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# In This Issue

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**SPECIAL FEATURES**



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# Massive Investments in Crude to Chemicals Refineries as Main Driver of Current Chinese Supremacy in the Global Petrochemical Market

Dr. Marcio Wagner da Silva

**Introduction and Context**

The current scenario presents 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, but as presented in Figure 1, there is a significant trend of reduction in the transportation fuels consumption, especially gasoline.

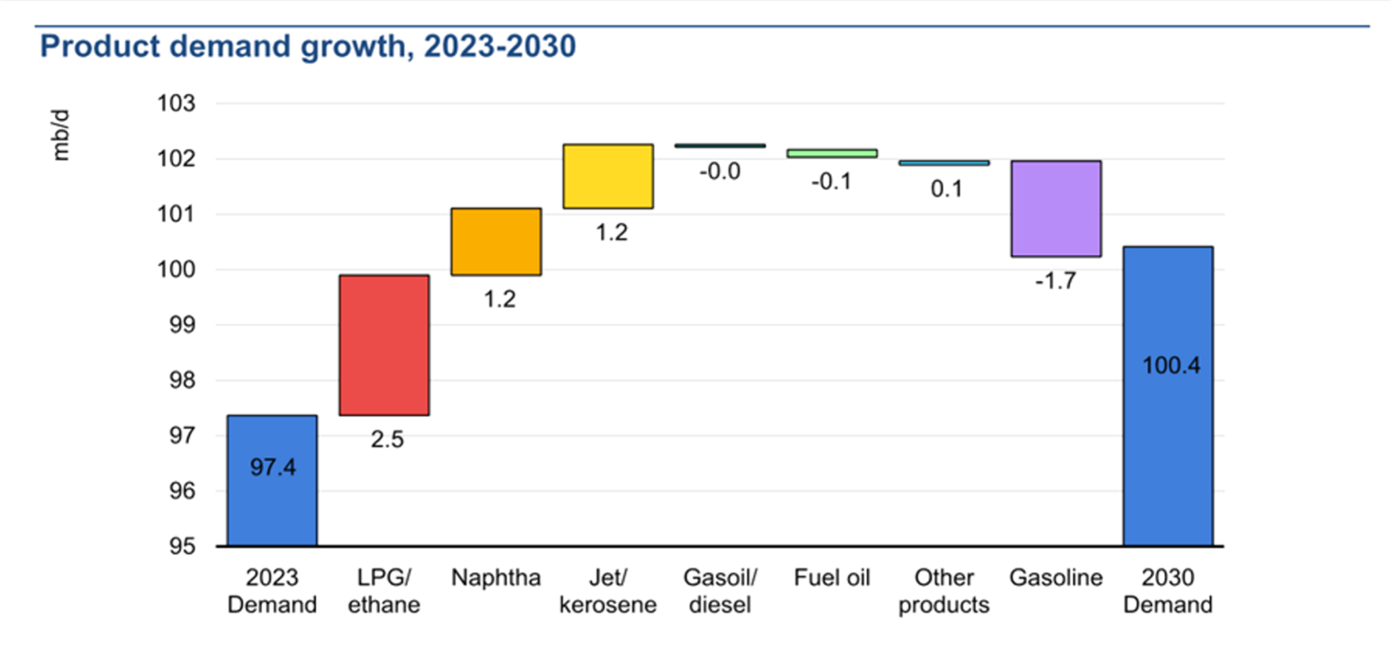



*Figure 1 – Global Oil Demand by Derivative (International Energy Agency, 2024)*

According to Figure 1, is expected a growing demand by petrochemicals while the transportation fuels tend to present falling consumption, in the gasoline case is expected a retraction in the market around 1,7 %. 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.

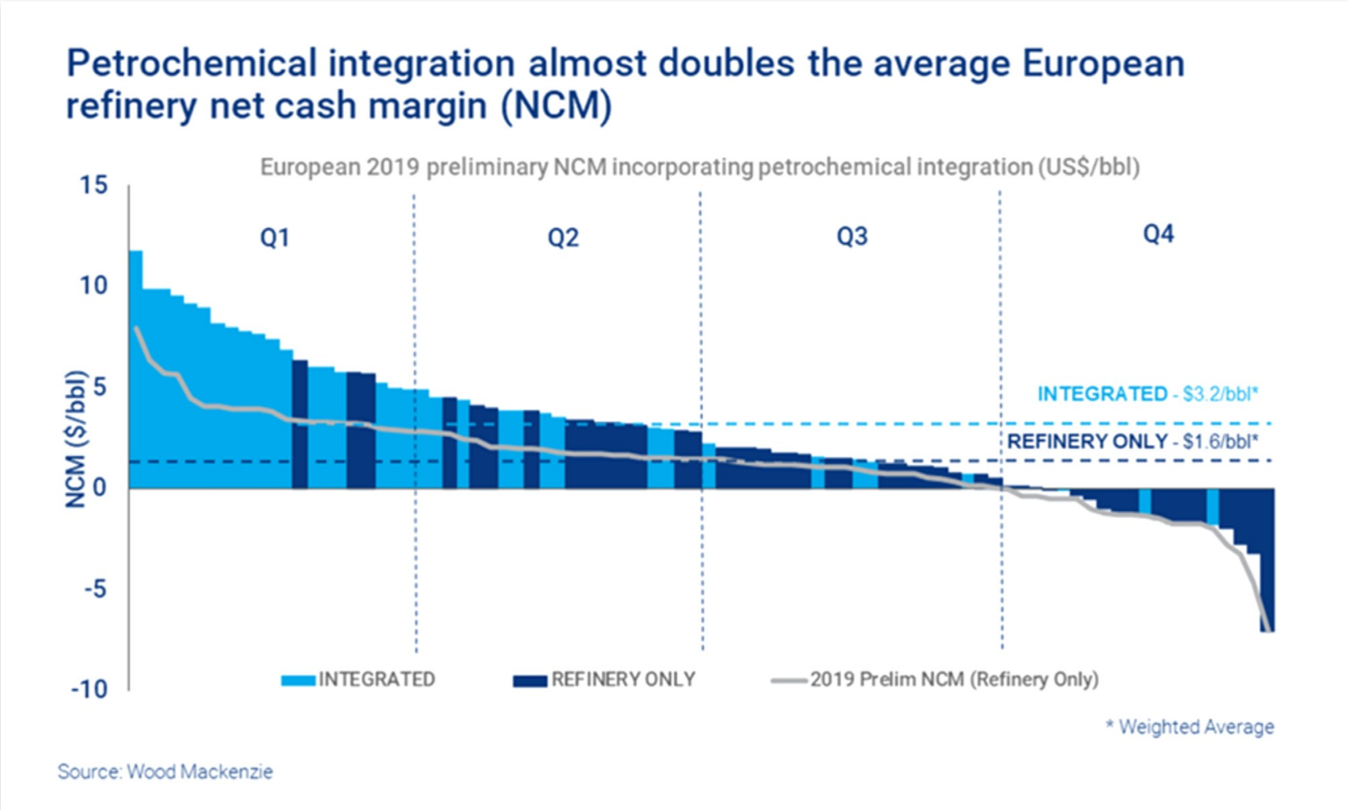



*Figure 2 – Refining Margins to Integrated and Non-Integrated Refining Hardware (Wood Mackenzie, 2020)*

*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. According to Figure 3, the demand for petrochemicals tends to rise in the next years and can be an attract way to refiners keep his protagonism in the market.

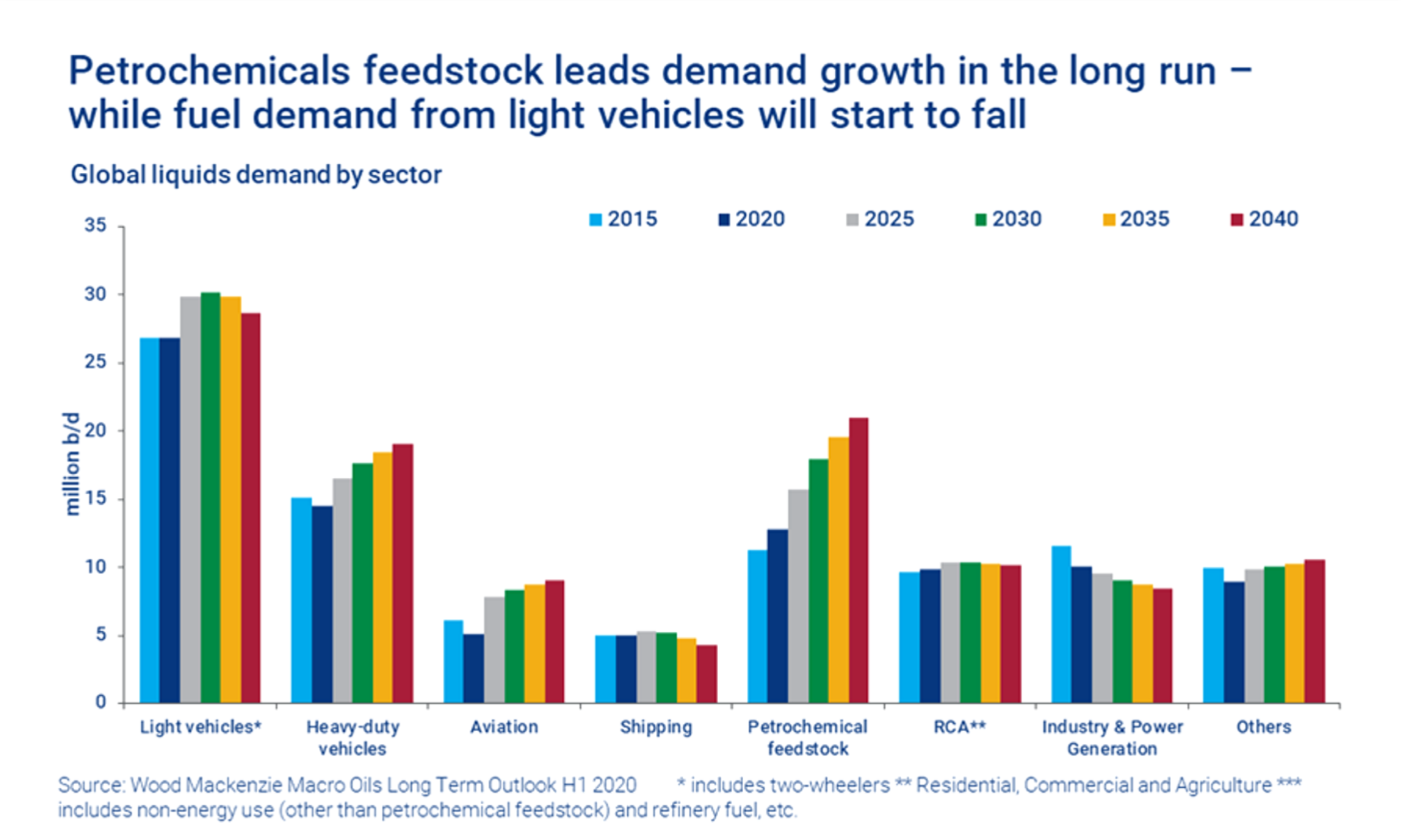



*Figure 3 – Growing Trend in the Demand by Petrochemical Intermediates (Wood Mackenzie, 2020)*

According to data presented in Figure 3, a significant growth in the market of petrochemical intermediates is expected, and a refining hardware capable of maximizing the yield of these derivatives can offer significant competitive advantage through closer integration with petrochemical assets and higher value addition to processed crude oil. Taking as example in the North American Market, it's possible to observe the falling demand by transportation fuels, as presented in Figure 4.

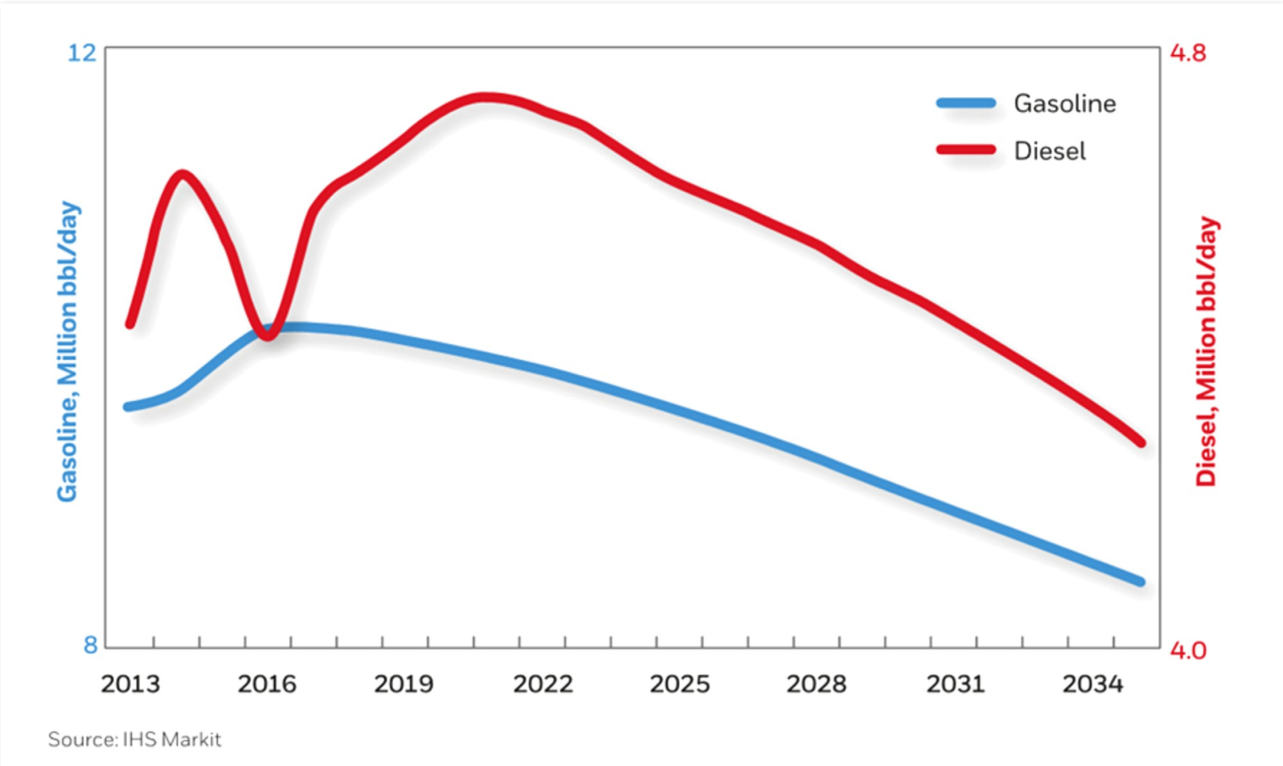



*Figure 4 – Transportation Fuels Demand for the North American Market (UOP Company, 2021)*

Another deep change in the downstream sector that reinforces the necessity of a high conversion refining hardware is the IMO 2020. Restrictive regulations like IMO 2020 raised, even more, the pressure over refiners with low bottom barrel conversion capacity once requires higher capacity to add value to residual streams, especially related to sulfur content that was reduced from 3,5 % (in mass) to 0,5 %. Refiners with easy access to low sulfur crude oils present relative competitive advantage in this scenario. These players can rely on relatively low cost residue upgrading technologies to produce the new marine fuel oil (Bunker) as carbon rejection technologies (Solvent Deasphalting, Delayed Coking, etc.), but they are the minority in the market. The most part of the players need to look for sources of low sulfur crudes, which present higher costs, putting under pressure his refining margins or look for deep bottom barrel conversion technologies to ensure more value addition to processed crude oils and avoid to loss competitiveness in the downstream market. For these refiners, deepest residue upgrading like hydrocracking technologies can offer great operational flexibility, despite the high capital spending. In this scenario, with necessity to higher value addition to bottom barrel stream and growing market of petrochemicals, refiners with adequate bottom barrel conversion capacity can achieve great competitive advantage in the downstream industry.

Based on description above it’s possible to apply the article published by W. Chan Kim and Renée Mauborge called “Blue Ocean Strategy” in Harvard Business Review, to classify the competitive markets in the downstream industry. In this article the authors define the conventional market as a red ocean where the players tend to compete in the existing market focusing on defeating competitors through the exploration of existing demand, leading to low differentiation and low profitability. The blue ocean is characterized by looking for space in non-explored (or few explored markets), creating and developing new demands and reaching differentiation. This model can be applied (with some specificities once is a commodity market) to the downstream industry, considering the traditional transportation fuels refineries and the petrochemical sector.

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

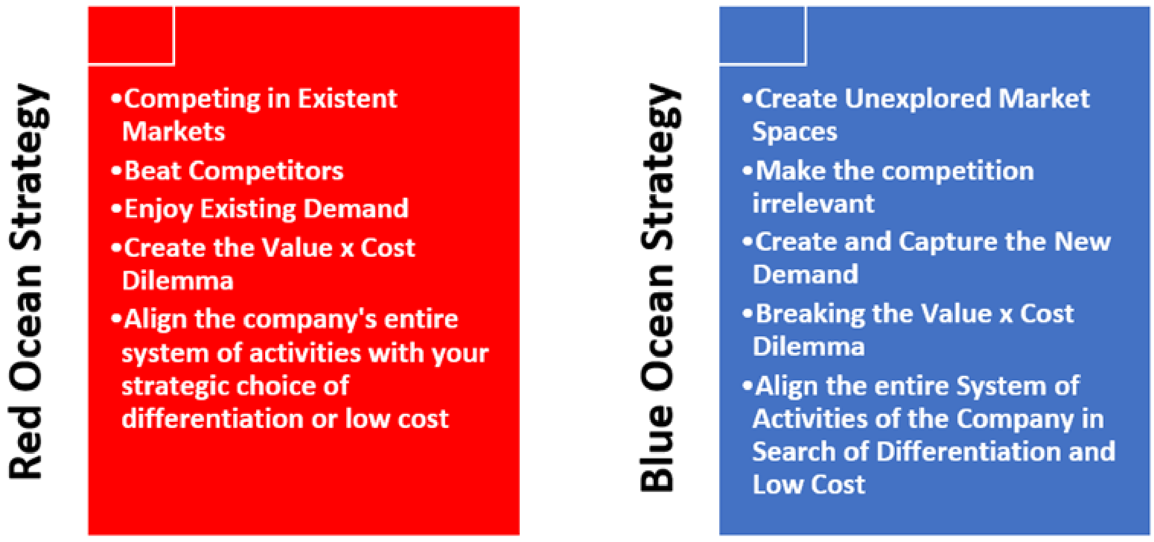



*Figure 5 – Differences between Blue and Red Ocean Strategies (KIM & MAUBORGNE, 2004)*

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

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

Considering 2025 as the base year, the petrochemical market size reached a total value of USD 700,10 billion with an expected compound annual growth rate (CAGR) of 6,11 % between 2025 and 2034 as presented in Figure 6.

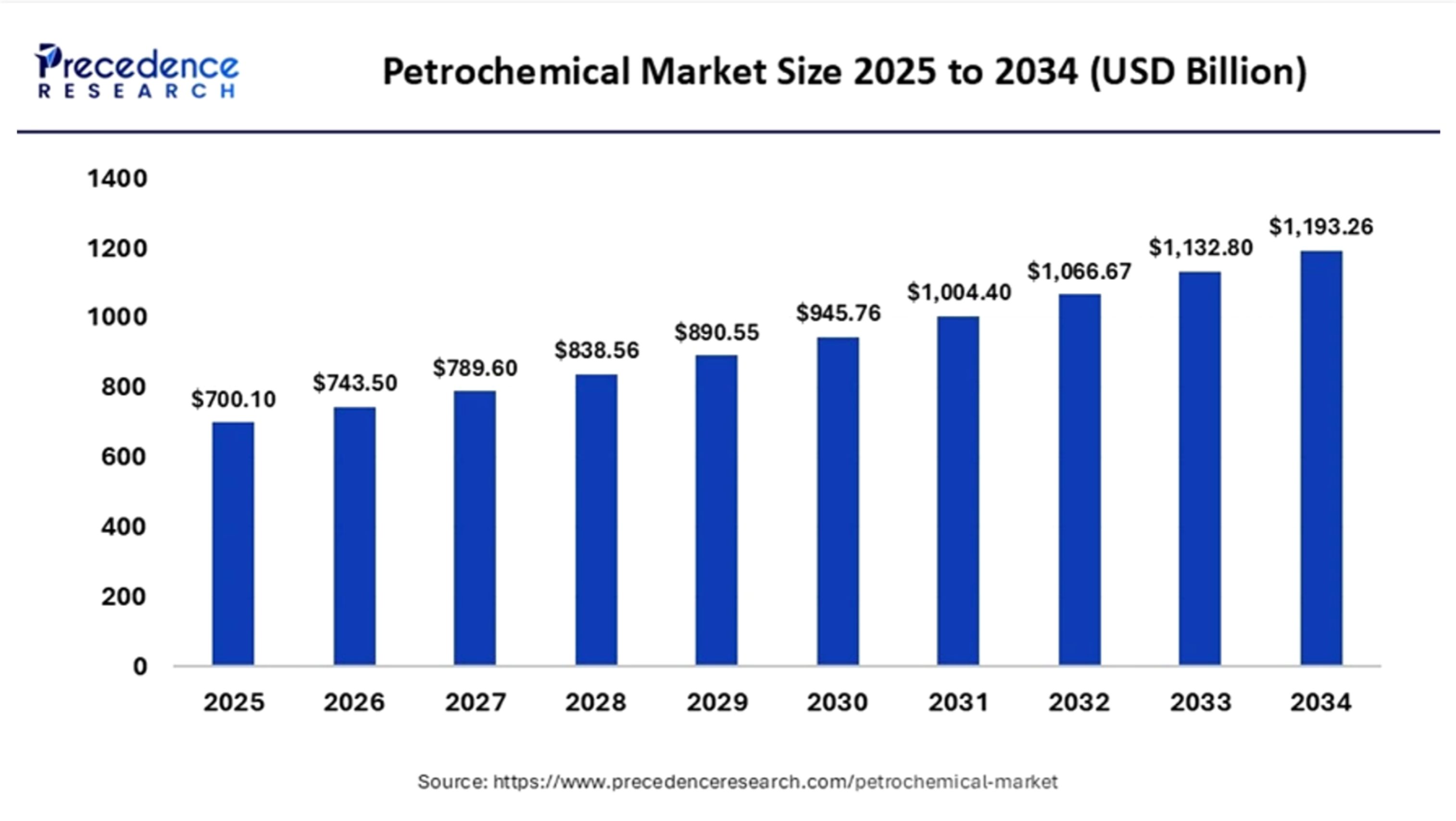



*Figure 6 – Petrochemical Market Size Forecast 2023-2033 (Precedence Research, 2025)*

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

Considering just the aromatics solvent market (Benzene, Toluene, and Xylenes) the CAGR expected between 2021 and 2030 is 4,80 % leading the aromatics solvent market size to reach USD 8,1 billion in 2030 still according to Precedence Research data.

