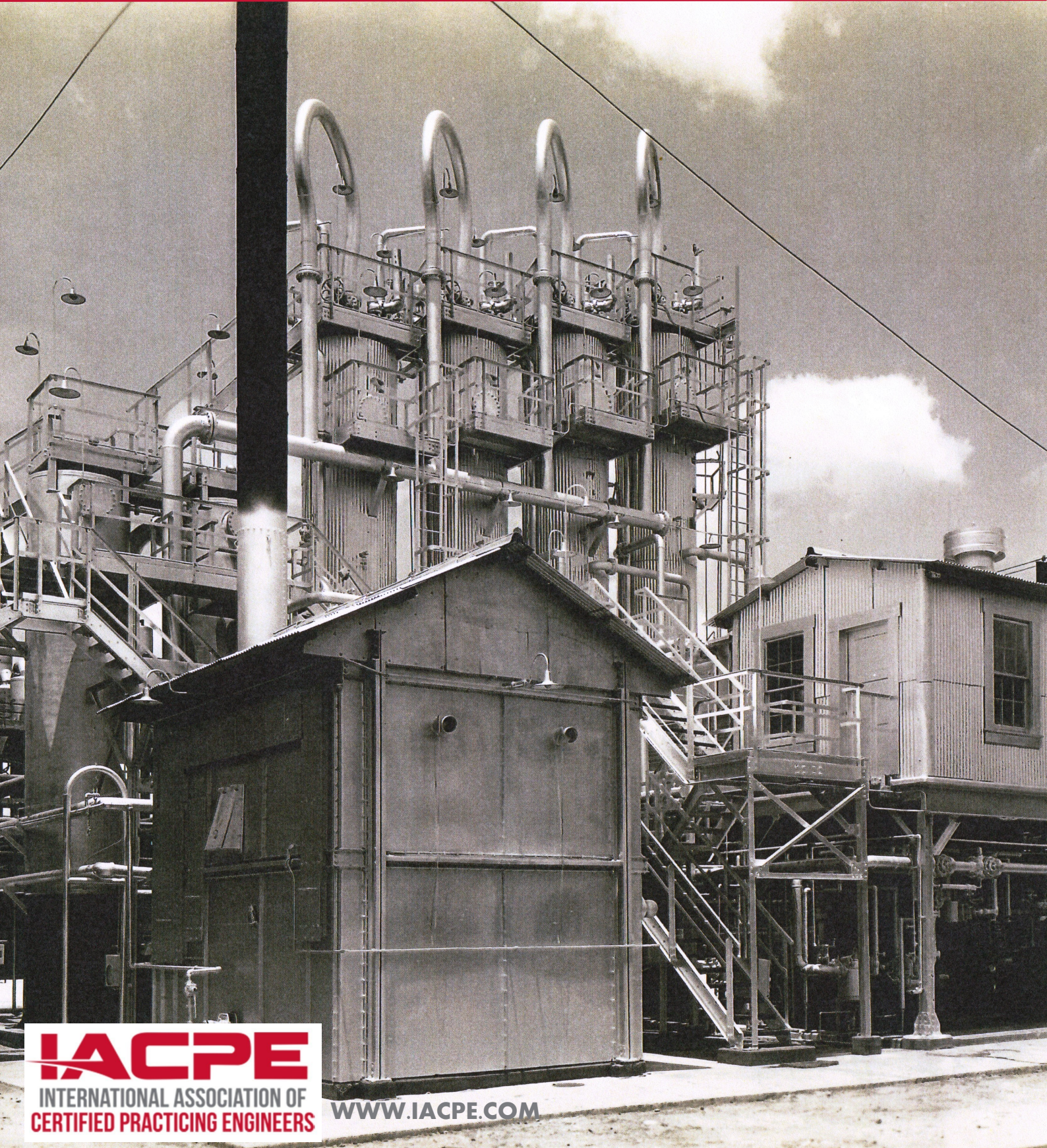


ENGINEERING PRACTICE

VOLUME 10 NUMBER 42

JANUARY 2024



IACPE
INTERNATIONAL ASSOCIATION OF
CERTIFIED PRACTICING ENGINEERS

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Can you trust your simulation software for troubleshooting?

Anand Govindarajan, Ph.D., Upasana Manimegalai Sridhar, Ph.D.
Ralph Weiland, Ph.D.

Abstract

Troubleshooting is a demanding task because, when a problem arises, a solution is usually urgently needed. Not only is the experience of the engineer essential, but also a rigorous process of root cause analysis and a reliable simulation tool are equally important to make well thought-out, sound recommendations.

This article discusses the following specifics regarding amine systems:

- How root cause analysis can be carried out
- What is data? And what is not?
- Issues with data reliability
- Data required for benchmarking

Subsequently, this article discusses some of the common issues and suggested mitigation measures in amine systems for dealing with:

- Failure to meet product quality
- Heat stable salts incursion
- Solvent losses
- Foaming
- Corrosion

For each of these issues, this article discusses the use of a simulation tool — ProTreat® — and how it is meant to be used in the context of troubleshooting, and the technically correct way to interpret the results. Finally, this work discusses a case study from an actual refinery fuel gas treating application using MDEA as the solvent. Root cause analysis principles and the benchmarking guidelines discussed here are used as an illustrative example.

Introduction to troubleshooting

Any problem occurring during operations can cause a lower quality product or loss of production and that can represent a significant cost. The issue is even more important as oil refiners and gas producers must protect their margin. Troubleshooting is a demanding task because, when a problem arises, its solution is urgently needed; no doubt it will catch the attention of management, and the engineers will be put under a lot of pressure.

With time being of the essence, not only is the experience of the engineer essential, but also a rigorous process of root cause analysis and a reliable simulation tool are equally important to make well thought-out and sound recommendations. This paper looks at troubleshooting in the context of amine systems and is meant to serve as a general guide for the use of simulation in troubleshooting.

Root Cause Analysis in Amine Systems

In engineering, root cause analysis is a series of steps used in problem solving and is a collective term that describes a wide range of techniques, tools and approaches to identify the causes of a problem. The process is well understood and goes through a few essential steps which can be generalized as follows:

- Definition of the problem
- Data collection and benchmarking
- Identification of potential root causes
- List of plausible explanations
- Testing corrective actions
- Implementation and monitoring

Table 1 provides a high-level summary based on the generalized troubleshooting steps in relation to amine systems.

Table 1: Summary of generalized troubleshooting steps

Troubleshooting step	Details
Define problem	Some typical problems associated with amine systems (illustrative, not exhaustive) <ul style="list-style-type: none"> • Off-spec product • Heat stable salts • Excessive solvent losses • Foaming • Corrosion
Data collection and benchmarking	<ul style="list-style-type: none"> • Temperature and pressure • Flow • Solvent analysis • Operating guidelines • System design limits
Identify potential root causes	Important to understand the cause effect relationships between multiple variables affecting the process and enlist all of them.
List of plausible explanations	Each root cause may have a possible explanation, and a series of steps that could explain why the problem occurs. Such explanations should be guided by data, scientific explanation, and engineering judgement. The role of experience is important, no doubt, but experience alone may not have all the answers.
Testing corrective actions	The usage of a simulation tool plays an important role in allowing testing of corrective action(s) and increases the level of confidence for the implementation of a solution. Process simulations should tell an engineer how the process is supposed to operate, given the system configuration and a set of operating parameters. Once again, far from treating process simulations as black box models, they should support deductive reasoning and not be viewed in isolation. Here, the role of experience in the utility of a simulation is important.
Implementation and monitoring	Once reasonable confidence has been established by testing corrective actions, and risks evaluated, the implementation of the proposed solution may occur. Subsequently, the unit should continue to be monitored to ensure desired objectives are met.

Data collection and benchmarking

Good detective work begins with collection of good data. Good troubleshooters do not accept data at face value, they validate the data and make decisions based upon the data, rather than speculation. When it comes to plant data, it is common to have errant data measurement as amine plants corrode or plug. Process simulators can be tuned to match some of these data. In many cases, process simulators allow arbitrary tuning of parameters in the guise of allowing flexibility to the user or allowing a photographic replication of the plant. Not only is such tuning devoid of scientific principles, in addition, if the simulation is tuned based on bad plant data it can lead to inconsistent predictions and incorrect conclusions. So, in troubleshooting validation is

critical and it is a reasonable guiding principle to treat any piece of data as error prone until proven innocent. This brings us to the question - how does one ensure that plant data available on hand is good (reliable)? Some of the obvious preliminary checks include questions such as –

- Are the valves lined up to put the flow where it should go?
- Are all instruments measuring correctly?
- Are the laboratory analyses reproducible and accurate?

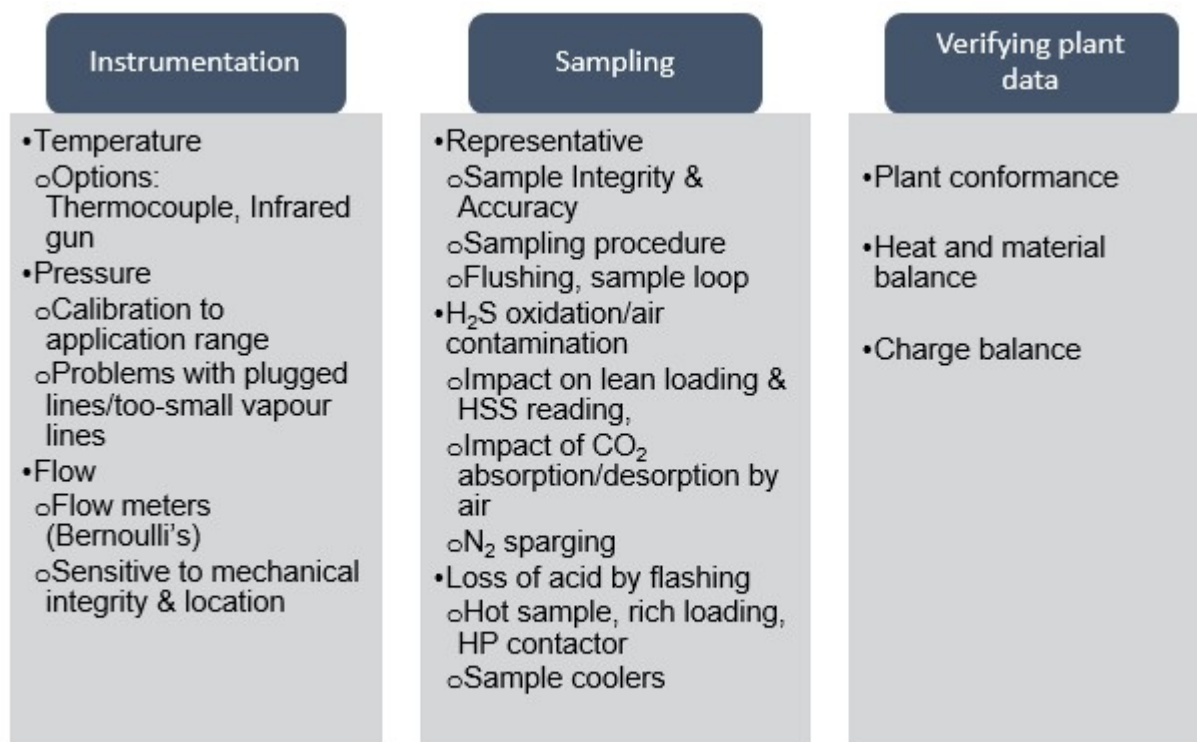


Figure 1: Troubleshooting preliminaries.

Temperature & pressure measurements

Figure 1 details some preliminary issues related to data collecting in amine systems. Starting with instrumentation – the level of trust in instrumentation is inversely proportional to instrument complexity. Collecting temperature data can be relatively simple. Options include thermocouple devices, or infrared guns. Temperature measurement devices can be miswired or improperly installed. Pressure measurements, once again simple, yet are prone to incorrect installations. Lines can be plugged, or too small diameter vapour lines can contain liquid legs trapped by joints, or incorrectly sloped vapour lines accumulating liquid. Also, it is critical to check that pressure transducers are properly calibrated to the range of pressure for the application.

Flow measurements

Measuring flow is essential, and it can be hard sometimes to diagnose malfunctions of flowmeters. The most common type of flowmeter uses an obstruction on a pipe to generate pressure difference which is then measured (based on Bernoulli's equation to infer velocity from measured pressure difference). Portable clamp-on ultrasonic flow meters are also available and at the more complex end of the spectrum, mass flow rate can be measured, even for two-phase flows. Flow meters are sensitive to mechanical integrity and solid deposit/scale can cause some problems. Their location is obviously important as you want to avoid wrong

measurement if these are in a region of turbulence like downstream of a flow control device. There are other issues as well with flow meters including incorrect calculations on the DCS, corroded primary elements, defective transmitters and so on. It may sometimes be advisable to have two flow measurements and ensure they agree. In any troubleshooting exercise, correct flow measurements are essential - direct visual inspection and simple hand calculations to ensure the DCS is producing the correct output will save significant embarrassments.

Sampling

An amine sample must be representative of the amine that is circulating in the system. It is essential to follow proper sampling procedures to maintain sample integrity and accuracy. Often samples are drawn from a long sample line which may contain stagnant liquid and not be truly representative of the circulating amine. Sample integrity can be preserved by flushing the sampling line to avoid any stagnant material or better, by having a sample loop on the system and N₂ blanketing. Care must be taken to avoid air ingress that can cause H₂S oxidation or absorption of CO₂ into the amine sample. In addition, loss of acid gas can occur by flashing as well when sample is taken from the hot section, such as the rich section or a high-pressure contactor – so it is important to cool such samples immediately by running through a cooler. More discussion on sampling is available in (Weiland, et al. 2018).

Table 2: Preliminary sanctity check requirements for plant data

Unit operation/item	Comments
Absorber	<ol style="list-style-type: none"> Does acid gas removed as calculated by gas analysis and gas flow rates agree with acid gas picked up by solvent (by solvent flow rate and loading change)? Is column energy balance consistent with heat released from the heat of absorption and known inlet and outlet gas and liquid temperatures and flows? <p>In both cases above, the material and energy balance should close within the accuracy of the basic measurements involved, if not something is wrong.</p>
Regenerator	<ol style="list-style-type: none"> Dry acid gas flow from the overhead condenser must agree with the acid gas flow rate to the column in the rich amine minus the residual acid gas in the lean amine. Is reboiler duty consistent with the measured reflux flow rate and top tray temperature? <p>In both cases above, it is important to account the thermal state of the rich amine feed. In case there is feed flashing, it is important to understand the contribution of the flashing vapour flow to the overheads.</p>
Charge balances (A more comprehensive discussion is available (Hatcher and Weiland 2009))	<ol style="list-style-type: none"> All solution analysis should be checked for obeying the charge balance whenever possible. Overlooking this may lead to incorrect simulations based on wrong solution analysis that do not even obey the charge balance. A charge balance spreadsheet is available free of cost at Optimized Gas Treating Inc's website https://www.ogtr.com/free_downloads

Verifying plant data

Table 2 provides a summary of the preliminary data checks that should be carried out to ensure that the data conform to material, component, and energy balances.

Data required for benchmarking

Knowing now that our data satisfy initial sanctity tests, Table 3 is a check list that covers the

physical properties, and composition of different streams. Gathering these data over multiple stable regions of operation will be useful for simulation benchmarking of the amine plant. Do note that Table 3 assumes that the hardware information related to the columns such as number of installed trays / packing height, type of the internals, column diameter, etc. are known.

Table 3: Amine system data required for benchmarking

Temperature and Pressure	<ul style="list-style-type: none"> Feed gas Sweet gas Rich amine at bottom of absorber Lean amine at top of absorber Rich amine at the exit of L/R exchanger or control valve outlet (consistent location for T and P) Lean amine leaving reboiler Acid gas leaving the regenerator Lean amine at exit of L/R exchanger Condenser Reboiler Column overhead going into condenser Column bottoms going into reboiler
IC amine analysis	<ul style="list-style-type: none"> Lean amine
pH	<ul style="list-style-type: none"> Report temperatures where pH readings are taken Lean amine Rich amine
Gas flows and composition analysis	<ul style="list-style-type: none"> Gas flows and composition analysis Feed Gas to absorbers Treated gas from absorber Acid gas leaving regenerator reflux drum
Other flows	<ul style="list-style-type: none"> Lean amine flow Reflux flow Reboiler steam flow (also furnish T/P for thermal condition)
Acid Gas Loadings (H ₂ S and CO ₂)	<ul style="list-style-type: none"> Lean amine (at the outlet of lean cooler)

Common issues and suggested measures

This section attempts to layout some of the common issues and their causal factors. This is by no-means an exhaustive list. In addition, the direction of the deviation (high/low/erratic) and potential solutions are also listed in Table 4.

Expectations from simulation

Simulation tools are often trusted as a gospel without a finer understanding of the tool being used. In general, a set of input parameters are given to the simulator. The simulator then calculates how the process is supposed to

Table 4: Common Issues, causal factors, and potential solutions in amine units

Issue	Causal factor	Solution
Off-spec product (sweet gas)	Rich loading (high)	Increase circulation rate and/or amine strength
	Rich loading (maxed)	Increase amine strength
	Lean loading (high)	Check regenerator for poor stripping performance Check L/R exchanger for contamination/leaks
	Lean temperature (high/low)	Check amine cooler
	Absorber/Regenerator DP (high)	Fouled trays – cleanout maybe necessary
	Absorber/Regenerator DP (low)	Check for mechanical damages
	Absorber/Regenerator DP (erratic)	Foaming may be occurring – troubleshoot
Heat stable salt incursion (Hatcher, Miglani and Govindarajan 2020)	Regenerator temperature (low)	Increase heat to reboiler
	Contaminants from upstream processing	Some HSS precursors can be removed through water washes of inlet hydrocarbon gas (HCl and SO ₂ for instance)
	Impurities in make-up water	Clean makeup water to be ensured (boiler feed water quality or better)
	Piping mis-lineups	Ensure flow is going where it is meant to
	Excessive caustic addition for neutralization	Use enough to liberate the amine to react with the acid gas. A reliable simulation should be able to tell you.
Solvent losses	Oxygen ingress	Blanket tanks, eliminate leaks in vacuum gas gathering equipment etc.
	Mechanical losses	Improve operation and maintenance practices
	Physical entrainment / vaporization	Water wash to retain the solvent. Check for foaming losses.
Foaming	Degradation	Thermal reclaiming maybe necessary
	Incursion of surface-active ingredients	Source prevention is critical as most filtration methods are ineffective
	Hydrocarbon ingress	Check the functioning of the inlet separator Purge or drain the skimmed hydrocarbon from the flash drum Controlling the lean amine temperature (typical rule of thumb is to maintain a lean amine temperature at 10 F above the feed gas, to prevent hydrocarbon condensation) Purge reflux
Corrosion	Iron sulphide	Cartridge filters maybe necessary to ensure colloidal particles that increase the surface viscosity are eliminated
	Rich loading (high)	Upper limited for carbon steel (CS) to be maintained between 0.4-0.5 mol/mol maximum (subject to other guidelines) Schedule reclaiming
	Velocity (high)	Lean amine velocity in CS: 7-10 ft/s (2.13-3.05 m/s) Rich amine velocity in CS: 3-5 ft/s (0.91-1.52 m/s) Ensure gradual changes in process conditions Manage throughput
<i>Additional guidelines on corrosion are available in other sources listed in the references.</i>		

behave. A simulator may be able to solve a problem through deductive reasoning. However, it is very important to note that, the simulator cannot tell whether the piping and valves are lined up as per requirement and if the mechanical integrity is intact, the cleanliness of the amine unit and the solvent and whether the instruments are calibrated correctly and if lab test results can be trusted.

Thinking that sitting in front of a computer, working on simulations will solve all your problems, is simply not true. Simulations must be regarded as a mean to the end, not the only means to the end. Lastly, it begs reinforcing that simulations tuned to bad plant data are not better than the data themselves (Alvis, Hatcher and Weiland 2015).

About ProTreat®

The case study that follows will use ProTreat®, which is one of OGT's simulation products. OGT Simulation Software began with gas treating in 1992 and has been strictly mass and heat transfer rate-based right from the beginning. For 30 years OGT has led the way in this revolutionary modern technology and, after witnessing its power, others have followed. Today, most simulators claim some mass transfer rate-based capabilities, but only ProTreat is fully rate-based in the true meaning of the word (OGT Inc. 2010) and allows you to simulate treating using single, multiple, and specialty amines, non-amine systems, amines mixed with a physical solvent, sour water stripping, and glycol dehydration in columns containing

a vast range of trays, random packing and structured packing in absorbers, regenerators, and quench towers.

The latest gas treating addition to ProTreat simulates hybrid solvent systems in which part of a standard aqueous amine is replaced with an organic nonreactive component. Currently ProTreat can simulate up to three aqueous generic amines with sulfolane as the organic additive. Other combinations are under development. Some of ProTreat's applications and capabilities are that it can be used for acid gas removal, acid gas enrichment, tail gas treating and sour water stripping processes. Glycol dehydration, caustic mercaptan removal, LNG and ammonia can also be simulated using ProTreat. Asset integrity can be estimated using the corrosion prediction capabilities. ProTreat relates to actual hardware and tells what the plant should be doing and contains the largest commercially available database of solvents, both generic solvents and formulated solvents including INEOS, Dow and Eastman

proprietary solvents. ProTreat is a versatile tool that is flexible and integrable with downstream SRU and is a CAPE-OPEN plug.

Globally over 100 companies including the largest producers, refiners, solvent vendors, internals manufacturers, licensors, EPC, engineering, and research companies use ProTreat as their de facto simulator for gas treating.

How ProTreat® helps in troubleshooting amine systems

Issue	What ProTreat® can do
Off-spec product (sweet gas)	Affords reliable investigation of various process parameters on the performance of the amine system. ProTreat's mass transfer rate model is rich in detail and is a faithful mirror of the real world. In the language of process control, a true mass transfer rate simulation uses a distributed parameter model. Consequently, such models are mechanistic, detailed, and fully predictive. Mass transfer performance predictions have been validated against a large amount of full-scale plant performance data, and regeneration columns are simulated just as accurately as absorbers. ProTreat is 100% mass transfer rate-based and uses detailed chemistry and mass transfer calculations to predict performance of real equipment without guesswork.
Heat stable salt incursion	Use solvent analysis information to predict the impact of HSS on treatment. Also, enter estimated HSS based on guidelines to simulate impact on treatment. ProTreat contains (probably) the industry's largest database of HSS.
Solvent losses	Simulate the impact of water washes, degradation products, specify liquid carry-over in top vapour stream, vapor carry-under in bottom liquid streams
Foaming	Use the advanced features for columns to increase or decrease the interfacial area factors to understand the potential extent of foaming in the tower.
Corrosion	Setup virtual corrosion coupons to study the impact of loading, temperature, velocity, amine strength, and speciation on corrosion, for a variety of metallurgies, roughness, fittings, and sizes.

Case study

The case study deals with an amine unit that is a standard absorber-regenerator circuit of a fuel gas unit. The absorber contains 30 Nutter float valve trays with 8 ft. diameter and 62% active area. The regenerator has a diameter of 6 ft. on top and 8.5 ft at the bottom with three sections total; first section is packed with 10ft. of 2 in. metal pall rings, second section has 2 Koch flexi trays with 84% active area and the final bottom section has 18 Koch flexi trays with 8.5 ft diameter and 56.4% column active area. The feed gas flow is 38.2 MMSCFD of fuel gas at 38 C and 13.3 kg/cm² that contains (in mol%) 39% CH₄, 14.9 % C₂H₆, 2% C₃H₈, 3.1 % C₄H₁₀, 1.2% CO₂, 16% H₂S, 10.9% H₂S and the rest being N₂, CO, and water vapor. MDEA with a strength of 42.6 wt% is being used to selectively remove the H₂S in the feed gas at 193 cum/h (850 USGPM). According to the plant, all the columns, pumps, and exchangers were at about 70-80% of their design capacity, so in essence there was no bottleneck. The treatment objective was < 4 ppmv H₂S and 16.5% CO₂ slip into the treated gas. However, the unit was currently producing around 22-25 ppmv H₂S leak and 20.5% CO₂ slip.

First, a simulation model was set up in Pro-Treat® based on the temperature, pressure, and flow data from the plant instruments. Amine analysis (lean) indicated the presence of degradation products and HSS, whereas the simulation used clean amines. On updating the simulation to account for the degradation and HSS, the simulation was able to nearly replicate the plant performance with 20.5% CO₂ slip and 24 ppmv H₂S leak. Importantly this was done without any arbitrary tuning – an important element in simulation benchmarking. Next the lean and rich loading of the absorber was examined and found to comply with the guidelines. Based on the mitigation measures for off-spec treating, a few actions that can be taken to re-establish satisfactory treating are – adjusting solvent circulation (step A), strength (step B), reboiler steam (step C), and lean amine temperature (step D). The following observations are then made regarding each of the possible actions in isolation:

- **Step A – adjust solvent circulation** - The temperature profile in figure 2 below indicates a well circulated absorber, so changing the circulation rate would not be necessary.

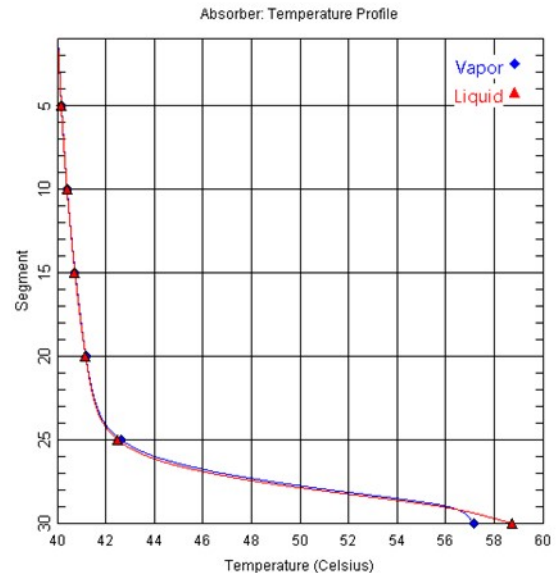


Figure 2: Temperature profile of the absorber

- **Step B – increase solvent strength** - The concentration profile in figure 3 clearly indicates that the absorber is lean end pinched. Therefore, increasing solvent strength will not be of use, and will worsen treating.

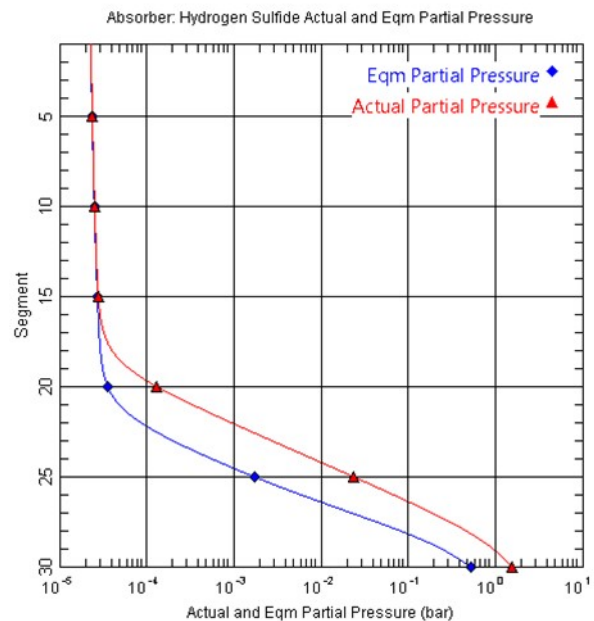


Figure 3: Actual vs equilibrium partial pressure of H₂S in the absorber

- **Step C – increase reboiler steam (to lower lean loading)** - Stripping the solvent to lower lean loadings than current levels may provide better treating. This is however subject to having room in the reboiler and regenerator hydraulic capacity.
 - Examining the hydraulics of the regenerator, indicated that

regenerator was at about 56 % jet flooding, and the reboiler was operating only at about 65% of capacity. This then is one of the obvious knobs to turn.

