send water or mixture of spent caustic to quench oil tower as reflux.

During HSS mode when only steam as a furnace effluent, reflux of quench oil tower should be higher side to maintain the temperature because due to the steam partial pressure of hydrocarbon reduces and stripping effect happen.

PROBLEM-3

The bottom temperature of the quench of the tower is very sensitive and critical because it is responsible for the Viscosity of quench oil. For normal hydrocarbon or liquid, viscosity decreases with an increase in temperature but for quench oil, it is not the case. Due to increases in temperature polymerization reaction starts and increases the viscosity of the quench oil. An increase in viscosity leads to plugging and pumping problems. At lower temperatures also viscosity reduces as well as less heat recovery.

The bottom temperature of the tower should be maintained as per design otherwise it creates problems. But most of the plants have an issue with the instrument (temperature sensor and indicator). Quench oil tower bottom temperature sensor should be in the liquid sump of the Quench oil so it mostly crosses from skirt support of the tower and according to the diameter of tower or skirt support its length vary.

Figure-2 shows the length of the temperature sensor when column diameter is 5m and temperature thermo-well pipe length is 1m.

Mostly all temperature thermowells length is 300mm to 700mm in the whole plant. If we use a smaller length thermo-well unknowingly then this temperature sensor detects the temperature of the stagnant quench oil in the pipe of the thermo-well. This lower detected temperature may confuse the operator to increase the temperature to meet the design value. This over increment in the bottom temperature might increase the polymerizing reaction and result to an increment in viscosity.

To avoid this type problem, check and review at the design stage or arrange the longer temperature sensor.

Temperature sensor should be 300-350mm inside the shell of tower and direct contact with the liquid of bottom sump.

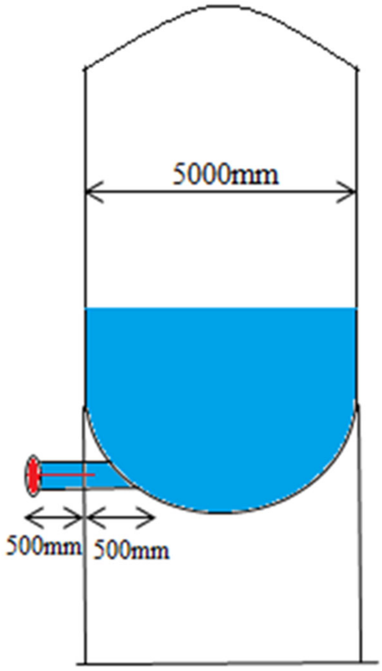

Figure 2: Thermo-well length vs Pipe length.

PROBLEM-4

During the inspection and installation of packing of the tower, people walk on the packing which may crush the packing or compress it which may create liquid hold-up or higher pressure drop across the packing. Binding of the packing is also possible. Packing is supported by the beam (I- beam) or support grid (hold-down, holdup grid) which can also reduce the opening of the packing3.

I- beam is also a cause of mal-distribution of vapor due to blockage. To avoid the mal-distribution problem I-beam should have holes throughout (horizontal) to equalize the vapor.

Once I have seen an event, where a temperature sensor was located in between the packing bed and installed after the packing installation. When the temperature sensor was installed, packing was compressed. Because packing covered the nozzle and thermowell was forcefully installed.

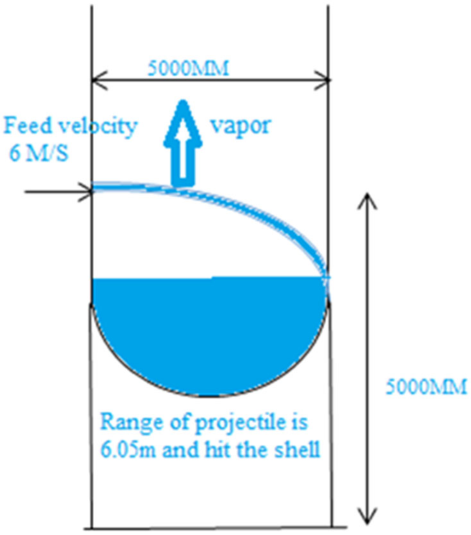

Figure 3: Feed velocity and Dynamics

PROBLEM-5

The high momentum of feed from the furnace to quench oil tower may entrain the liquid to packing and coke particles can trap in the packing which can choke the packing. Once the gap of packing is reduced or chocked then capacity reduces and column may premature flood. Feed to quench oil tower is the dual-phase (mixture of liquid and gas) which may have higher velocity due to mixture of gas and can hit on the shell of the tower or instrument tapping(Level or pressure tapping at the bottom of the tower). High velocity or momentum of the feed can also erode the shell and make thinner the shell of the tower.(Figure-3)

There should be extra shell thickness in front of the feed nozzle or a momentum reduction baffle should be installed.

PROBLEM-6

Any upset in the cracked gas compressor/ furnace or any fluctuation in the feed of quench oil tower may lead to the Quench oil carryover or entrainment above the total draw-off tray. Once quench oil entrain above the total draw-off tray then quench oil will accumulate over there and contaminates the draw-off with heavier.

To avoid this problem, a pipeline from draw-off to below total draw-off tray with distributor should be there. This distributor can work as a wash spray and avoid quench oil entrainment.

CONCLUSION

- Temperature profile to be maintained during start-up and normal operation based on dew point calculation.
- Water entry is to be avoided by maintaining interference levels in the oil-water separator and steam trap in the steam-out line.
- All columns bottom temperature pipe and thermo-well length to be checked before start-up.
- Binding of packing or crushing during inspection to be avoided and I-beam should have holes (horizontal) throughout.
- Based on the momentum of feed and to avoid entrainment, an impingement plate or wedge baffle is to be provided.
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REFERENCES

Kister H.Z., “What Caused Tower Malfunctions in the Last 50 years” Vol 81, Part A, Jan -2003 Institute of Chemical Engineers.

Kister H.Z., “Distillation Troubleshooting” John Wiley & Sons Inc. publication(2006)

Lieberman N.P., “Process Design for Reliable Operation” Gulf Publishing Company (1998).

Lieberman N.P., Lieberman E.T., “A Working Guide to Process Equipment” McGraw-Hill Education(Fourth edition 2014)

Kister H.Z., “Distillation Operation” McGraw-Hill Inc. (NY -1990)

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