Considering exclusively the propylene market, the forecasts are even more encouraging for investments in on purpose propylene production routes. Figure 7 presents the projection to propylene market size for the next years.

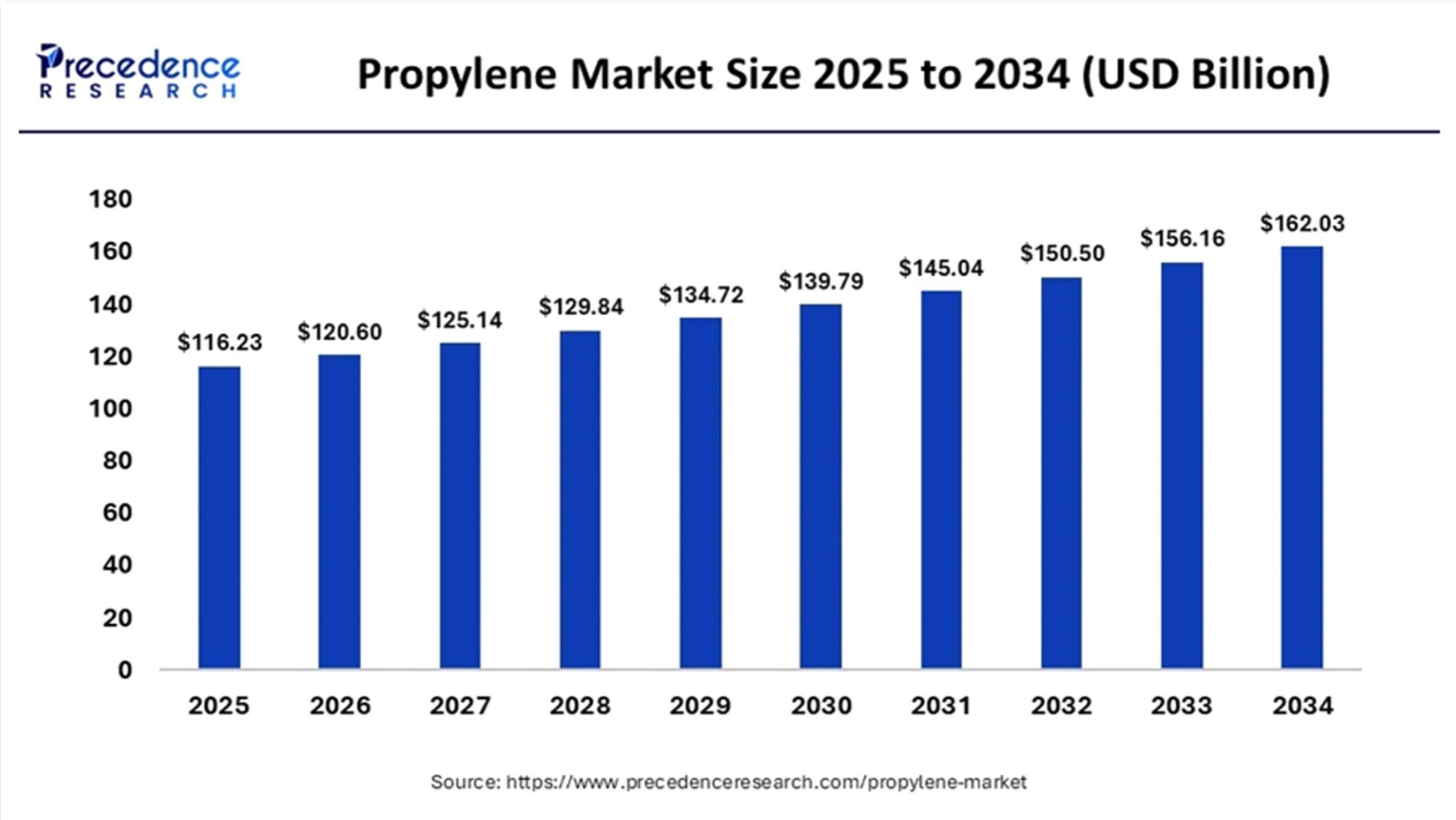



*Figure 7 – Evolution of Propylene Market Size for the next years (Precedence Research, 2025)*

According to Figure 7, the propylene market can reach higher than 162 billion USA dollars in 2034 with an annual rate of 3,76 % with Asia being the bigger market as expected.

**Synergies between Refining and Petrochemical Assets – Petrochemical Integration**

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

Table 1 – Refining and Petrochemical Industry Characteristics

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

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

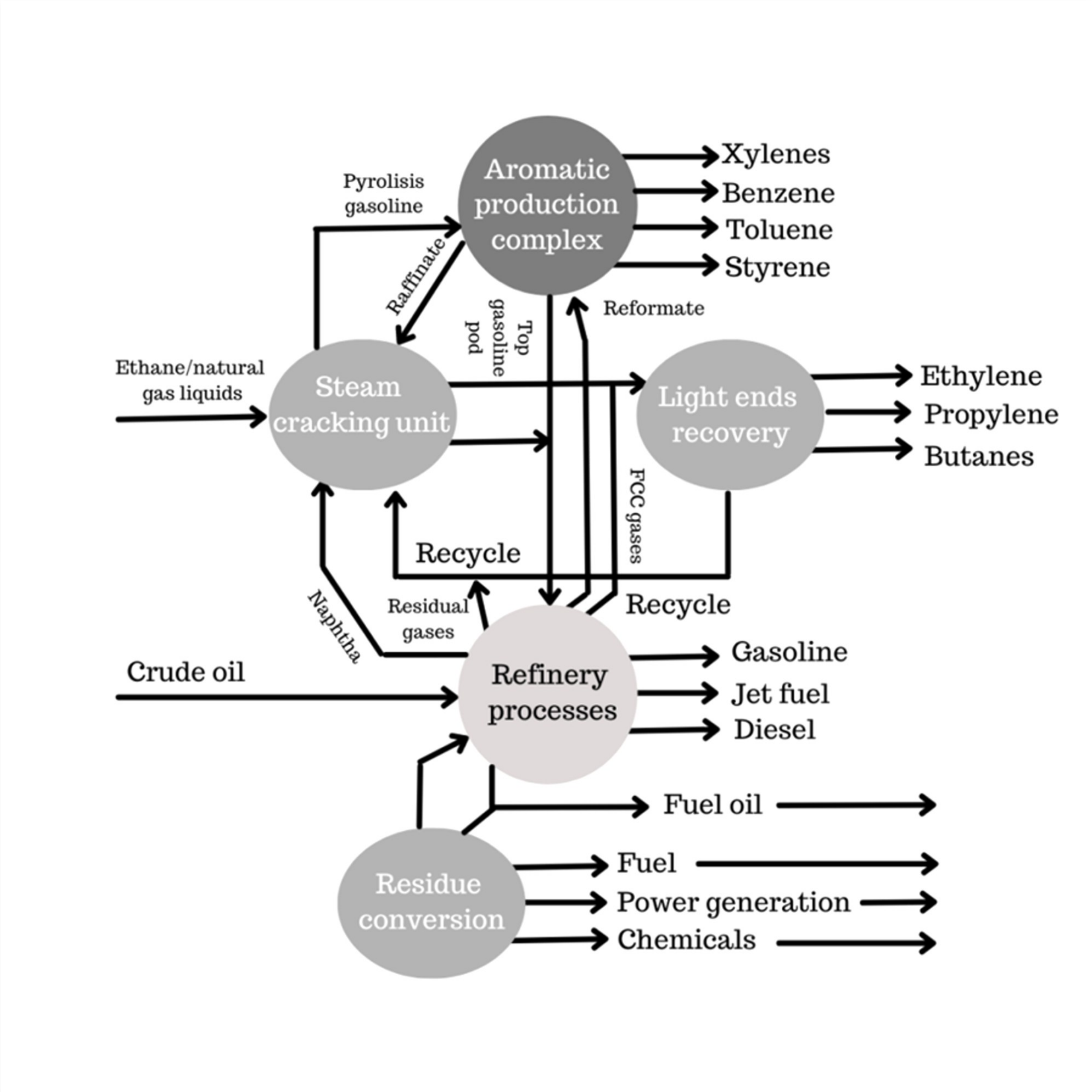



*Figure 8 – Synergies between Refining and Petrochemical Processes*

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

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

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

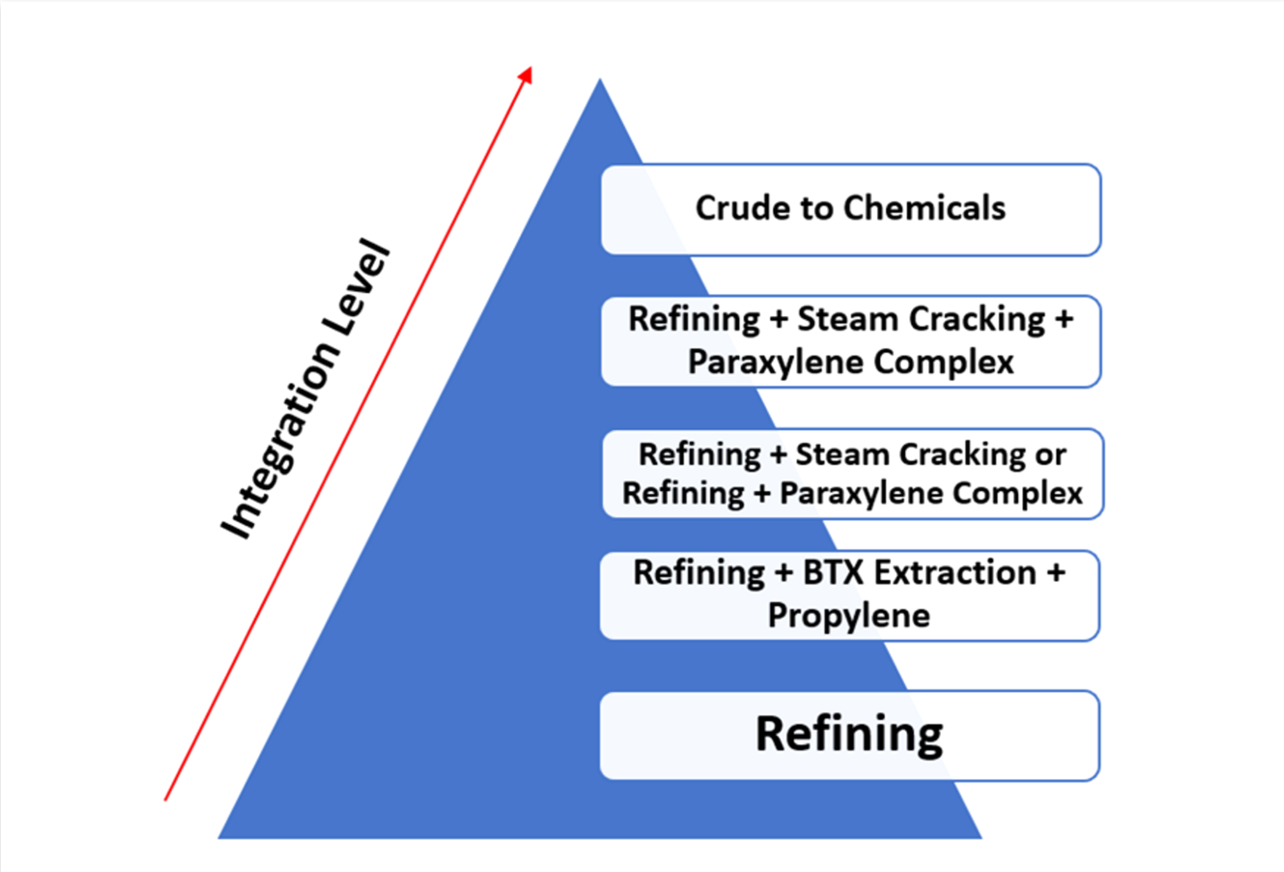



*Figure 9 – Petrochemical Integration Levels (IHS Markit, 2018)*

According to the classification proposed, the crude to chemicals refineries is considered the maximum level of petrochemical integration where the processed crude oil is totally converted into petrochemical intermediates like ethylene, propylene, and BTX. Considering the current scenario of the downstream industry, the crude to chemicals refineries can create the necessary differentiation to allow the players to reach the Blue Ocean Strategy.

**The Crude Oil to Chemicals Refining Assets**

Due to the increasing market and higher added value as well as the trend of reduction in transportation fuels demand, some refiners and technology developers has dedicated their efforts to develop crude to chemicals refining assets. One of the big players that have been invested in this alternative is the Saudi Aramco Company, the concept is based on the direct conversion of crude oil to petrochemical intermediates as presented in Figure 10.

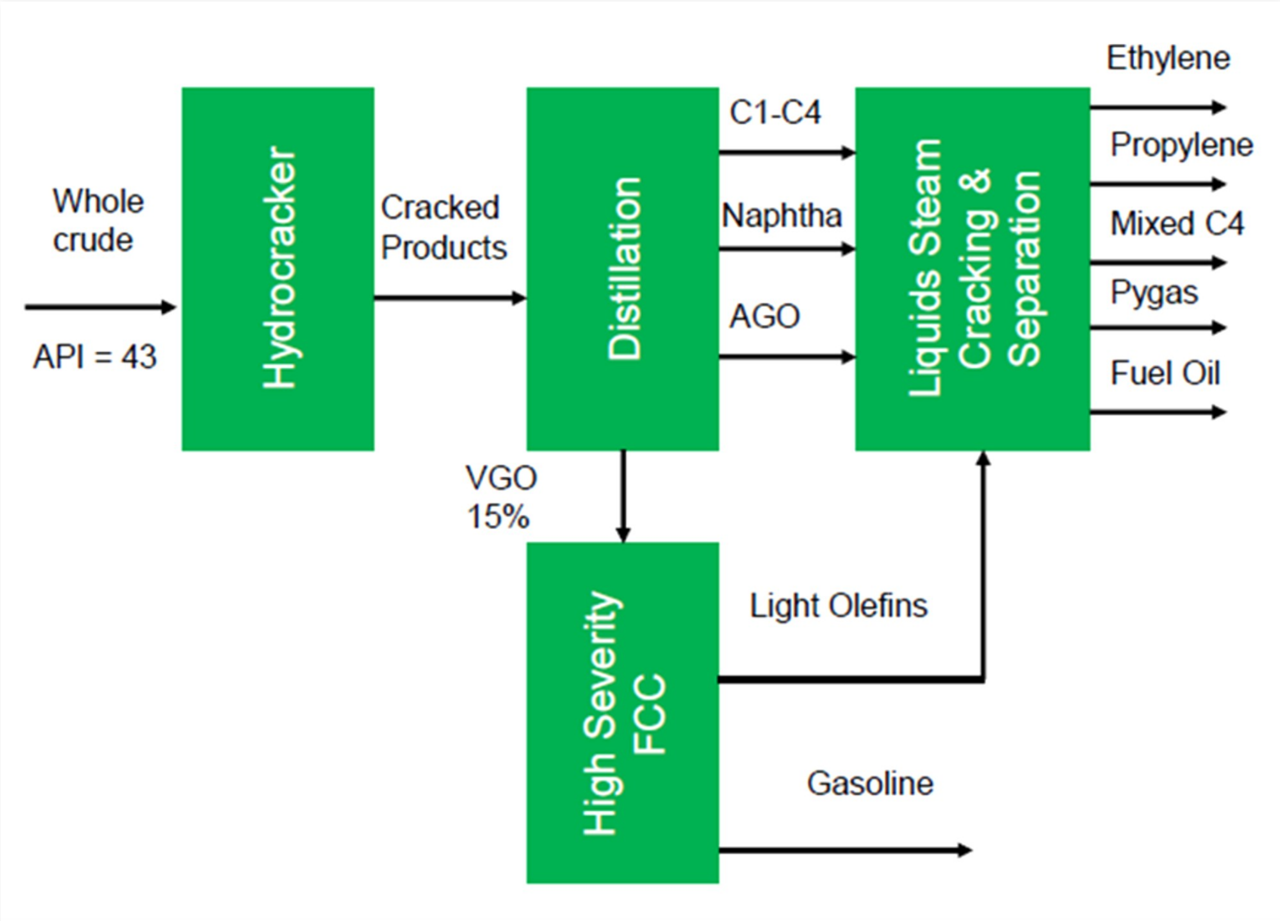



*Figure 10 – Saudi Aramco Crude Oil to Chemicals Concept (IHS Markit, 2017)*

The process presented in Figure 8 is based on the quality of the crude oil and deep conversion technologies like High Severity or petrochemical FCC units and deep hydrocracking technologies. The processed crude oil is light with low residual carbon that is a common characteristic in the Middle East crude oils, the processing scheme involves deep catalytic conversion process aiming to reach maximum conversion to light olefins. In this refining configuration, the petrochemical FCC units have a key role to ensure high added value to the processed crude oil.

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 plants, as presented in Figure 9. Other available technologies are the HS-FCC™ process commercialized by Axens Company, and INDMAX™ process licensed by Lummus Company. The basic process flow diagram for HS-FCC™ technology is presented in Figure 11.

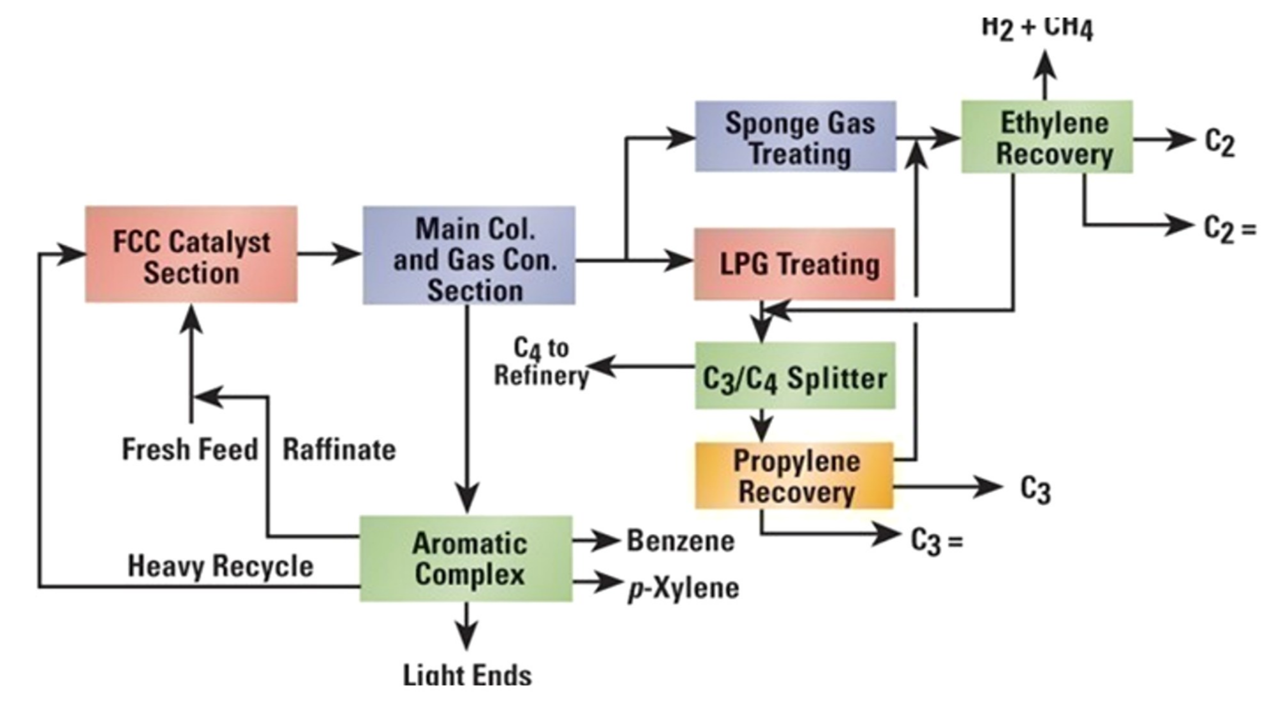



*Figure 11 – PetroFCC™ Process Technology by UOP Company.*

It's important to consider that the technology presented in Figure 11 is based on Petrochemical FCC units that present especial design due to the most severe operating conditions.