- o Upon increasing the reboiler steam ratio by about 15% from 82 kg steam/m³ of solvent (0.68 lb. steam/gal) to 96 kg steam/m³ of solvent (0.8 lb. steam/gal) – the H₂S leak was down to 10 ppmv and CO₂ slip is around 20.3%. This was a move in the right direction with H₂S, but not yet with CO₂. We have not made it to the required spec yet. There was also no hydraulic limitation apparent in the regenerator on account of the increased reboiler duty. An important outcome was the lower lean loading from increasing the reboiler duty.

Step D - The lean amine temperature was about 46°C. The trim cooler would allow this to be reduced to around 40°C. Here the expectation is that the lower lean temperature will lower the H₂S backpressure and allow for better treatment. Simulation

indicated 17.5 ppmv H₂S and 17.3 % CO₂ slip. There was a stronger influence on the CO₂ slip by lowering lean amine temperature than in Step C (increasing reboiler steam). H₂S however was more sensitive to the Step C (increasing reboiler steam).

- Combination of Step C and Step D – resulting treating shows H₂S at about 7 ppmv and CO₂ slip at 17.2%

A quick comparative summary is in table 5.

Starting from about 24 ppmv H₂S and 20.5% CO₂ slip, the series of investigations has been able to move the plant directionally towards the desired treating of < 4 ppmv H₂S and 16-17% CO₂ slip. However, the plant was still not back

to the desired treating levels. One obvious option was to further increase reboiler steam, but the refinery was undertaking serious steam saving efforts, so plant personnel were not particularly keen on this. On further discussion with plant personnel, the plant mentioned that they had a mobile HSS removal skid come in and clean up and removed some HSS and degradation product a few months ago. At the face of it, this does not ring any alarm bells, however, the plant was asked to furnish a lean sample analysis before and after the HSS removal skid was brought in (see table 6 below) –

Table 5: Comparative summary of changes in KPIs

KPI	Base case	Step C (Increased reboiler steam rate)	Step D (Decreased lean amine temperature)	Step C + D together
H ₂ S (ppmv)	24	10	17.5	7
CO ₂ slip (%)	20.5	20.3	17.3	17.2
H ₂ S loading in rich amine	0.46	0.453	0.454	0.452
CO ₂ loading in rich amine	0.026	0.026	0.026	0.0026
H ₂ S loading in lean amine	0.004	0.001	0.004	0.001
CO ₂ loading in lean amine	Trace	Trace	Trace	Trace

Table 6: Comparative analysis of HSS

	2 months ago	Now	
Degradation product/HSS	Value	Value	Units
DEA	2	0.2	wt. %
Thiocyanate	0.638	0.0447	wt. %
Formate	0.8		wt. %
Sodium ion	0.18	0.1212	wt. %
MDEA	42.7	42.61	wt. %

As observed from table 6, there was more HSS and nearly 10 times more DEA degradation product two months ago than now. The plant removed the HSS and higher degradation product due to conventional wisdom that higher amounts of these is bad for treating and corrosion.

So, at this point – the simulation run was once again updated with the HSS and degradation products from two months ago. The treated gas in the simulation produced 3.7 ppmv H₂S and 16.5% CO₂ slip. The lean loading is lower than the current operations (which is after cleaning out the HSS and degradation products), which might spring a surprise. HSS and degradation products have complex behaviour and there are several situations where a small quantify of HSS can benefit operations by allowing strip to leaner solvent, when the absorber is lean end pinched. The lower CO₂ slip can also be attributed to the higher amounts of DEA which is a secondary amine and tends to pick up more CO₂. High HSS and degradation products are not always bad, it is important to make a careful assessment with the right tools, following the right fundamental engineering principles. Given that it is not practically possible to estimate the buildup of HSS, there is no way to revert to levels of HSS that were doing the job 2 months ago. Therefore, the plant at this point was left with only option to increase reboiler steam ratio further from 96 kg steam/m³ of solvent to 108 kg steam/m³ of solvent to re-establish treating back to required levels (<4 ppmv H₂S and 16.5% CO₂ slip).

Conclusions

This paper was developed to be used as a general guide for using process simulation tools such as ProTreat® for troubleshooting.

- Troubleshooting can lead to consequences that are costly if not done correctly. Correctly we mean the inputs, lab analysis, constraints, and deductive reasoning from our end.
- It is always good to do a thorough root cause analysis and use a robust and versatile simulator to model and look at what is happening to the process.
- The simulator used for troubleshooting should not require adjusting arbitrary parameters to match the plant data; It should be based on mass transfer rate principles and detailed solution chemistry to be able to accurately model the process by accounting for contaminants, HSS, charge balances and solution chemistry.

- The use of ProTreat® allows troubleshooting exercises to proceed without the need to be concerned about the reliability of the simulation. Availability of reliable simulation tools with process engineering teams should not be a haggling point, they are the right way to carry out engineering.

References

Alvis, R Scott, Nathan A. Hatcher, and Ralph H. Weiland. 2015. "Expectations from Simulation." Laurence Reid Gas Conditioning Conference. Norman.

Hatcher, Nathan A., Manu S. Miglani, and Anand Govindarajan. 2020. "Design Guidelines for Heat Stable Salts Levels in Amine Systems." SulGas Conference. Mumbai.

Hatcher, Nathan, and Ralph Weiland. 2009. "Acid Gas Loading Error Analysis." Brimstone - STS Sulphur Symposium, Vail, Colorado.

OGT Inc. 2010. "Conflicted Meanings of "Rate Based"." The CONTACTOR, Vol 4, Issue 3.

Weiland, Ralph H., Nathan A. Hatcher, Clayton E. Jones, Steven M. Fulk, and Simon A. Weiland. 2018. "Accuracy of measured lean amine H₂S loading." PTQ Gas.










Avoiding Environmental Cracking in Amine Units API Recommended Practice 945 fourth edition, September 2022

TrayHeart

Tower Internals Design



TrayHeart is a professional software that performs hydraulic calculations for all types of tower trays, random and structured packings and liquid distributors. The development of **TrayHeart** started in 1998 and was continued jointly by universities, companies of the chemical industry and tower internals suppliers. **TrayHeart** ...

-  is based on multiple calculation models and large databases of packings, float valves, fixed valves, bubble caps, and liquid distributor templates
-  is a supplier-independent tool. There are no preferred product placements or promoted designs
-  considers static dimensions, manways and fastenings
-  offers an interactive 3D-view for all designs
-  can be used for single stage, profile and data validation calculations
-  has a unique, logical and multi-lingual user interface, with multiple input and output options
-  applies hundreds of online queries to check the feasibility and limits of the calculated designs
-  is a well introduced software many companies have relied on for more than 20 years
-  has extensive documentation and is licensed on annual basis

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Pinch Analysis: A Retrofit Approach, An Opportunity To Optimize Process & Energy

Syed Abbas Raza

Pinch Analysis provides techniques for heat integration to design the Heat Exchanger Network (HEN) not only for grass root design (figure 1a, 1b and 1c), it provides an opportunity to retrofit an existing system as well to optimize the process. These techniques are interactive and user-friendly. Retrofit study is more complex as compared to Grass Root Design [1].

This article is focused on the techniques for retrofit and discusses the options available to optimize the system. Generally, the Retrofit may be done for following reasons in a refinery heat exchanger network (HEN):

- To optimize the energy efficiency
- To increase throughput

- To reduce emissions
- Process modifications

Previously, the retrofit was carried out by changing the grass root design of the network towards the ideal network (figure 3). But the problem associated with this approach is large number of modifications in existing network. A better approach is to evolve the network from the existing structure to identify only the most critical, and therefore cost-effective changes to network structure, such that the existing arrangement can be modified in such a way that the ideal curve can be approached [1]. Below figure 2 shows the methodology for retrofit approach.

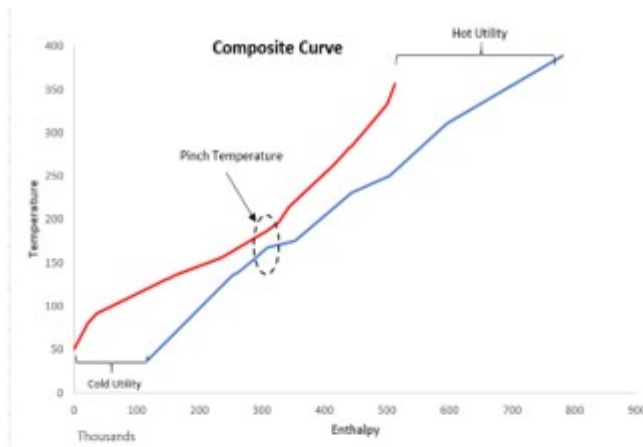


Figure 1a: Composite Curves

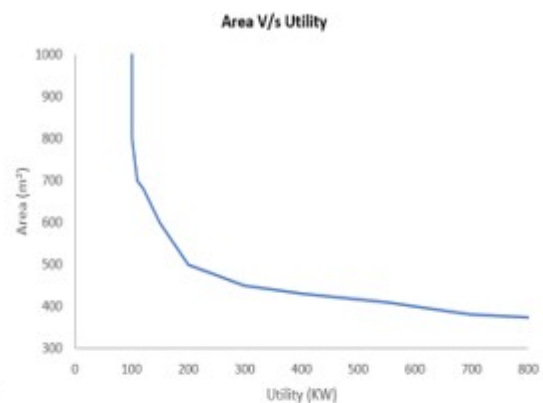


Figure 1b: Area versus Utility Curve

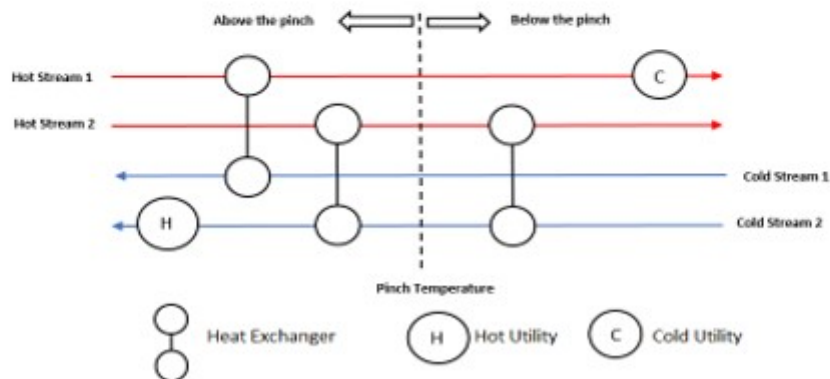


Figure 1c: Grid Diagram showing Heat Exchangers, Hot and Cold Utilities.

The main purpose of the retrofit analysis is to keep the design as similar as possible to the original one, which implies suggesting as few modifications as possible.

Network Pinch

Tjoe and Linhoff [2] proposed the first pinch retrofit method and presented the idea of Area efficiency ($\alpha = A_{\text{idea}}/A_{\text{existing}}$). Alpha (α) represents how close the existing area to target (ideal) area is. Lower values of α shows more cross pinching heat and unoptimized system (figure 4).

The difference between Process Pinch and Network Pinch; Process Pinch is the function of process streams only while Network Pinch is the function of both process streams and HEN topology [3]. However, topology modifications keep the process pinch unaffected while network pinch is changed.

The Network Pinch Analysis is used to determine the structural elements of the network, which prohibit the topology changes. Exchanger minimum approach (EMAT) is the minimum approach temperature which is the bottleneck in heat recovery (figure 5) whereas the ΔT_{min} the minimum temperature difference between hot and cold process streams only (Process Pinch). In retrofit studies, pinching matches are identified by minimum value of EMAT in the HEN and are referred as Network Pinch. Network Pinch Techniques are applied to overcome these bottlenecks [4].

For retrofit, EMAT and ΔT_{min} are treated as same to achieve and set the energy targets. This approach is defined and adopted by Uday V. Shenoy "Heat Exchanger Network Synthesis, Process Optimization by Energy and Resource Analysis" [4].

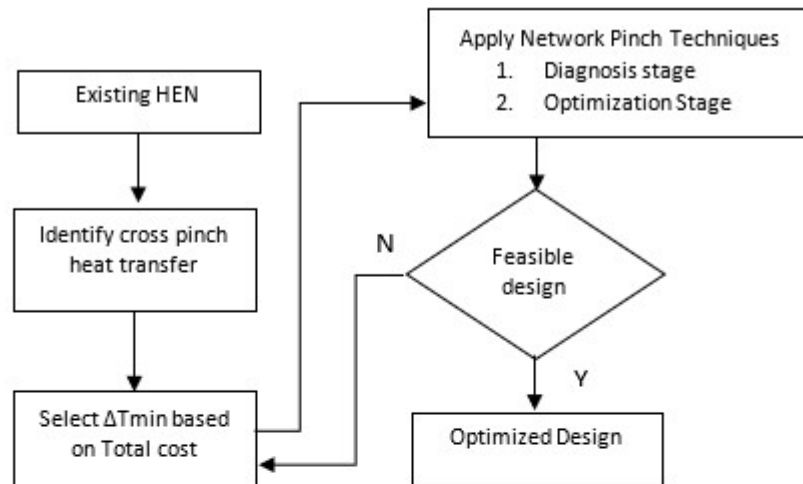


Figure 3. Sequence of Retrofit approach

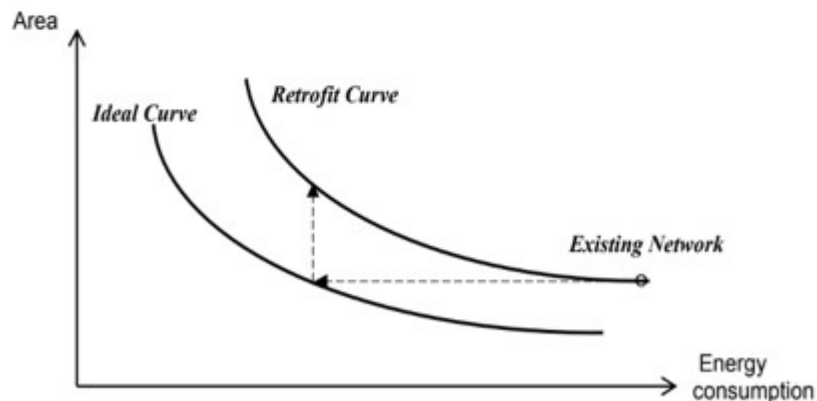


Figure 4. Ideal and Retrofit curve for Retrofit Pinch Analysis

After setting the targets for the heat recovery that will decide the Pinch Temperature, network pinch techniques are applied to carry out the retrofit to increase the heat recovery. Below figures 3 and 4 plots the composite curve with heat exchangers. Figure 3 shows the pinching match after heat recovery.

Network Pinch Techniques

Pinching Matches creates a bottleneck in heat recovery. Network pinch techniques are applied to overcome these bottlenecks. Asante and Zhu [5] proposed following network pinch techniques in the form of topology modifications capable to overcome this network pinch (bottlenecks).

- Re-sequencing
- Re-piping

- Addition of new heat exchanger
- Stream split
- Generally, these techniques are applied in same sequence. As a rule, to overcome the network pinch heat to be moved from below to above the pinch.
- Re-Sequencing
- Moves heat from below to above the pinch to de-bottleneck HEN.
- Streams remain the same.
- Scope of heat recovery increases through utility path.

Below figures 6a, 6b, 6c and 6d will define the re-sequence activities to overcome network pinch.

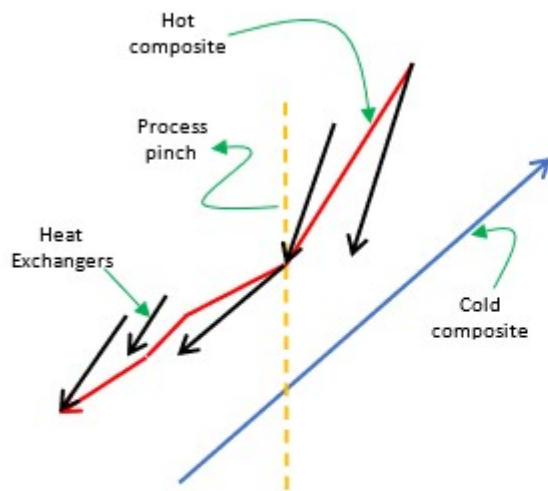


Figure 4. Composite curve with Process Pinch and heat exchangers

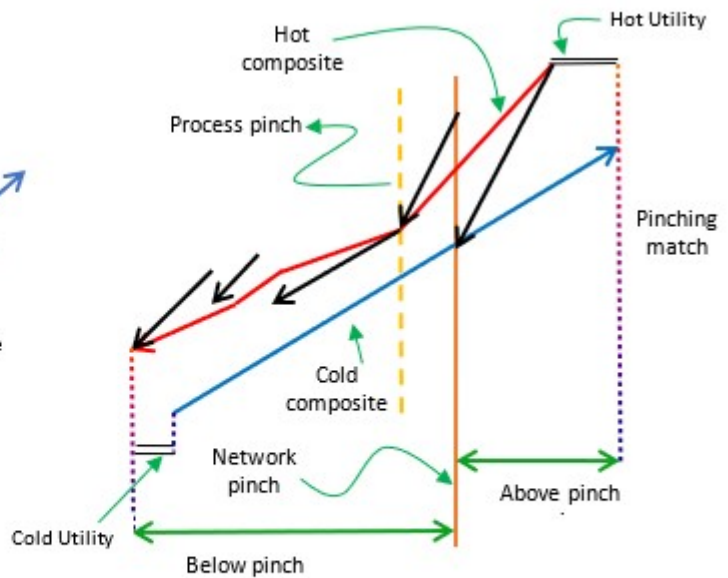


Figure 5. Network Pinch with Pinching

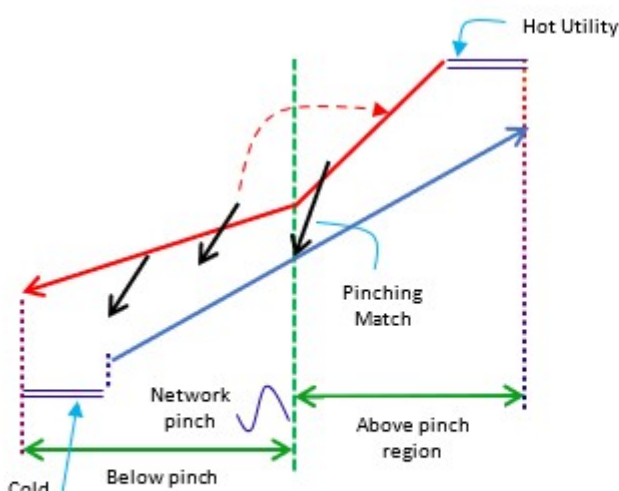


Figure 6a: Composite curve showing

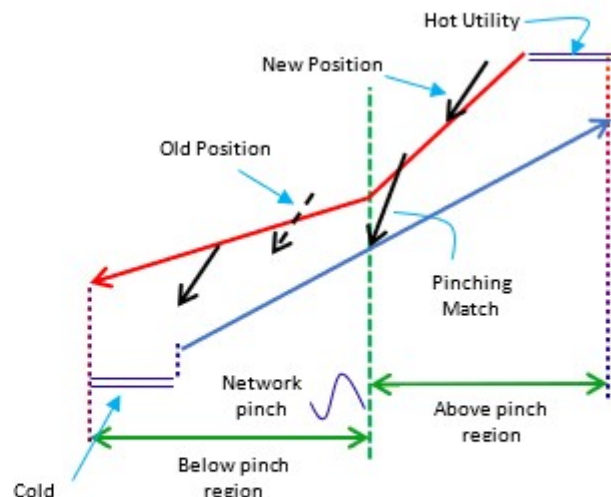


Figure 6b: Moving heat from below to above the pinch by Re-sequencing

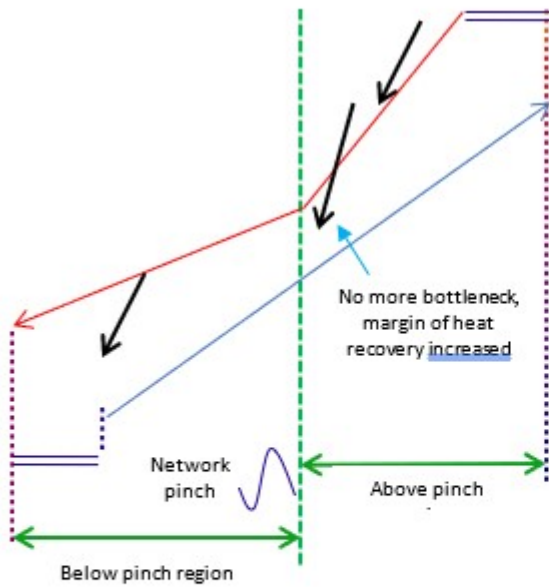


Figure 6c: No more bottleneck in increasing

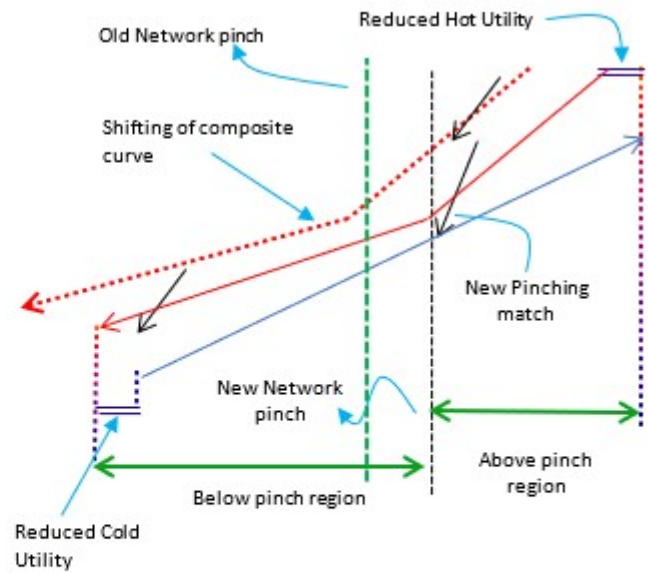


Figure 6d: Increasing heat recovery and reduced utilities with new pinching match

This re-sequence is shown in below grid diagram figure 7a and 7b to increase heat recovery in the loop shown.

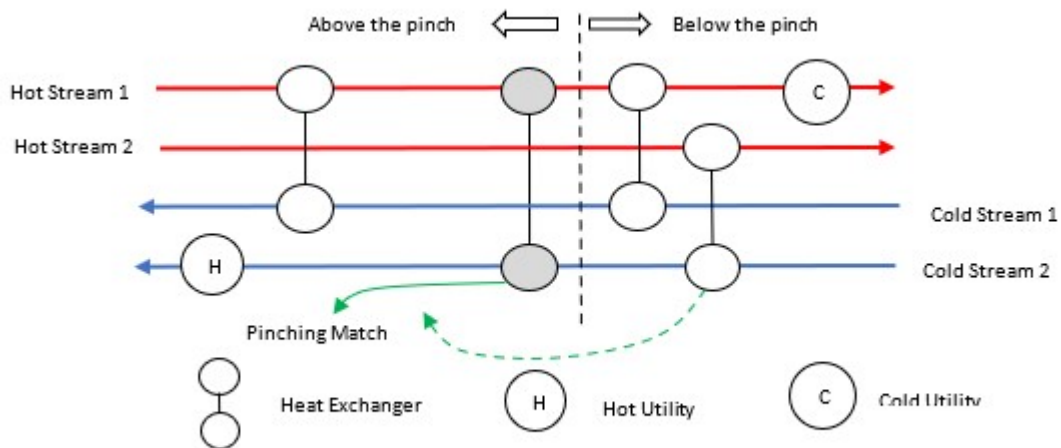


Figure 7a: Grid Diagram showing heat exchangers and

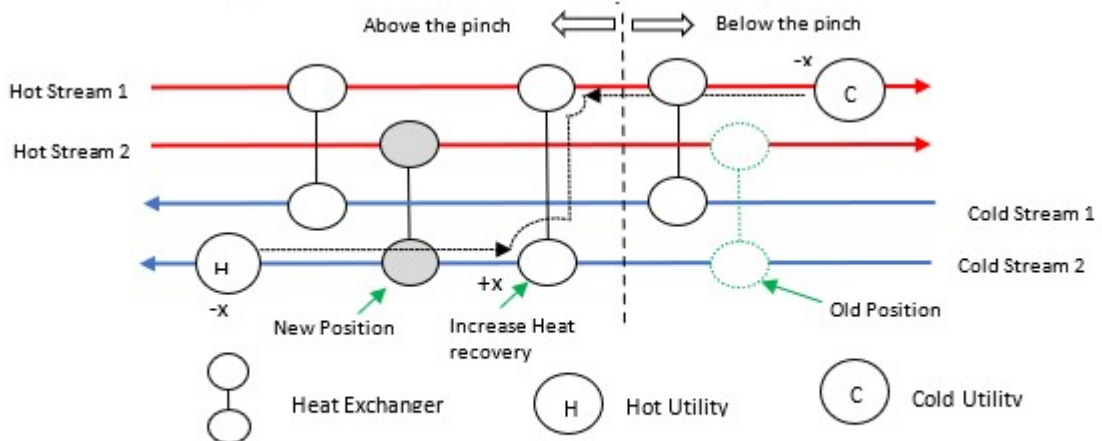


Figure 7b: Grid Diagram showing resequencing of heat exchangers from below to above the pinch.

- Re-piping: It is like re-sequencing however the streams are not the same like in re-sequencing. Streams can be different depending on the requirement like material etc. Re-piping provides more flexibility than re-sequencing.
- Addition of New Heat Exchanger: If a new match is inserted such that the heat duty on the hot stream adjacent to the pinching match is decreased and replaced by the

new match, then the position of the pinching match can be changed such that it is no longer pinching. This introduces scope to exploit the utility path to reduce the utility consumption of the network, until it is again pinched. The network is now pinched again, but at a lower utility consumption. Figure 8a, 8b, 8c and 8d illustrates it.