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

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 12 presents a block diagram showing a case study demonstrating how the petrochemical FCC unit, in this case the INDMAX™ technology by Lummus Company, can maximize the yield of petrochemicals in the refining hardware. Another technology dedicated to maximizing olefins from residue is the R2P™ process, developed by Axens.

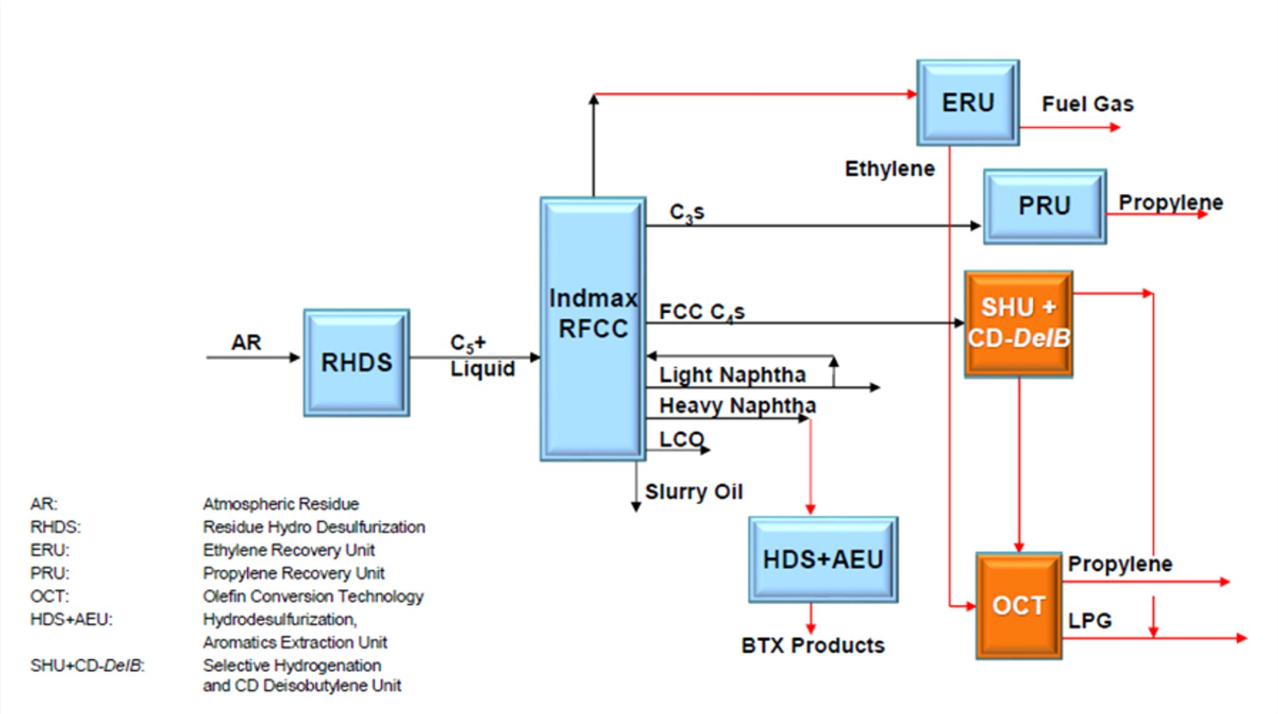



*Figure 12 – Olefins Maximization in the Refining Hardware with INDMAX™ FCC Technology by Chevron Lummus Global Company (SANIN, A.K., 2017)*

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.

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

C + ½ O2 → CO (Partial Combustion) ΔH = - 27 kcal/mol

C + O2 → CO2 (Total Combustion) ΔH = - 94 kcal/mol

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

Another key refining technology to crude oil to chemicals refineries is the hydrocraking units. Despite the high performance, the fixed bed hydrocracking technologies can be economically effective to treat crude oils directly cue to the possibility of short operating lifecycle. Technologies that use ebullated bed reactors and continuum catalyst replacement allow higher campaign period and higher conversion rates, among these technologies the most known are the H-Oil and Hyvahl™ technologies developed by Axens Company, the LC-Fining Process by Chevron-Lummus, and the Hycon™ process by Shell Global Solutions. These reactors operate at temperatures above 450 oC and pressures until 250 bar. Figure 13 presents a typical process flow diagram for a LC-Fining™ process unit, developed by Chevron Lummus Company while the H-Oil™ process by Axens Company is presented in Figure 14.

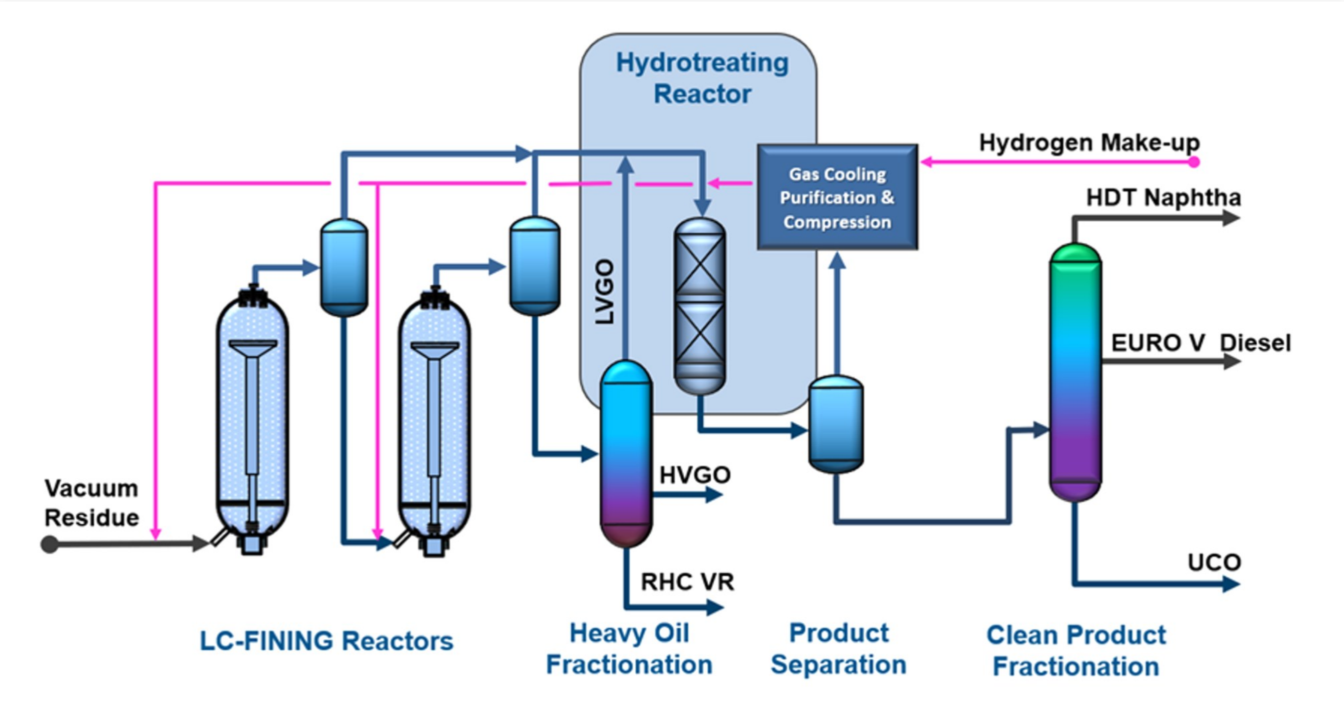



*Figure 13 – Process Flow Diagram for LC-Fining™ Technology by CLG Company (MUKHERJEE & GILLIS, 2018)*

Catalysts applied in hydrocracking processes can be amorphous (alumina and silica-alumina) and crystalline (zeolites) and have bifunctional characteristics, once the cracking reactions (in the acid sites) and hydrogenation (in the metals sites) occur simultaneously.

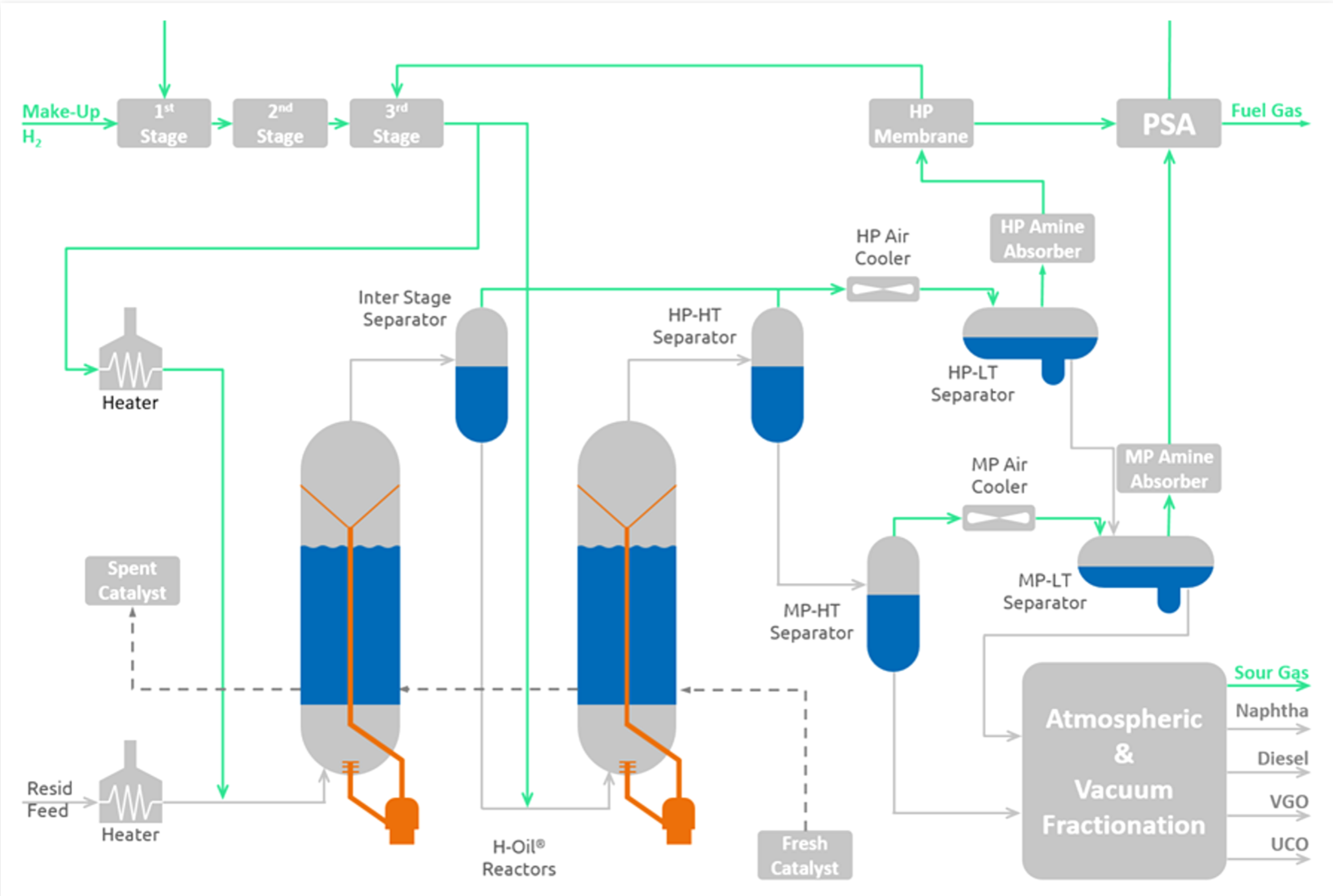



*Figure 14 – Process Flow Diagram for H-Oil™ Process by Axens Company (FRECON et. al, 2019)*

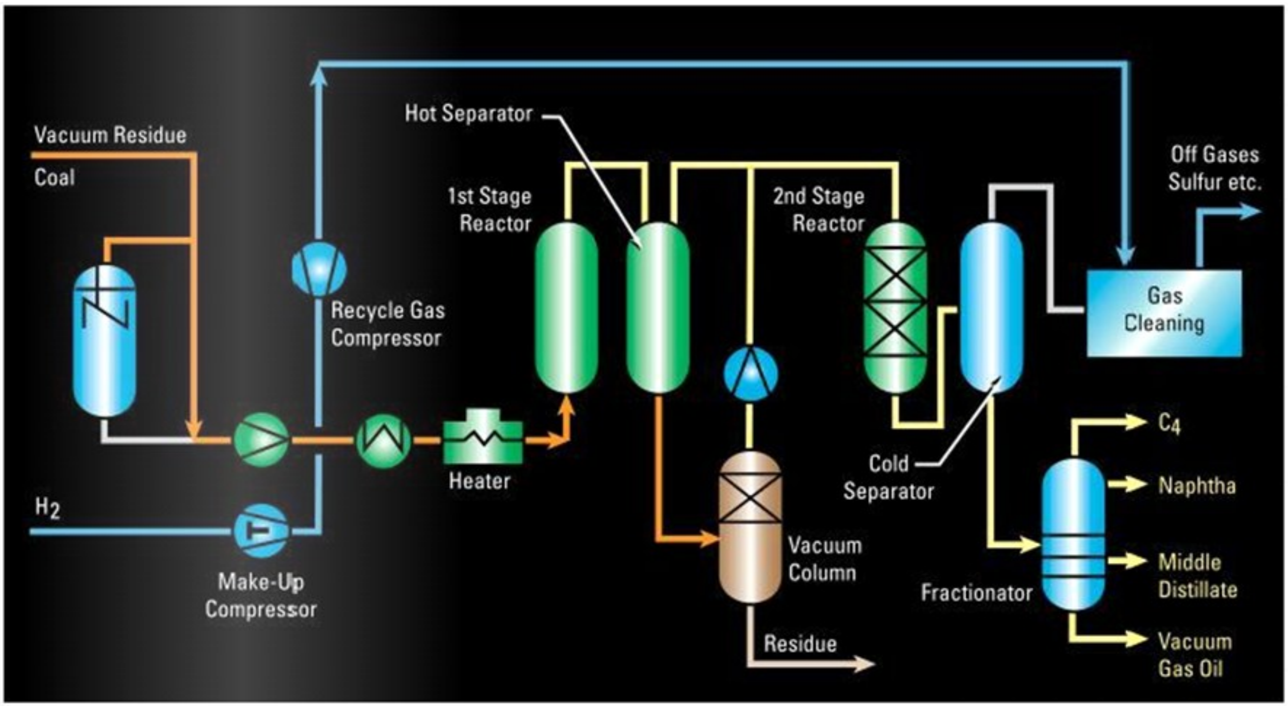



*Figure 15 – Basic Process Arrangement for VCC™ Slurry Hydrocracking by KBR Company (KBR Company, 2019)*

An improvement in relation to ebullated bed technologies is the slurry phase reactors, which can achieve conversions higher than 95 %. In this case, the main available technologies are the HDH™ process (Hydrocracking-Distillation-Hydrotreatment), developed by PDVSA-Intevep, VEBA-Combicracking Process (VCC)™ commercialized by KBR Company, the EST™ process (Eni Slurry Technology) developed by Italian state oil company ENI, and the Uniflex™ technology developed by UOP Company. Figure 15 presents a basic process flow diagram for the VCC™ technology by KBR Company.

In the slurry phase hydrocracking units, the catalysts in injected with the feedstock and activated in situ while the reactions are carried out in slurry phase reactors, minimizing the reactivation issue, and ensuring higher conversions and operating lifecycle. Figure 16 presents a basic process flow diagram for the Uniflex™ slurry hydrocracking technology by UOP Company.

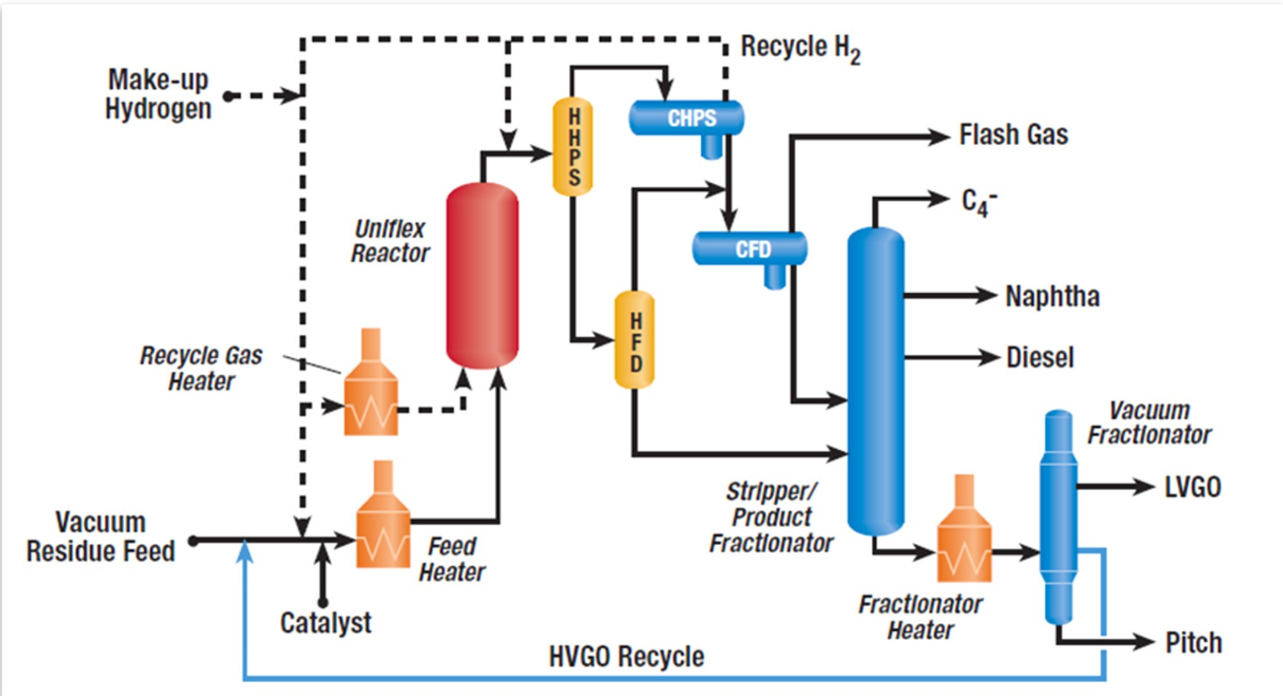



*Figure 16 – Process Flow Diagram for Uniflex™ Slurry Phase Hydrocracking Technology by UOP Company (UOP Company, 2019)*

Other commercial technologies to slurry hydrocracking process are the LC-Slurry™ technology developed by Chevron Lummus Company and the Microcat-RC™ process by Exxon Mobil Company.

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

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 %). Figure 17 presents a typical steam cracking unit applying naphtha as raw material to produce olefins.