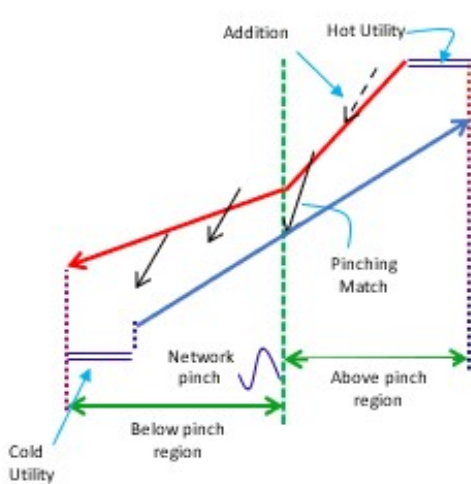


Figure 8a: Composite curve showing

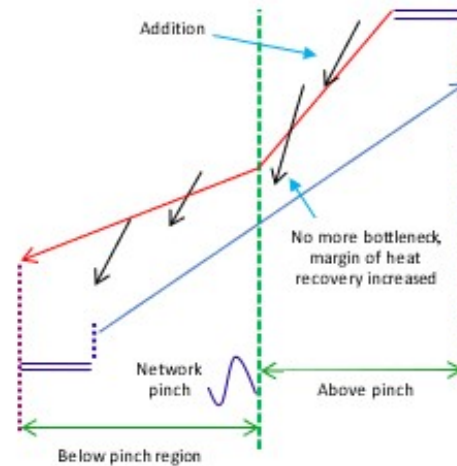


Figure 8b: Moving heat from below to above the pinch by addition of heat exchanger

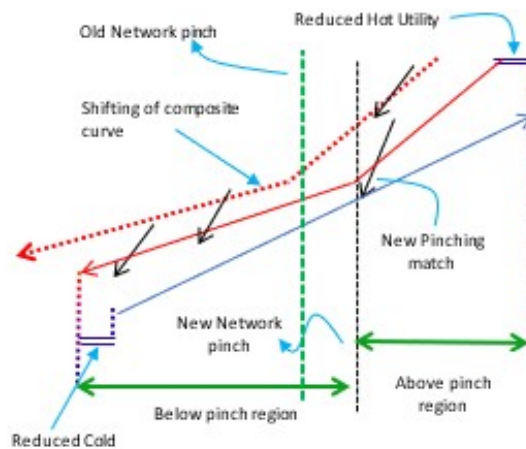


Figure 8c: Increasing heat recovery and reduced utilities with new pinching match

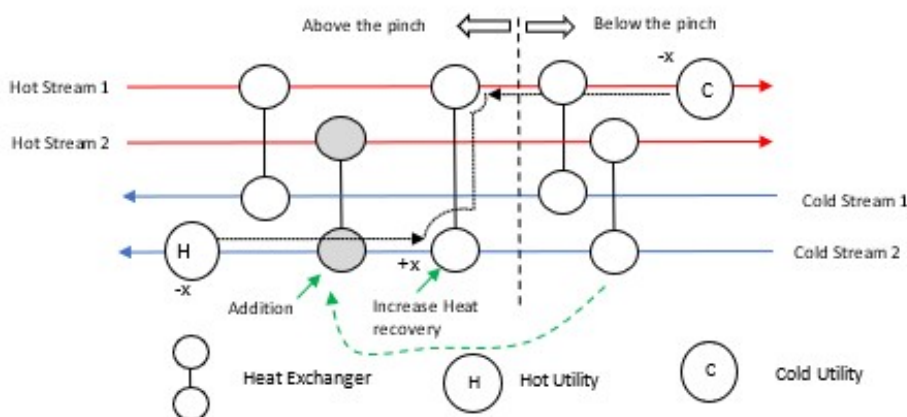


Figure 8d: Grid Diagram showing addition of heat exchangers and loop for heat

- Stream Splitting: In this case, two matches are pinched simultaneously. By introducing a stream split, the cold stream profiles in the two pinched units are now such that one of the pinching matches is no longer pinched. This means that there is scope to exploit a utility path and reduce the energy consumption. Below figures 9a, 9b, 9c, 9d and 8e illustrates it.

In general, the heat to be moved from below the network pinch to above the network pinch to

overcome the network pinch (bottleneck). Figure 10 shows this approach.

After each modification of retrofit an optimization is carried out to optimize the area and utility as per the set targets [1]. Figure 11 shows the area and utility changes after modifications and optimization for each step. Finally, after no more heat recovery is feasible, optimization is stopped.

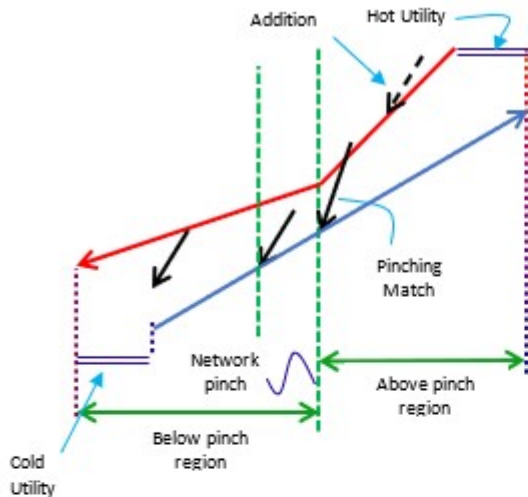


Figure 9a: Composite curve showing

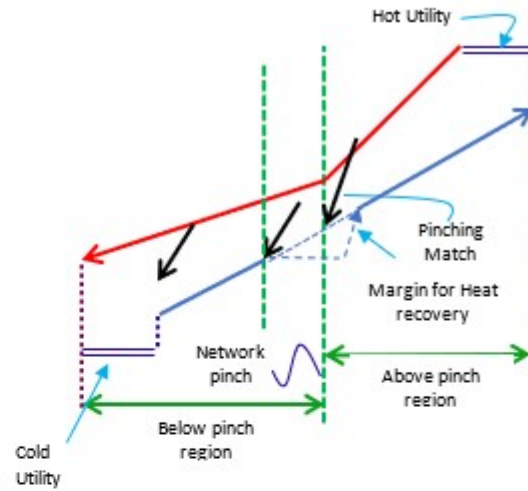


Figure 9b: Moving heat from below to above the pinch by Stream splitting

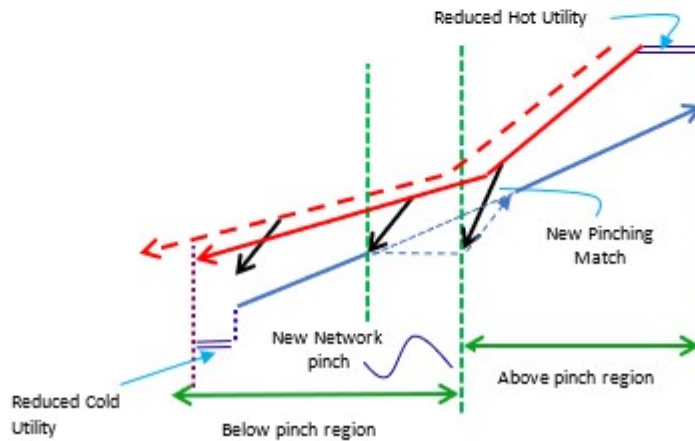


Figure 9c: Increasing heat recovery and reduced utilities with new pinching match

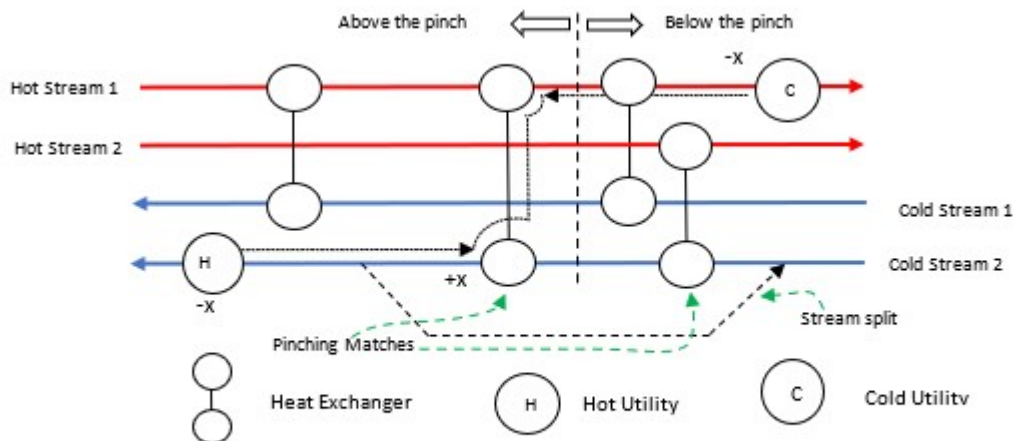


Figure 9d: Grid Diagram showing stream split with pinching match

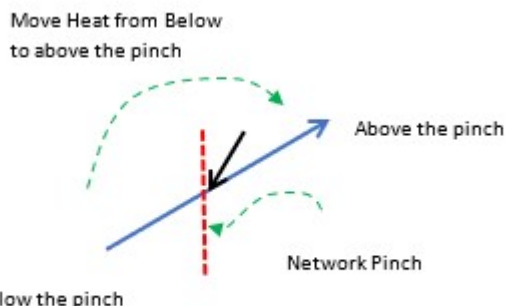


Figure 10: General rule to overcome network pinch

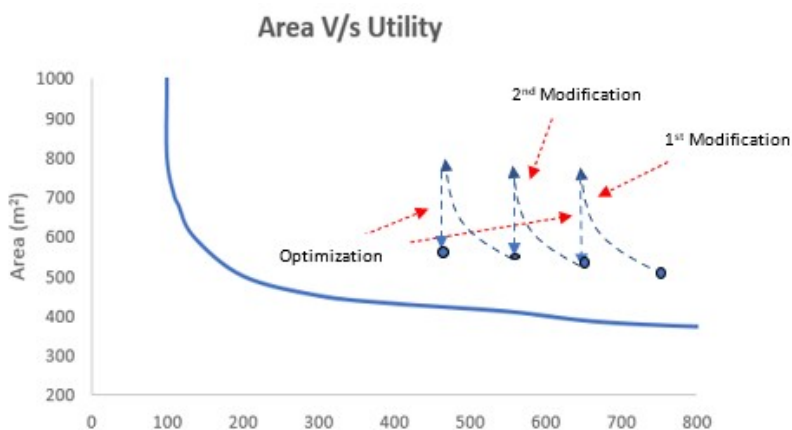


Figure 11: Area and utility curve showing the retrofit modifications and optimization

Case Study

The crude pre-heat --- -train of Crude Distillation Unit of 30,000 BPD crude capacity has been selected for a retrofit study to increase heat recovery. After identifying the cross pinch heat transfer and bottlenecks in heat

recovery (Network Pinches), the network pinch techniques are applied to increase heat recovery. Figure 12 shows the flow sheet of existing process, Figure 17, 18 shows the retrofit methodology and retrofit design after modifications.

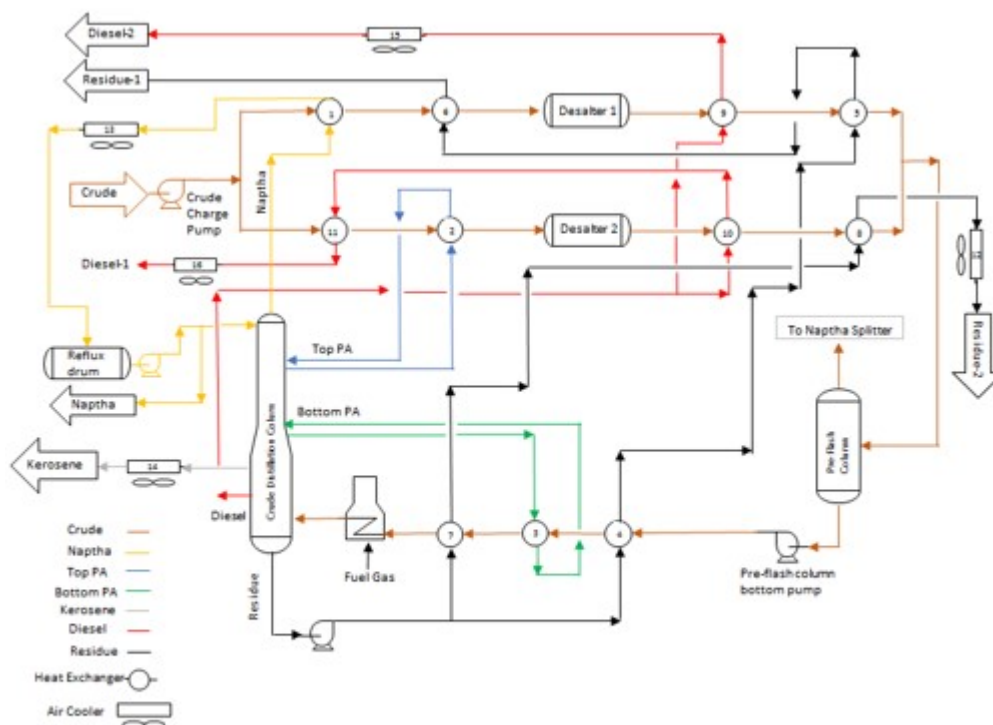


Figure 12: Flow Sheet of Existing CDU Process

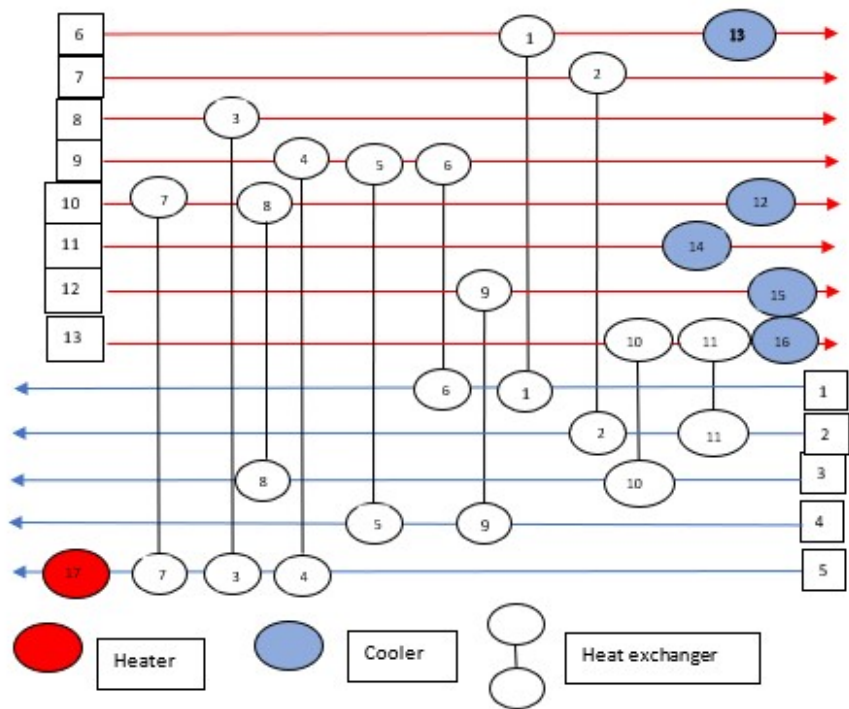


Figure 13: Grid Diagram of existing HEN of CDU showing heat exchangers, Heater and coolers

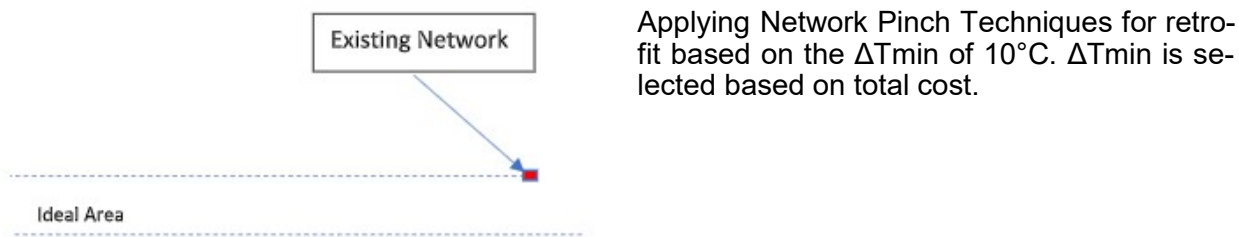


Figure 14: Existing Area and Idea Area curve

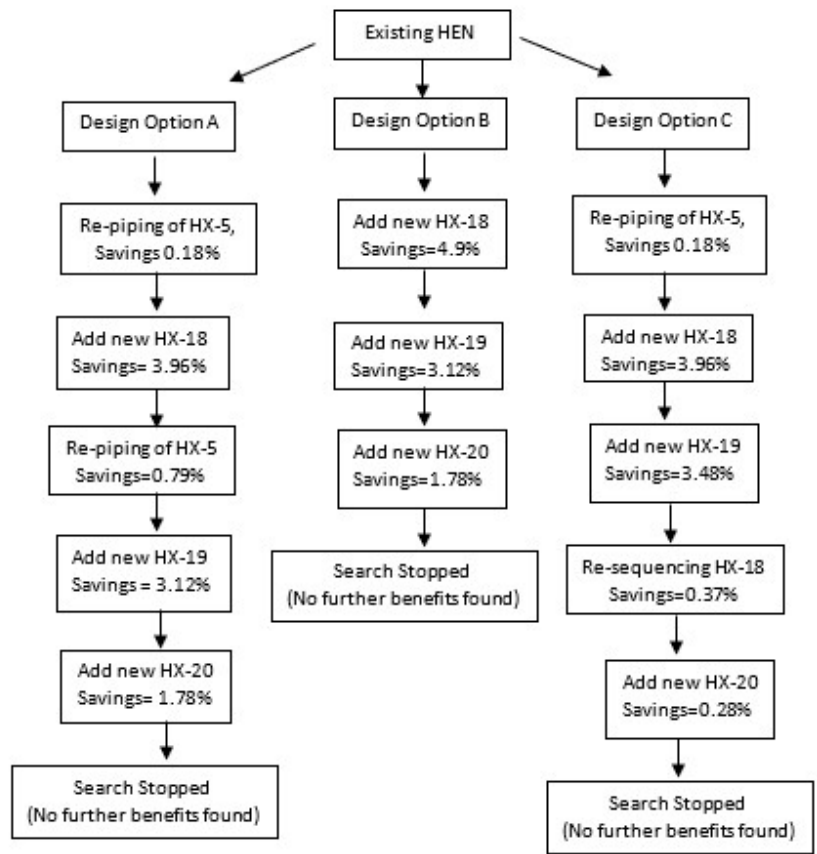


Figure 15: Network Pinch Diagnostic Techniques applied to Existing HEN

Comparison of 3 options diagnosed

Design	Δt_{min}	Investment	Savings	Area	ΔA	Existing Hot Utility	Target Hot Utility	Achieved Hot Utility	Energy Savings Achieved	Pay Back Period
	$^{\circ}C$	Million US\$	Million US\$/y	m^2	m^2	KW	KW	KW	%	Years
Existing	28			3962		14232				
Option A	10	1.2	1.53	5354	1440	-	9929	11482	64	0.83
Option B	10	1.2	1.53	5354	1440	-	9929	11482	64	0.83
Option C	10	1.27	1.53	5354	1440	-	9929	11482	64	0.96

Change in Area and utility after Retrofit

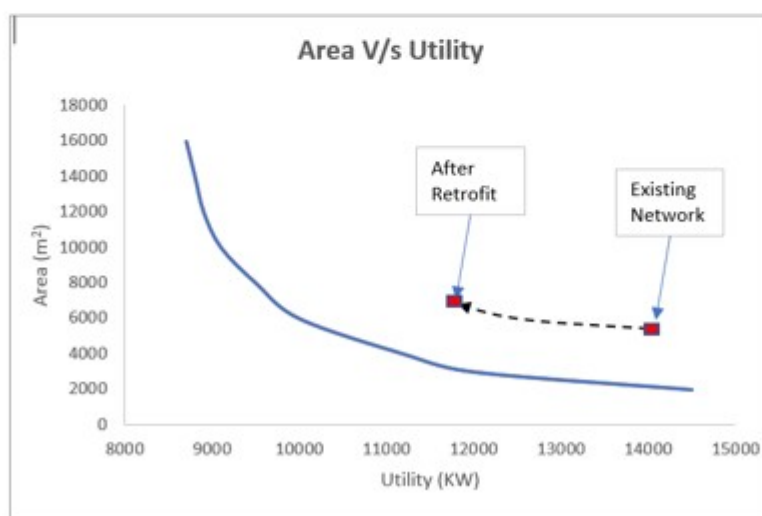


Figure 16: Improvement in Existing HEN after Retrofit

Endings

Out of 3 different options for network retrofit, option B is selected based on the minimum capital, minimum modifications, and less payback time. Option A has same payback time however the extent of modifications is higher than option B and cost of re-piping a heat exchanger would also increase cost of option A. figure 14 shows the increase in area efficiency after retrofit as the utility is reduced which slight increase in area.

The proposed retrofit design achieved total energy savings of 5,500 KW i.e. 2,750 KW hot and 2,750 KW cold utility which is approximately 64% savings of target utilities. The total utility cost savings obtained is 1.53million \$/ year which equivalent to 23% reduction in utility bill with existing bill. The cost of re-piping and re-sequencing are counted which can also contribute to cost which can slightly increase the payback. Below figure 17 and 18 show the modified HEN after retrofit.

References

- Adel S. Ashaibani and Iqbal M. Mujtaba, "Minimisation of fuel energy wastage by improved heat exchanger network design – an industrial case study", *Asia-Pac. J. Chem. Eng.*, vol. 2, pp. 575–584, Oct. 2007.
- Mamdouh A. Gadalla, Zarkoolujic, Peter J. Jansens, Megan jobson and Robin Smith, "Reducing CO2 Emissions and Energy Consumption of Heat-Integrated Distillation Systems", *Environmental Science & Technology* Vol. 39, No. 17, Jul. 2005.
- Robin Smith, "The nature of chemical process design and integration" in *Chemical Process Design and Integration*, John Wiley & Sons Ltd. UK, 2005, pp. 7-11.
- Amiya K. Jana, "Heat integrated distillation operation", *App. Energ.* Vol. 87, pp. 1477–1494, Nov. 2009.

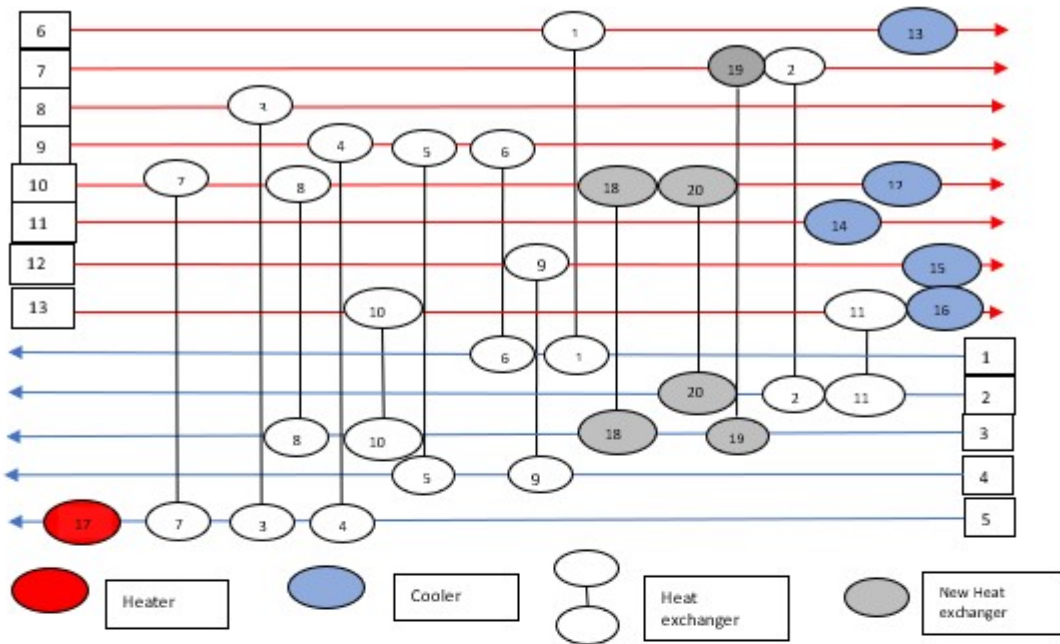


Figure 17: Modified Grid Diagram of HEN of CDU after retrofit

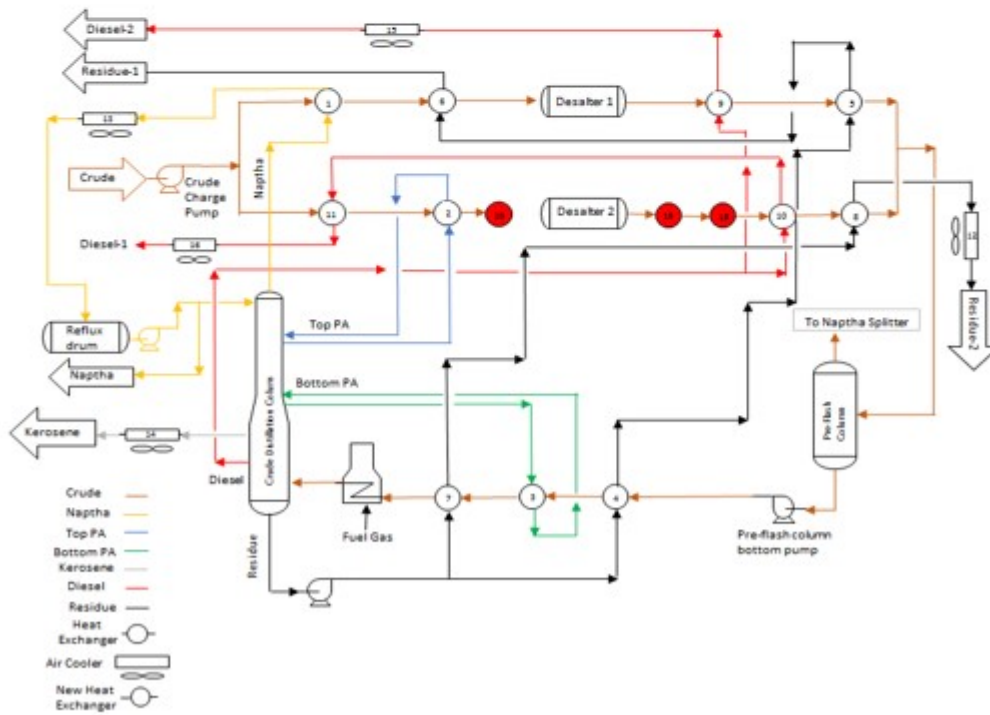


Figure 18: Modified CDU Process Flow sheet after Retrofit

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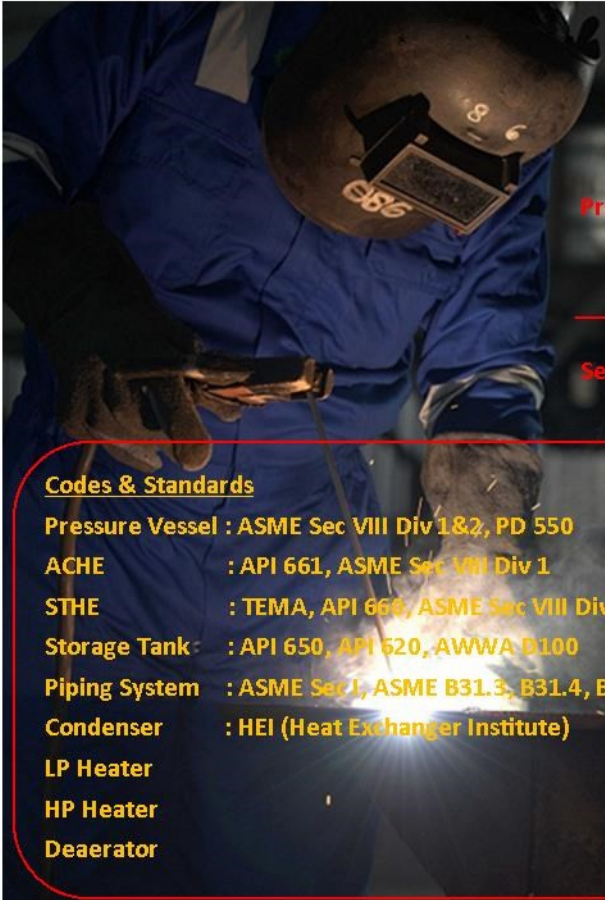
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Naphtha to Chemicals Refining Routes as Strategy to Face the “Red Ocean” of Gasoline Market

Dr. Marcio Wagner da Silva

Introduction and Context

The current scenario present great challenges to the crude oil refining industry, prices volatility of raw material, pressure from society to reduce environmental impacts and refining margins increasingly lower. The newest threat to refiners is the reduction of the consumer market, in the last years became common, news about countries that intend to reduce or ban the production of vehicles powered by fossil fuels in the middle term, mainly in the European market. Despite the recent forecasts, the transportation fuels demand is still the main revenues driver to the downstream industry, as presented in Figure 1, based on data from Wood Mackenzie Company.

According to Figure 1, is expected a growing demand by petrochemicals while the transportation fuels tend to present falling consumption. Still according to Wood Mackenzie data,

presented in Figure 2, due to the higher added value, the most integrated refiners tend to achieve higher refining margins than the conventional refiners which keep the operations focused on transportation fuels.

NCM = Net Cash Margins

The improvement in fuel efficiency, growing market of electric vehicles tends to decline the participation of transportation fuels in the global crude oil demand. New technologies like additive manufacturing (3D printing) have the potential to produce great impact to the transportation demands, leading to even more impact over the transportation fuels demand. Furthermore, the higher availability of lighter crude oils favors the oversupply of lighter derivatives that facilitate the production of petrochemicals against transportation fuels as well as the higher added value of petrochemicals in comparison with fuels.

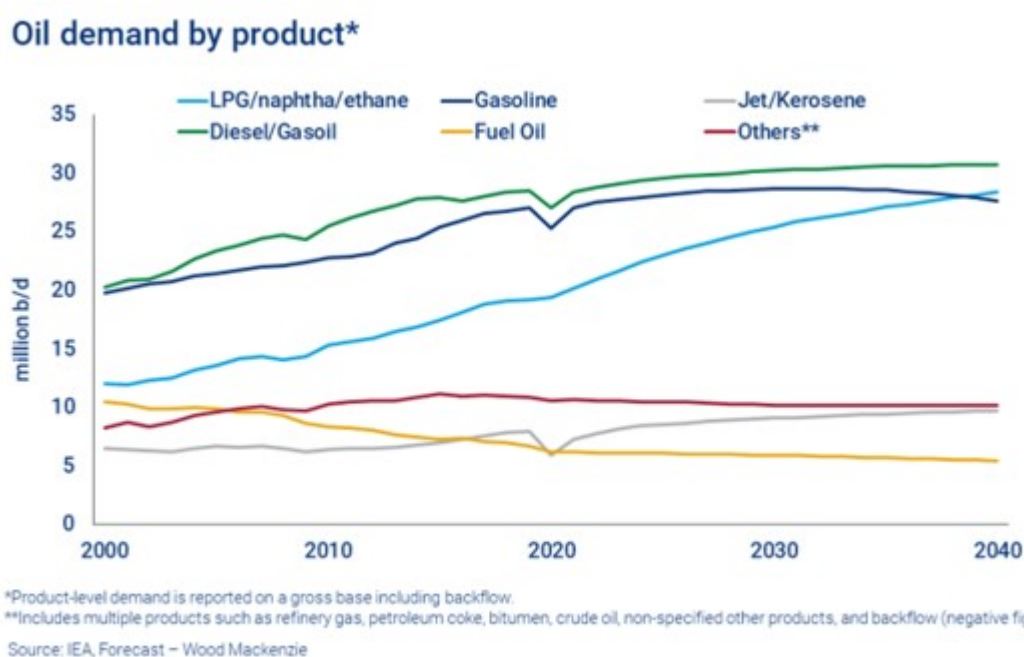
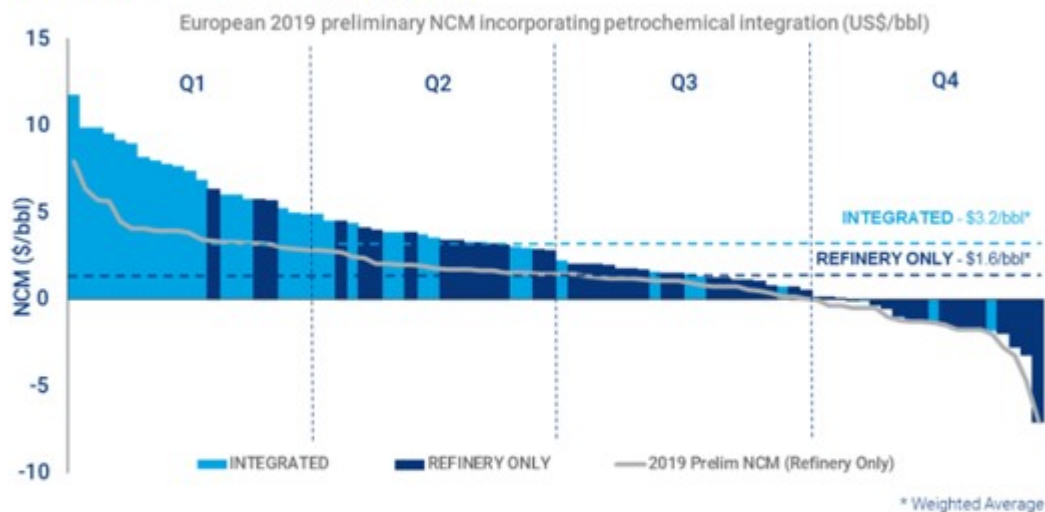


Figure 1 – Global Oil Demand by Derivative (Wood Mackenzie, 2020)

Petrochemical integration almost doubles the average European refinery net cash margin (NCM)



Source: Wood Mackenzie

Figure 2 – Refining Margins to Integrated and Non-Integrated Refining Hardware (Wood Mackenzie, 2020)

Figure 3 presents an overview of the trend of growing to the petrochemical market in short term.

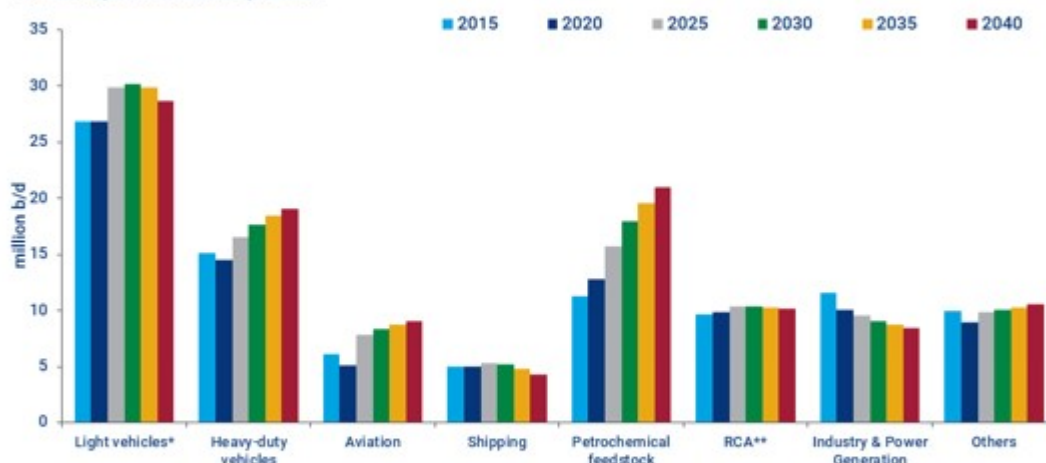
As presented in Figure 3, some of the most promising petrochemical intermediates are the aromatics benzene and p-xylene. The maximization of aromatics in the refining hardware is possible through the installation of catalytic reforming technologies associated with a separation unit.

Beyond the aromatics production, in markets with surplus of gasoline, some alternatives like blend the heavier fraction of naphtha with

diesel and jet fuel can be an interesting strategy, but this alternative presents limitations due to the middle distillates specifications like volatility and Reid Vapor Pressure (RVP). In this case, technologic routes capable of managing naphtha molecules aiming to direct these streams to petrochemical intermediates can ensure closer integration with petrochemical assets as well as higher added value to refiners. Some markets already are facing the gasoline surplus, in these cases, directing naphtha to petrochemicals against gasoline can be an attractive way to ensure competitiveness to refiners. It's interesting to quote

Petrochemicals feedstock leads demand growth in the long run – while fuel demand from light vehicles will start to fall

Global liquids demand by sector



Source: Wood Mackenzie Macro Oils Long Term Outlook H1 2020 * includes two-wheelers ** Residential, Commercial and Agriculture *** includes non-energy use (other than petrochemical feedstock) and refinery fuel, etc.

Figure 3 – Growing Trend in the Demand by Petrochemical Intermediates (Wood Mackenzie, 2020)

Regional crude oil distillation unit (CDU) capacity and integrated refinery capacity (million b/d)

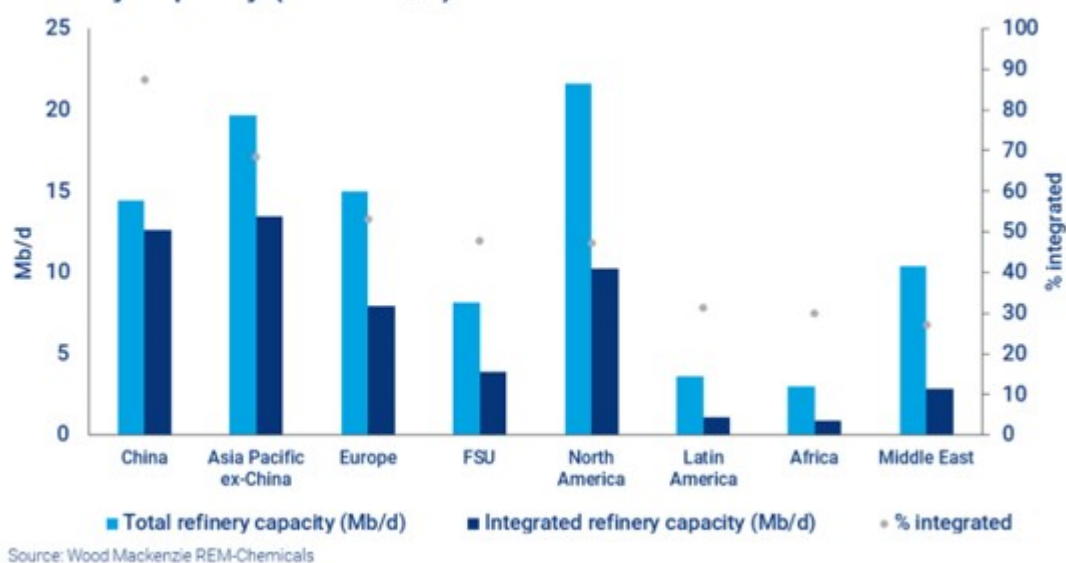


Figure 4 – Crude Oil Distillation Capacity and Integrated Refinery Capacity for Each Continent (Wood Mackenzie, 2023)

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

Figure 4 shows that the Asian players have a superior integration capacity of their refining assets in comparison with another continents, as mentioned above, this can be translated in a significant competitive advantage to the

Asian players and a great potential of competitive imbalance of the downstream market considering the recent forecasts which indicates growing demand for petrochemicals. Furthermore, it's possible to see the power of the China in the Asian and global downstream market.

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

Petrochemical sector drives demand growth in the medium term

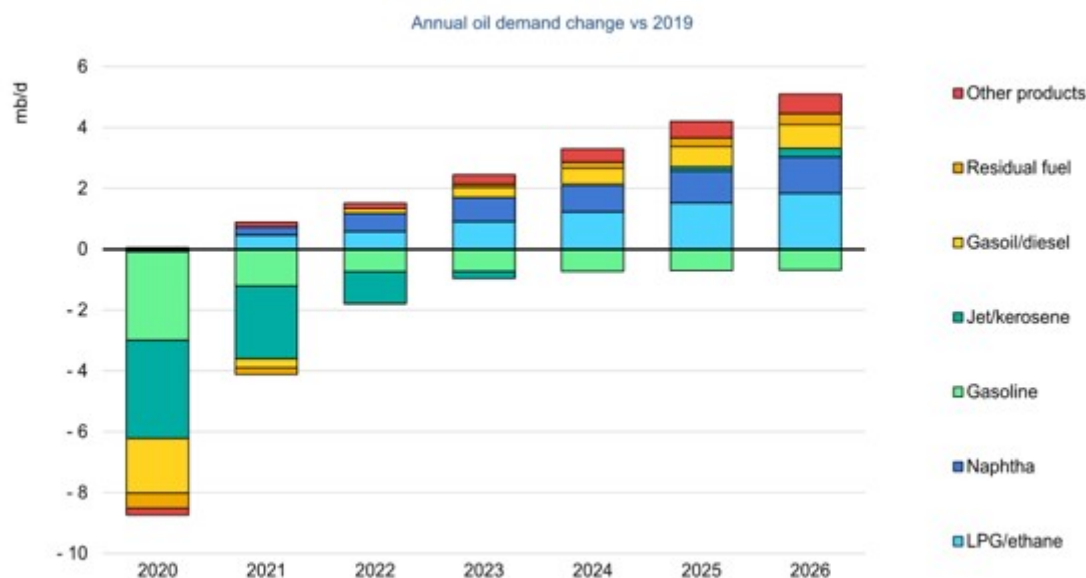


Figure 5 – Growth of Petrochemicals as Driver for Crude Oil Consumption (IEA, 2021)

Gasoline's future is outside the OECD

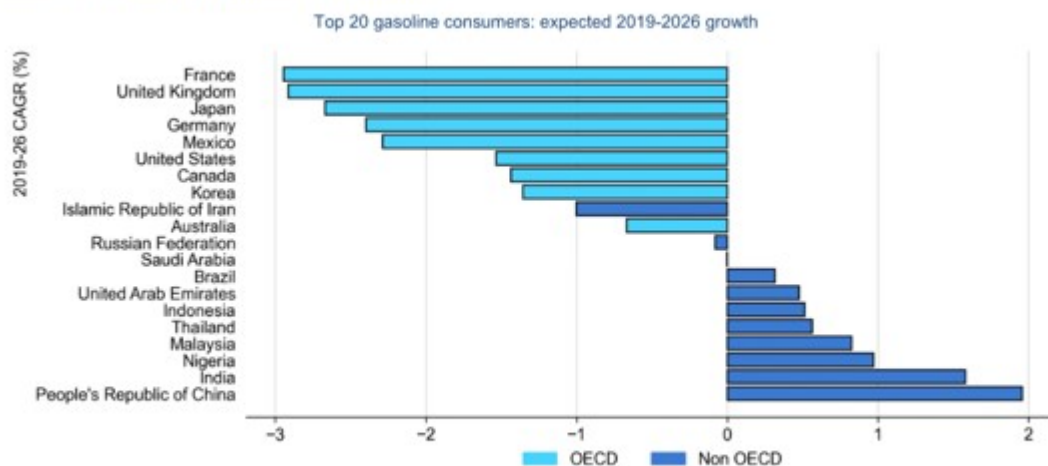


Figure 6 – Growth of Gasoline Demand for the Next Years (IEA, 2021)

Mackenzie, the Asian Market presents the major concentration of integrated refining plants.

According to the recent forecasts, the Asian market tends to respond by 90 % of the expected growth of world crude oil consumption considering the 2019 to 2026 period and the most part of this growth is related to petrochemicals demand. According to data from Asian Downstream Insights (2021) is expected an annual growth of 4,25 % in the petrochemicals demand in the Asian Market considering next five years, this scenario is leading the players of the Asian downstream sector to lead the world efforts to petrochemical

integration, this is translated into high capital spending into crude to chemicals refineries, especially in China.

As presented in Figure 5, the petrochemicals demand tends to drive the crude oil demand for the next years.

Additionally, it's important to quote that the gasoline demand will be sustained by the in developing economies, as presented in Figure 6.

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

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

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

Due to his characteristics, the transportation fuels market can be imagined like the red ocean, where the margins tend to be low and under high competition between the players with low differentiation capacity. On the other side the petrochemicals sector can be faced like the blue ocean where few players are able to meet the market in competitive conditions, higher refining margins, and significant differentiation in relation to refiners dedicated to transportation fuels market. Figure 7 presents the basic concept of blue ocean strategy in comparison with the traditional red ocean where the players fight to market share with low margins.

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

Can be difficult to some people to understand the term "differentiation" in the downstream industry once this is a market that deal with commodities, but the differentiation here is

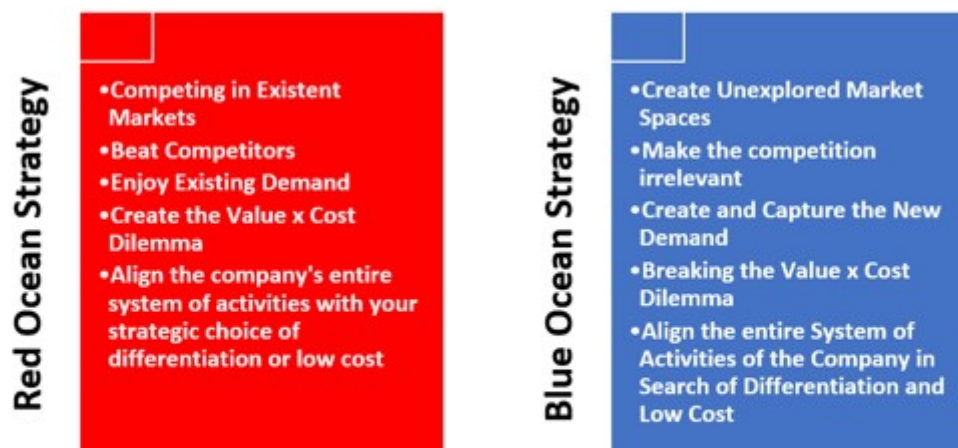


Figure 7 – Differences between Blue and Red Ocean Strategies (KIM & MAUBORGNE, 2004)

related to the capacity to reach more added value to the processed crude oil and as presented above, nowadays this is translated in the capacity to maximize the petrochemicals yield, creating differentiation between integrated and non-integrated players. In other words, it's possible to adapt the strategy to ensure more added value to the processed crude leaving the "red ocean" of transportation fuels enjoying the growing market of petrochemicals.

Higher Added Value to the Naphtha – Petrochemical Integration Concept

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

As aforementioned, the petrochemical industry has been growing at considerably higher rates

when compared with the transportation fuels market in the last years, additionally, represent a noblest destiny and less environmental aggressive to crude oil derivatives. The technological bases of the refining and petrochemical industries are similar which lead to possibilities of synergies capable of reducing operational costs and add value to derivatives produced in the refineries.

Figure 8 presents a block diagram that shows some integration possibilities between refining processes and the petrochemical industry.

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

Table 1 – Refining and Petrochemical Industry Characteristics

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

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

high once the severity of the process requires the use of material with noble metallurgical characteristics.

The IHS Markit Company proposed a classification of the petrochemical integration grades, as presented in Figure 9.

According to the classification proposed, the crude to chemicals refineries is considered the maximum level of petrochemical integration where the processed crude is totally converted into petrochemical intermediates like ethylene, propylene, and BTX.

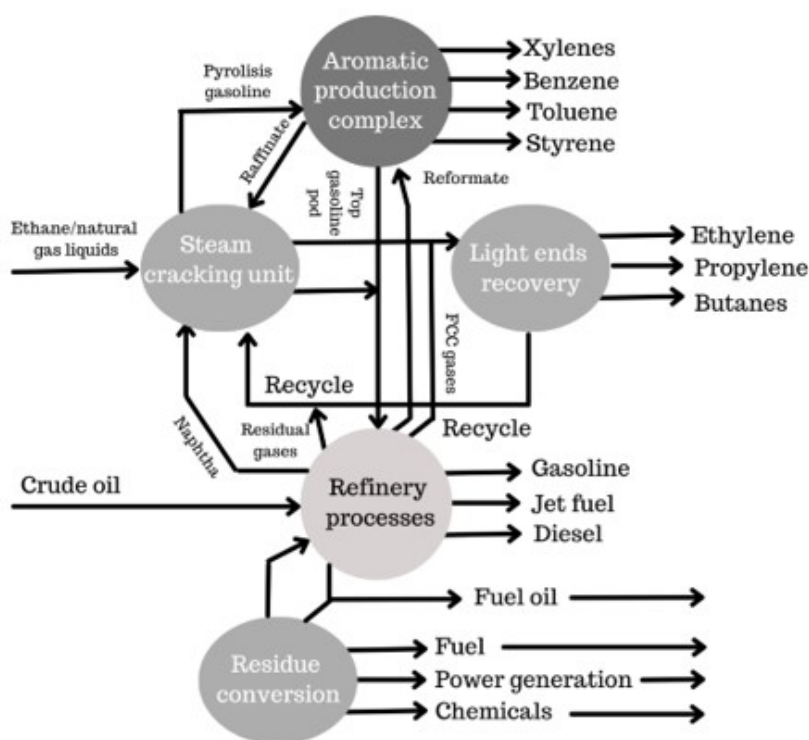


Figure 8 – Synergies between Refining and Petrochemical Processes

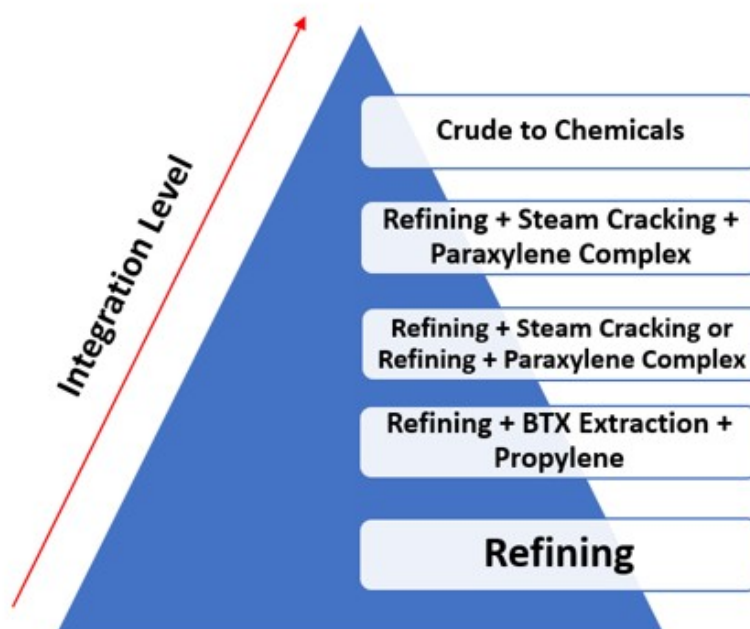


Figure 9 – Petrochemical Integration Levels (IHS Markit, 2018)

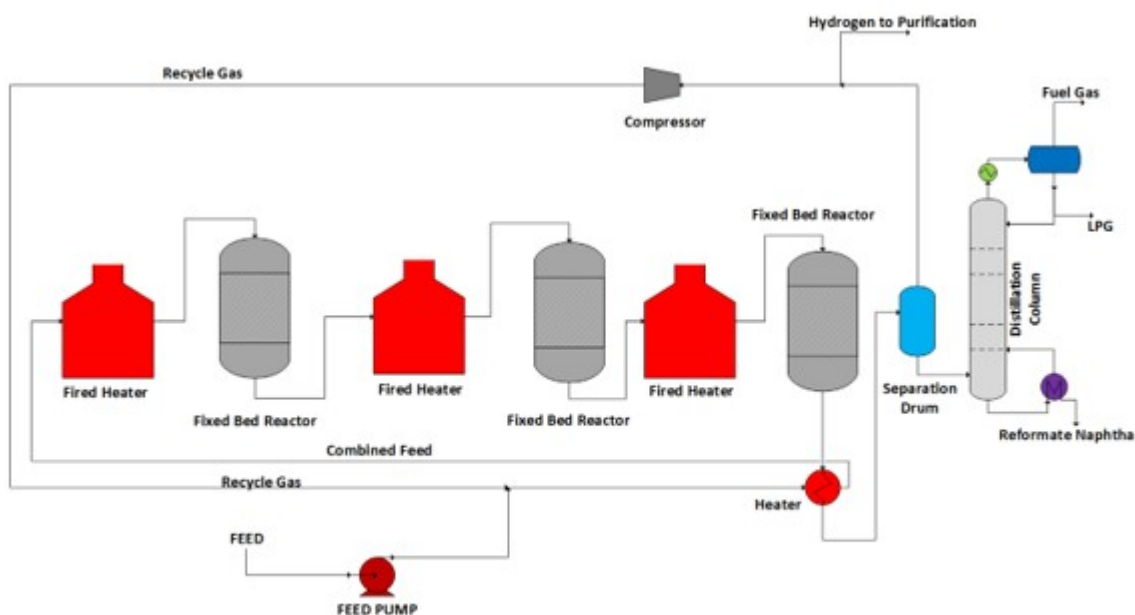


Figure 10 – Typical arrangement to Semi-regenerative Catalytic Reforming Process Unit

Catalytic Reforming Technologies – Naphtha to Aromatics

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

A typical feedstock to the catalytic reforming unit is the straight run naphtha, however, in the last decades due to the necessity to increasing the refining margin through installation of bottom barrel units, hydrotreated coke naphtha stream has been consumed like feedstock in the catalytic reforming unit.

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

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

- Naphthene Compounds dehydrogenation;
- Parafinns Isomerization;

- Isomerization of Naphthene Compounds;
- Paraffins Dehydrocyclization;

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

Figure 10 presents a basic process flow diagram for a typical semi-regenerative catalytic reforming unit.

The naphtha feed stream is blended with recycle hydrogen and heated at a temperature varying 500 to 550 oC before to enter in the first reactor, as the reactions are strongly endothermic the temperature fall quickly, so the mixture is heated and sent to the second reactor and so on. The effluent from the last reactor is sent to a separation drum where the phases liquid and gaseous are separated.

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

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

The reformate has a high aromatics content

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

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

Due to the process severity, the high coke deposition rate on the catalyst and consequently the quick deactivation leaves short operational campaign periods to semi-regenerative units that employ fixed bed reactors.

To solve this problem some technology licensors developed catalytic reforming process with continuous catalyst regeneration steps.

The process Aromizing™ developed by Axens company apply side by side configurations to the reactors while the CCR Platforming™ developed by UOP apply the configuration of stacked reactors to catalytic reforming process with continuous catalyst regeneration. Figure 11 presents a flow diagram to Aromazing™ catalytic reforming unit.

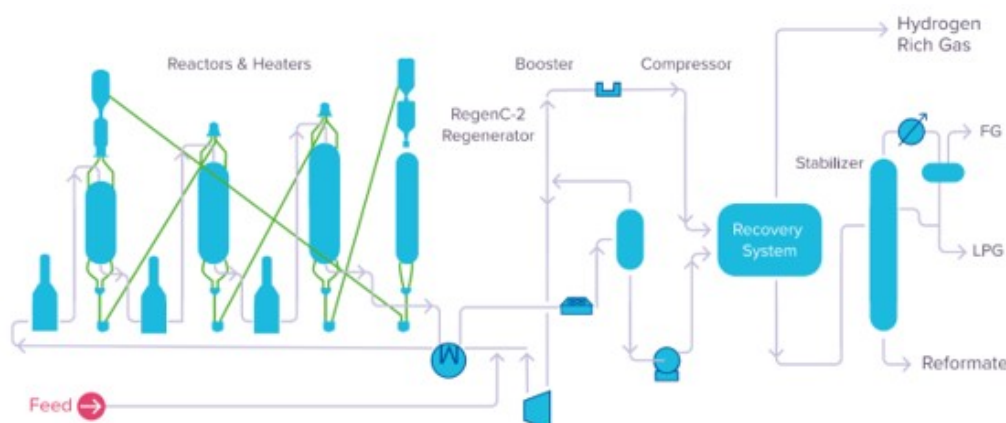


Figure 11 – Aromizing™ Reforming Technology by Axens Company

Both technologies are commercial and some process plants with these technologies are in operation around the world. Figure 12 presents a basic process flow diagram to CCR Platforming™ developed by UOP Company.

In the regeneration section the catalyst is submitted to processes to burn the coke deposited during the reactions and treated with chlorinated compounds to reactivate the acid function of the catalyst.

Despite the higher capital investment, the rise in the operational campaign and higher flexibility in relation to the feedstock to be processed in the processing unit can compensate the higher investment in relation of the semi-regenerative process.