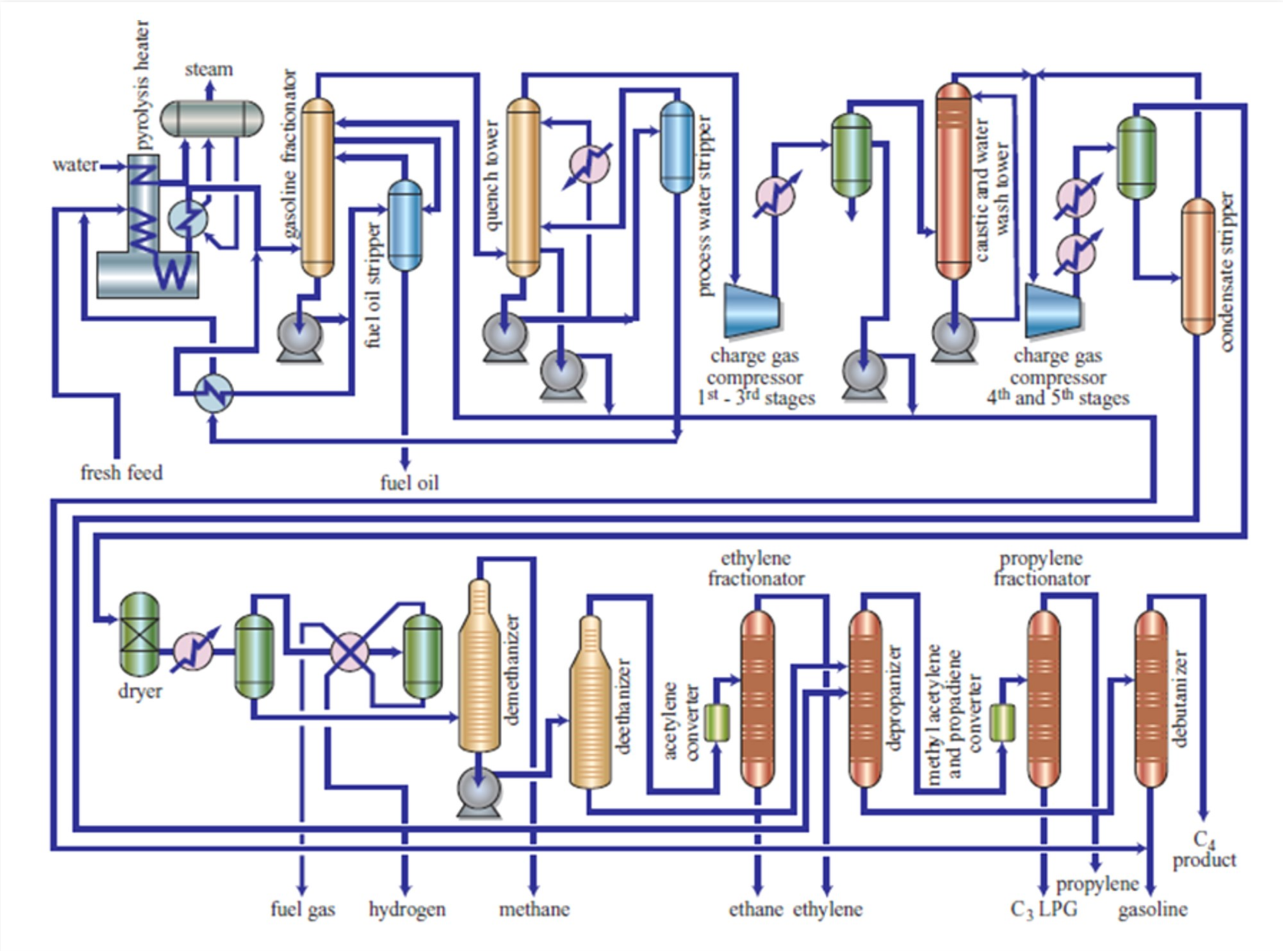



*Figure 17 – Typical Naphtha Steam Cracking Unit (Encyclopedia of Hydrocarbons, 2006)*

Due to his relevance, great technology developers have dedicated their efforts to improve 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™ 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 are carried out under high temperatures, between 500 oC to 700 oC according to the characteristics of the feed (inlet temperature). For heavier feeds like gas oil, lower temperature is applied aiming to minimize the coke formation, the combination of high temperatures and low residence time are the main characteristic of the steam cracking process.

As quoted above, some technology developers are dedicating their efforts to develop commercial crude to chemicals refineries. Figure 18 presents the concept of crude to chemicals refining scheme by Chevron Lummus Company.

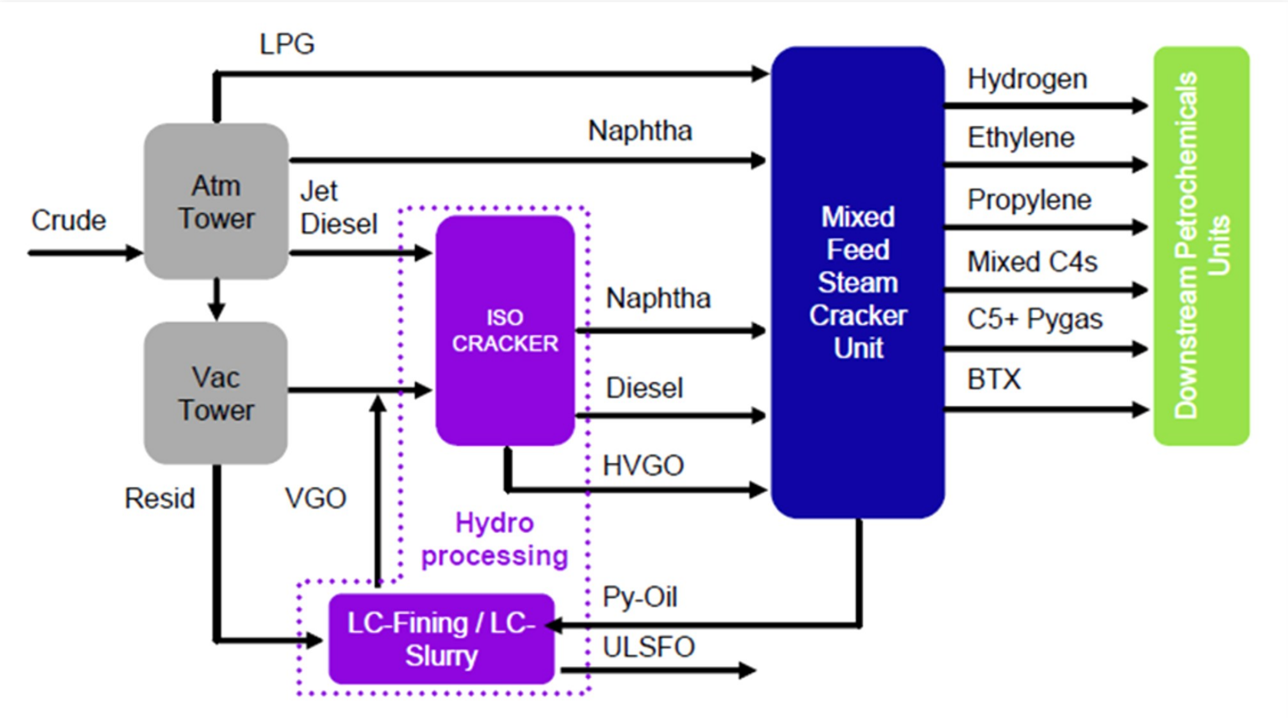



*Figure 18 – Crude to Chemicals Concept by Chevron Lummus Company (Chevron Lummus Global Company, 2019)*

Another crude to chemicals refining arrangements is proposed by Chevon Lummus Company, applying the synergy of residue upgrading strategies to maximize the petrochemical intermediates production, Figure 19 presents a crude to chemicals arrangement relying on delayed coking unit.

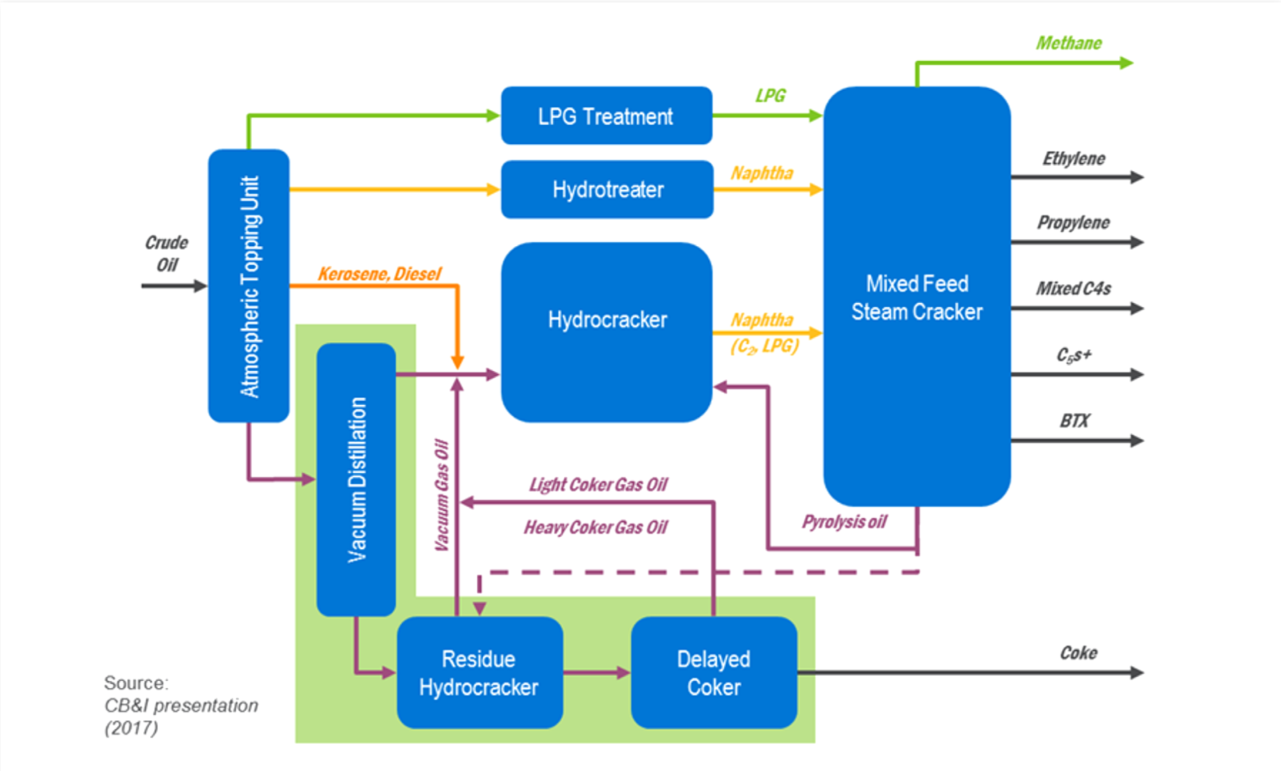
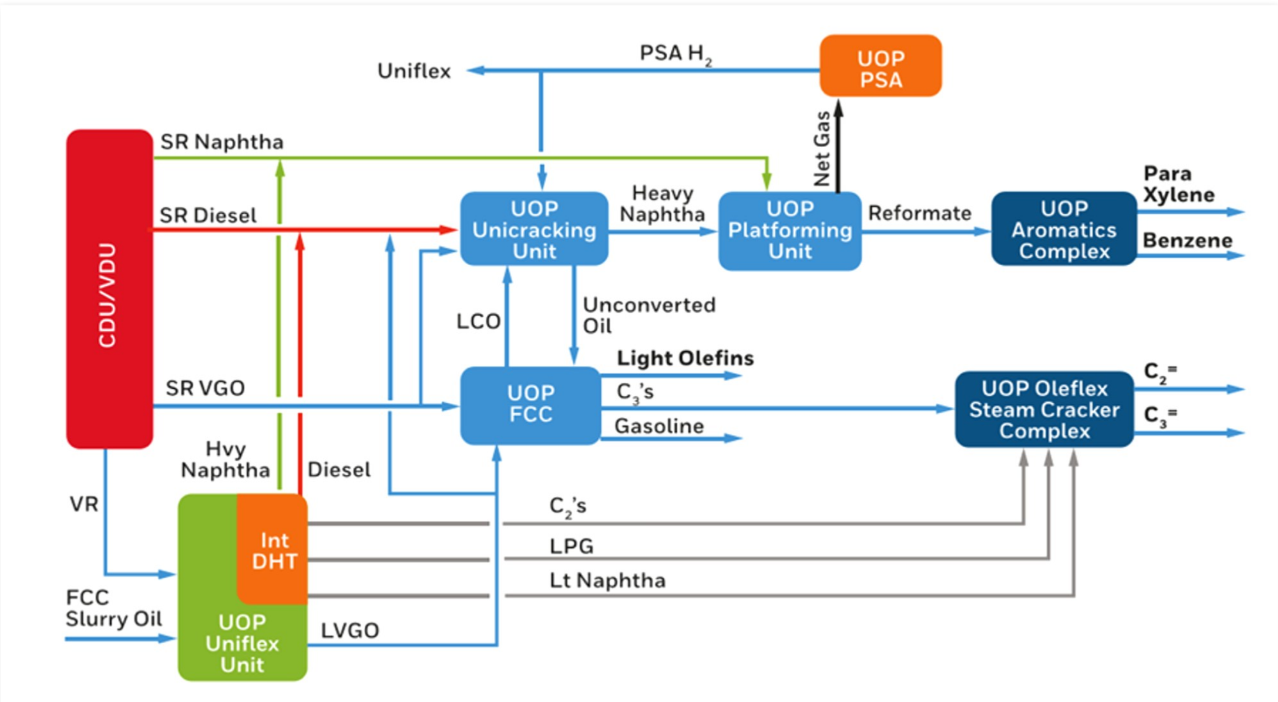



*Figure 19 – Crude to Chemicals Concept by Chevron Lummus Company (Nexant Company, 2018)*

Another great refining technology developers like UOP, Shell Global Solutions, ExxonMobil, Axens, and others are developing crude to chemicals technologies, reinforcing that this is a trend in the downstream market. Figure 20 presents a highly integrated refining configuration capable of converting crude oil to petrochemicals developed by UOP Company.




*FiFigure 20 – Integrated Refining Configuration Based in Crude to Chemicals Concept by UOP Company.*

As presented in Figure 20, the production focus changes to the maximum adding value to the crude oil through the production of high added value petrochemical intermediates or chemicals to general purpose leading to a minimum production of fuels.

As aforementioned, big players as Saudi Aramco Company have been made great investments in COC technologies aiming to achieve even more integrated refineries and petrochemical plants, raising considerably his competitiveness in the downstream market. Major technology licensors like Axens, UOP, Lummus, Shell, ExxonMobil, etc. have been applied resources to develop technologies capable to allow a closer integration in the downstream sector aiming to allow refiners to extract the maximum added value from the processed crude oil, an increasing necessity in a scenario where the refining margins are under pressure.

Is expected that some of these capital investments was postponed due to the economic crisis provoked by the COVID-19 pandemic, but these data reinforce the trend in the market, it's interesting to quote that close to 64 % of the global crude to chemicals investments are made by Asian players. Considering just the petrochemical complexes focused on PX (Para Xylene), we have total capital investments around 87 US billion dollars presented in Figure 21.

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

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

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


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
For more information:
www.welchem.com
service@welchem.com
WELCHEM
PROCESS TECHNOLOGY

Figure 22 presents a comparison between the petrochemicals yields of traditional refineries, a benchmark integrated refinery and crude to chemicals complexes, according to data from Wood Company.

Analyzing Figure 22 it's possible to note the higher added value reached in crude to chemicals refineries when compared even with highly integrated refineries. Figure 23 presents an example of how a crude to chemicals refineries can reach very high petrochemicals yield, in this case, considering the Hengli Petrochemical Complex in China.

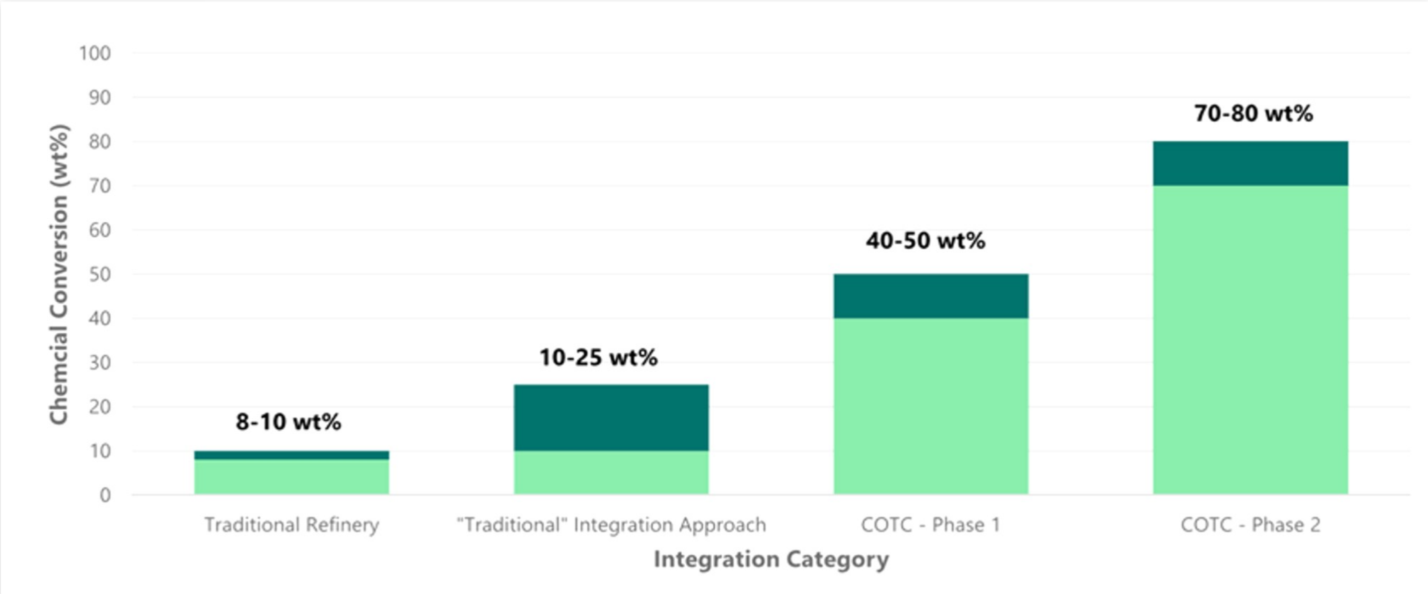



*Figure 22 – Petrochemicals Yield Comparison (Wood Company, 2024)*

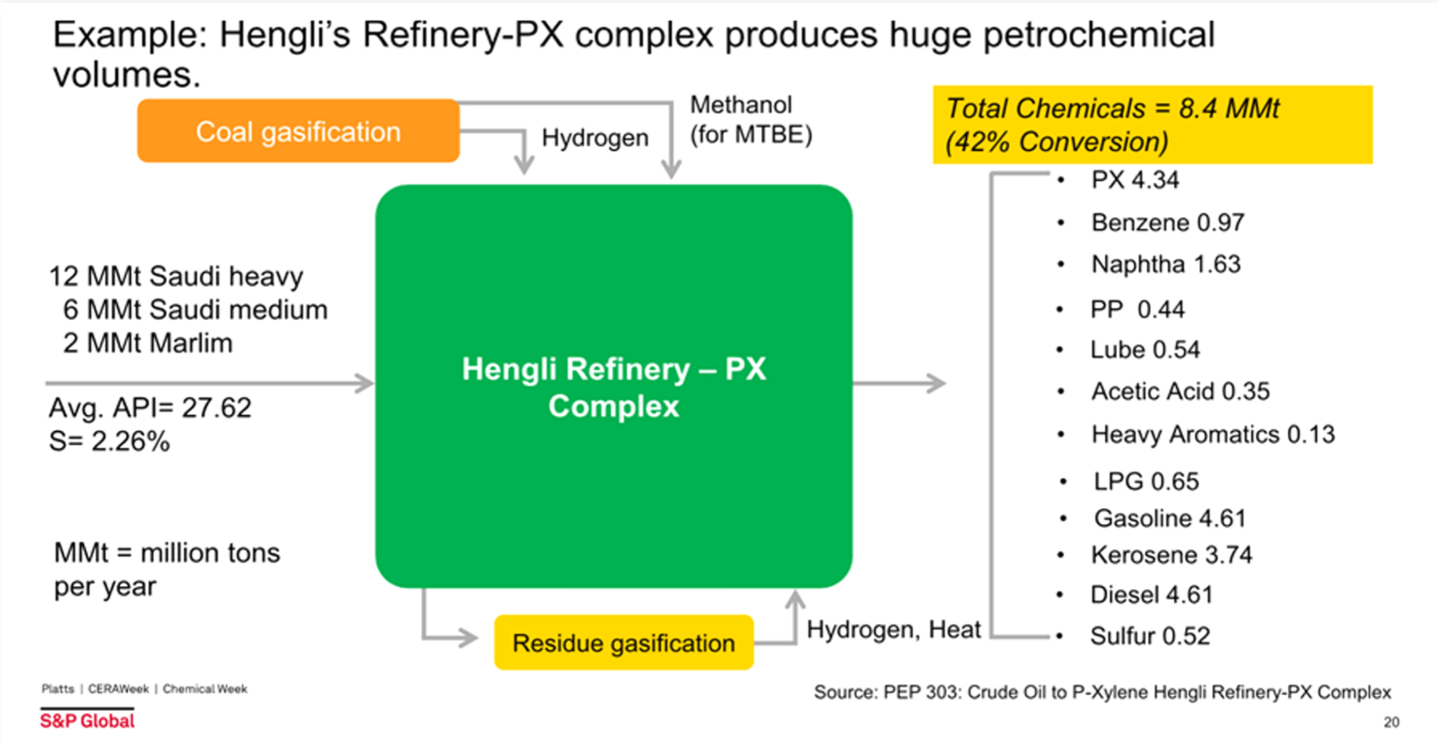



*Figure 23 – Petrochemicals Yield for the Hengli Crude to Chemicals Complex (S&P Global Commodity Insights, 2024)*

It's interesting to quote 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 24 present a comparison between the relation of crude oil distillation capacity and the integrated refinery capacity for each continent.