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

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

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

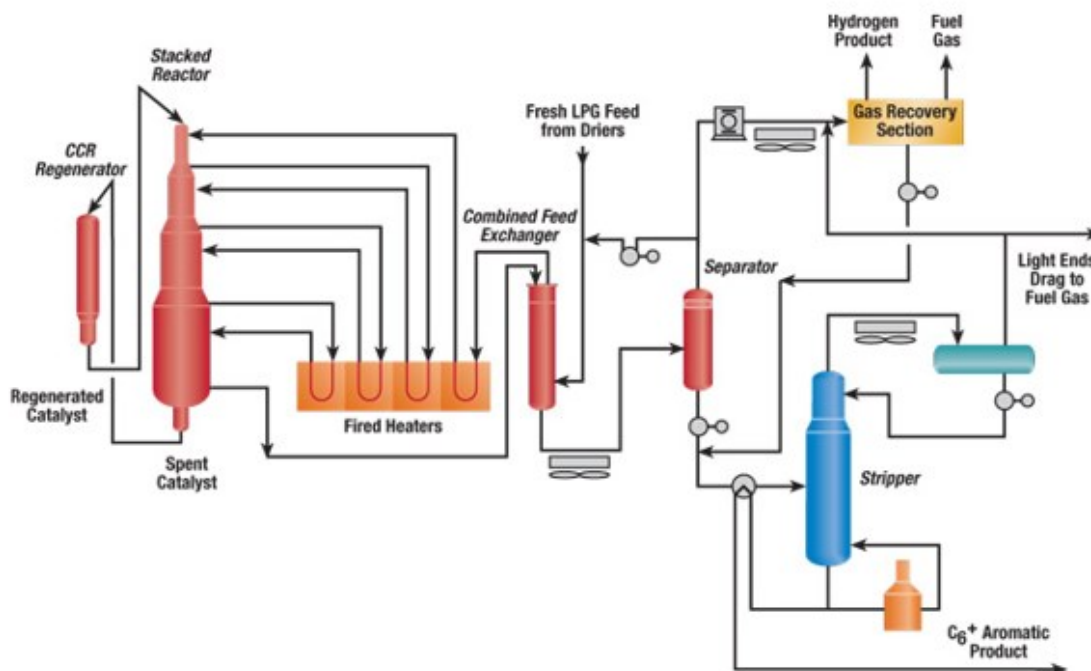


Figure 12 – CCR Platforming™ Reforming Technology by UOP Company. Source: Honeywell UOP - www.uop.honeywell.com

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

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

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

The xylenes blend is fed to a distillation column where the ethylbenzene is separated in arrangement of the xylenes isomers.

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

Ortho-xylene is separated from heavy aromatics in another distillation column while the Meta and Para-xylene are fed to a crystallization process, where is obtained a stream with a high concentration in Meta-xylene and the residual stream is directed to an isomerization unit, aiming to promote the conversion of residual Meta and Ortho-xylenes in Para-xylene. The aromatics production units are normally optimized to maximize the Para-xylene production because this is a petrochemical intermediate with higher interest, this compound is raw material to produce terephthalic acid that is used to produce PET (Polyethylene terephthalate). Figure 13 presents the chemical

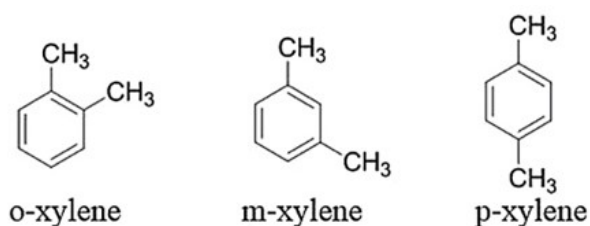


Figure 13 – Chemical Arrangement of the Xylene Isomers

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

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

The UOP Company developed an integrated aromatics complex aiming to maximize the production of benzene and p-xylene, which lead to a higher profitability to the refiner. A UOP Aromatics Complex scheme is presented in Figure 14.

Other companies have attractive and efficient technologies to produce high purity aromatics, the Axens Company license an aromatics

production complex also based on separation and conversion processes, called ParamaX™ that can be optimized to produce P-xylene. This process is presented in Figure 15.

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

As aforementioned, the capital investment to installation of aromatics production complexes is high, however, the obtained products have high added value and rely on great demand, and even the compounds with low interest can be commercialized with high margin. In countries with easy access to light oil reserves as Saudi Arabia and United States (Tight Oil) the installation of these process plants is even more economically attractive. As presented in Figure 14, the main reactions carried out in the aromatics production process aiming to improve the yield of benzene and xylenes are the toluene transalkylation presented in Figure 16 and the toluene disproportionation, presented in Figure 17.

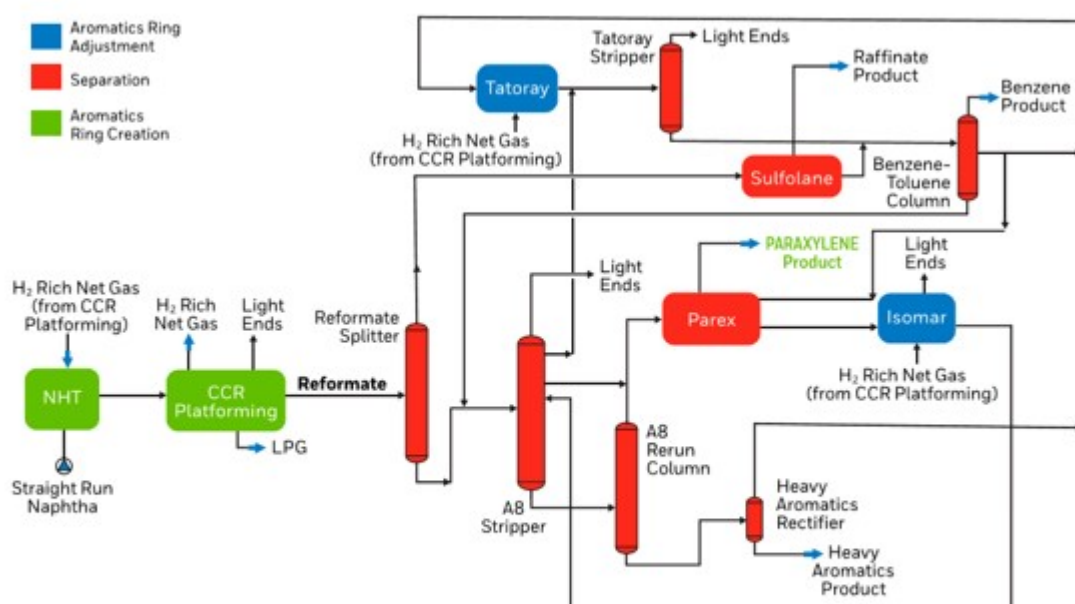


Figure 14 – Aromatics Complex by UOP Company. Source: Honeywell UOP - www.uop.honeywell.com

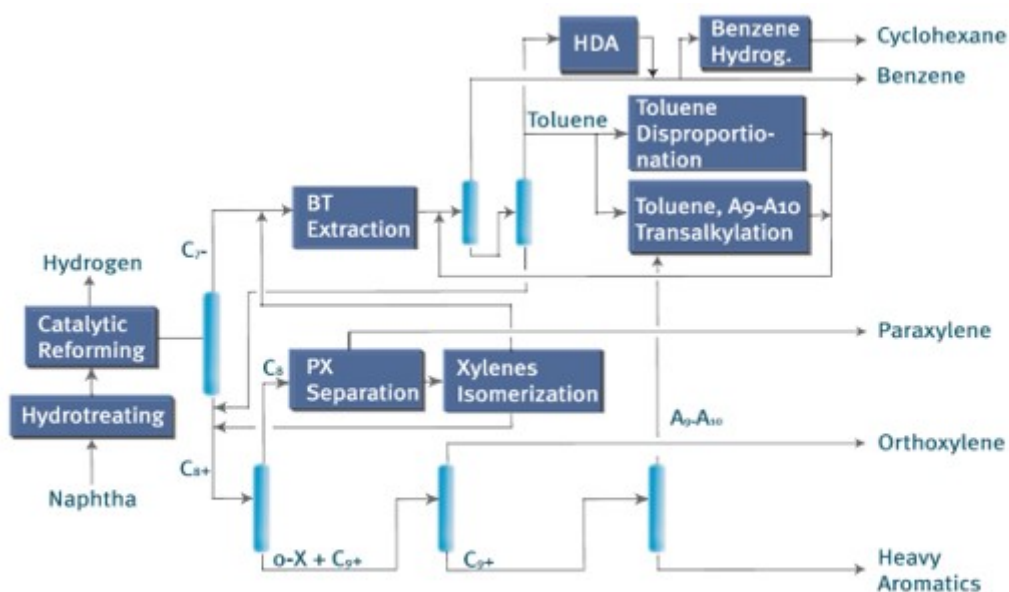


Figure 15 – Schematic Process Flow Diagram for ParamaX™ technology, by Axens Company.

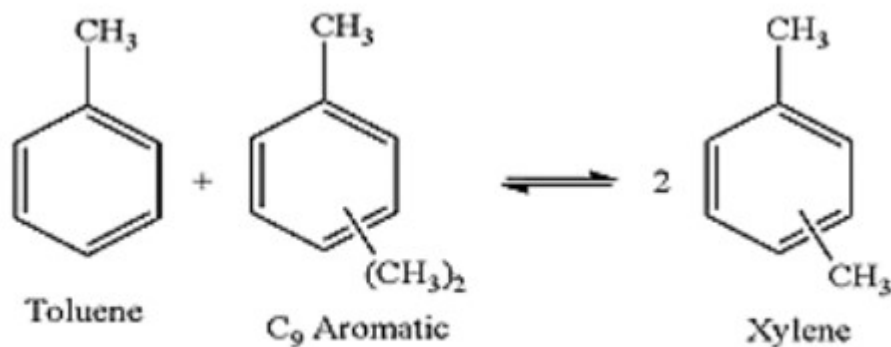


Figure 16 – Toluene Transalkylation Reaction

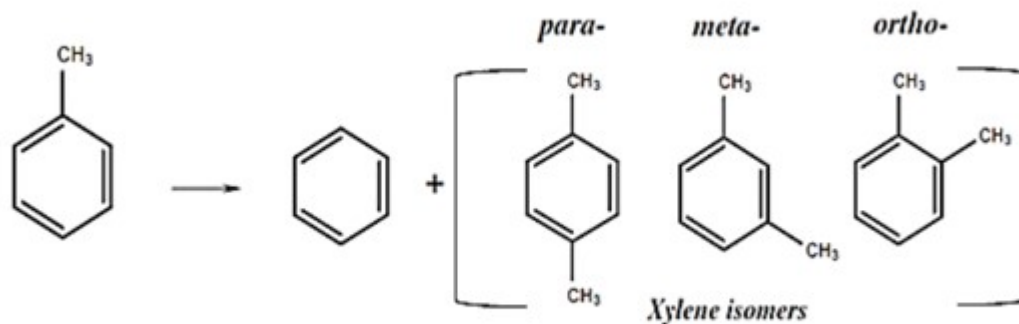


Figure 17 – Toluene Disproportionation

Fluid Catalytic Cracking Technologies – The Maximum Olefins Operation Mode

According to the market demand, the FCC units can be optimized to produce the most demanded derivatives, refiners facing gasoline surplus markets can operate the processing unit in maximum olefins operation mode, to minimize the production of cracked naphtha.

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

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

In the maximum LPG operation mode, the main restrictions are the cold area processing capacity, metallurgic limits in the hot section of the unit, treating section processing capacity as well as the top systems of main fractionating column. In markets with falling demand for transportation fuels, this is the most common FCC operation mode.

Through changing the reaction severity, it is possible to maximize the production of petrochemical intermediates, mainly propylene in conventional FCC units, as shown in Figure 18.

The use of FCC catalyst additives such as ZSM-5 can increase unit propylene production by up to 9,0%. Despite the higher operating costs, the higher revenues from the higher added value of derivatives should lead to a positive financial result for the refiner, according to current market projections. A relatively common strategy also applied to improve the yield of LPG and propylene in FCC units is the recycling of cracked naphtha leading to an over cracking of the gasoline range molecules.

The Petrochemical FCC Alternative – Complying with Current Market Demand

As quoted earlier, in markets with high demand for petrochemicals, the petrochemical FCC can be an attractive alternative to refiners aiming to ensure higher added value to bottom barrel streams. An example of FCC technology developed to maximize the production of petrochemical intermediates is the PetroFCC™ process by UOP Company, this process combines a petrochemical FCC and separation processes optimized to produce raw materials to the petrochemical process

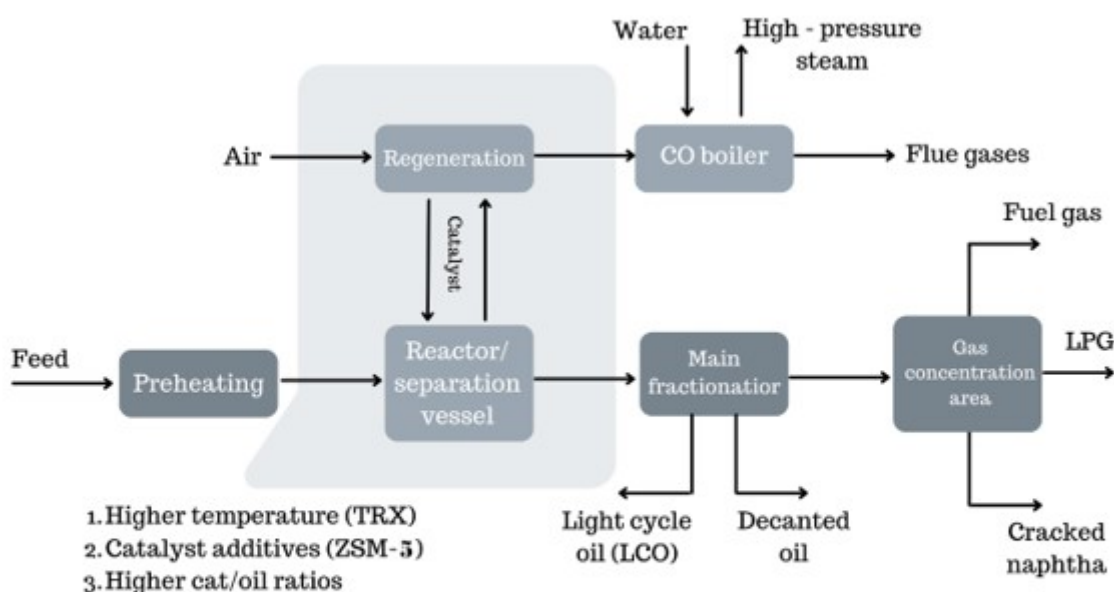


Figure 18 - Optimization of Process Variables in FCC Units to Improve the Yield of Petrochemicals Intermediates.

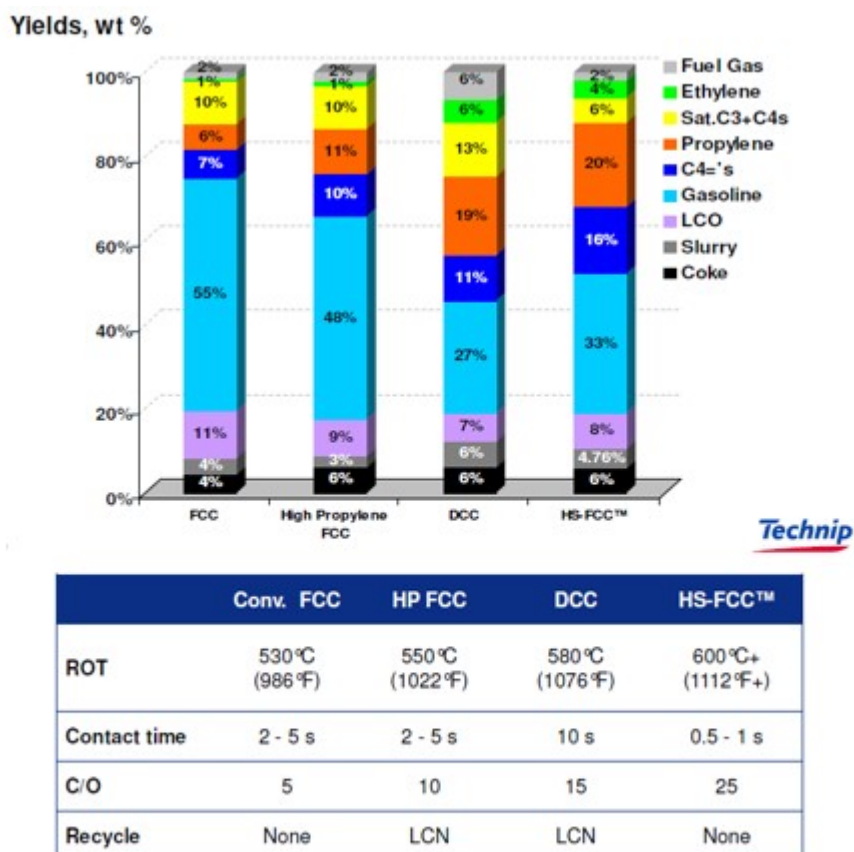


Figure 19 – Comparative Study between Conventional FCCs and Petrochemical FCC (HS-FCC™)

plants. Other available technologies are the HS-FCC™ process commercialized by Axens Company, and INDMAX™ process licensed by Lumus Company.

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

Figure 19 presents the results of a comparative study, carried out by Technip Company, showing the yields obtained by conventional FCC units, optimized to olefins (FCC to olefins), and the HS-FCC™ designed to maximize the production of petrochemical intermediates.

It's observed a higher reaction temperature (TRX) and a cat/oil ratio five times higher when are compared the conventional process units and the petrochemical FCC (HS-FCC™), leading to a growth of the light olefins yield (Ethylene + Propylene + C4+'s) from 14 % to 40%.

The installation of petrochemical catalytic cracking units requires a deep economic study taking into account the high capital investment and

higher operational costs; however, some forecasts indicate growth of 4,0 % per year to the market of petrochemical intermediates until 2025. In this scenario can be attractive the capital investment aiming to raise the market share in the petrochemical sector, allowing then a favorable competitive positioning to the refiner, through the maximization of petrochemical intermediates. Figure 20 presents a block diagram showing a case study demonstrating how the petrochemical FCC unit, in this case the INDMAX™ technology by Lumus Company, can maximize the yield of petrochemicals in the refining hardware.

In refining hardware with conventional FCC units, further than the higher temperature and catalyst circulation rates, it's possible to apply the addition of catalysts additives like the zeolitic material ZSM-5 that can raise the olefins yield close to 9,0% in some cases when compared with the original catalyst. This alternative raises the operational costs, however, as aforementioned can be economically attractive considering the petrochemical market forecasts.

Among another petrochemical FCC technologies, it's possible to quote the Maxofin™ process developed by KBR Company and the

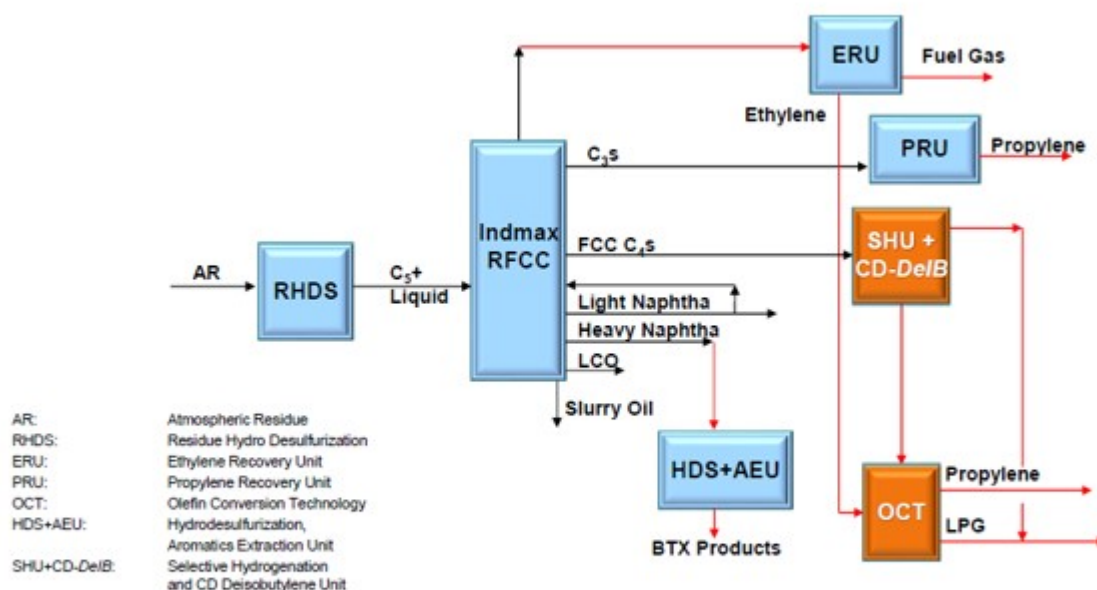


Figure 20 – Olefins Maximization in the Refining Hardware with INDMAX™ FCC Technology by Lummus (SANIN, A.K., 2017) © Copyright 2023 by Lummus Technology, LLC. All rights reserved. 2023

SCC™ technology developed by Lummus Company.

Due to the higher production of light olefins, mainly ethylene, another important difference between conventional and petrochemical FCC units is related to the gas recovery section, while in conventional FCC is applied absorber columns, in petrochemical units is applied cryogenic processes though refrigeration cycles in similar conditions which are applied in steam cracking units.

The cryogenic processes applied to olefins recovery raises, even more, the capital requirement to petrochemical FCC units when compared with conventional FCCs, despite this, the growing market for petrochemicals and falling demand for transportation fuels, tends to compensate the higher investment.

Propylene Recovery Section

The growing demand by petrochemicals lead some refiners to install propylene recovery units aiming to allow the maximization of light olefins yield in his refining hardware. Among the light olefins, the propylene is one of the most relevant petrochemical intermediate due to the high demand and added value.

The propylene can be applied as intermediate to the production some fundamental products, for example:

- Acrylonitrile;
- Propylene Oxide;

- Cumene;
- Acrylic Acid;
- Polypropylene;

Propylene can be produced through conventional processes like Steam Cracking and Fluid Catalytic Cracking (FCC) or through directed processes like metathesis of ethylene and butane, propane dehydrogenation, olefins cracking, Methanol to Olefins processes (MTO), among others. Currently a major part of the propylene market is supplied by steam cracking units, but close to 28 % of the global propylene demand is from the separation of LPG produced in Fluid Catalytic Cracking Units (FCC).

Normally, the LPG produced in FCC units contains close to 30 % of propylene and the added value of the propylene is close to 2,5 times of the LPG. According to the local market, the installation of propylene separation units presents an attractive return over investment. Despite the advantage, a side effect of the propylene separation from LPG is that the fuel stays heavier leading to specifications issues, mainly in colder regions, in these cases alternatives are to segregate the butanes and send this stream to gasoline pool, add propane to the LPG or add LPG from natural gas. It's important to consider that some of these alternatives reduce the LPG offer, which can be a severe restriction according to the market demand.

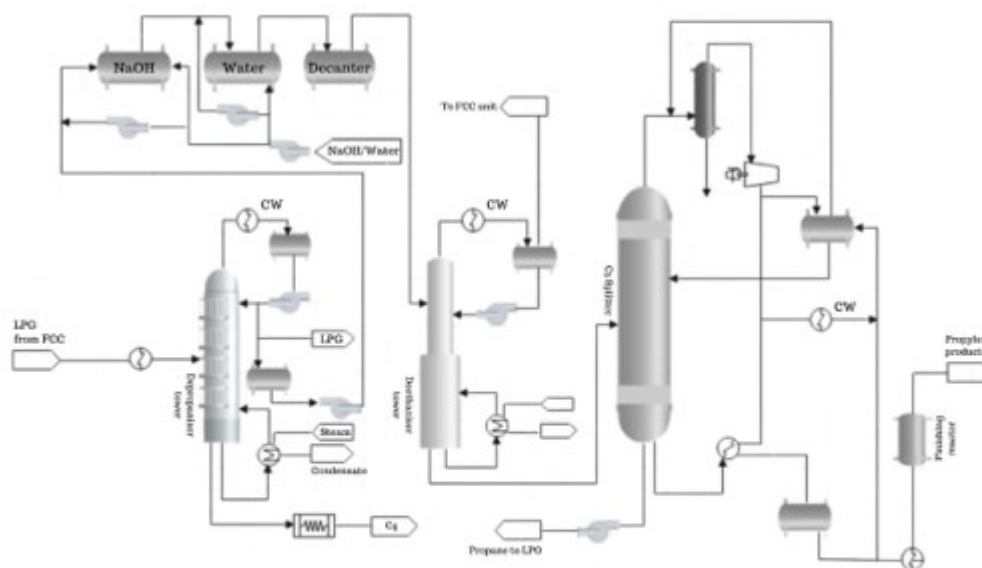


Figure 21 – Typical Process Flow Diagram for an FCC Propylene Separation Unit Applying Heat Pump Configuration

A great challenge in the propylene production process is the propane and propylene separation step. The separation is generally hard by simple distillation because the relative volatility between propylene and propane is close of 1.1. This fact generally conducts distillation columns with many equilibrium stages and high internal reflux flow rates.

There are two technologies normally employed in propylene-propane separation towers that are known as Heat-Pump and High-Pressure configurations.

The high-pressure technology applies a traditional separation process that uses a condenser with cooling water to promotes the condensation of top products, in this case, it's necessary to apply sufficient pressure to promote the condensation of products in the ambient temperature. Furthermore, the reboiler uses steam or another available hot source. The adoption of high-pressure separation route requires a great availability of low-pressure steam in the refining hardware, in some cases this can be a restrictive characteristic and the heat pump configuration is more attractive, despite the higher capital requirements.

The application of heat pump technology allows decrease the operating pressure by close of 20 bar to 10 bar, this fact increase the relative volatility propylene-propane, making the separation process easier and, consequently, reducing the number of equilibrium stages and internal

reflux flow rate required for the separation.

Normally, when the separation process by distillation is hard (with relative volatilities lower than 1.5) the uses of heat pump technology show more attractive.

Furthermore, some variables need to be considered during the choice of the best technology for the propylene separation process like availability of utilities, temperature gap in the column and installation cost.

Normally, propylene is produced in the refineries with to specifications. The polymer grade that is most common and has higher added value with a purity of 99,5 % (minimum) this grade is directed to polypropylene market. The chemical grade where the purity varies between 90 to 95% is normally directed to other uses. A complete process flow diagram for a typical propylene separation unit applying heat pump configuration is presented in Figure 21.

The LPG from FCC unit is pumped to a depropanizer column where the light fraction (essentially a mixture of propane and propylene) is recovered in the top of the column and sent to a deethanizer column while the bottom (butanes) is pumped to LPG or gasoline pool, according to the refining configuration. The top stream of the deethanizer column (lighter fraction) is sent back to FCC where is incorporated to refinery fuel gas

pool, or in some cases can be directed to petrochemical plants to recover the light olefins (mainly ethylene) present in the stream while the bottom of the deethanizer column is pumped to the C3 splitter column, where the separation of propane x propylene is carried out. The propane recovered in the bottom of the C3 splitter is sent to LPG pool where the propylene is sent to propylene storage park. The feed stream passes through a caustic wash treating aiming to remove some contaminants that can lead to deleterious effect to petrochemical processes, an example is the carbonyl sulfide (COS) that can be produced in the FCC (through the reaction between CO and S in the Riser).

Naphtha Steam Cracking Process – More Olefins and Less Fuels

The Steam cracking process has a fundamental role in the petrochemical industry, nowadays the most part of light olefins light ethylene and propylene is produced through steam cracking route. The steam cracking consists of a thermal cracking process that can use gas or naphtha to produce olefins, in this review we will describe the naphtha steam cracking process.

The naphtha to steam cracking is composed basically of straight run naphtha from crude oil distillation units, normally to meet the requirements as petrochemical naphtha the stream needs to present high paraffin content (higher than 66 %).

Due to his relevance, great technology developers have dedicated their efforts to improve the steam cracking technologies over the years, especially related to the steam cracking furnaces. Companies like Stone & Webster, Lummus, KBR, Linde, and Technip develop technologies to steam cracking process. One of the most known steam cracking technologies is the SRT™ process (Short Residence Time), developed by Lummus Company, that applies a reduce residence time to minimize the coking process and ensure higher operational lifecycle. Another commercial technology dedicated to optimizing the yield of ethylene is the SCORE™ technology developed by KBR and ExxonMobil Companies which combines a selective steam cracking furnace with high performance olefins recovery section.