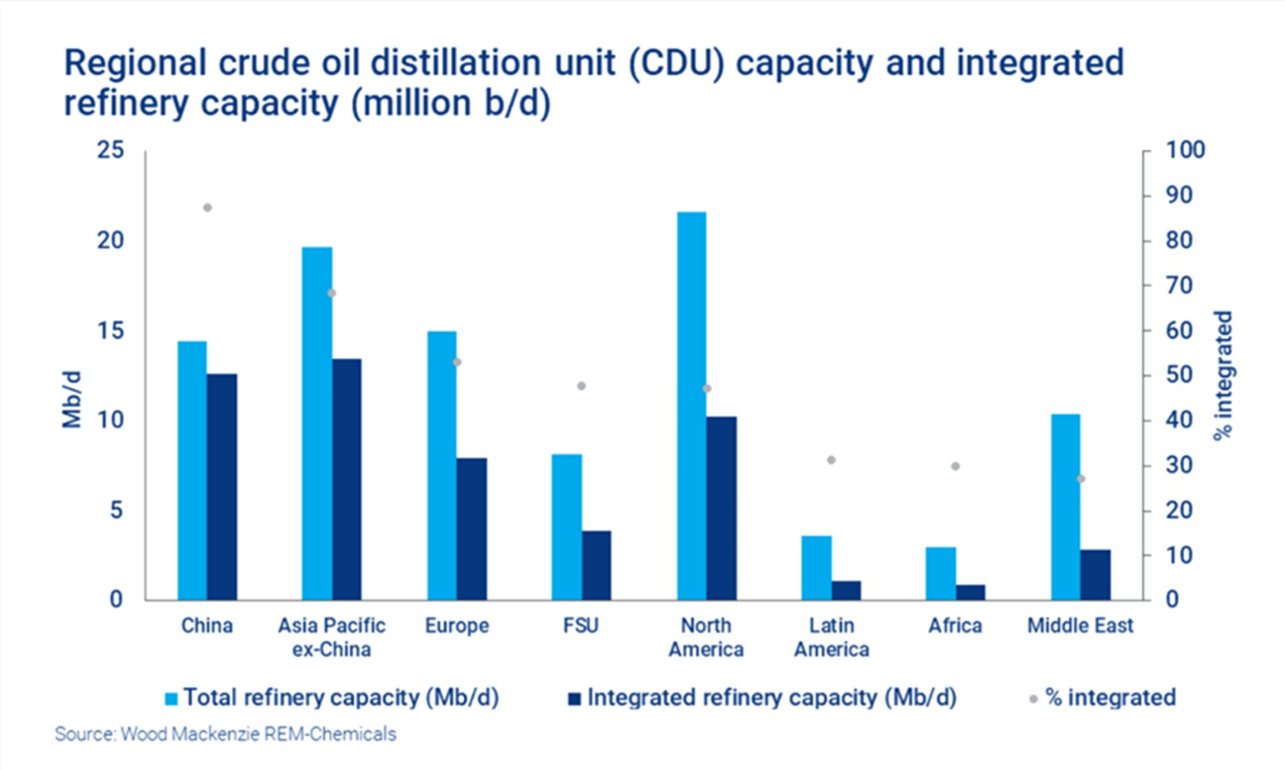



*Figure 24 – Crude Oil Distillation Capacity and Integrated Refinery Capacity for Each Continent (Wood Mackenzie, 2023)*

Figure 24 shows that the Asian players have a superior integration capacity of their refining assets in comparison with another continents, as mentioned above, this can be translated in a significant competitive advantage to the Asian players and a great potential o competitive imbalance of the downstream market considering the recent forecasts which indicates growing demand for petrochemicals. Furthermore, it’s possible to see the power of the China in the Asian and global downstream market.

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

Recently, Lummus Company announced the implementation of your proprietary crude to chemicals technology, called TC2C™ (Thermal Crude to Chemicals) by a big player of the downstream industry in the Asian market, reinforcing the growing trend of crude to chemicals in the Asian continent. The TC2C™ process can reach a yield of 70% in mass of high value petrochemicals from light crudes as informed by the licensor. Figure 25 presents a block diagram for this technology.

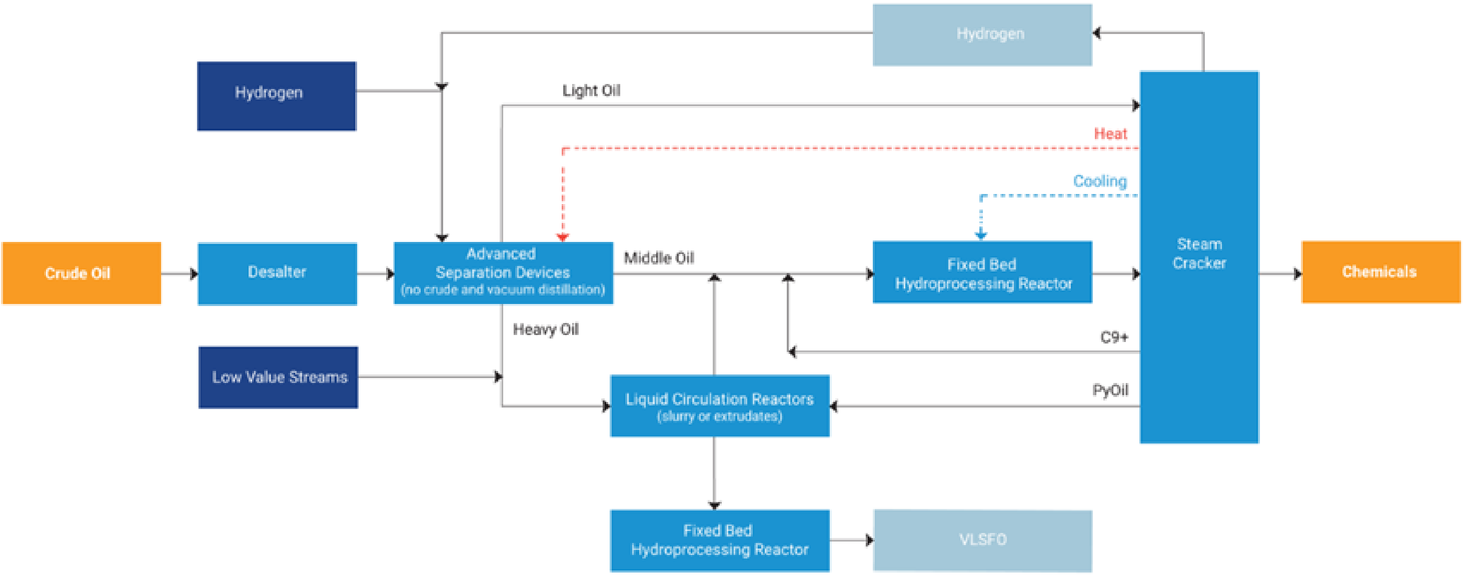



*Figure 25 – Block Diagram for the TC2C™ Crude to Chemicals Technology by Lummus Company*

**Available Crude to Chemicals Processing Routes**

Nowadays, there are three technically available routes that are considered to capital investments to crude to chemicals refining complexes. Figure 26 presents the concepts based on the information of S&P Global Commodities Insights Company.

The conventional routes consider the processing of crude oil in a conventional crude oil refinery, producing petrochemical intermediates like naphtha which is supplied to a petrochemical asset like a steam cracking unit. The ExxonMobil route is based on the direct feed of selected crude oils, normally light and low contaminants crudes, to petrochemical assets, while the Chinese enterprise Hengli Zhejiang Shenghong Henyi project consider the feed of mixed crude oil slate to a crude to PX (Para-Xylene) complex to ensure the domestic Chinese market that present high demand by light aromatics (BTX). A conventional highly integrated refining hardware is capable to achieve 15 to 20 % of petrochemicals yield while a crude to chemicals refinery can reach up to 70 % as presented in Figure 22.

As aforementioned, the Aramco/Sabic concept is based on a high complexity refining hardware to convert selected crude oil (light) to maximize the yield of petrochemical intermediates, mainly light olefins.

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 a transitive period where, as presented in Figure 1, the transportation fuels are responsible for a 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 competitive mistake.

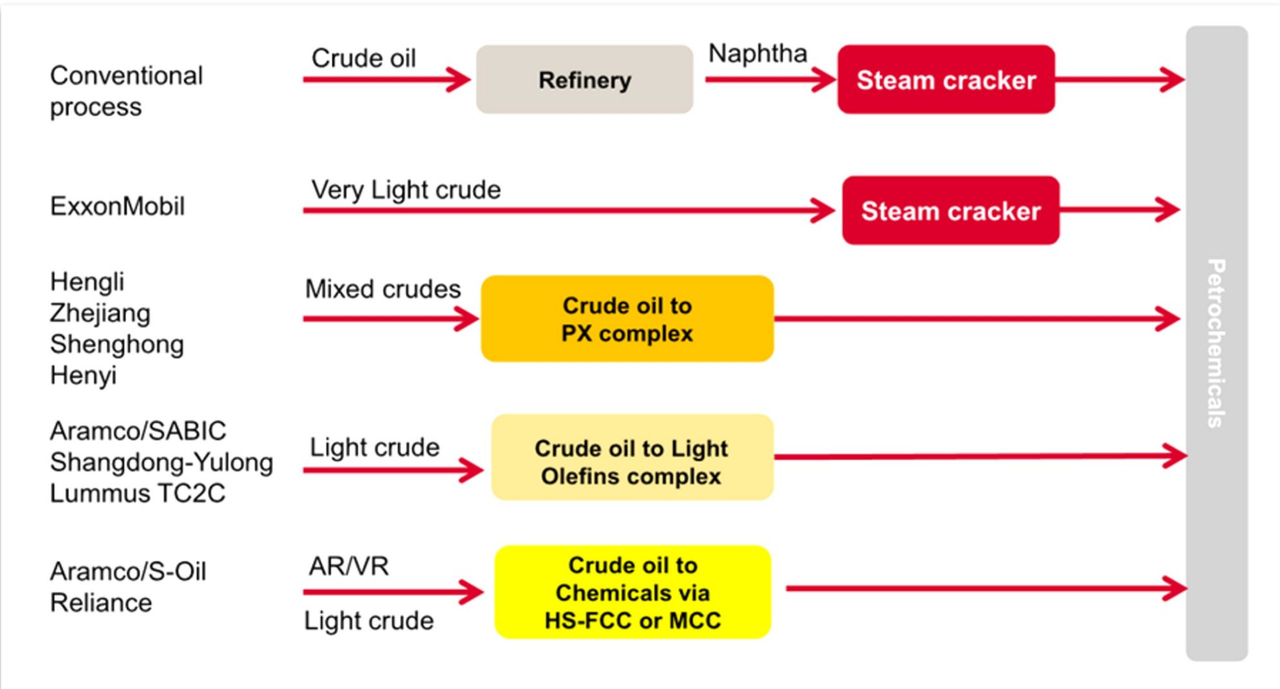



*Figure 26 – Crude to Chemicals Concepts (S&P Global Commodities Insights Company, 2024)*

## Conclusion

Nowadays, it is still difficult to imagine the global energetic matrix free of fossil transportation fuels, especially in developing economies. Despite this fact, recent forecasts, and growing demand by petrochemicals as well as 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.

Based on description above it’s possible to apply the article published by W. Chan Kim and Renée Mauborge called “Blue Ocean Strategy” in Harvard Business Review, to classify the competitive markets in the downstream industry. In this article the authors define the conventional market as a red ocean where the players tend to compete in the existing market focusing on defeating competitors through the exploration of existing demand, leading to low differentiation and low profitability. The blue ocean is characterized by looking for space in non-explored (or few explored markets), creating and developing new demands and reaching differentiation. This model can be applied (with some specificities once is a commodity market) to the downstream industry, considering the traditional transportation fuels refineries and the petrochemical sector.

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

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

In the extreme side of the petrochemical integration trend, there are zero fuels refineries, as quoted above, it’s still difficult to imagine the downstream market without transportation fuels, but it seems a serious trend and the players of the downstream sector need to consider the focus change in their strategic plans like opportunity and threat. As discussed above, even the players with less capital power can take actions to maximize the petrochemicals yield in their refining hardware. Despite this scenario, disruption is still a hard work in the case of downstream industry, but the crude to chemicals refining assets can produce a competitive imbalance in the market, especially due to the concentration of capital investments in the Asian market. The downstream industry has a history of adaptation of crude consumption patterns through the years and the crude to chemicals refining assets represents an evolution aiming to maximize the added value to the processed crude, reaching petrochemicals yield higher than 40 %.

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. Recently one of the biggest petrochemical players, SABIC Company, announces the intention to make investments in a new crude to chemicals refinery with capacity of 400.000 barrels per day and the SATORP Company (A joint venture between Total Energies and Aramco companies) announced USS 11 billion dollars in capital investments in the Amiral petrochemical complex to promote closer integration with Jubail refinery (Saudi Arabia), reinforcing the trend of closer integration between refining and petrochemical assets in order to maximize the added value to the processed crude.

Less integrated refiners tend to compete in a kind of red ocean market where the refining margins tend to be lower due to the lower added value to the crude oil like transportation fuels, high sulfur fuel oil, and asphalt. Despite this, and according to the characteristics of the local markets, it's possible to achieve economic sustainability, in this case, capital discipline and operational efficiency are even more important for these players.

In our point of view the current oversupply crisis is a natural consequence of the massive capital investments made by the Chinese players in crude to chemicals refineries which give great production capacity and scale economy able to reduce drastically the production costs for the Chinese players in comparison with another players. Considering the trend among the refiners to promote closer integration between refining and petrochemical assets due to the hostile scenario for fossil fuels in the last years, the players with capital power made significant changes in their refining assets to maximize petrochemicals in detriment of fuels, with a refining park with great potential to produce petrochemicals the Chinese get the "control" of the petrochemical market.

**References**

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

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

CHARLESWORTH, R. ***Crude oil to Chemicals (COTC) – An Industry Game Changer?*** IHS Markit Company, Presented in 14th GPCA Forum, 2019.

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

COUCH, K. ***The Refinery of the Future – A Flexible Approach to Petrochemicals Integration.*** Honeywell UOP Company, Presented in 12th Asian Downstream Summit, 2019.

COUCH, K.; CARP, S. - ***Advances in Crude to Chemicals.*** Honeywell UOP Company PTQ Magazine, 2021.

CUI, K. – ***Why Crude to Chemicals is the Obvious Way Forward.*** Wood Mackenzie, 2019.

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

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

FAISCA, N. – ***From Crude Oil to Chemicals: Back to the Fundamentals.*** Nexant Company, 2018.

FRECON, J.; LE BARS, D.; RAULT, J. – ***Flexible Upgrading of Heavy Feedstocks.*** PTQ Magazine, 2019.

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.

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.

LIU, S.; KOTS, P.A.; VANCE, B.C.; DANIELSON, A.; VLACHOS, D.G. ***Plastic Waste to Fuels by Hydrocracking at Mild Conditions,*** Science Advances, 2021.

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

MUKHERJEE, U.; GILLIS, D. – ***Advances in Residue Hydrocracking.*** PTQ Magazine, 2018.

MULDOON, B.S. – ***Profit Pivot Points in a Crude to Chemicals Integrated Complex*** – Presented at Ethylene Middle East Technology Conference, 2019.

MURPHY, J. J.; PAYN, C.F. – ***Oil to Chemicals: New Approaches***. PTQ Magazine, 2019.

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

***Refinery-Petrochemical Integration (Downstream SME Knowledge Share).*** Wood Mackenzie Presentation, 2019.

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.

ZHANG, Z. ***Crude oil to chemicals: Challenges and Opportunities in a Sustainable World*** - Wood Company Presentation at METECH 2024, 2024.

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

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# Process Design of Industrial De-Super Heaters

**Jayanthi Vijay Sarathy**

In industrial applications, while superheated steam is more useful for power generation due to its higher sensible heat, allowing steam turbines to extract more work, saturated steam is also required for heating applications. Towards this, a portion of the superheated steam from the steam generation system (steam boiler) is brought down close to its saturation point, i.e., desuperheating.

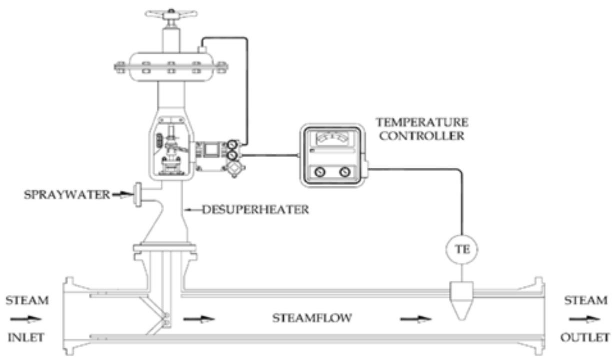



*Figure 1. Desuperheater schematic [1]*

A typical desuperheating system consists of an inlet pipe carrying superheated steam, a flow control valve which supplies cooling water to "quench" the superheated steam, a temperature controller and sensor that detects the desuperheated steam to vary the cooling water supply.

Desuperheaters can be broadly classified as direct and indirect contact type desuperheaters. Shell and Tube heat exchangers are an example of direct contact desuperheaters whereby either cooling air or water is used to cool the superheated steam until saturation. Whereas in direct contact desuperheaters, the cooling medium is often the same as the liquid or vapour required to be desuperheated. In case of steam, a measured amount of cooling water is sprayed into the incoming superheated steam at a constant pressure to reduce the sensible heat.

The following article covers the process design aspects of direct contact desuperheaters.

## General Notes

1. Superheated steam is primarily a combination of enthalpy of water, latent heat of vapourization at saturation, and enthalpy of superheat. Between them, the bulk of the energy is concentrated in the latent heat of vapourization component.

2. The set temperature of desuperheated steam at TE location [Fig 1] is typically kept atleast 60C above the saturation [2]. This is because as desuperheated steam passes through the discharge piping, there would be heat losses to the surroundings, causing steam condensation at saturation point.

3. Since the spray nozzle supplying cooling water needs to atomize the liquid, a strainer of mesh size of ~0.25 mm can be provided upstream of the supply line, to prevent the atomizer from clogging due to debris [2].

4. For satisfactory evaporation, the ratio of cooling water mass flow to desuperheated mass flow should be < 15% [2].

5. Since the downstream temperature sensor (TE) detects the temperature of the desuperheated steam to adjust the cooling water flow, sufficient transition length must be provided for complete mixing and

the steam flow to fully develop. For this, a minimum of 25 feet [7.62 m] must be provided. Alternatively, a distance of 15D to 20D can be considered.

6. The ensuing desuperheated steam velocity should be in the range of 25 to 40 m/s for fixed nozzle types to limit noise, erosion and corrosion [3]. For discharge lines, longer than 50 m, a pressure drop check needs to be performed.

7. For fixed type spray nozzles, the turndown ratio is of the order of 4:1 and upto 40:1 for variable spray type atomizers.

**Case Study**

A stream of 100,000 kg/h of superheated steam at 10 bara and 3000C is required to be quenched with cooling water at 15 bara and 1500C. Towards this, the cooling water flow rate is to be estimated with a suitable flow control valve (FCV) to deliver desuperheated steam at 1860C, as measured at the downstream temperature element (TE). The TE then accordingly provides a signal to the FCV to alter the cooling water flow. The piping material chosen is ASTM A108 Grade B with schedule 80 wall thickness. As per vendor's recommendation, the discharge piping length is 15 m.

**Assumptions**

Prior to initiating calculations, the following assumptions are made,

1. No insulation is provided on the desuperheated line. This adds to the heat losses and reduces the required spray / cooling water supply flow via the FCV.

2. The wall temperature of the desuperheater discharge piping is assumed to be at a constant temperature of 300C.

3. The cooling water supplied is taken to be debris free with no dissolved salts. The enthalpy and saturation data is based on steam tables IAPWS IF-97.

4. Any pressure losses after mixing of superheated steam and cooling water is neglected. Therefore the ensuing desuperheated steam is also taken to be at the same pressure as that of superheated steam, i.e., 10 bara.