The cracking reactions occur in the furnace tubes, the main concern and limitation to operating lifecycle of steam cracking units is the coke formation in the furnace tubes. The reactions carry out under high temperatures, between 500 oC to 700 oC according to the characteristics of the feed. For heavier feeds like gas oil, is applied lower temperature aiming to minimize the

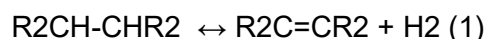
coke formation, the combination of high temperatures and low residence time are the main characteristic of the steam cracking process. Despite be possible to operate with naphtha, nowadays the steam cracking operators have chosen to operate with ethane or LPG against naphtha due to the competitive prices related to the new sources of NGL (Natural Gas Liquid), despite this trend over the last years, in markets where is observed a gasoline surplus, naphtha can still an attractive alternative as feedstock to steam crackers.

According to some forecasts, the demand by propylene will raise from 130 million metric tons in 2020 to around to 190 million metric tons in 2030. Facing the increasingly light feed to refineries and steam cracking units which tends to favor the ethylene production in detriment of propylene, the propylene demand tends to be supplied by on-purpose propylene production routes like propane dehydrogenation, methanol to olefins (MTO), and olefins metathesis.

Light Paraffin Dehydrogenation Technologies

Another alternative to improve the yield of light olefins in the refining hardware is to apply paraffin dehydrogenation technologies. Light paraffin is normally commercialized as LPG or gasoline and presents reduced added value when compared with light olefins.

Dehydrogenation process involves the hydrogen remove from paraffinic molecule and consequently hydrogen production, according to the reaction (1):



The dehydrogenation reactions have strongly endothermic characteristics, and the reactions conditions include high temperatures (close to 600 oC) and mild operating pressures (close to 5 bar). The catalyst normally applied in the dehydrogenation reactions are based on platinum carried on alumina (others active metals can be applied). Figure 22 shows a schematic process flow diagram for a typical dehydrogenation process unit.

The main processes that can produce streams rich in light paraffin are physical separation processes such as LPG from atmospheric distillation and units dedicated to separate gases from crude oil.

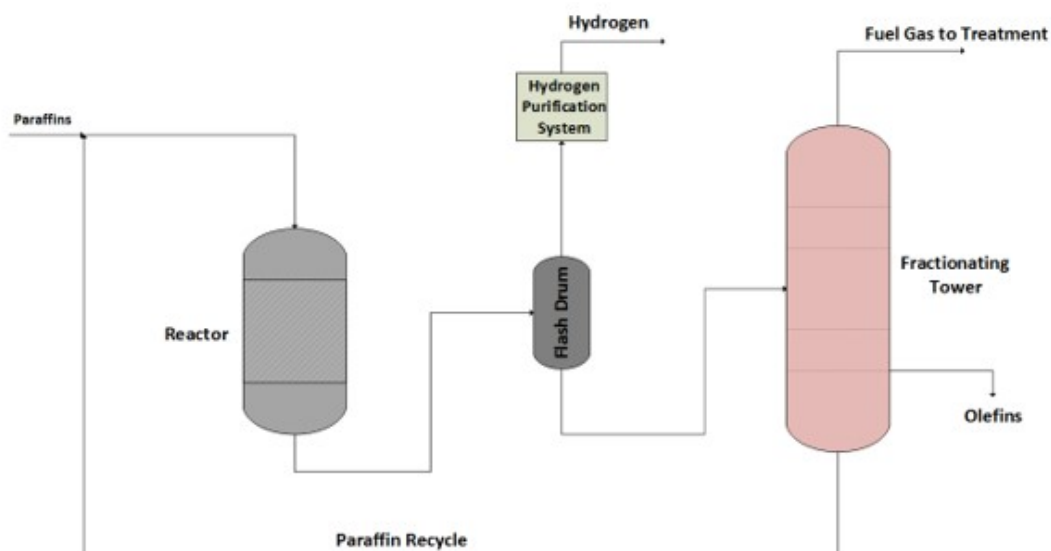


Figure 22 – Process Flow Diagram for a Typical Light Paraffin Dehydrogenation Process Unit.

The feed stream is mixed with the recycle stream before to enter to the reactor, the products are separated in fractionating columns and the produced hydrogen is sent to purification units (normally PSA units) and, posteriorly sent to consumers units as hydrotreating and hydrocracking, according to refining scheme adopted by the refiner. Light compounds are directed to the refinery or petrochemical complex fuel gas pool, after adequate treatment while the olefinic stream is directed to petrochemical intermediates consumer market.

During the dehydrogenation process there is a strong tendency to coke deposition on the catalyst surface and, periodically is carried out the regeneration of the catalytic bed through controlled combustion of the produced coke. Some process arrangements present two reactors in parallel aim to optimize the processing unit operational availability, in these cases while one reactor is in production the other is in the regeneration step.

Due to the growing market and high added value of light olefins, great technology developers have been dedicated their efforts to develop paraffin dehydrogenation technologies. The UOP company developed and commercialize the OLEFLEX™ that is capable to produce olefins from paraffin dehydrogenation with a continuous catalyst regeneration process, despite the higher initial investment, this technology can minimize the unavailability period to regenerate the catalyst. Figure 23 presents a basic process flow diagram for the OLEFLEX™ technology by UOP Company. Another paraffin dehydrogenation technology from UOP Company is the PACOL™ process.

Another available technology is the CATOFIN™ process, licensed by Lummus Company, as aforementioned, in this case, is applied two reactors in parallel, as presented in Figure 24.

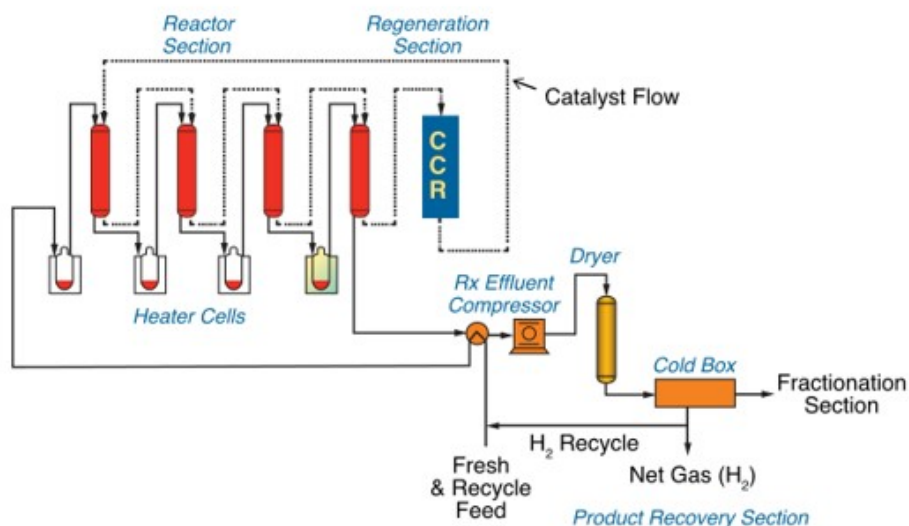


Figure 23 – Basic Process Flow Diagram for the OLEFLEX™ Technology by UOP Company. Source: Honeywell UOP - www.uop.honeywell.com

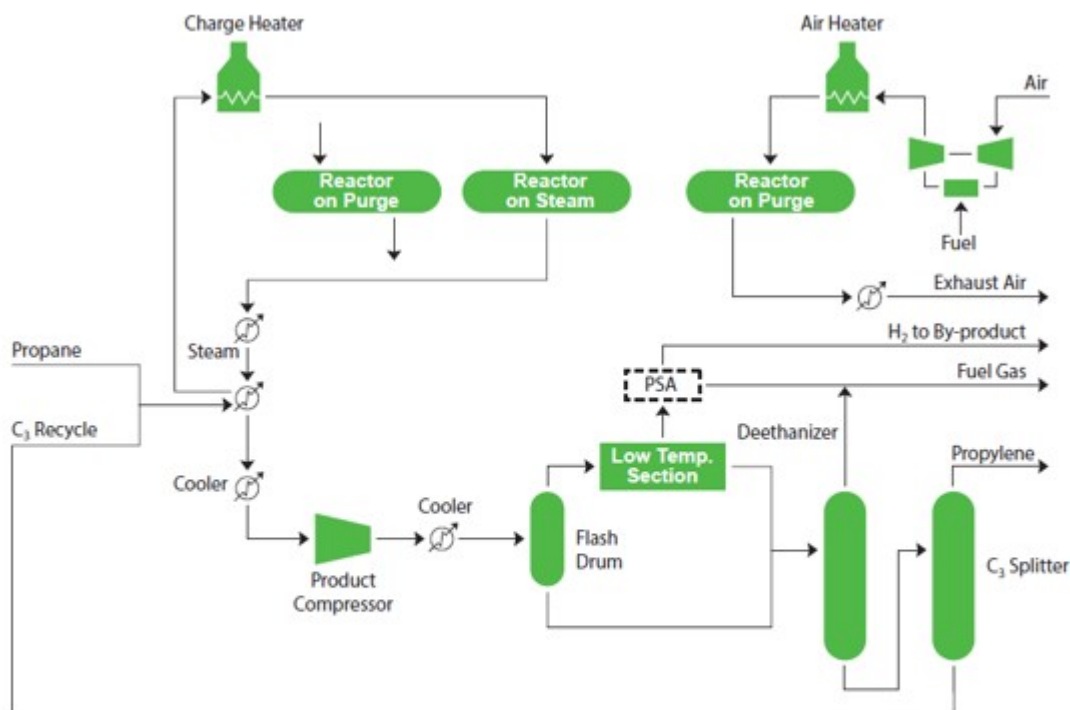
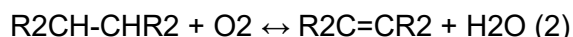


Figure 24 – Simplified Process Scheme to CATOFIN™ Dehydrogenation Technology, by Lummus Company. © Copyright 2023 by Lummus Technology, LLC. All rights reserved. 2023

Others dehydrogenation technologies available are the processes STAR™ commercialized by ThyssenKrupp-Uhde Company and the process FBD™ by SnamProgetti Company.

Due to his chemical characteristics, olefinic compounds can be employed in the production of a large quantity of interest products as polymers (polyethylene and polypropylene) propylene oxide and oxygenated compounds production intermediates (MTBE, ETBE, etc.).

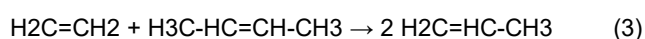
As a process of high energy consumption, there is a great variety of research in the sense of developing more active and selective catalysts that reduce the need for energetic contribution to the dehydrogenation process. One of the main variations of the dehydrogenation process is the process called oxidative dehydrogenation that occurs according to reaction 2.



This reaction is strongly exothermic, and this is the main advantage in relation to the traditional dehydrogenation process, due to the high risk of paraffin combustion against the dehydrogenation reaction.

Olefins Metathesis

The olefins metathesis process involves the combination of ethylene and butene to produce propylene as presented in reaction 3.



The main technology licensors for olefins metathesis processes are the Lummus Company, and IFP (Institut Français du Pétrole). Figure 25 presents a basic process flow arrangement for the OCT™ technology, developed by Lummus Company.

The economic viability of olefins metathesis units relies on the price gap between propylene and ethylene as well as the ethane availability in the market.

The Combination of FCC and Steam Cracking Units – Maximum Olefins Yield

As aforementioned, maximize the light olefins yield in the refining hardware can be an attractive way to ensure competitiveness in the downstream market according to the recent forecasts. The combination of FCC and steam cracking units in the refining hardware can be an alternative to achieve this goal. Table 2 presents a comparison between steam cracking and FCC technologies.

The characteristics of the FCC and steam cracking units allows high yield of olefins in the refining without competition for feedstocks, once the FCC is a bottom barrel conversion technology based in carbon rejection that applies mainly gasoil as feed stream while the steam cracking process produces mainly ethylene through thermal cracking of ethane and high paraffinic naphtha.

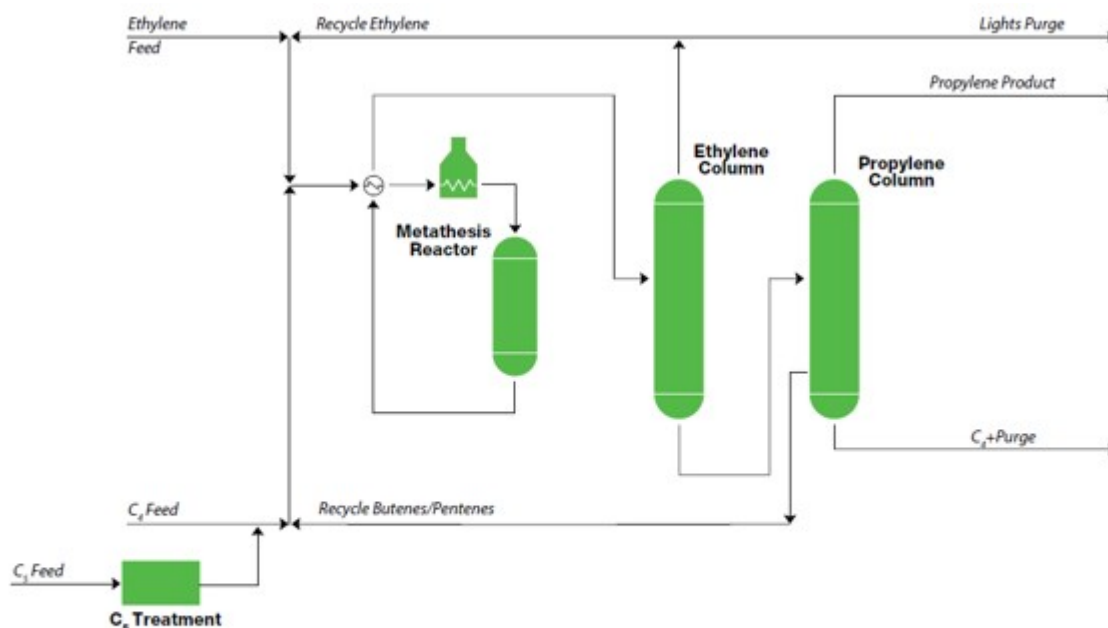


Figure 25 – Process Flow Diagram for OCT™ Olefins Metathesis Technology by Lummus Company. © Copyright 2023 by Lummus Technology, LLC. All rights reserved. 2023

Table 2 – FCC and Steam Cracking Comparison

Characteristic	FCC Technologies	Steam Cracking
Mechanism	Catalytic Process	Thermal Process
Residence Time (s)	1 to 5	0,1 to 0,5
Reaction Temperature (oC)	500 to 650 oC	700 to 900
Feedstock	Heavier Molecules (Gasoil)	Lighter Molecules (Naphtha or Ethane)
Main Production Focus	Propylene	Ethylene

The yield of propylene in the steam cracking units relies on the feedstock quality, being higher in units processing naphtha. In the last years, some refiners are adopting the ethane as main feedstock due to his lower prices, this fact reduces the propylene offer from steam crackers, raising the relevance of the propylene from FCC units to ensure the market supply. This fact has been the main driver to the growing of propylene on purpose technologies like propane dehydrogenation, methanol to olefins, and metathesis. Despite this recent trend, the steam cracking units remain the main propylene source to the market with close to 48 % of the market.

An example of refining configuration relying on FCC and steam cracking units is presented in Figure 26.

Considering the recent trend of reduction in transportation fuels demand followed by the growth of petrochemicals market makes the synergy between FCC and steam cracking units an attractive way to maximize the petrochemicals production in the refining hardware.

Although the advantages presented by closer integration between refining and petrochemical assets, it's important to understand that the players of downstream industry are facing with a transitive period where, as presented in Figure 1, the transportation fuels are responsible by great part of the revenues. In this business scenario, it's necessary to define a transition strategy where the economic sustainability achieved by the current status (transportation fuels) needs to be invested to build the future (maximize petrochemicals). Keeping the eyes only in the future or only in the present can be a strategic mistake.

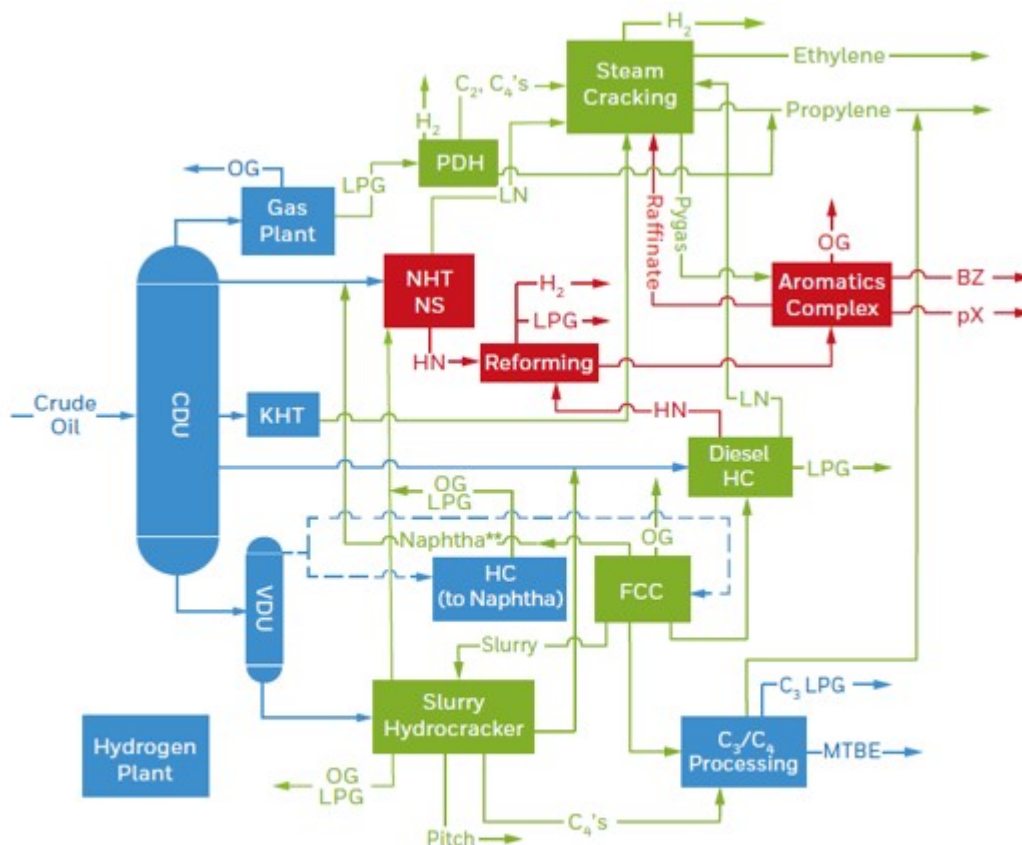


Figure 26 – Integrated Refining Scheme Base on FCC and Steam Cracking Units (UOP, 2019) Source: Honeywell UOP - www.uop.honeywell.com

Conclusion

Nowadays, is still difficult to imagine the global energetic matrix free of fossil transportation fuels, especially in developing economies. Despite this fact, recent forecasts, growing demand by petrochemicals, and the pressure to minimize the environmental impact produced by fossil fuels creates a positive scenario and acts as main driving force to closer integration between refining and petrochemical assets, in the extreme scenario the zero fuels refineries tend to grow in the middle term, especially in developed economies.

The synergy between refining and petrochemical processes raises the availability of raw material to petrochemical plants and makes the supply of energy to these processes more reliable at the same time ensures better refining margin to refiners due to the high added value of petrochemical intermediates when compared with transportation fuels. The development of crude to chemicals technologies reinforces the necessity of closer integration of refining and petrochemical assets by the brownfield refineries aiming to face the new market that tends to be focused on petrochemicals against transportation fuels, it's important to note the competitive advantage of the refiners from Middle East that

have easy access to light crude oils which can be easily applied in crude to chemicals refineries. As presented above, closer integration between refining and petrochemical assets demands high capital spending, despite this fact, the installation of refining units capable to add value to the naphtha can be a significant competitive advantage among the refiners, especially those players inserted in market with gasoline surplus. The advantage of the naphtha to chemicals routes can be exemplified through the growing “propylene gap” present in the text where the refiners capable to maximize the propylene yield both to on-purpose or traditional production (FCC of Steam Cracking) can enjoy a significant competitive advantage in the market.

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

References

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

ALBAHAR, M.Z. Selective Toluene Disproportionation over ZSM-5 Zeolite. PhD Thesis – University of Manchester, 2018.

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

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

Encyclopedia of Hydrocarbons (ENI), Volume II – Refining and Petrochemicals (2006).

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

GELDER, A. Refinery-Petrochemical Integration Disrupts Gas-Based Cracker Feedstock Advantage, Wood Mackenzie, 2023.

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

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

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

LAMBERT, N.; OGASAWARA, I.; ABBA, I.; REDHWI, H.; SANTNER, C. HS-FCC for Propylene: Concept to Commercial Operation. PTQ Magazine, 2014.

MALLER, A.; GBORDZOE, E. High Severity Fluidized Catalytic Cracking (HS-FCC™): From concept to commercialization – Technip Stone & Webster Technical Presentation to REF-COMM™, 2016.

MUKHERJEE, M.; VADHRI, V.; REVELLON, L. Step-Out Propane Dehydrogenation Technology for the 21st Century. The Catalyst Review, 2021. OYEKAN, S.O. Catalytic Naphtha Reforming Process. 1st ed. CRC Press, 2019.

Refinery-Chemicals Integration: How to Benchmark Success. Wood Mackenzie Article, 2020.

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

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

TALLMAN, M. J.; ENG, C.; SUN, C.; PARK, D. S. - Naphtha Cracking for Light Olefins Production. PTQ Magazine, 2010.

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

YOUSSEF, F.; ADRIAN, M. H.; WENZEL, S. – Advanced Propane Dehydrogenation, PTQ Magazine, 2008.

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

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How to...STRUCTURED PACKINGS

How to design and calculate Structured Packings

Dr.-Ing. Volker Engel

Due to their construction, structured packings can provide a very large surface area for mass transfer while maintaining a low pressure drop. After the introduction of sheet metal packing in the 1980s and the initial euphoria to displace all other column internals, nowadays they have found their firm place for several applications.

History

Structured packings are – compared to random packings – quite new internals, even though they were invented in the 1960s. The first patent was filed 1962 by Sulzer for corrugated wire gauze packing. In 1972 the first sheet metal packing was presented by Sulzer at the international fair ACHEMA at Frankfurt/Germany.

When the first patents expired in the 1980s, other suppliers began to develop new structured packings. Fig. 1 shows the history of the development of structured packings by different suppliers and their packing brand names.

Principle

The basic principle of a structured packing is to grant surface by corrugated material (wire gauze as well as sheet). The packing is built by corrugated elements, where every second sheet is flipped. This results in channels

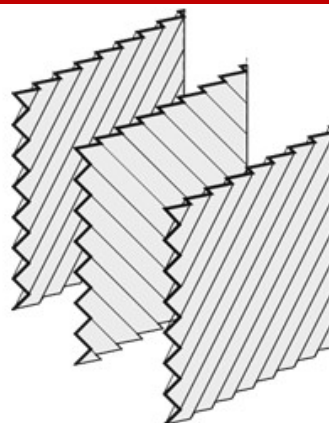


Fig. 2: Corrugated sheets of a structured packing

between the sheets (Fig. 2). The sheets are grouped to blocks (to pass the tower's man-hole) and installed in the tower so that it is filled with layers of structured packings.

Packing Parameters

The corrugated elements are normally designated (also called, 'knurled' or 'patterned') for optimal wetting and for steady renewal of the border layer of the liquid film. And they are often perforated for connecting the compartments of both sides of the elements for minimizing gas maldistribution.

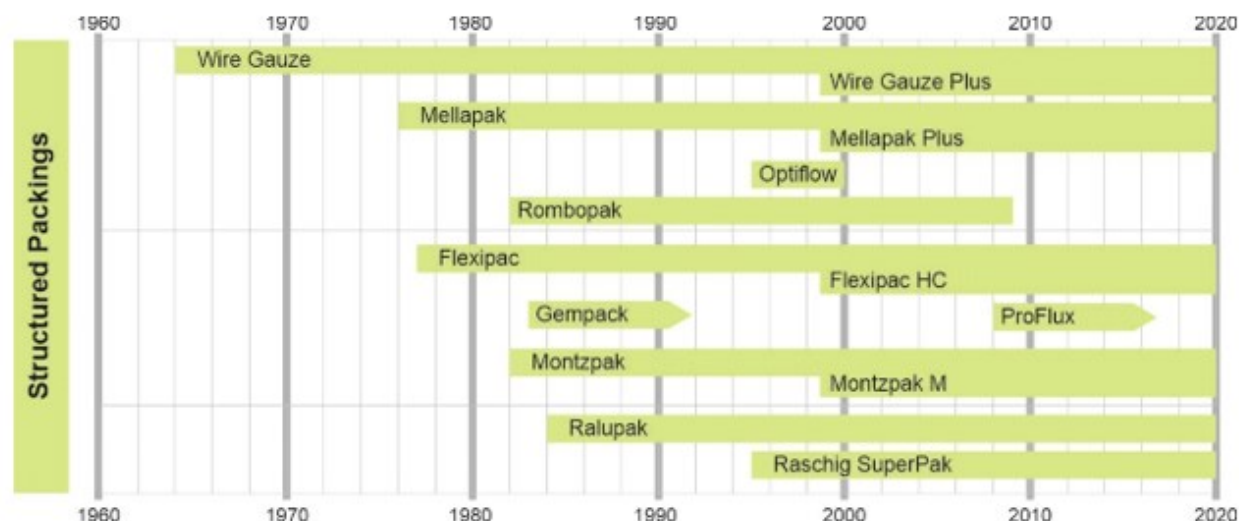


Fig. 1: History of structured packing development

The corrugation amplitude of the sheets steers the number of sheets per block and by this the supplied surface area of the packing. The standard of such a specific surface area (SSA) is $250 \text{ m}^2/\text{m}^3$. The material thickness of such a packing is normally 0.1 mm (this results in a bulk weight of about $95 \text{ kg}/\text{m}^3$).

For higher specific area (for special applications up to $1.000 \text{ m}^2/\text{m}^3$) the material thickness can be less, because the statics can be fulfilled by more sheets per block.

For packings with small SSA, the material thickness should be higher due to static needs. A rule of thumb (Firth's law) is shown in Eq.(1):

$$s_{min} = \frac{0,05m}{a_{geo}} \quad (1)$$

For the bottom layer (e.g. above chimney trays or gas inlets) as well as for the top layer (mechanical stress due to installation of the distributor) a strong material thickness should be chosen.

For heat transfer and heavy duty services there are structured packings with specific areas of $40 \dots 170 \text{ m}^2/\text{m}^3$. Some of these packings are made from special bended elements (material

thickness up to 2 mm); those packings are called grid packings.

The corrugation angle (measured from the horizontal) steers the capacity of the structured packings (Fig. 3). There are two standard angles:

$\sim 60^\circ$ is for high throughput, $\sim 45^\circ$ is for high efficiency (higher holdup). Sulzer introduced the suffix X for 60° corrugation angle and Y for 45° .

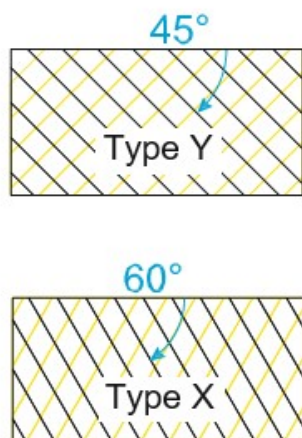


Fig. 3: Corrugation angle

Most packings are made of metal. The parameters shown apply to this material. For special applications ceramics, plastics, glass or carbon material is used. Due to the respective material properties and production processes, different parameters apply to these materials.

Naming

The common nomenclature of packings is the (metric) specific surface area combined with the corrugation angle symbol (X or Y). For high performance packings (see later in this article) there may be an indicator of the shape of corrugation at the nomenclature, too. Most of the suppliers stick to this naming convention.