5. All pipe sizes are based on ASME B36.10M. The flow control valve sizing for liquids is based on ANSI/ISA S75.01 [6]. In the current undertaking the FCV upstream and downstream piping are assumed to be of the same size, i.e., NPS 2 inch [OD 60.3 mm with wall thickness (WT) of 5.54 mm as per ASME B36.10M]

**Methodology**

To estimate the spray / cooling water mass flow rate required, the set temperature above the saturation temperature [$T_{sat}$] of the desuperheated stream is initially assumed, and then iteratively solved until the temperature at the TE is 1860C. Therefore, taking a set temperature of 250C above $T_{sat}$ of the desuperheated steam at 10 bara,

Enthalpy [H] of superheated steam at 10 bara and 3000C

$$h_s = H_{10\ bara,300°C} = 3{,}051.7\ kJ/kg \qquad (1)$$

Enthalpy [H] of cooling water at 15 bara and 1500C

$$h_{cw} = H_{15\ bara,150°C} = 632.9\ kJ/kg \qquad (2)$$

Saturation Temperature [Tds,sat] of desuperheated steam at 10 bara,

$$T_{ds,sat,10\ bara} = 179.9°C \qquad (3)$$

Including the initial set temperature of 250C, the desuperheated steam temperature [Tds] at the outlet of the spray nozzle,

$$T_{ds} = 179.9 + 25 = 204.9°C \qquad (4)$$

Enthalpy of desuperheated steam at 10 bara and 204.90C,

$$h_{ds} = H_{10\ bara,204.9°C} = 2840\ kJ/kg \qquad (5)$$

The cooling water supply temperature is now estimated as,

$$m_{cw} = 100{,}000 \times \frac{[3051.7 - 2840]}{[2840 - 632.9]} \qquad (6)$$

$$m_{cw} = 9590\ kg/h \qquad (7)$$

The total desuperheated flow rate is,

$$m_{ds} = m_s + m_{cw} = 100{,}000 + 9{,}590 \ kg/h \quad (8)$$

$$m_{ds} = 109{,}590 \, kg/h \quad (9)$$

**Desuperheated Steam Discharge Pipe Size**

Based on the desuperheated mass flow rate [$m_{ds}$] estimated as above, the discharge piping size can be calculated accordingly. A preliminary pipe sizing for a desuperheated steam flow rate of 109,590 kg/h yields,

$$\rho_{ds,10\ bara,204°C} = 4.791 \ kg/m^3 \quad (10)$$

$$Q_{ds} = \frac{m_{ds}}{\rho_{ds}} = \frac{109590}{4.791 \times 3600} = 6.3539 \ m^3/s \quad (11)$$

Computing the preliminary velocity for NPS 18 inch, 20 inch and 22 inch, 80 Sch pipe sizes,

$$NPS\ 18" \rightarrow OD\ 457.2\ mm, WT\ 12.7\ mm \quad (12)$$

$$NPS\ 20" \rightarrow OD\ 508\ mm, WT\ 12.7\ mm \quad (13)$$

$$NPS\ 22" \rightarrow OD\ 558.8\ mm, WT\ 12.7\ mm \quad (14)$$

The preliminary velocities for NPS 18" is,

$$V_{ds} = \frac{Q_{ds}}{A_{ds}} = \frac{6.3539}{\frac{\pi}{4} \times \left[\left(\frac{457.2}{1000}\right) - \left(2 \times \frac{12.7}{1000}\right)\right]^2} = 43.5\ m/s \quad (15)$$

The preliminary velocities for NPS 20" is,

$$V_{ds} = \frac{Q_{ds}}{A_{ds}} = \frac{6.3539}{\frac{\pi}{4} \times \left[\left(\frac{508}{1000}\right) - \left(2 \times \frac{12.7}{1000}\right)\right]^2} = 34.7\ m/s \quad (16)$$

The preliminary velocities for NPS 22" is,

$$V_{ds} = \frac{Q_{ds}}{A_{ds}} = \frac{6.3539}{\frac{\pi}{4} \times \left[\left(\frac{558.8}{1000}\right) - \left(2 \times \frac{12.7}{1000}\right)\right]^2} = 28.4\ m/s \quad (17)$$

Between the 18", 20" and 22" NPS pipe sizes, the NPS 20", Sch 80 pipe is chosen, considering it is smaller with lower material cost, and that the calculated steam velocity is between the acceptable range of 25 m/s and 40 m/s.

**Temperature Estimation at TE**

Taking the pipe material properties as follows, the heat transfer rate is calculated to estimate the temperature at TE sensing point.

| | | |
|---|---|---|
| Selected Material Grade | ASTM A106 Gr B | |
| Selected Pipe NPS | 20 | inch |
| Pipe Outer Diameter [OD] | 508.0 | mm |
| Pipe Wall Thickness [WT] | 12.7 | mm |
| Pipe Inner Diameter [ID] | 482.6 | mm |
| Thermal conductivity [$k_{pipe}$] | 45.5 | W/m.K |
| Pipe Density [$\rho_{pipe}$] | 7850 | kg/m³ |
| Steam Velocity [$V_{ds}$] | 34.7 | m/s |
| Pipe Wall Temperature [$T_w$] | 30.0 | $^0$C |

*Table 1. Desuperheater Piping Details*

Taking the and desuperheated steam data at 10 bara as follows,

| | | |
|---|---|---|
| Steam Pressure [$P_{ds}$] | 10 | bara |
| Steam Temperature [$T_{ds}$] | 204.9 | $^0$C |
| Steam Density [$\rho_{ds}$] | 4.7910 | kg/m³ |
| Steam Dynamic Viscosity [$\mu_{ds}$] | 0.0161 | cP |
| Steam Specific Heat [$C_{p,ds}$] | 2.393 | kJ/kg.K |
| Steam Thermal conductivity [$k_{ds}$] | 0.0363 | W/m.K |
| Required Steam T at TE | 186 | $^0$C |

*Table 2. Desuperheated Steam Properties*

The internal heat transfer coefficient [$h_i$] using Dittus-Boelter correlation is calculated as,

$$Nu = 0.023 \times Re^{0.8} \times Pr^{0.3} \quad (18)$$

Where,

Re = Reynolds Number [-]

Pr = Prandtl Number [-]

Nu = Nusselt Number [-]

Estimating the dimensionless numbers,

$$Re = \frac{D_i \times V_{ds} \times \rho_{ds}}{\mu_{ds}} = \frac{482.6 \times 34.7 \times 4.791}{1000 \times 0.0000161} \quad (19)$$

$$Re \approx 4,986,976 \quad (20)$$

$$Pr = \frac{C_{p,ds} \times \mu_{ds}}{k_{ds}} = \frac{2393 \times 0.0000161}{0.0363} \approx 1.0613 \quad (21)$$

$$Nu = 0.023 \times 4986976^{0.8} \times 1.0613^{0.3} \quad (22)$$

$$Nu \approx 5343 \quad (23)$$

The internal heat transfer coefficient [h] is,

$$h = \frac{k_{ds} \times Nu}{D_i} = \frac{0.0363 \times 5343}{0.4826} \approx 402\ W/m^2K \quad (24)$$

For a discharge piping length of 15 metres as suggested by the vendor, the temperature at TE sensing point is,

$$\frac{\pi h D_o}{m_{ds} C_p} = \frac{\pi \times 402 \times 0.508 \times 3600}{109590 \times 2.393 \times 1000} = 0.0088\ 1/m \quad (25)$$

$$T(15m) = 30 + \left[[204.9 - 30] \times e^{-0.0088 \times 15}\right] \quad (26)$$

$$T(15m) = 183.3°C \quad (27)$$

From the above estimate it is seen that the desuperheated steam temperature at TE is 183.30C, which is short of the required desuperheated steam temperature of 1860C by 186 - 183.3 = 2.30C.

**Iteration 2 for TE location Temperature**

Therefore, performing another iteration by increasing the set temperature at TE to 28.30C, the desuperheated temperature at the point of mixing after the cooling water atomizer is,

$$T_{ds} = 179.9 + 28.3 = 208.2°C \quad (28)$$

Enthalpy of desuperheated steam at 10 bara and 208.20C,

$$h_{ds} = H_{10\ bara, 208.2°C} = 2847.9\ kJ/kg \quad (29)$$

The cooling water supply is now estimated as,

$$m_{cw} = 100,000 \times \frac{[3051.7 - 2847.9]}{[2847.9 - 632.9]} \quad (30)$$

$$m_{cw} = 9,202\ kg/h \quad (31)$$

The total desuperheated flow rate is,

$$m_{ds} = m_s + m_{cw} = 100,000 + 9,202\ kg/h \quad (32)$$

$$m_{ds} = 109,202\ kg/h \quad (33)$$

The desuperheated steam properties at 10 bara and 208.20C is,

| | | |
|---|---|---|
| Steam Pressure [$P_{ds}$] | 10 | bara |
| Steam Temperature [$T_{ds}$] | 208.2 | $^0C$ |
| Steam Density [$\rho_{ds}$] | 4.7497 | kg/m³ |
| Steam Dynamic Viscosity [$\mu_{ds}$] | 0.0162 | cP |
| Steam Specific Heat [$C_{p,ds}$] | 2.372 | kJ/kg.K |
| Steam Thermal conductivity [$k_{ds}$] | 0.0365 | W/m.K |
| Required Steam T at TE | 186 | $^0C$ |

*Table 3. Desuperheated steam properties [208.20C]*

Checking for the desuperheated steam piping velocity in the NPS 20 inch, Sch 80 pipe,

$$V_{ds} = \frac{Q_{ds}}{A_{ds}} = \frac{\left[\frac{109202}{4.7497}\right]}{\frac{\pi}{4} \times \left[\left(\frac{508}{1000}\right) - \left(2 \times \frac{12.7}{1000}\right)\right]^2} \approx 35\ m/s \quad (34)$$

The calculated piping velocity for the selected NPS 20", Sch 80 pipe is 35 m/s. This is within the 25 m/s to 40 m/s velocity range and acceptable.

Calculating for the desuperheated steam temperature at TE placed at a piping distance of 15 metres from the atomizer, the dimensionless numbers are,

$$Re = \frac{D_i \times V_{ds} \times \rho_{ds}}{\mu_{ds}} = \frac{482.6 \times 34.9 \times 4.7497}{1000 \times 0.0000162} \quad (35)$$

$$Re \approx 4,925,880 \quad (36)$$

$$Pr = \frac{C_{p,ds} \times \mu_{ds}}{k_{ds}} = \frac{2372 \times 0.0000162}{0.0365} \approx 1.0557 \quad (37)$$

Or,

$$Nu = 0.023 \times 4925880^{0.8} \times 1.0557^{0.3} \quad (38)$$

$$Nu \approx 5,282 \quad (39)$$

The internal heat transfer coefficient [h] is,

$$h = \frac{k_{ds} \times Nu}{D_i} = \frac{0.0365 \times 5282}{0.4826} \approx 399.5\ W/m^2K \quad (40)$$

For a discharge piping length of 15 metres as suggested by the vendor, the temperature at TE sensing point is,

$$\frac{\pi h D_o}{m_{ds} C_p} = \frac{\pi \times 399.5 \times 0.508 \times 3600}{109202 \times 2.372 \times 1000} = 0.00886\ 1/m \quad (41)$$

$$T(15m) = 30 + [[208.2 - 30] \times e^{-0.00886\times 15}] \quad (42)$$

$$T(15m) = 186.02°C \approx 186°C \quad (43)$$

Therefore, for the selected NPS 20", Sch 80 discharge piping, desuperheated steam at 10 bara, mixing temperature of 208.20C after the atomizer, and desuperheated steam flow rate of 109,202 kg/h, the temperature profile along the pipe length can be plotted as follows,

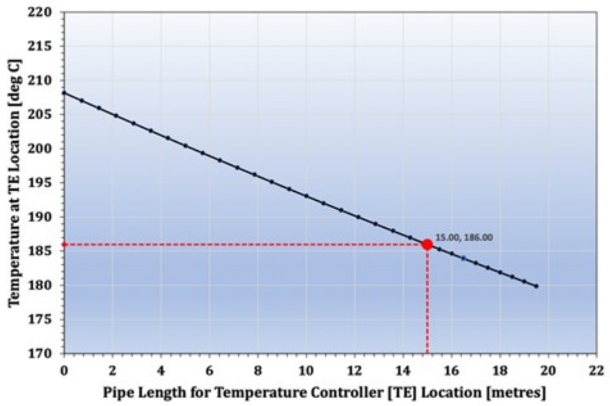



*Figure 2. TE location along discharge piping*

The % of spray / cooling water mass flow to the desuperheater steam mass flow is,

$$\%\ CW\ to\ DS = \frac{9202}{109202} \times 100 = 8.4\% \quad (44)$$

The %CW to DS mass flow rate is less than 15% for satisfactory performance.

## Cooling Water Control Valve Sizing

The cooling water control valve on the supply line can be sized as per ANSI/ISA S75.01 standard. Taking cooling water properties,

| | | |
|---|---|---|
| Spray water mass flow [$m_{cw}$] | 9,202 | kg/h |
| Spray water Density [r] | 917.6 | kg/m³ |
| Spray water Vol. Flow [q] | 44.15 | gpm |
| Vapour Pressure [$P_v$] | 4.76 | bara |
| Critical Pressure [$P_c$] | 217.7 | bara |
| $\rho_1/\rho_0$ [Sp Gravity at Flowing T] | 0.92 | - |

*Table 4. Cooling water properties – 15 bara, 1500C*

The control valve chosen is 1.5 inch with linear trim characteristics [6]. The upstream and downstream piping of the control valve is taken as NPS 2 inch, Sch 80 [OD 60.3 mm, WT 5.54 mm]. For the selected 1.5" control valve, the valve coefficient [Cv] is 39.2.

The steps to calculate the spray / cooling water control valve parameters are as follows,

**Step 1: Calculate Piping Geometry ($F_p$) & Liquid Pressure Recovery Factor ($F_{LP}$)**

$$F_P = \left[1 + \frac{\sum K}{N_2}\left(\frac{C_V}{d^2}\right)^2\right]^{-1/2} \quad (45)$$

Where,

$F_p$ = Piping geometric Factor [-]

$N_2$ = Constant [Value = 890]

$C_v$ = Valve Coefficient [GPM/Öpsi]

d = Control Valve Size [inch]

The value of $F_p$ is dependent on the fittings such as reducers, elbows or tees that are directly attached to the inlet & outlet connections of the control valve. If there are no fittings, $F_p$ is taken to be 1.0. The term SK is the algebraic sum of the velocity head loss coefficients of all the fittings that are attached to the control valve & is estimated as,

$$\sum K = K_1 + K_2 + K_{B1} - K_{B2} \quad (46)$$

Where,

$K_1$ = Upstream fitting resistance coefficient [-]

$K_2$=Downstream fitting resistance coefficient [-]

$K_{B1}$ = Inlet Bernoulli Coefficient [-]

$K_{B2}$ = Outlet Bernoulli Coefficient [-]

Where,

$$K_{B1} = 1 - \left(\frac{d}{D_1}\right)^4 \quad (47)$$

$$K_{B2} = 1 - \left(\frac{d}{D_2}\right)^4 \quad (48)$$

Where,

$D_1$ = Pipe Inlet Diameter [in]

$D_2$ = Pipe Outlet Diameter [in]

If the upstream and downstream piping are of equal size, then, $K_{B1} = K_{B2}$, and therefore, are dropped from the ΣK equation. The most commonly used fitting in control valve installations is the short-length concentric reducer. The equations for these fittings are,

$$K_1 = 0.5 \times \left[1 - \left(\frac{d^2}{D_1^2}\right)\right]^2 \text{, for inlet reducer.} \quad (49)$$

$$K_2 = 1.0 \times \left[1 - \left(\frac{d^2}{D_2^2}\right)\right]^2 \text{, for outlet reducer} \quad (50)$$

If the concentric reducers installed on either side of the control valve are identical, then

$$\sum K = K_1 + K_2 = 1.5 \times \left[1 - \left(\frac{d^2}{D^2}\right)\right]^2 \quad (51)$$

The liquid Pressure Recovery Factor ($F_{LP}$) is calculated as,

$$F_{LP} = \left[\frac{K_1 + K_{B1}}{N_2}\left(\frac{C_v}{d^2}\right)^2 + \frac{1}{F_L^2}\right]^{-1/2} \quad (52)$$

**Step 2: Calculate Pressure Drop Required for Sizing ($\Delta P_{Sizing}$)**

To estimate the ΔP required for sizing [$\Delta P_{sizing}$], first the liquid critical pressure ratio ($F_F$) is calculated. Therefore,

$$F_F = 0.96 - 0.28\sqrt{\frac{P_v}{P_c}} \quad (53)$$

Where,

$F_F$ = Liquid Critical Pressure Ratio [-]

$P_v$ = Vapour Pressure [psia]

$P_c$ = Critical Pressure [psia]

Using the value of $F_F$, $\Delta P_{choked}$ is calculated as,

$$\Delta P_{choked} = \left[\frac{F_{LP}}{F_p}\right]^2 [P_1 - F_F P_v] \quad (54)$$

If $\Delta P_{Valve} \leq \Delta P_{Choked}$, then $\Delta P = \Delta P_{Sizing}$

Else, Repeat calculations for next size

**Step 3: Calculate required control valve $C_v$**

The required control valve $C_v$ is calculated as,

$$C_v = \frac{Q}{N_1 F_p \sqrt{\frac{\Delta P_{Sizing}}{[\rho_1/\rho_0]}}} \quad (55)$$

Where, $N_1$ = Constant [Value = 1.0]

Upon calculating the required $C_v$, it is needed to check if the calculated $C_v$ is within the $C_v$ limit of the selected control valve. If not, the next size of control valve is chosen and the calculations are repeated. To arrive at accurate predictions for $C_v$ of the selected size, the calculations are repeated by re-inserting the calculated $C_v$ and control valve size (d) value into the $F_p$ equation, to calculate the new value of $F_p$, and further continued to estimate the final value of $C_v$.

Applying the above steps, the required $C_v$ of the control valve is calculated by estimating the valve coefficients first followed by checking if choked flow exists.

$$K_{B1} = 1 - \left(\frac{d}{D_1}\right)^4 = 1 - \left[\frac{1.5}{1.938}\right]^4 = 0.641 \quad (56)$$

$$K_{B2} = 1 - \left(\frac{d}{D_2}\right)^4 = 1 - \left[\frac{1.5}{1.938}\right]^4 = 0.641 \quad (57)$$

$$K_1 = 0.5 \times \left[1 - \left(\frac{1.5^2}{1.938^2}\right)\right]^2 = 0.08 \quad (58)$$

$$K_2 = 1.0 \times \left[1 - \left(\frac{1.5^2}{1.938^2}\right)\right]^2 = 0.161 \quad (59)$$

$$\sum K = 0.08 + 0.161 + 0.641 - 0.641 = 0.241 \quad (60)$$

The selected control valve is 1.5" valve with $C_v$ of 39.2 and FL of 0.82 [6].

$$F_p = \left[1 + \frac{0.241}{890}\left(\frac{39.2}{1.5^2}\right)^2\right]^{-1/2} = 0.9613 \quad (61)$$

$$F_{LP} = \left[\frac{0.08 + 0.641}{890}\left(\frac{39.2}{1.5^2}\right)^2 + \frac{1}{0.82^2}\right]^{-1/2} \approx 0.76 \quad (62)$$

Taking 0.5 bar ΔP across the control valve,

$$P_1 = (15 - 1.01325) \times 14.5 = 202.8\ psig \quad (63)$$

$$P_2 = 202.8 - (0.5 \times 14.5) = 195.5\ psig \quad (64)$$

$$\Delta P_{sizing} = 202.8 - 195.5 = 7.3\ psi \quad (65)$$

$$F_F = 0.96 - 0.28\sqrt{\frac{P_v}{P_c}} = 0.96 - 0.28\sqrt{\frac{69.03}{3157}} \quad (66)$$

$$F_F = 0.919 \quad (67)$$

$$\Delta P_{choked} = \left[\frac{0.76}{0.9613}\right]^2 [202.8 - (0.919 \times 69.03)] \quad (68)$$

$$\Delta P_{choked} = 87\ psi \quad (69)$$

Since $\Delta P_{sizing} \leq \Delta P_{Choked}$, then $\Delta P_{sizing} = 7.3$ psi

$$C_v = \frac{\frac{9202}{917.6} \times 4.4028675}{1 \times 0.9613 \times \sqrt{\frac{7.3}{0.92}}} = 16.4\ gpm/\sqrt{psi} \quad (70)$$

Reinserting the calculated $C_v$ value of 16.4, the value of $F_p$, $F_{LP}$ and new $C_v$ is re-computed iteratively,

| Iteration | $F_P$ | $F_{LP}$ | $C_v$ |
|---|---|---|---|
| 1 | 0.96 | 0.76 | 16.4 |
| 2 | 0.99 | 0.81 | 15.8 |
| 3 | 0.99 | 0.81 | 15.8 |
| 4 | 0.99 | 0.81 | 15.8 |
| 5 | 0.99 | 0.81 | 15.8 |
| 6 | 0.99 | 0.81 | 15.8 |
| 7 | 0.99 | 0.81 | 15.8 |
| 8 | 0.99 | 0.81 | 15.8 |
| 9 | 0.99 | 0.81 | 15.8 |
| **10** | **0.99** | **0.81** | **15.8** |

*Table 5. FP, FLP & Cv Iterations*

Therefore, for a cooling water flow of 9,202 kg/h, the $C_v$ required is 15.8 gpm / √psi, and the % opening of the control valve is,

$$\%\ C_v = \frac{15.8}{39.2} = 40.4\ \% \quad (71)$$

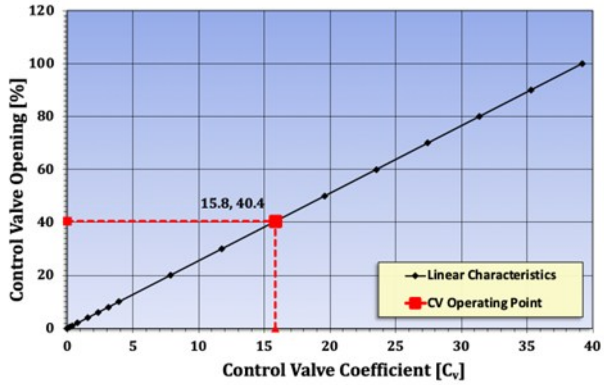



*Figure 3. % Spray water control valve opening*

**Spray / Cooling Water Piping Velocity**

Calculating the liquid velocity in the upstream and downstream cooling water NPS 2” piping,

$$V_{cw} = \frac{Q_{cw}}{A_{cw}} = \frac{\left[\frac{9202}{917.6}\right]}{\frac{\pi}{4} \times \left[\left(\frac{60.3}{1000}\right) - \left(2 \times \frac{5.54}{1000}\right)\right]^2} \approx 1.5\ m/s \quad (72)$$

The calculated spray / cooling water velocity of 1.5 m/s is within the velocity limit of 5 m/s for liquid lines.

**Appendix**

**Appendix A – Cooling Water Temperature**

To reduce the superheated steam temperature a spray of cooling water is injected into the stream. Performing a heat and mass balance,

$$[m_s h_s + m_{cw} h_{cw}] = [m_s + m_{cw}] \times h_{ds} \quad (73)$$

$$m_{cw} h_{cw} - m_{cw} h_{ds} = m_s h_{ds} - m_s h_s \quad (74)$$

$$m_{cw}[h_{cw} - h_{ds}] = m_s[h_{ds} - h_s] \quad (75)$$

$$m_{cw} = m_s \times \frac{[h_{ds} - h_s]}{[h_{cw} - h_{ds}]} = m_s \times \frac{[h_s - h_{ds}]}{[h_{ds} - h_{cw}]} \quad (76)$$

Where,

$m_s$ = Mass flow of superheated steam [kg/h]

$m_{cw}$ = Mass flow of cooling water [kg/h]

$h_s$ = Enthalpy of superheated steam [kg/h]

$h_{cw}$ = Enthalpy of cooling water [kg/h]

$h_{ds}$ = Enthalpy of desuperheated steam [kg/h]

## Appendix B – Temperature drop in piping for fully developed flow

To estimate the temperature drop along the desuperheater discharge piping, the assumption made is that the flow is fully developed. Therefore, the wall temperature is fairly uniform along the length of the pipe wall. This also means the difference between the fluid mean temperature and pipe wall temperature follows an exponential distribution along the pipe length. No insulation is provided since the heat losses through the discharge piping also aids in cooling the steam to the required desuperheated temperature, and reduces the cooling water load. The heat balance is therefore,

$$m_{ds}C_p\frac{dT}{dx} = -hA[T_f - T_w] \qquad (77)$$

$$\frac{dT}{dx} = -\frac{\pi hDL}{m_{ds}C_p}[T_f - T_w] \qquad (78)$$

For L = 1 m and taking $T_f = T(x)$,

$$\int \frac{dT}{[T(x) - T_w]} = -\frac{\pi hD}{m_{ds}C_p}\int dx \qquad (79)$$

$$ln\left[\frac{T(x) - T_w}{T_f - T_w}\right] = -\frac{\pi hD}{m_{ds}C_p}\int_0^{L(x)} dx \qquad (80)$$

$$\left[\frac{T(x) - T_w}{T_f - T_w}\right] = e^{-\frac{\pi hD}{m_{ds}C_p}L(x)} \qquad (81)$$

$$T(x) = T_w + \left[[T_f - T_w] \times e^{-\frac{\pi hD}{m_{ds}C_p}L(x)}\right] \qquad (82)$$

Where,

$T(x)$ = Temperature along Pipe length [m]

L = Pipe length [m]

h = Internal heat transfer Coefficient [W/m2.K]

D = Outer diameter (OD) of pipe [m]

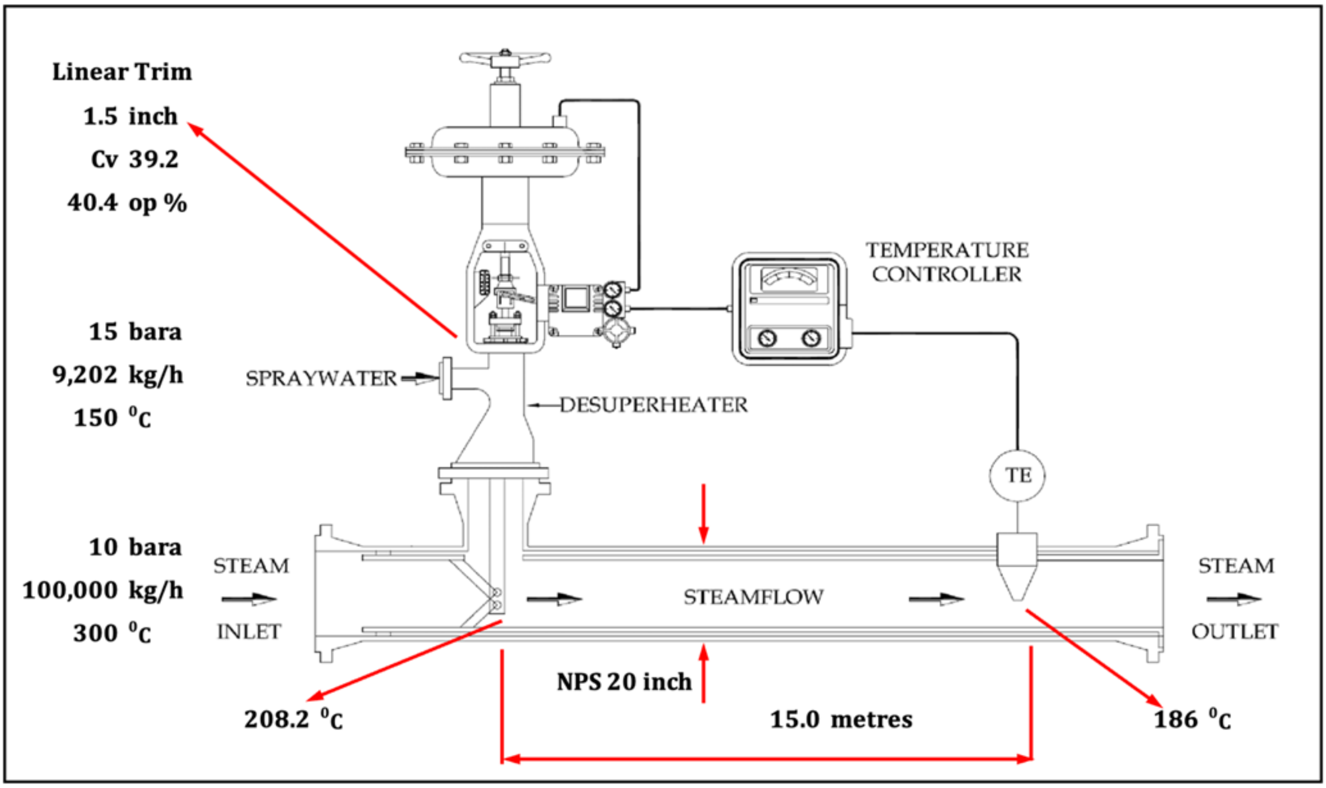
$m_{ds}$ = Desuperheated steam mass flow [kg/s]

$T_f$ = Flowing Temperature after mixing [K]

$T_w$ = Wall Temperature [K]

$C_p$ = Specific heat capacity [J/kg.K]

## Appendix C – Results Summary



# Process Design of Industrial De-Super Heaters

| Super heated Steam Conditions | | |
|---|---|---|
| Superheated Steam Pressure [$P_s$] | 10.0 | bar(a) |
| | 1,000 | kPa(a) |
| Superheated Steam Temperature [$T_s$] | 300 | °C |
| Enthalpy of Superheated Steam [$h_{s,g}$] | 3,051.7 | kJ/kg |
| Superheated Steam Mass Flow rate [$m_s$] | 100,000 | kg/h |
| **Spray Water Supply** | | |
| Spray Water Pressure [$P_{cw}$] | 15.0 | bar(a) |
| | 1,500 | kPa(a) |
| Spray Water Temperature [$T_{cw}$] | 150.0 | °C |
| Enthalpy of Spray Water [$h_{cw}$] | 632.9 | kJ/kg |
| **De-superheated Steam** | | |
| Desuperheated Steam Pressure [$P_{ds}$] | 10.0 | bar(a) |
| | 1,000 | kPa(a) |
| Saturation Temperature at 10 bara [T_ds] | 179.9 | °C |
| Superheat Temperature above Saturation T | 28.3 | °C |
| Desuperheated Steam T [$T_{ds}$] at point of mixing | 208.2 | °C |
| Enthalpy of Desuperheated Steam [$h_{ds,g}$] | 2,847.9 | kJ/kg |
| **Spray Water Requirements & Desuperheated Steam Flow** | | |
| Desuperheated Steam Flow rate [$m_{ds}$] | **109,202** | kg/h |

| Spray Water Control Valve Sizing [ANSI/ISA S75.01] | | |
|---|---|---|
| Spray Water Supply [$m_{cw}$] | **9,202** | kg/h |
| Spray water Density [ρ] | 917.6 | kg/m$^3$ |
| Spray water Vol. Flow [q] | 44.15 | gpm |
| Vapour Pressure [$P_v$] | 69.03 | psi |
| Critical Pressure [$P_c$] | 3,157 | psi |
| $\rho_1 / \rho_0$ [Sp Gravity at Flowing T] | 0.920 | - |
| **Control Valve Details** | | |
| **Control Valve Characteristics** | **Linear** | |
| **Size, *d* [inch]** | **$C_v$ @100%** | **$F_L$** |
| 1 | 20.6 | 0.84 |
| 1.5 | 39.2 | 0.82 |
| 2 | 72.9 | 0.77 |
| 3 | 148 | 0.82 |
| 4 | 236 | 0.82 |
| 6 | 433 | 0.84 |
| 8 | 846 | 0.87 |
| Selected Control Valve Size | 1.5 | in |
| $C_v$ of Selected Valve | 39.2 | - |
| $F_L$ of Selected control valve | 0.82 | - |

| | | |
|---|---|---|
| Control Valve Inlet [$P_1$] | 202.8 | psig |
| Control Valve Outlet [$P_2$] | 195.6 | psig |
| Control Valve Pressure Drop [ΔP] | 7.3 | psi |
| U/S Pipe Size [OD] | 60.3 | mm |
| U/S Pipe Wall Thickness [WT] | 5.54 | mm |
| U/S Pipe Inner Diameter [ID] | 1.938 | in |
| D/S Pipe Size [OD] | 60.3 | mm |
| D/S Pipe Wall Thickness [WT] | 5.54 | mm |
| D/S Pipe Inner Diameter [ID] | 1.938 | in |
| $K_1$ [Resistance Coefficient of U/s fitting] | 0.080 | - |
| $K_2$ [Resistance Coefficient of D/s fitting] | 0.161 | - |
| $K_{B1}$ [Inlet Bernoulli Coefficient] | 0.641 | - |
| $K_{B2}$ [Outlet Bernoulli Coefficient] | 0.641 | - |
| ΣK [Sum of Resistances] | 0.241 | - |
| U/S Liquid Velocity [$V_{Cv,in}$] | 1.5 | m/s |
| D/S Liquid Velocity [$V_{Cv,out}$] | 1.5 | m/s |
| Required Control Valve $C_v$ | **15.8** | gpm/√psi |
| % Opening of Selected Valve | **40.4** | % |

| Choke Conditions Check | | |
|---|---|---|
| Critical pressure ratio [$F_F$] | 0.919 | - |
| Choked ΔP [$\Delta P_{Choke}$] | 87.0 | psi |
| Sizing ΔP [$\Delta P_{Sizing}$] | 7.3 | psi |
| Flow Condition Check | **Subcritical** | - |

**$F_P$ & $F_{LP}$ Calculations**

| Iteration | $F_P$ | $F_{LP}$ | $C_v$ |
|---|---|---|---|
| 1 | 0.96 | 0.76 | 16.4 |
| 2 | 0.99 | 0.81 | 15.8 |
| 3 | 0.99 | 0.81 | 15.8 |
| 4 | 0.99 | 0.81 | 15.8 |
| 5 | 0.99 | 0.81 | 15.8 |
| 6 | 0.99 | 0.81 | 15.8 |
| 7 | 0.99 | 0.81 | 15.8 |
| 8 | 0.99 | 0.81 | 15.8 |
| 9 | 0.99 | 0.81 | 15.8 |
| 10 | 0.99 | 0.81 | 15.8 |
| **11** | **0.99** | **0.81** | **15.8** |

| De-super heater Piping Details | | |
|---|---|---|
| Selected Material Grade | ASTM A106 Gr B | - |
| Selected NPS of Desuperheated Steam | 20 | inch |
| Pipe Outer Diameter [OD] | 508.0 | mm |
| Pipe Wall Thickness [WT] | 12.7 | mm |
| Pipe Inner Diameter [ID] | 482.6 | mm |
| Thermal conductivity [$k_{pipe}$] | 45.5 | W/m.K |
| Pipe Density [$\rho_{pipe}$] | 7850 | kg/m$^3$ |
| Desuperheated Steam Velocity [$V_{ds}$] | 34.9 | m/s |
| Pipe Wall Temperature [$T_w$] | 30.0 | °C |
| **De-super heater Discharge Pipe Length** | | |
| Condition 1: [15D to 20D length] [Taking 20D] | 10.16 | m |
| Condition 2: Minimum 25 ft or 7.62 m | 7.62 | m |
| Vendor Recommendation | 15.00 | m |
| Selected Pipe Length [L] for Temperature Control [TE] | **15.00** | m |
| **De-super heater Steam Properties** | | |
| Desuperheated Steam Pressure [$P_{ds}$] | 1,000 | kPa(a) |
| Desuperheated Steam Temperature [$T_{ds}$] | 208.2 | °C |
| Desuperheated Steam Density [$\rho_{ds}$] | 4.7497 | kg/m$^3$ |
| Desuperheated Steam Dynamic Viscosity [$\mu_{ds}$] | 0.0162 | cP |
| | 0.0000162 | kg/m.s |
| Desuperheated Steam Specific Heat [$C_{p,ds}$] | 2.372 | kJ/kg.K |
| Desuperheated Steam Thermal conductivity [$k_{ds}$] | 0.0365 | W/m.K |
| Saturation Temperature at 1000 kPa(a) [T_ds] | 179.9 | °C |
| Required Desuperheated Steam Temperature at TE | 186.0 | °C |
| Superheat T above Saturation T at mixing point | 28.3 | °C |
| % of Cooling Water mass flow to Desuperheated mass flow | 8.4 | % |
| **Heat Transfer for Fully Developed Flow** | | |
| Reyonolds Number [Re] | 4,925,879 | - |
| Prandtl Number [Pr] ['=Pr_pT(C13,C14)] | 1.0557 | - |
| Nusselt Number [Nu] | 5,282 | - |
| Internal heat transfer coefficient [h] | 399.5 | W/m$^2$.K |
| $\pi h D_o / m_{ds} C_{p,ds}$ | 0.00886 | 1/m |

**De-super heater Discharge Piping**

| | | | |
|---|---|---|---|
| Steam Velocity Lower Limit | 25 | m/s | Desuperheated Steam Mass Flow Rate: 109,202 kg/h |
| Steam Velocity Upper Limit | 40 | m/s | Desuperheated Steam Volumetric Flow Rate: 6.3865 m$^3$/s |

| NPS | OD | Sch XS WT | ID | Fluid Velocity | Allowable V | Selected Pipe Size | | Selected ID | WT |
|---|---|---|---|---|---|---|---|---|---|
| inch | mm | mm | mm | m/s | m/s | inch | mm | mm | mm |
| **1** | 33.4 | 4.55 | 24.3 | 13770.9 | | | | | |
| **1.25** | 42.2 | 4.85 | 32.5 | 7698.5 | | | | | |
| **1.5** | 48.3 | 5.08 | 38.14 | 5590.0 | | | | | |
| **2** | 60.3 | 5.54 | 49.22 | 3356.5 | | | | | |
| **2.5** | 73 | 7.01 | 58.98 | 2337.6 | | | | | |
| **4** | 114.3 | 8.56 | 97.18 | 861.0 | | | | | |
| **6** | 168.3 | 10.97 | 146.36 | 379.6 | | | | | |
| **8** | 219.1 | 12.7 | 193.7 | 216.7 | | | | | |
| **10** | 273.1 | 12.7 | 247.7 | 132.5 | | | | | |
| **12** | 323.9 | 12.7 | 298.5 | 91.3 | | | | | |
| **14** | 355.6 | 12.7 | 330.2 | 74.6 | | | | | |
| **16** | 406.4 | 12.7 | 381 | 56.0 | | | | | |
| **18** | 457.2 | 12.7 | 431.8 | 43.6 | | | | | |
| **20** | 508.0 | 12.7 | 482.6 | 34.9 | 34.9 | **20.0** | **508.0** | **482.6** | **12.7** |
| **22** | 558.8 | 12.7 | 533.4 | 28.6 | 28.6 | | | | |
| **24** | 609.6 | 12.7 | 584.2 | 23.8 | | | | | |
| **26** | 660.4 | 12.7 | 635 | 20.2 | | | | | |
| **28** | 711.2 | 12.7 | 685.8 | 17.3 | | | | | |
| **30** | 762 | 12.7 | 736.6 | 15.0 | | | | | |
| **32** | 812.8 | 12.7 | 787.4 | 13.1 | | | | | |
| **34** | 863.6 | 12.7 | 838.2 | 11.6 | | | | | |
| **36** | 914.4 | 12.7 | 889 | 10.3 | | | | | |
| **38** | 965.2 | 12.7 | 939.8 | 9.2 | | | | | |
| **40** | 1016 | 12.7 | 990.6 | 8.3 | | | | | |
| **42** | 1066.8 | 12.7 | 1041.4 | 7.5 | | | | | |
| **44** | 1117.6 | 12.7 | 1092.2 | 6.8 | | | | | |
| **46** | 1168.4 | 12.7 | 1143 | 6.2 | | | | | |
| **48** | 1219.2 | 12.7 | 1193.8 | 5.7 | | | | | |
| | | | | | **V [m/s]** | **NPS [in]** | **OD [mm]** | **ID [mm]** | **WT [mm]** |
| **Selected Pipe Parameters** | | | | | **34.9** | **20.0** | **508.0** | **482.6** | **12.7** |

| L(x) | T(x) |
|---|---|
| m | °C |
| 0 | 208.2 |
| 0.71 | 207.1 |
| 1.43 | 205.9 |
| 2.14 | 204.8 |
| 2.86 | 203.7 |
| 3.57 | 202.6 |
| 4.29 | 201.5 |
| 5.00 | 200.5 |
| 5.71 | 199.4 |
| 6.43 | 198.3 |
| 7.14 | 197.2 |
| 7.86 | 196.2 |
| 8.57 | 195.1 |
| 9.29 | 194.1 |
| 10.00 | 193.1 |
| 10.71 | 192.0 |
| 11.43 | 191.0 |
| 12.14 | 190.0 |
| 12.86 | 189.0 |
| 13.57 | 188.0 |
| 14.29 | 187.0 |
| **15.00** | 186.0 |
| 15.50 | 185.3 |
| 16.00 | 184.6 |
| 16.50 | 183.9 |
| 17.00 | 183.3 |
| 17.50 | 182.6 |
| 18.00 | 181.9 |
| 18.50 | 181.2 |
| 19.00 | 180.6 |
| 19.50 | 179.9 |

### References & Further Reading


1. https://www.pneuconvalves.com/wp-content/uploads/2019/04/DESUPERHEATER-SERIES-800-n.pdf
2. https://www.spiraxsarco.com/learn-about-steam/desuperheating/basic-desuperheating-theory?sc_lang=en-GB
3. https://www.spiraxsarco.com/learn-about-steam/steam-distribution/pipes-and-pipe-sizing?sc_lang=en-GB#:~:text=As%20a%20general%20rule%2C%20a,m/s%20for%20saturated%20steam.
4. https://www.womicsteel.com/news/a106-gr-b-nace-pipe-technical-data-sheet/
5. https://www.spiraxsarco.com/learn-about-steam/desuperheating/other-types-of-desuperheater?sc_lang=en-GB#:~:text=The%20steam%20flow%20turndown%20ratio,more%20than%20several%20metres%20long.
6. Emerson's Control Valve Handbook, 6th Ed


### Author

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# SAF Production: A Future Sustainable Air Fuel (SAF)

## (An Innovative Feedstock Plastic Waste and Spent Coffee Grounds)

**Hamid Reza Seyed Jafari**


## 1. Abstract

Refineries In an era of growing environmental concerns and the need for sustainable alternatives to fossil fuels, the search for renewable and eco-friendly energy sources is crucial. SAF, a sustainable alternative to conventional aviation fuel, is gaining attention due to its potential to reduce greenhouse gas emissions and reliance on fossil fuels. Sustainable aviation fuel (SAF) can be produced through multiple technological pathways, including Hydroprocessed Esters and Fatty Acids (HEFA) and Alcohol-to-Jet (ATJ), which are currently the most mature and commercially deployed options. Other pathways such as Fischer–Tropsch (FT) and Power-to-Liquid (PtL) enable the production of synthetic jet fuel from biomass, waste, or renewable electricity and captured $CO_2$. While these technologies offer significant potential to reduce aviation greenhouse gas emissions, their widespread adoption is constrained by cost, feedstock availability, and scalability challenges.