Packing Layout (Installation Plan)

The packing supplier must plan the block layout to fit the tower cross sectional area (and

to fit the manhole dimension). To ensure a good packing performance, there shouldn't be any gaps between the blocks or to the tower shell.

To achieve this goal, there has to be a so-called wall wiper at the outer elements of the packing to close the gap to the tower shell. For small towers the wall wipers are normally tack-welded to the blocks. For large tower diameters they

are added after installing an entire packing layer (Fig. 4).

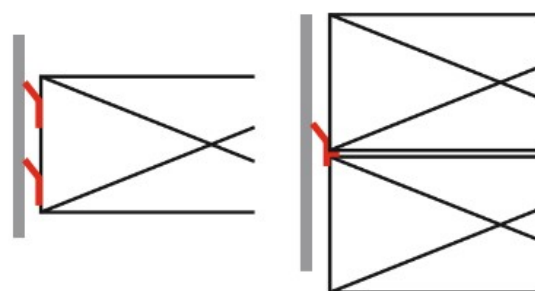


Fig. 4: Variants of wall wipers
left: Tack-welded to block
right: loose band installed between layers

To avoid gaps within the packing, there are so-called filler sheets delivered by the packing manufacturer to close gaps between the blocks. These gaps are not resulting from poor manufacturing, but from column tolerances (tower diameter and tower out-of-roundness)!

In order to avoid aligned joints of packing blocks, each layer of the packing is rotated by

90°. This assumes that the manufacturer has well defined the dimensions and joints of the blocks. Fig. 5 shows a poor and a well-designed packing layout.

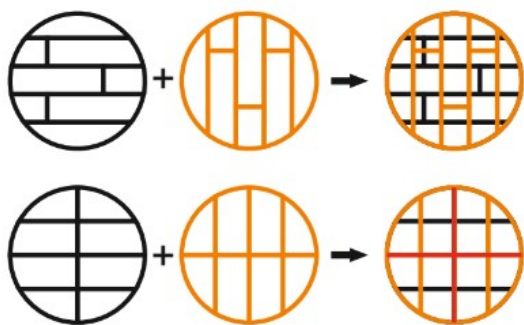


Fig. 5: Packing block layout
upper: well designed layout
lower: poor layout with aligned joints of layers

Layer Effects

Liquid is flowing from the top to the bottom of each layer of the packing. At each contact point of the corrugated sheets, the liquid can swap to the adjacent sheet. By this the liquid flows – at least at small and moderate gas loads – vertically downwards, continuously changing sheets instead of following a channel. The gas flows upwards in counter-current to the liquid.

At each layer interface, liquid and gas must sort its way to the next layer blocks (Fig. 6). Because each layer is rotated by 90°, the structure

changes at each interface. While the liquid and gas flow within each layer is quite smooth, the change to the next layer is the hydraulic bottleneck of each structured packing. To minimize this effect and to optimize the capacity of the packings, there were several approaches of the suppliers. All these variants have in common, that the corrugation angle at the bottom

of each layer is higher and the liquid velocity therefore is higher. These modified packing types apply to the Y-type and are called “2nd generation” or “high performance” packings.

Related Internals

To achieve the best performance of structured packings, the liquid distributor as well as the support grids must meet the performance requirements. The necessity of a good liquid distribution is higher than for random packings, because there is less cross distribution in a structured packing due to the sheets. The liquid distributor has to irrigate the packings channels. The higher the specific surface area of the packings, the higher the drip point density of the distributor must be. Furthermore, the free open area for the gas flow of the distributor must be maximized.

Because of the relevance of the liquid distributor as well as the support grid, the suppliers will only guarantee performance and efficiency, if all internals are designed and fabricated by them.

Hydraulic calculation

The qualitative pressure drop characteristic over the gas load of a structured packing is shown in Fig. 7. It is in principle the same diagram as for random packings.

The gas load is normally expressed by the F-Factor:

$$F = u_G \cdot \sqrt{\rho_G} \quad (2)$$

Another common expression for structured packings is the gas capacity factor c_G :

$$c_G = u_G \cdot \sqrt{\frac{\rho_G}{\rho_L - \rho_G}} \quad (3)$$

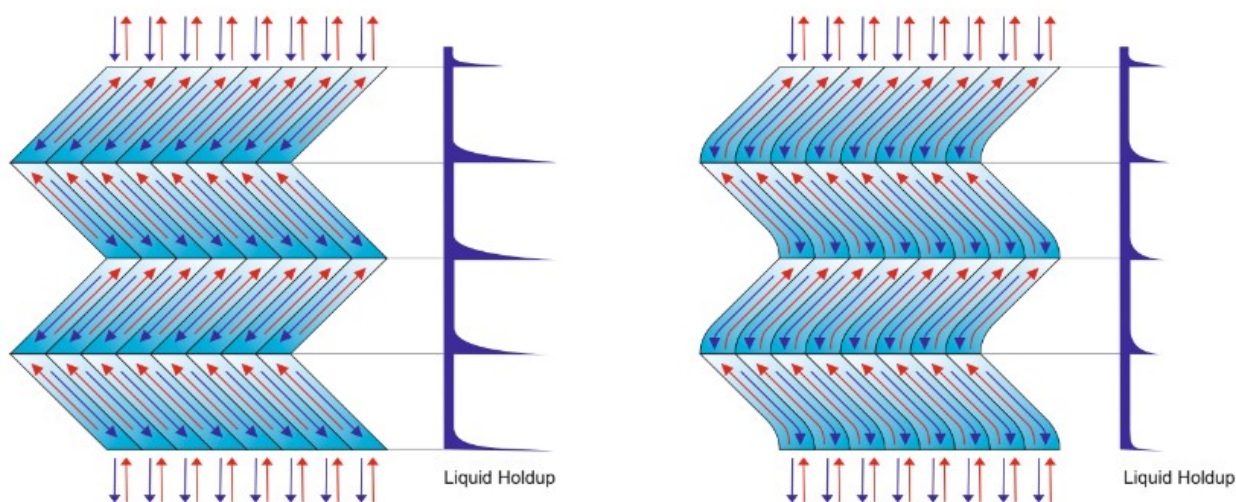


Fig. 6: Effect of layer interface on liquid holdup
left: standard packings, right: modified packings

In the same structure the liquid capacity factor c_L is calculated:

$$c_L = u_L \cdot \sqrt{\frac{\rho_L}{\rho_L - \rho_G}} \quad (4)$$

The gray line in Fig. 7 shows the dry pressure drop. The dark-blue lines represent the pressure drop with increasing liquid flow rate. Even at the presence of liquid the lines in the diagram stay straight for most of the range, as there is almost no interaction between gas and liquid. Because of the reduced void fraction of the packing by the liquid, the gas velocity is increasing with higher liquid loads and therefore the pressure drop is higher. Therefore, the pressure drop curves are parallel to the dry pressure drop.

By increasing liquid and/or gas load, the so-called loading point is reached, where liquid is accumulated by the counter-current gas: liquid holdup rises (see Fig. 8), and the pressure drop increases disproportionately.

By further increase of liquid and/or gas load, the liquid is accumulated to such an extent that a liquid layer is formed. In structured packings this happens at the interface between the layers, as discussed earlier.

For calculating structured packing hydraulics there are two design approaches:

The first one is for random packings and is preferably used for applications without pressure drop issues. The maximum throughput is defined by flooding, where all liquid flow is blocked by the gas. For the design case, you will try to get in the region of the loading point.

In many applications pressure drop is the driving force for design. For this case, the second approach should be applied [Spiegel/Meier 1992]: In this approach 12 mbar/m is set to be equivalent to 100% capacity. Based on this capacity definition the design is calculated.

Maximum values of capacity are about 70%, the design capacity value above 45%.

To calculate the hydraulics of structured packings there are several possibilities:

Most of the suppliers of structured packings offer public software for calculating their packings. None of the models implemented in this software are published.

There are some published models in literature [Mersmann 2000; Billet/Schultes 1999; Engel/Stichlmair 1999; Spiegel/Meier 1992; Suess/Spiegel 1992 and others]. All available models are implemented in the TrayHeart software.

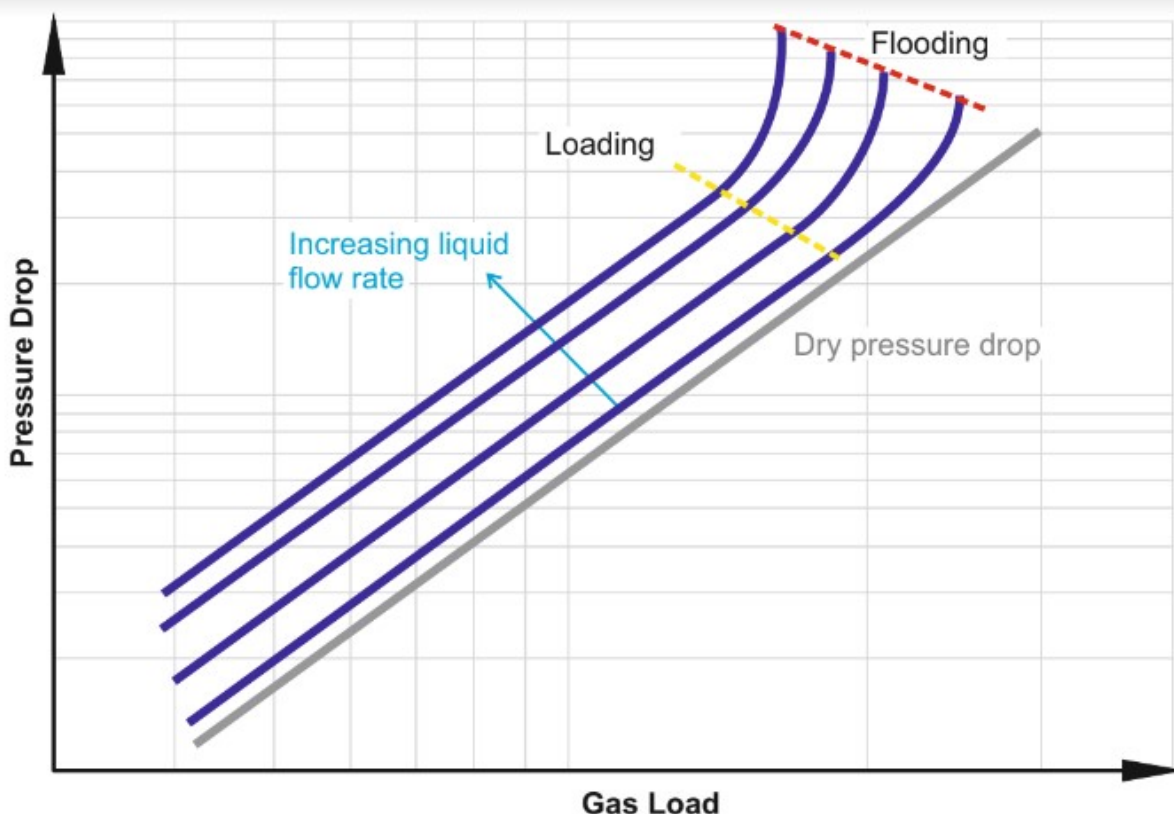


Fig. 7: Qualitative characteristic of pressure drop in packings

Packing Pros and Cons

The main advantages of structured packings are the low pressure drop as well as the high efficiency due to the high geometric surface area. For the separation of temperature-sensitive products, the small liquid holdup is also an advantage.

In the beginning of their development, the structured packings were thought to take over the complete field of all applications (random packings as well as trays)! You will meet a notable number of revamped tray towers with structured packings. But there are still many trayed towers and random packings in operation, which is related to the following disadvantages of structured packings:

- **Fouling:** It is nearly impossible to clean a structured packing. This is not only about cleaning a de-installed block, but de-installing as an action: When the blocks and layers are compacted by fouling, it is hard work to split up the elements for de-installation. By this they normally are damaged, and cleaning is no longer useful.
- **Handling:** The installation must be done very carefully. If there is any damage to the interface area of the layers, the capacity as well as efficiency is limited. De-installing a

packing for cleaning / tower maintenance is only applicable for thick material variants.

- **Corrosion:** If there is any risk of corrosion, even a small corrosion rate leads to a relevant weakening of the (very thin) base material. In result the statics of the affected layers of the packings is damaged, the layer compacts and blocks the gas and liquid flow.
- **Inspection:** A structured packing is a "black box". There is little chance to get an idea of its state. For mechanical inspections of the tower shell, the packing must be de-installed.

Because of the risk of damage at de- and re- installation, the structured packings are normally replaced by new ones. These costs of operational lifetime are – compared to trays – considerably higher.

- **Packing fires:** Due to their high surface area and the problems by getting them clean at shutdown, there is the risk of self-ignition due to pyrophoric substances (e.g. iron sulfide FeS) or hydrocarbons. By this a metal fire can arise, which results in very high damages. Although this risk is well known, it happens again and again [Ender&Laird, 2003].

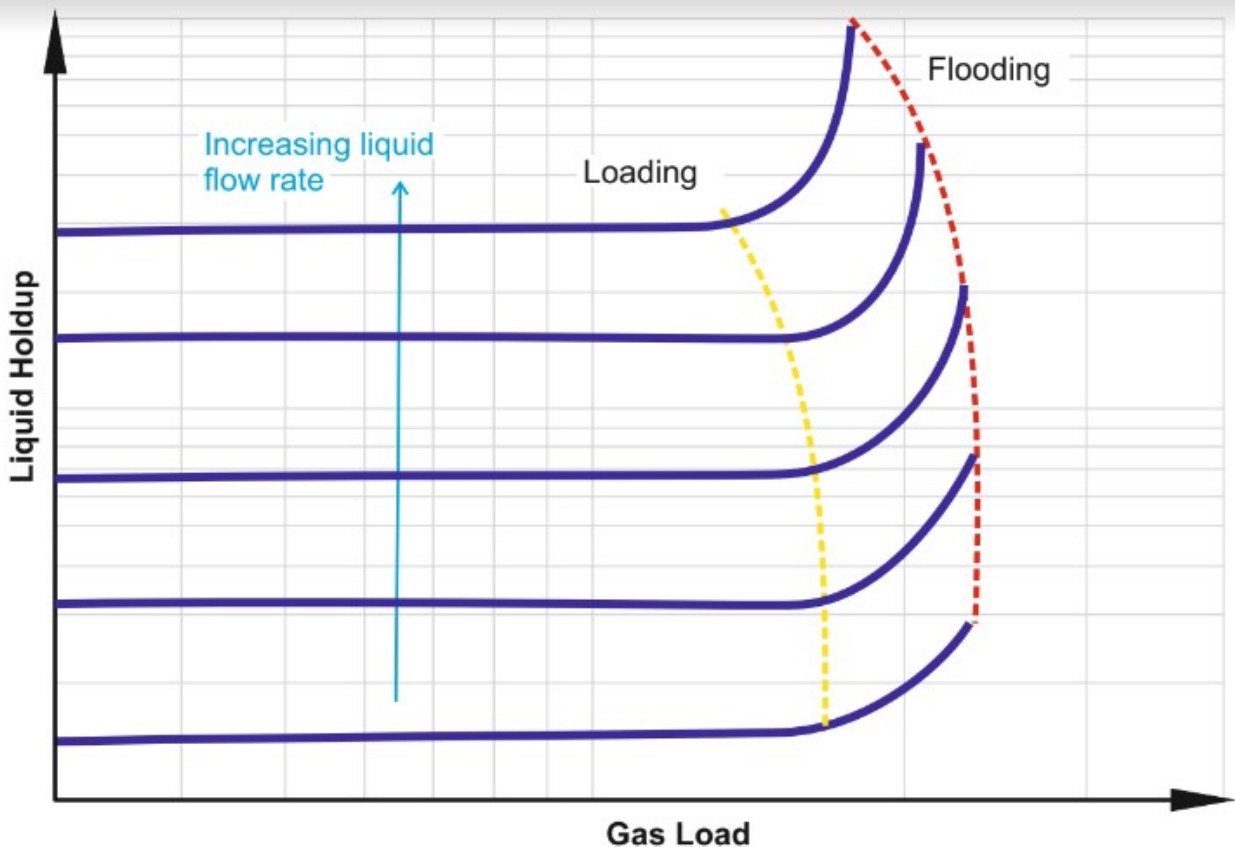


Fig. 8 Qualitative characteristic of liquid holdup

Author

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

References

Billet, R.; Schultes, M.: Prediction of mass transfer columns with dumped and arranged packings- Update Summary of the Calculation Method of Billet and Schultes, Trans IChemE, Vol 77, Part A, September 1999

Ender, Ch., Laird, D.: Minimize the Risk of Fire During Column Maintenance. Chemical Engineering Progress (CEP), 09/2003

Engel, V.: Fluidynamik in Packungskolonnen für Gas-Flüssig-Systeme. Fortschritt-Berichte VDI, Reihe 3, Nr. 605, Düsseldorf (1999)

Engel, V.; Stichlmair, J.G.; Geipel, W.: Fluid Dynamics of Packings for Gas-Liquid Contactors. Chem.Eng.Tech. (2001), Vol. 24, 459-462

Mersmann, A.; Druckverlust und Flutpunkt in berieselten Packungen, in VDI-Wäremeatlas; 9. Auflage (2000); VDI-Verlag

Spiegel, L.; Meier, W.: A generalized pressure drop model for structured packings, ICHIME Symposium Series No. 128 (1992)

Spiegel, L.: 50 Years of Structured Packing. EFCE WP Fluid Separation (June 2013)

Stichlmair, J.G.; Bravo, J. L.; Fair, J. R.: General Model for Prediction of Pressure Drop and Capacity of Countercurrent Gas/Liquid Packed Columns, Gas Separation & Purification (1989) 3; 19-28

Suess, P.; Spiegel, L.: Hold-up of Mellapak structured packings, Chemical Engineering and Processing, 31 (1992) 119-124

WelChem Process Technology: TrayHeart Software. Tower Internals Calculation Software. Internet: www.welchem.com; Info: service@welchem.com

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Estimating Temperature Rise in Pump Recycle Lines

Jayanthi Vijay Sarathy

The following article covers how much liquid temperature rises during pump operations. The head loss in incompressible flow, represents, mechanical energy being converted to thermal energy due to frictional resistance to flow. In an actual pump loop we would have mechanical losses due to pipe surface friction, friction due to turbulence, presence of valve fittings such as measurement gauges, restriction orifices, valves and pipe bends.

In the present undertaking, a sample case study is presented with a thermodynamic estimate of the liquid temperature rise and what should be the minimum design temperature of the associated piping and its associated fittings.

General Notes & Assumptions

1. The heat losses from the pump system are negligible and hence neglected.
2. The liquid density [ρ] and specific heat capacity [C_p] is assumed to be constant across the pump loop.
3. In the current undertaking, a restriction orifice [RO] is present in the recycle loop which constantly recirculates light oil into the pump suction to ensure the operating point is maintained above the minimum flow rate limit of 80%, as per API 610.
4. The restriction orifice [RO] is taken to have a discharge coefficient of 0.62, i.e., an efficiency of 62%.
5. Ideally, throttling across the RO can be isenthalpic, and RO flow is expected to cool the liquid. However due to frictional losses in the piping and associated fittings, there would be heat generation which is absorbed by the liquid followed by a temperature rise.

Governing Relationships

To estimate the temperature rise, the mechanical energy loss is expressed as thermal energy loss by relating the head loss with the specific heat capacity and fluid temperature as follows,

$$H_L = C_p \times \Delta T \quad (1)$$

Where,

H_L = Head Loss [kJ/kg]

C_p = Specific Heat Capacity [kJ /kg.°C]

ΔT = Liquid Temperature Rise [°C]

Note: The head loss in metres can be expressed in energy terms by the conversion 1 kJ/kg = 102.04 metres.

For the following pump schematic, an energy balance is performed as follows,

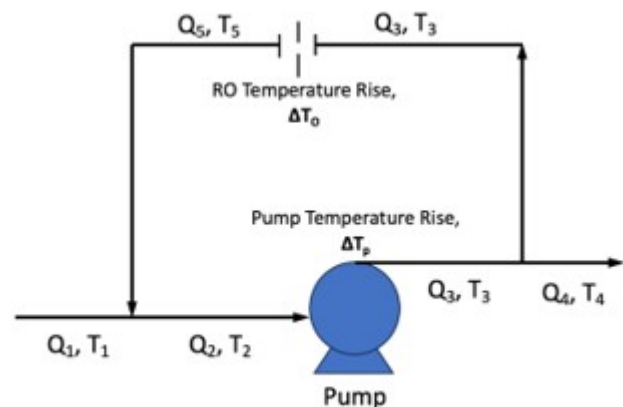


Figure 1. Pump Loop Schematic

Applying an energy balance,

$$\Delta T_P = \frac{H_L}{c_p \times \eta_p} \quad (2)$$

Where, η_p = Pump Efficiency

$$\Delta T_O = \frac{H_L}{c_p \times \eta_o} \quad (3)$$

Where, η_o = RO Efficiency

$$T_3 = T_2 + \Delta T_P \quad (4)$$

$$T_5 = T_3 + \Delta T_O \quad (5)$$

$$Q_1 T_1 + Q_5 T_5 = [Q_1 + Q_5] T_2 \quad (6)$$

Substituting eq. 4 and eq.5 in eq. 6, the RO Outlet temperature [T_5] is estimated as,

$$T_2 = T_5 - [\Delta T_P + \Delta T_O] \quad (7)$$

$$T_2 = \frac{Q_1 T_1 + Q_5 T_5}{Q_1 + Q_5} \quad (8)$$

$$T_5 - [\Delta T_P + \Delta T_O] = \frac{Q_1 T_1 + Q_5 T_5}{Q_1 + Q_5} \quad (9)$$

$$T_5 = \frac{Q_1 T_1 + Q_5 T_5}{Q_1 + Q_5} + [\Delta T_P + \Delta T_O] \quad (10)$$

$$T_5 = \frac{[Q_1 T_1 + Q_5 T_5] + [\Delta T_P + \Delta T_O] \times [Q_1 + Q_5]}{Q_1 + Q_5} \quad (11)$$

$$[Q_1 + Q_5] T_5 = [Q_1 T_1 + Q_5 T_5] + [\Delta T_P + \Delta T_O] \times [Q_1 + Q_5] \quad (12)$$

$$[Q_1 T_5 + Q_5 T_5] = [Q_1 T_1 + Q_5 T_5] + [\Delta T_P + \Delta T_O] \times [Q_1 + Q_5] \quad (13)$$

$$Q_1 T_5 = Q_1 T_1 + [\Delta T_P + \Delta T_O] \times [Q_1 + Q_5] \quad (14)$$

$$Q_1 [T_5 - T_1] = [\Delta T_P + \Delta T_O] \times [Q_1 + Q_5] \quad (15)$$

$$T_5 = T_1 + \frac{[\Delta T_P + \Delta T_O] \times [Q_1 + Q_5]}{Q_1} \quad (16)$$

Rearranging eq. 5, the pump discharge temperature becomes [T_3],

$$T_3 = T_5 - \Delta T_O \quad (17)$$

Rearranging eq. 4, the pump inlet temperature after the recirculation take-off point [T_2],

$$T_2 = T_3 - \Delta T_P \quad (18)$$

Case Study

A pump loop operates with light oil at a feed flow rate [Q_1] of 75 m³/h from the supply vessel and 500C [T_1] with a pump head of 1,013 metres. The recycle flow is 20% recycle flow via a restriction orifice.

The heat capacity of the light oil is 1.9317 kJ/kg.0C. The pump efficiency is 90% and the orifice efficiency is 62%. Estimate the temperature rise in the pump loop. The pump

performance curves is shown in fig 2.

Therefore,

$$\text{Recycle flow} = 0.2 \times 75 = 15 \text{ m}^3/\text{h} \quad (19)$$

$$\text{Pump Suction flow} = 75 + 15 = 90 \text{ m}^3/\text{h} \quad (20)$$

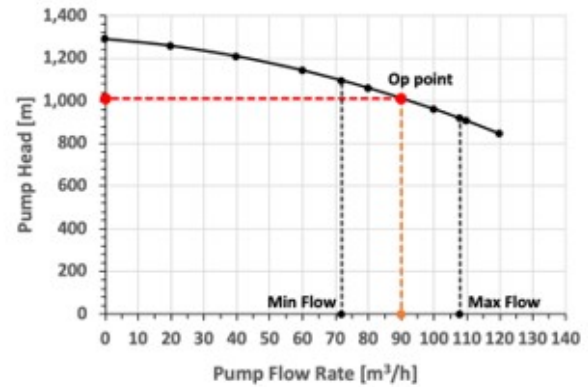


Figure 2. Pump Performance Curves

Rise in Pump and RO temperature,

$$\Delta T_P = \frac{1.013}{1.9317 \times 102.04 \times 0.9} = 5.7^\circ\text{C} \quad (21)$$

$$\Delta T_O = \frac{1.013}{1.9317 \times 102.04 \times 0.62} = 8.3^\circ\text{C} \quad (22)$$

$$T_5 = 50 + \frac{[5.7 + 8.3] \times [75 + 15]}{75} = 66.8^\circ\text{C} \quad (23)$$

$$T_3 = 66.8 - 8.3 = 58.5^\circ\text{C} = \sim 60^\circ\text{C} \quad (24)$$

$$T_2 = 58.5 - 5.7 = 52.8^\circ\text{C} = \sim 53^\circ\text{C} \quad (25)$$

From the above estimates, the recycle line's minimum design temperature must be at least 670C and the pump discharge line minimum design temperature must be at least 600C.

Appendix A: Pump Head, 'm' to 'kJ/kg'

To convert fluid head delivered by a pump from meters to kJ/kg, the following conversions can be made.

$$1J = 1N.m \quad (26)$$

Multiplying and dividing by 1000 and 'kg' and taking, $1 \text{ kg} = 9.8$

$$1 \frac{\text{kJ}}{\text{kg}} = \frac{1000 \text{ N.m}}{\text{kg}} = \frac{1000 \text{ N.m}}{9.8 \text{ N}} = 102.04\text{m} \quad (27)$$

Appendix B: Pump Schematic Calculations

References

"Pipe Flow: A Practical and Comprehensive Guide", Donald C. Rennels and Hobart M. Hudson

https://www.engineeringtoolbox.com/specific-heat-fluids-d_151.html

http://home.zcu.cz/~kovarikp/MT/freestudy/FLUID_MECHANICS_D203/FLUID_MECHANICS_D203.pdf

Author



Vijay Sarathy holds a Master's Degree in Chemical Engineering from Birla Institute of Technology & Science (BITS), Pilani, India and is a Chartered Engineer from the Institution of Chemical Engineers, UK. His expertise over 10 years of professional experience covers Front End Engineering, Process Dynamic Simulation and Subsea/Onshore pipeline flow assurance in the Oil and Gas industry. Vijay has worked as an Upstream Process Engineer with major conglomerates of General Electric, ENI Saipem and Shell.

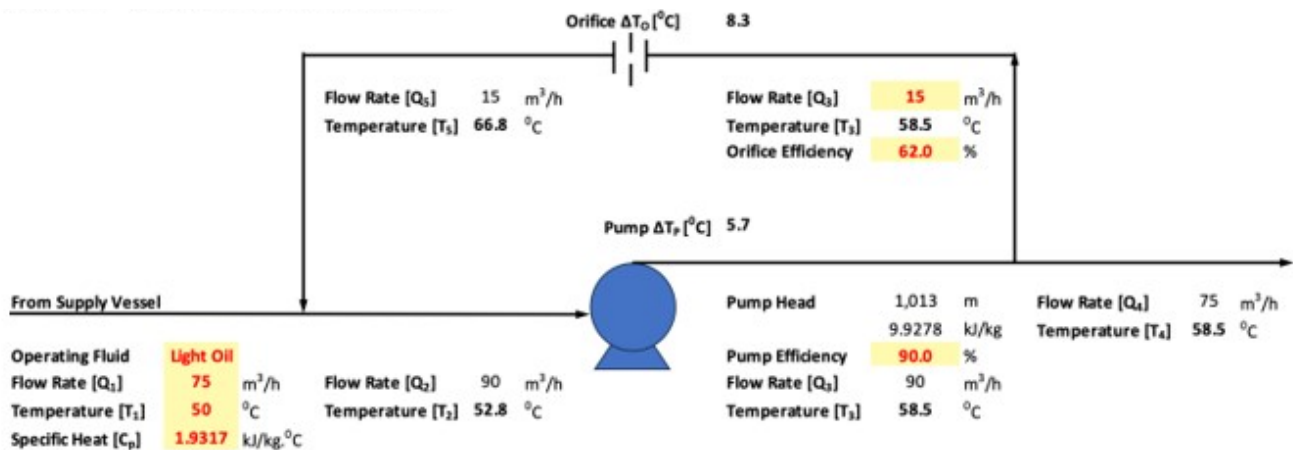
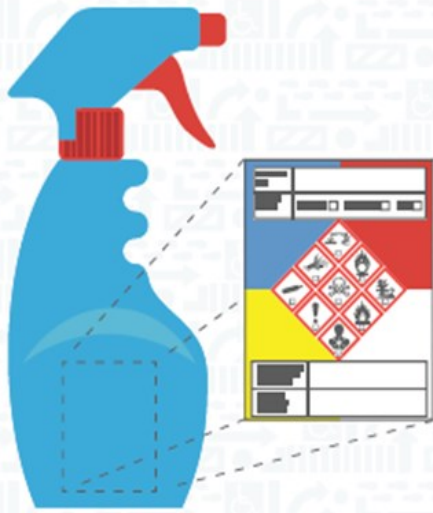


Figure 3. Pump Schematic Calculations



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Reformer Platinum Based Catalyst Acidic Site Chloriding

Shahzeb H.M Ismail, Arsalan Saleem

Framework

The insight for the topic is to elaborate the effects of employing two different chloriding agents used to promote the acidic site reactions over Pt-Rh based Reformer catalyst in contrast to PDC (Propylene di-chloride) & PERC (Perchloroethylene) that results in octane boosting from regular Heavy Treated Naphtha. The base for the abovementioned catalyst consists of alumina (Al_2O_3) having metallic & acidic sites, Metallic site doesn't need the chloriding agent (based on platinum crystallites) while acidic site requires a promoter to execute the certain mandatory & obligatory reactions like isomerization, hydrocracking, dehydrocyclization & dealkylation of aromatics. The catalyst comprises 0.24 to 0.26 weight percent Platinum and 0.48 to 0.52 weight percent rhenium disposed on an alumina spheroidal carrier. Fixed bed catalysts are usually formulated as extrudates having particular shapes, diameters, and lengths intended to optimize the catalyst performance for a given application. The influence of catalyst shape, gas flow rate, and liquid flow rate on hydrodynamics through different industrial catalyst beds that comprised of spherical, cylindrical, trilobe, and quadrilobed.

Forewords

Typically, all Fixed bed reformers facilities in a refinery requires dual function catalysts each add up to the octane boosting that can further be blended to formulate Gasoline grade fuel. To get catalyst acidic site triggered it needs to be dosed with chloride on surface that can promote the necessary reaction.

Fixed Bed Platformer consists of Platinum promoted by Rhenium on base Alumina to provide surface area where complex & desired reactions take place depending on certain conditions that need to be maintained. Exothermic reactions take place on acidic sites that causes mainly isomerization & cracking & hydro

cyclization up to some extent for their controlled execution Chloride need to be injected that amplify the reactions while water contributes to activate & create slot for Chloride to get attached to base Al.

Water is often present in the reactant stream e.g. a naphtha feed typically contains 3-5 ppm by volume of water. Likewise, certain amount of water is required to activate the chloride on the catalyst base and increase its acidic nature. The catalyst must have some acidic character to promote isomerization of reactant molecules which is required along with dehydrogenation for successful reforming reactions to occur. However, if the system contains too much water initially the catalyst is activated to such a degree that it possesses too much acidic character which leads to hydrocracking of the reactant molecules (leading to coke formation on the catalyst surface). The high levels of water can wash the chloride from the surface of the catalyst and may result in corrosion problems. Both scenarios are detrimental to the reforming process.

Residual water must be maintained in the hydrotreated naphtha feed to Reformer. An alumina-supported reforming catalyst requires moisture to activate the acid function and provide homogeneous chloride content over the whole catalyst bed. When the environmental atmosphere for reactor is too wet, however, chloride in the catalyst may leach off and thus deteriorate its acidic behavior. Therefore, water control must be performed along with chloride control to maintain a proper chloride-water balance in the Reactor environment. As per my experience, the chloride content on the catalyst should be kept in the range of 0.9-1.2 wt% for bimetallic catalysts. To meet this requirement, you should maintain an environment of 1-5 ppm of hydrogen chloride and 15-20 ppm of water should be available in the circulating gas over the bimetallic reforming catalyst.

Comparative Sources Of Chlorine For Reactor

As elaborated in prior paragraph the importance of having a chloride source for the desired nature of product & catalyst integrity, we will be discussing the two widely used sources (PDC & PERC) & their competitive behaviors when employed.

Perchloroethylene C₂Cl₄ :

It's a generic name given to a solvent (Tetra-Chloroethylene) having non-flammable nature, relatively inert, insoluble in water & apparently more stable than other chlorinated solvents that can serve the purpose.



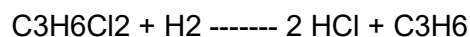
- It has relatively high boiling point & considerable density that can also be a merit while diluting the concentration & mixing with HTN.
- The conversion rate is experimentally determined to be >90% & the reduction for PCE precede oxidation reactions as Chlorine gains Hydrogen & forms HCl that in other case can lead to catalyst poisoning.
- It is preferable to be low-cost operation that can be used in large field operations.
- Density is more compatible for diluting with HTN i.e., 1.62 g/lit
- Non-Flammable having very less vapor pressure (14 mmHg) & having no flashing problems.
- Boiling point of 121 0C makes it more favorite for carrier to disperse over the bed when phase changes.

However, it has also major disadvantages like

- It is toxic in nature and irritating for eye & skin, can cause drowsiness & severe migraine.
- It decomposes at high temperatures (greater than 600°F) to form gases such as hydrogen chloride, chlorine and phosphine which are very corrosive.
- Additionally, Chloride treaters are required at downstream of lighter stream like off gases & LPG to remove HCl.
- Need a constant water injection in order to be dispersed maintaining H₂O/Cl balance or can cause coking and excessive cracking, leading to losses.

Propylene Dichloride C₃H₆Cl₂ :

1,2-Dichloropropane is an organic compound classified as a chlorocarbon. It is a colorless, flammable liquid with a sweet odor. It is obtained as a byproduct of the production of epichlorohydrin, which is produced on a large scale.



- Having high vapor pressure (40 mmHg) & lower molecular weight makes it considerable to be well blend within recycle gas.
- Having lower boiling point of 95 0C makes sure that it is completely vaporized within the pre heating stage.
- Soluble in water that is added benefit to be used for varying concentration.
- Density is gradient towards carrier i.e. 1.156 g/lit (HTN / Condensate), which provides better intermixing & serves the dilution purposes.

However, it also has some disadvantages like

- Decomposition is relatively at lower temperature that causes coking & corrosive gases, impacting purity of Hydrogen.
- Ability to recombine with carry over olefines to form long chain polymers & short straight chain paraffins.
- Requires more quantity as compare to PERC, means more cost for operation.
- Toxic in nature, effect lungs, kidney & eyes causing irritation if inhaled.
- Nearly 40% more of the solvent is required as compare to PERC.
- High vapor pressure needs special care to be taken when handling and transferring.

Relative Consumption

Stoichiometrically, we can analyze the quantity of both solvents required in order to promote the catalyst acidic function for 1ppm, keeping other parameters constant.

- PDC

$$V = 105 \text{ lit}$$

$$M = \text{Density} * V$$

$$M = 1.156 * 105$$

$$M = 121.38 \text{ gm} = 0.12138 \text{ Kg} = 0.26 \text{ lb}$$

$$0.26 \text{ lb PDC (considering 100 \% purity)}$$

Volumetric flowrate for HC = 1000 bbl/hr
= 159000 lit/hr

Mass of Hydrocarbon = $0.7455 * 159000$
= 118534 kg
= 260775 lb

by PPM = $0.26 / 260775 * 1000000 = 1$ ppm

*For 1 ppm of injection we need 105 Lit i.e.
0.26 lb of pure PDC for 1000 BBL HC feed

- PERC

$V = 75$ lit

$M = \text{Density} * V$

$M = 1.620 * 75$

$M = 121.4 \text{ gm} = 0.121 \text{ Kg} = 0.26 \text{ lb}$

0.26 lb PERC (considering 100 % purity)

Volumetric flowrate for HC = 1000 bbl/hr
= 159000 lit/hr

Mass of Hydrocarbon = $0.7455 * 159000$
= 8534 kg
= 260775 lb

by PPM = $0.26 / 260775 * 1000000 = 1$ ppm

*For 1 ppm of injection, we need 75 Lit i.e.,
0.26 lb of pure PERC for 1000 BBL HC feed

*Hence, we need 30 lit more of PDC as compare to PERC for having same ppm in feed, which is 40 % additional. While the same amount of Volume would have given 1.44 pm of Perc in Feed.

Comparative Study

Comparative study was conducted to analyze the behavior when both the solvents were utilized, some points are discussed briefly to cater a wise decision for choosing the best for the catalyst.

- Observed much better dispersing when PERC is employed, causing a byproduct CH₄ that easily carries away into Fuel gas system.
- Purity of H₂ is marginally compromised when PERC is employed excessively in cases, causing system pressure to be continuously rejected.
- Less quantity requirement of PERC as compared to PDC giving more Chloride moles per mole of PERC.

- Need a mixing agitator or impeller to keep concentration constant throughout make up vessel for PERC (if diluted)
- PDC byproduct is LPG grade Carbon that can certainly crack & causes coking on bed making the active area less responsive.
- More quantity is required to give certain moles of chloride as compared to PERC
- PDC Do not need a mixing auxiliary equipment.
- PERC has a key role in regeneration of fixed bed catalyst employed during oxidation to disperse the metal crystallites, while PDC in this scenario can lead to excessive cracking & coking due to high regeneration temperature & low pressure.
- PERC produces a by product methane that is a nominal part of Natural gas & can be used as a source in fuel gas system, although it variates the purity of Hydrogen.
- PDC by product is 3- carbon chain compound with double bond that consumes hydrogen at 230-250 oC temperature and form a saturated straight chain paraffin that cracks & reduces the hydrogen purity of System.
- A Common relation of PDC at temperature & pressure is shown in figure 1.

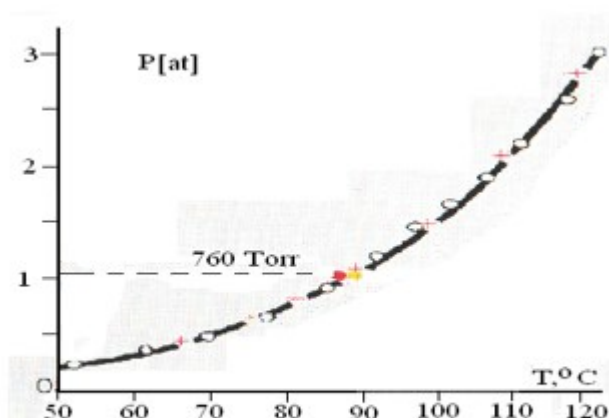


Figure. #1

Experiencing Pdc Vs Perc:

For a fixed Bed Reformer with Pt-Rh catalyst based on Alumina, different Chlorine sources were employed, and the behavior was examined on solvents. It is observed that regardless of the quantity factor where more of the PDC was required as compare to PERC,

extensive cracking was observed in case of PDC as most of the propylene saturated at 238 Celsius to a straight chain paraffin and causes cracking resulted in losses while traces of HCl was found in Recycle gas even at high injection rates, also the cracking causes the downfall in purity, which was tried to be compensated with more naphthenic feed, although not much effective it was, later it was investigated that higher pressure & temperature causes the saturation of by product of PDC i.e. Propylene which cracked decreases the hydrogen purity. Additionally, the saturates in reformat increased up to 55 %

Whereas, PERC shown no such behavior as nominal injection rates were adjusted giving necessary acidic site reactions, also the saturates were minimized to 38 % on same feed stock, although the purity was little effected but still a controlled water – chloride balance is much favorable and gave much better results in terms of losses to cracking & Octane.

Closure Notes

With reference to above mentioned elaboration, it can be extracted that PERC is by far a better promoter than PDC, although you can use PDC where a more naphthenic feed stock is available that helps in equating the metallic reactions & can add up in enhancing Hydrogen purity.

Furthermore, Perchloroethylene isn't flammable as discussed it decomposes at 316 C to form HCl & phosgene that is corrosive by nature so rigorous amount of Hydrogen is needed with condenser outlet temperatures not lower than 105 oC & ideally at 115-125 oC to avoid corrosion effects. Whereas Propylene dichloride starts decomposing at 220 oC to propene that saturates under Hydrogen atmosphere where some part of it further decomposes to Ethylene, Methane & Hydrogen.

References

<https://en.wikipedia.org/wiki/1,2-Dichloropropane>

<https://en.wikipedia.org/wiki/Tetrachloroethylene>

https://www.atamanchemicals.com/perchloroethylene_u29989/#:~:text=Perchloroethylene%20will%20not%20burn%2C%20but,up%20of%20gases%20and%20vapours.

<https://www.stenutz.eu/chem/solv6.php?name=propylene+dichloride>

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Generations

Karl Kolmetz, KLM Technology Group

According to a Chinese proverb, "Wealth does not pass three generations" -- the first generation builds the wealth; the second generation is inspired to preserve it by witnessing the hard work of their parents; and the third generation, having never witnessed the work that went into the creation of this wealth, squanders it.

Middle East proverb "My grandfather rode a camel, my father rode a camel, I drive a Mercedes, my son drives a Land Rover, his son will drive a Land Rover, but his son will ride a camel,"

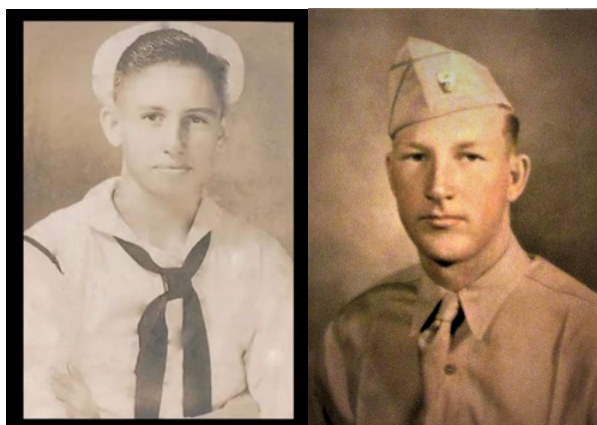
Afghan proverb "The grandfather was born in a tent. He remembers living in a tent. He will lead the people well. The father was born in a tent. He remembers both living in a tent and living in a palace. He will lead the people well. The son was born in a palace. He has never lived in a tent. He will not lead the people at all."



It is clear that we are seeing this in the world today. My father was born in 1926. In 1929 the great depression hit and in 1934 his father died. He dropped out of school in the 7th grade to farm so the family would have food to eat. It is clear from the family pictures they were undernourished. My father is in the middle – very thin.



At 17 he joined the Navy to fight in WW2. (Pictured below on the left.)



My mother's brother, WW Barnes was a very successful fisherman in Panama City Florida, he joined the Army at the age of 26 and died in France in WW2. (Pictured above on the right.)

This is called the greatest generation. They survived the great depression, fought in WW2, and then came home and built the USA into a great economic power. My father and his brothers only knew that life was hard, and you had to work hard to survive. My father and his brothers worked an average of 70 to 80 hours per week.

My father worked 40 hours a week as a welder in a chemical plant in Pensacola, Florida. When he came home, we would work on houses in the evenings and every Saturday. Every Sunday we would go to church twice.

My generation is now attempting to maintain the wealth, but we are not doing a very good job. I would work about 60 hours per week in a refinery and then went to night school two nights per week for eleven years to get an engineering degree. Still hard working, but not nearly as hard as my father and his brothers. I am approaching 70, my generation is basically completed. But in general, my generation did not work as hard as I did.

Today's generation rarely works forty hours per week but expects everything that my father had while working 80 hours per week. They have not seen a major depression, not been hungry, not fought in a major war, and certainly never worked hard. This being born in a palace leads to some interesting facts.

1. Drug Addition - Fifteen percent of the USA has an addictive personality with drugs and alcohol dependency. Every time I fly into Singapore the airline is required to announce that Singapore has a policy of death to drug traffickers. I do not really like that statement, but the thing I like less than that statement is that 107,081 people died in USA of drug overdose in 2022.
2. Mental Health - Ten percent of the USA has mental issues, if you travel to any major city in the USA, you will think it is also higher than 10%.
3. Trainability - Average world IQ is 85 to 90. The US Military will not allow anyone to join that is below 82 IQ, as they are functionally untrainable. So about 40% of the world's population is functionally untrainable. Details of IQ measurement are controversial to some, but however you measure it, the increasing complexity of society is making it harder for the 'average' person to keep up.
4. Honesty - Given the choice to accept a bribe, or reject the bribe and do the right thing, a very high percentage of the people in the world will take the money. There is a direct correlation between corruption and the living standard of the population. The IQ may have a part to play here – they are just not smart enough not to take the money – they are not smart enough to understand they are hurting themselves and their family.

5. First World? There is a divorce rate of over 50% in most of the 1st world nations – hard to call any country with a divorce rate of over 50% 1st in any measure. This is due to a lack of morals and forgiveness. Family is the building block of society. The family is the unit that ensures children are fed loved protected nurtured and raised in the virtues they need to become the responsible citizens of tomorrow, but family breakdown has become epidemic with nearly half of children experiencing the dissolution of their parents' relationship.

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

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

A reluctance to prosecute petty crimes like shoplifting and a failure to integrate immigrants have eroded social trust. Deindustrializing and globalization have ripped the economic heart out of many of our towns. As manufacturing has declined communities have been left empty, without a shared economic endeavor our families and our neighborhoods are caught in a spiral of decline and our nations well.

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

prosperity, and happiness, but the pursuit of good deeds, hard work, and truth in action. (1)

6. Democracy - As a child I was taught that a democracy was the best – now that I see how people vote, I am not sure I believe it anymore. They do not vote in their own best interest.

A democracy is always temporary in nature; it simply cannot exist as a permanent form of government. A democracy will continue to exist until the time that voters discover that they can vote themselves generous gifts from the public treasury. From that moment on, the majority always votes for the candidates who promise the most benefits from the public treasury, with the result that every democracy will finally collapse due to a loss of fiscal policy, which is always followed by a dictatorship.

The average age of the world's greatest civilizations from the beginning of history has been about 200 years. During those 200 years, these nations always progressed through the following sequence: From bondage to spiritual faith; From spiritual faith to great courage; From courage to liberty; From liberty to abundance; From abundance to selfishness; From selfishness to complacency; From complacency to apathy; From apathy to dependence; From dependence back into bondage. Alexander Fraser Tytler -- 1780

The US Founding Fathers understood this concept. They made a representative government, and limited who could vote. We have expanded voting rights, which is a good thing – if we train our citizens to vote wisely. Here we are failing – sometime the correct thing to vote for is not in our best interest, and we lack the will and the values to do the right thing. We can keep voting ourselves money – but the end is bondage. We have voted to legalize alcohol and some drugs – is this really the best given our addictive personalities?

Given this list – which is not good – what are some things we can do to improve the human condition.

1. We must have a war on drugs and drug traffickers – 15% of the population has addictive personalities (not my statistics, go search) – this is like leaving the sheep out for the wolves, we must have some shepherds. We need to go after the wolves hard and there may only be 35% capable shepherds. In Malaysia and Singapore, they put high taxes on alcohol, I would be in favor of this given the instabilities and low IQs.

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

3. Empathy – 65% of the population has challenges. We need to understand the challenges and help the less fortunate.

References

How can we begin to reverse Cultural Decline?
British Member of Parliament, Miriam Cates

Rock Bottom: Toward Lower Carbon Transport Alternatives, Especially Ethanol's Role in Reducing Aviation's Footprint

Ron Cormier

Hello again and Happy 2024 from central Mexico on the shores of Lake Chapala, where the weather is mild mostly year-around. Hopefully winter (for those of us north of the equator anyhow) is bearable, though television reports over the last month reveal such may prove brutal. Many relate real-time, seemingly outrageous, man-made weather events to climate change leveraged over the last 125 years of fossil-fuel refining and petrochemical manufacture. To that end, let's look at a well-known, but evolving transportation energy source that soon will be commercially "spreading its wings" into the skies.

Global refined oil product markets are on track to remain tight until mid-decade, with the global downstream industry facing significant near-, medium- and long-term challenges, according to a new report by S&P Global and the International Energy Forum (IEF). Global oil refining capacity dropped in 2023 for the first time in 20 years. In 2020-2023, pandemic-weakened margins, accelerated refinery closures, and motivated conversions to biofuels or distribution terminals, according to the S&P/IEF Oil Refining Industry Insights report.

Forecasts suggest that there are no quick fixes for the forces at play, creating a tight market for refined oil products now and into the future. The causes for the current market conditions include headwinds from the unprecedented restructuring following record refinery closures in 2020-2023, to sanctions on Russia, (one of the world's largest fossil fuel exporters), to China's strategic shift to reduce petroleum product exports. All of which comes even as operators are facing an uncertain future for refining demand, global transitions to lower-carbon energy sources, and possible paradigm shifts in demand-required platforms altogether.

Refining margins for key products ballooned in the summer of 2022 to a record \$35-50/bbl compared to the normal \$10/bbl - underscoring the severe bottlenecks in the sector. More

than 2 mb/d in net capacity was scheduled to come online by end-2023, but delays and startup challenges slowed progress somewhat. In addition, these are likely the last major greenfield fuel-oriented refineries to be built, as the energy transition will limit the need for conventional refinery capacity additions in the future. High refining margins in the past led to more investment, but that is not occurring going forward. The expectation that the energy transition could reduce the need for refineries has deterred investment. Such uncertainty always induces substantial planning/capital uncertainty.

Russia and China have available refining capacity, but due to the Russia/Ukraine war, western sanctions limit Russia's exports and domestic policies limit China's. The ethanol industry, traditionally known for providing high-octane renewable blend stock in gasoline for internal combustion engines, is now turning its expertise toward aviation, crafting new evolution in sustainable aviation fuel (SAF).

The Russia/Ukraine Impact

Russia's invasion of Ukraine is spurring an unprecedented reshuffling of global refined product trade, and more obstacles are on the horizon. Meanwhile, rerouting traditional flows of Russia's petroleum product exports is more complicated than for crude oil, according to the report. Russia is one of the largest global exporters of fuels, exporting a total of ~2.9 mb/d of refined products pre-invasion, of which the EU imported 2.2 mb/d while the US imported 0.4 mb/d.

The most likely alternative markets for Russian products, particularly diesel, are Africa and Latin America. However, the transit time for product deliveries from Russia to West Africa and Latin America is 25-30 days, compared with 8 days to Europe. Already high freight rates (exacerbated by strife and possible closure of the Strait of Hormuz, prompting

alternative old-school routes around the South African Cape of Good Hope), and a looming insurance ban means these trades may be uneconomic for both buyer and seller.

Further, an EU embargo on importing Russian crude via pipeline, and a ban on importing refined products of Russian origin took effect in 2023. There are several exemptions to the embargo, but it broadly cut 90% of “legal” Russian crude imports and nearly all product imports.

What About Additional Use of Ethanol Then...

Plug-in vehicle sales are expected to grow from 6.6 million in 2021 to 35.7 in 2030. This is expected to replace 4 mb/d of gasoline and diesel demand by the end of the decade, and cause hydrocarbon fuel’s share of transportation fuels consumption to plateau by 2028. Refining’s Rocky Road! While the refining industry is headed toward a less carbon-intensive future, it will be a rocky, uneven road, due in large part to an evolving and fractured decarbonization policy landscape.

Ethanol producers see SAF as a promising opportunity to grow their industry amid rising sales of electric vehicles (EVs). This strategic shift towards SAF production is a testament to ethanol’s versatility and potential to significantly reduce the aviation sector’s carbon footprint. Central to ethanol’s feasibility and wider use, is reducing its greenhouse gas (GHG) emissions relative to petroleum jet fuel and achieving a low carbon intensity (CI) score. A CI score that indicates low GHG emissions throughout the fuel’s lifecycle, from farming, transportation, production, and consumer use is crucial for environmental compliance and market viability.

The majority of the SAF used to date (in the U.S. and the European Union) is being produced via the hydrogenation of seed oils such as soybean oil and rapeseed oil, animal fats, distiller’s corn (DCO) and used cooking oil (UCO) – otherwise known as hydro-processed esters and fatty acids (HEFA).

While not at economies of scale like the HEFA route, alcohol-to-jet (ATJ) technology pathways are showing promise as a contributor to SAF production in the long term. In early 2023, the U.S. Environmental Protection Agency (EPA) approved the first Renewable Fuel Standard pathway for Lanzajet’s production of SAF from Brazilian un-denatured sugarcane ethanol. The EPA determined this pathway gives a 54-66 percent reduction in GHGs compared to petroleum jet fuel.

Pioneering research underway is pivotal in ensuring feedstock and production compliance,

facilitating technology and feedstock pathway registration, and conducting risk analysis and feasibility studies. Indirect land use calculations (iLUC), regulatory engagement, compliance management and carbon market navigation are invaluable for ethanol producers transitioning to SAF production. While the lifecycle emissions profile of corn ethanol doesn’t have an inherently low CI score, innovative land management and the addition of carbon capture and storage (CCS) can help further reduce its carbon footprint.

However, integrating ethanol into the SAF market demands a deep understanding of carbon pipeline logistics, managing carbon dioxide (CO₂) capture, transport and sequestration, which is critical for reducing overall emissions. Integrating sustainable farming practices like cover cropping, no-till farming and the use of green fertilizer could reduce the overall CI score of ethanol production even further.

The technical, regulatory, and public acceptance challenges in this domain necessitate a nuanced approach, combining industry knowledge with environmental stewardship.

The ethanol industry’s venture into SAF production is a bold step toward a more sustainable aviation sector. With the support of specialists like EcoEngineers, the industry is well-equipped to overcome the challenges of SAF production, proving ethanol’s role in the energy landscape is evolving to meet modern-day environmental challenges.

This shift underscores the industry’s adaptability and commitment to innovative solutions, paving the way for a more sustainable future in aviation fuel technology. The downstream industry is innovative and resilient; there is little doubt that successful companies will emerge leaner, more efficient, and better adapted to the evolving market and fuels of the future, be usage be on land or in the skies.

Until EPM’s March edition, I wish you all, our valued readers, the very best.

Sources

- Breaking Barriers: Ethanol’s Role in Reducing Aviation’s Carbon Footprint, EcoEngineers News, January 19, 2024. Article originally published in the January/February 2024 issue of Ethanol Today, By Kristine Klavers, Managing Director, Houston, Petroleum and Refining

- U.S. Treasury and IRS Release Guidance on Section 40B Sustainable Aviation
- S&P Global Commodity Insights, BLOG Sep 23, 2022, Global Fuels Market Outlook, Eleanor Budds, Kurt Barrow
- S&P Global and International Energy Forum...
- BLOG JAN 18, 2022, Oil Refinery Industry Insights, Stretched Sector Fuels Market Volatility, Sharing Insights Elevates Their Impact
- April 2023: Evolving global energy policies and energy transition efforts and challenges
- MAY 22, 2023, BLOG Global Power and Renewables Research Highlights,
- NOV 15, 2023 BLOG Economic Headwinds Halt Upstream Cost Inflation



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