An innovative feedstock to SAF production could be converting plastic waste and spent coffee grounds into SAF. This mixture feedstock not only addresses the global challenges of waste disposal but also provides a viable energy solution for the aviation industry. The co-pyrolysis of plastic waste and spent coffee grounds (SCG) and then hydrogenated upgrading technology could be a promising method to produce SAF and other valuable biofuels, leveraging low-cost waste materials for sustainable energy solutions that needs more investigators to survey it in detail for commercialization.


## 2. Overview of Sustainable Air Fuel (SAF)

Sustainable aviation fuel (SAF), is derived from biomass or waste materials and can replace conventional jet fuel. SAF is chemically similar to petroleum-based jet fuel but is produced using renewable sources. The aviation industry has been actively exploring SAF as part of its efforts to reduce its carbon footprint.

Sustainable aviation fuel (SAF) can be produced through several technological pathways that utilize different feedstocks and conversion processes. Hydroprocessed Esters and Fatty Acids (HEFA) is the most technologically mature pathway, converting waste oils, fats, and greases into drop-in jet fuel through hydrogenation and refining processes (International Civil Aviation Organization [ICAO], 2018). The Alcohol-to-Jet (ATJ) pathway converts alcohols such as ethanol or isobutanol—derived from biomass or waste streams—into aviation fuel via dehydration, oligomerization, and upgrading steps. Both HEFA and ATJ pathways are certified under international fuel standards and are currently deployed at commercial scale (International Air Transport Association [IATA], 2023).

Other important SAF production routes include Fischer–Tropsch (FT) synthesis and Power-to-Liquid (PtL) technologies. The FT process converts syngas produced from biomass or municipal solid waste into synthetic hydrocarbons, including jet fuel, using catalytic reactions

(ASTM International, 2022). PtL fuels, also known as e-fuels, are produced by combining green hydrogen generated through renewable electricity with captured carbon dioxide to synthesize liquid hydrocarbons. Although FT and PtL pathways offer significant long-term potential for deep decarbonization of aviation, their large-scale deployment is currently limited by high production costs and infrastructure requirements (ICAO, 2018) (Fuel, 2023) (Energy Conversion and Management, 2024), (Fan et al., 2023).

## 3. Review of Main Sustainable Aviation Fuel (SAF) Production Technologies

The aviation sector faces significant challenges in reducing greenhouse gasemissions due to limited alternatives to liquid hydrocarbon fuels. Sustainable aviation fuel (SAF) has emerged as the most viable short- and long-term solution for decarbonizing air transport without major changes to aircraft or infrastructure (ICAO, 2018). Several technological pathways have been developed for SAF production, among which Hydroprocessed Esters and Fatty Acids (HEFA), Alcohol-to-Jet (ATJ), Fischer–Tropsch Synthesis (FTS), and Power-to-Liquid (PtL) are the most prominent. These technologies differ substantially in terms of technological maturity, economic feasibility, feedstock availability, and commercialization potential.

- **Hydroprocessed Esters and Fatty Acids (HEFA)**

HEFA is currently the most mature and commercially deployed SAF production pathway. This technology converts lipid-based feedstocks such as waste cooking oils, animal fats, and vegetable oils into jet fuel through hydrotreatment and hydrocracking processes (Staples et al., 2018). As illustrated in the figure, HEFA ranks highest in terms of technological maturity, with multiple commercial-scale facilities operating worldwide.

From an economic perspective, HEFA is considered the most favorable SAF pathway at present. The process benefits from established refinery infrastructure, relatively simple processing steps, and lower capital investment compared to other SAF technologies (IATA, 2023). Consequently, HEFA is currently the only SAF pathway that can be considered commercially profitable under existing market conditions. The use of waste-based feedstocks also contributes to favorable life-cycle greenhouse gas emission reductions.

However, HEFA faces critical limitations related to feedstock availability. The global supply of waste oils and fats is inherently limited and insufficient to meet the long-term fuel demand of the aviation sector (IEA, 2020). Additionally, reliance on virgin vegetable oils raises concerns regarding land-use change, food security, and sustainability. Therefore, while HEFA is expected to dominate the SAF market until approximately 2030, it is unlikely to serve as a standalone long-term solution.

- **Alcohol-to-Jet (ATJ)**

The Alcohol-to-Jet pathway converts alcohols—primarily ethanol or isobutanol—into aviation fuel through dehydration, oligomerization, and hydrogenation steps (Pearlson et al., 2019). According to the figure, ATJ is considered a technically mature process, although less developed than HEFA in terms of large-scale deployment.

Economically, ATJ is currently classified as medium-performing, as production costs remain relatively high and profitability has not yet been fully achieved (IATA, 2023). One of the major challenges associated with ATJ is the competition for feedstocks, particularly when alcohols are derived from food crops such as corn or wheat. This raises sustainability and social concerns similar to those observed in first-generation biofuels.

Despite these challenges, ATJ offers notable advantages. The technology allows for a broader range of feedstocks, including lignocellulosic biomass and agricultural residues, and can leverage existing bioethanol infrastructure in many regions (Staples et al., 2018). As indicated in the figure, ATJ is expected to become more economically viable after 2030, particularly as HEFA feedstock constraints intensify and advanced alcohol production technologies mature.

- **Fischer–Tropsch Synthesis (FTS)**

Fischer–Tropsch synthesis is a well-established chemical process that converts synthesis gas (CO and $H_2$) into liquid hydrocarbons using catalytic reactions. In the context of SAF, syngas can be produced from biomass, municipal solid waste, sewage sludge, or algae through gasification (Dry, 2002). As shown in the figure, FTS is a proven technology, but its application for SAF production remains largely at the research and demonstration stage.

From an economic standpoint, FTS is currently considered poorly competitive, primarily due to high capital costs, complex process integration, and large-scale facility requirements (IEA, 2020). Consequently, FTS-based SAF has not yet been widely commercialized. However, the technology offers significant sustainability advantages, including extensive feedstock flexibility and minimal competition with food resources.

The figure highlights that FTS has strong long-term potential, particularly beyond 2030, when technological advancements and economies of scale may reduce costs. Additionally, integrating FTS with carbon capture and storage (CCS) could further enhance its climate benefits, making it a promising pathway for deep decarbonization of aviation.

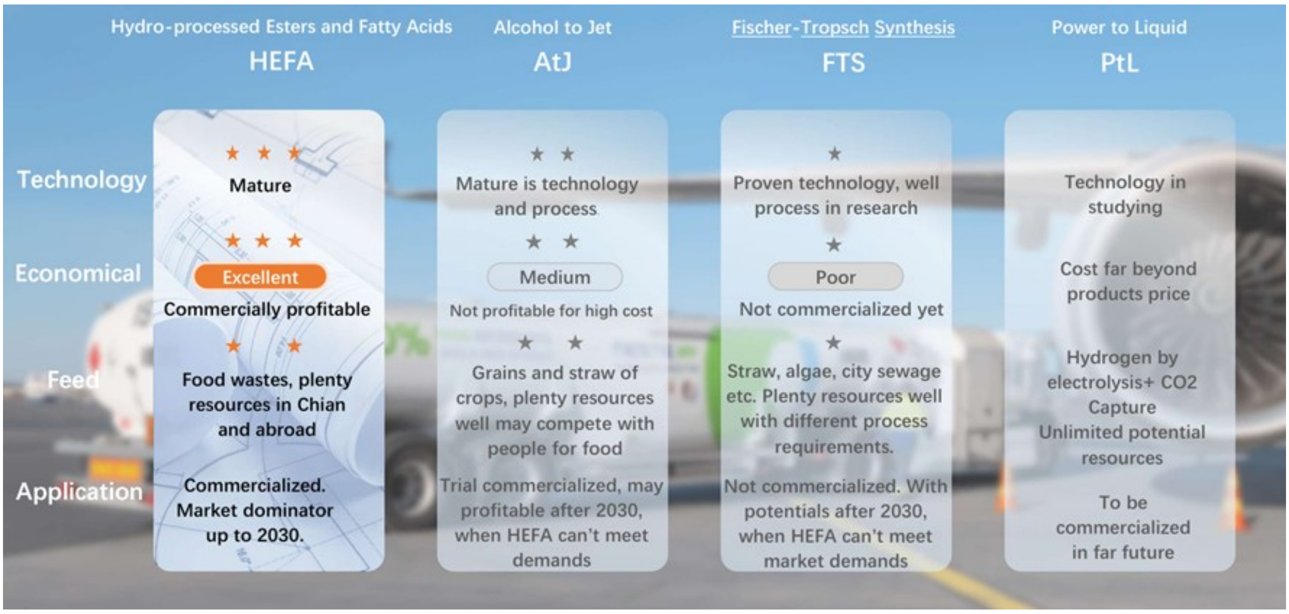


*Figure 1- SAF Production Technologies, (Yanxin, 2025)*

- **Power-to-Liquid (PtL)**

Power-to-Liquid, also referred to as electrofuels or e-fuels, represents the most advanced and future-oriented SAF pathway. In this technology, renewable electricity is used to produce green hydrogen via water electrolysis, which is then combined with captured $CO_2$ to synthesize liquid hydrocarbons suitable for aviation use (Schmidt et al., 2018). According to the figure, PtL is currently in the early research and development stage.

Economically, PtL faces substantial challenges. Production costs are significantly higher than those of conventional jet fuel and other SAF pathways, largely due to the high cost of renewable electricity and electrolysis systems (IEA, 2020). As indicated in the figure, PtL fuel costs remain far beyond current market prices, preventing near-term commercialization.

Nevertheless, PtL offers unmatched long-term advantages. It relies on virtually unlimited resources—renewable electricity, water, and captured $CO_2$—and does not depend on biomass or arable land. For this reason, PtL is widely regarded as a critical long-term solution for achieving net-zero emissions in aviation. Commercial deployment is expected only in the distant future, contingent upon major cost reductions and strong policy support.

- **Overall Comparison and Outlook SAF technologies**

In summary, the figure-1 and literature clearly indicate that HEFA will remain the dominant SAF technology in the short term, ATJ and FTS will likely play complementary roles in the medium term, and PtL represents the most promising long-term pathway for sustainable aviation. A diversified portfolio of SAF technologies, tailored to regional feedstock availability and supported by policy incentives, will be essential to achieving aviation decarbonization goals (Yanxin, 2025).

**4. Waste Materials for SAF Production**

- **Plastic Polymer Waste**

Plastic polymer waste has become one of the largest environmental issues in recent decades. Globally, over 300 million tons of plastic are produced annually, with much of it ending up in landfills or oceans. Plastic waste, especially from single-use plastics, represents a significant opportunity for recycling and conversion into valuable products, including biojet fuel.

- **Spent Coffee Grounds (SCG)**

Spent coffee grounds are another underutilized resource. Over 9 million tons of coffee grounds are produced annually, most of which are discarded. Coffee grounds contain high amounts of lipids, which are valuable for biofuel production. When combined with pyrolysis, spent coffee grounds can be converted into biojet fuel. (Fuel, 2023)

**5. Pyrolysis Technology: The Backbone of SAF Production**

- **Principles of Pyrolysis**

Pyrolysis is a thermal decomposition process that occurs in the absence of oxygen. It involves heating organic materials to high temperatures (typically between 350°C and 700°C), breaking down complex molecules into simpler compounds. Pyrolysis of plastic waste and spent coffee grounds can produce a range of products, including bio-oil, syngas, and char. Bio-oil, which is a liquid product, can be further refined to produce biojet fuel. (Fuel, 2023)

- **Pyrolysis to SAF Production**

In the context of SAF production, the bio-oil produced through pyrolysis can be upgraded and refined to meet the specifications required for aviation fuel. This bio-oil can be blended with traditional jet fuel or used as a direct replacement in some cases. Pyrolysis offers a sustainable solution to waste management while producing high-energy products like biojet fuel. (Fuel, 2023) (Fan et al., 2023) .It is shown in figure-2.

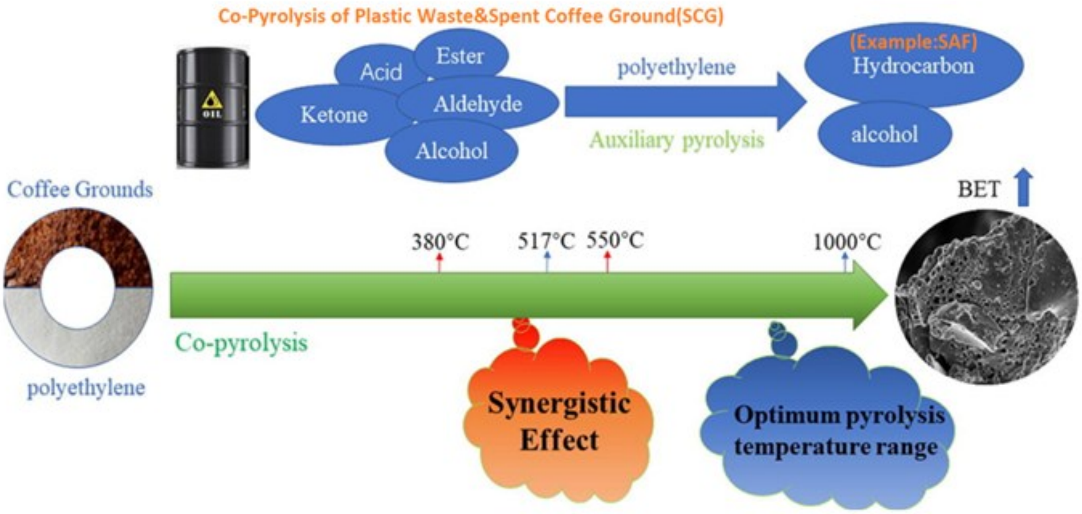


*Figure 2- pyrolysis of polymer waste and (SCG), (Fan et al., 2023)*

- **Preparation of Plastic Polymer Waste and Coffee Grounds**

Before undergoing pyrolysis, plastic polymer waste and coffee grounds need to be prepared. Plastic waste is typically sorted, cleaned, and shredded to ensure uniformity in size and composition. Coffee grounds are collected, dried, and sometimes pre-processed to reduce moisture content. These preparations optimize the efficiency of the pyrolysis process.

- **Pyrolysis Process**

Once prepared, the materials are heated in a pyrolysis reactor. The temperature and pressure inside the reactor are controlled to break down the organic materials into smaller molecules. The resulting of pyrolysis products are:

  - Bio-oil: A liquid that can be upgraded to biojet fuel.
  - Syngas: A gaseous by-product that can be used for energy or further processing.
  - Char: A solid by-product that can be used for various applications, including as a fuel source or soil amendment.

**Conversion to SAF**

Syngas or bio-oil undergo a refining process to to produce SAF. These processes may involve hydrodeoxygenation (HDO), catalytic cracking, and distillation to achieve the required fuel characteristics such as shown in figure-3.

Producing SAF also from plastic waste and spent coffee grounds via pyrolysis involves a few key steps. Since both materials have different compositions—plastic being a hydrocarbon-based polymer and coffee grounds being an organic biomass—they need to be co-pyrolyzed under optimized conditions. (Fuel, 2023)

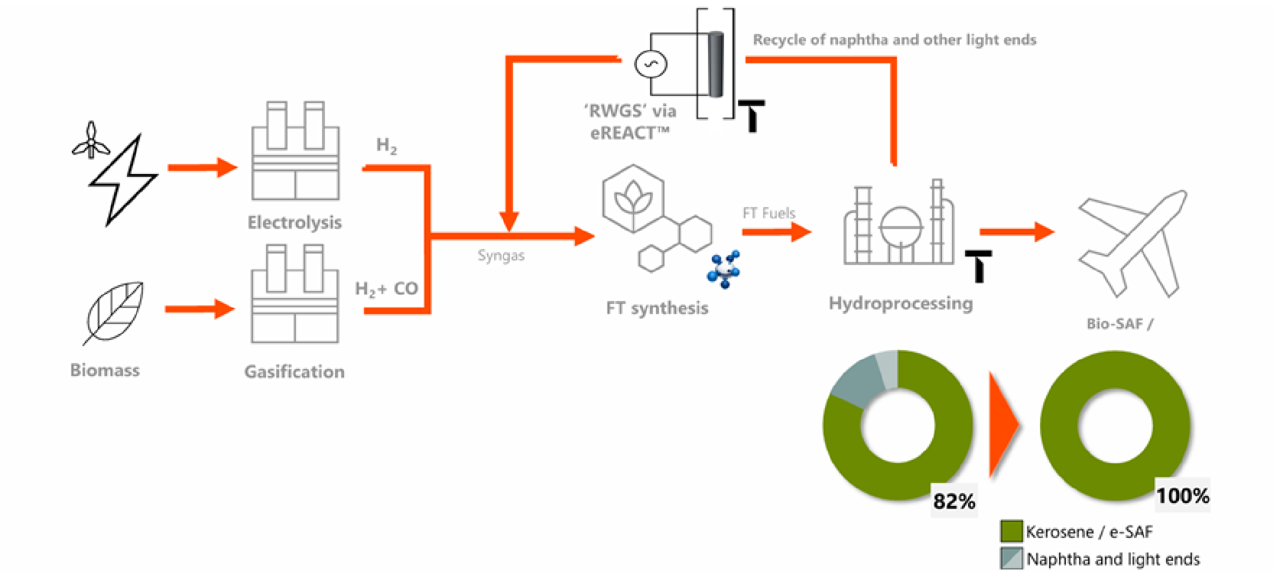


*Figure 3- Fischer Tropsch SAF Production Technology from syngas, (TOPSOE, SASOL, 2025).*

## A. Feedstock Preparation

Plastic Waste: Common polymer plastics especially polyethylene (PE), then polypropylene (PP), and polystyrene (PS) are ideal for pyrolysis since they break down

into liquid hydrocarbons. Plastics are shredded and pre-cleaned to remove contaminants. Utilization of spent polymer plastic as feedstock through pyrolysis can significantly reduce the volume of polymer waste typically sent to landfills.

Spent Coffee Grounds (SCG): Coffee grounds contain cellulose, hemicellulose, lignin, and oils, making them suitable for bio-oil production. They are dried to reduce moisture content, which improves pyrolysis efficiency. Utilization of SCG as feedstock through pyrolysis can significantly reduce the volume of organic waste typically sent to landfills. (Fuel, 2023)

Blending: The plastic waste and coffee grounds are mixed in optimal ratios (e.g., 50:50 or 60:40) to enhance liquid fuel yields.

Note :The optimum blending ratios of SCG and plastics significantly influence the yield. For instance, a study found a 25:75 ratio of SCG to PET produced the highest biochar yield compared to other ratios.

## B. Co-Pyrolysis of SCG and Plastic Waste Process Parameters (Fan et al., 2023)

- **Reaction Conditions:**
  - ◊ Temperature: 400–600°C
  - ◊ Heating Rate: Moderate (~10–50°C/min)
  - ◊ Atmosphere: Inert gas (e.g., nitrogen or argon)
- **Thermal Breakdown:**
  - ◊ Plastics decompose into hydrocarbon gases and liquid oil.
  - ◊ Coffee Grounds break down into bio-oil, syngas, and biochar.
  - ◊ The oxygenated compounds from biomass (SCG) help reduce impurities in the plastic-derived pyrolysis oil.

Note: Temperature and time are critical in determining the yield and quality of the bio-oil and biochar produced. Studies indicate that temperatures around 350 °C optimize biochar production while higher temperatures (e.g., 500 °C) are favorable for maximizing bio-oil yield.

## C. Product Separation & Refining

- **Liquid Oil Fraction:**
  - ◊ The pyrolysis oil from plastic is rich in hydrocarbons but contains waxes and heavy fractions, as shown in figure-4.
  - ◊ The bio-oil from coffee grounds has oxygenated compounds that need upgrading.
  - ◊ A hydrocracking or catalytic upgrading step improves the jet fuel fraction (C8–C16 hydrocarbons).
- **Gaseous Products:** Some gases (e.g., methane, ethylene) can be used as an energy source to power the pyrolysis unit.
- **Solid Biochar:** The biochar from SCG can be used in soil enhancement or as an activated carbon material. (Chemical Engineering World, 2024.)

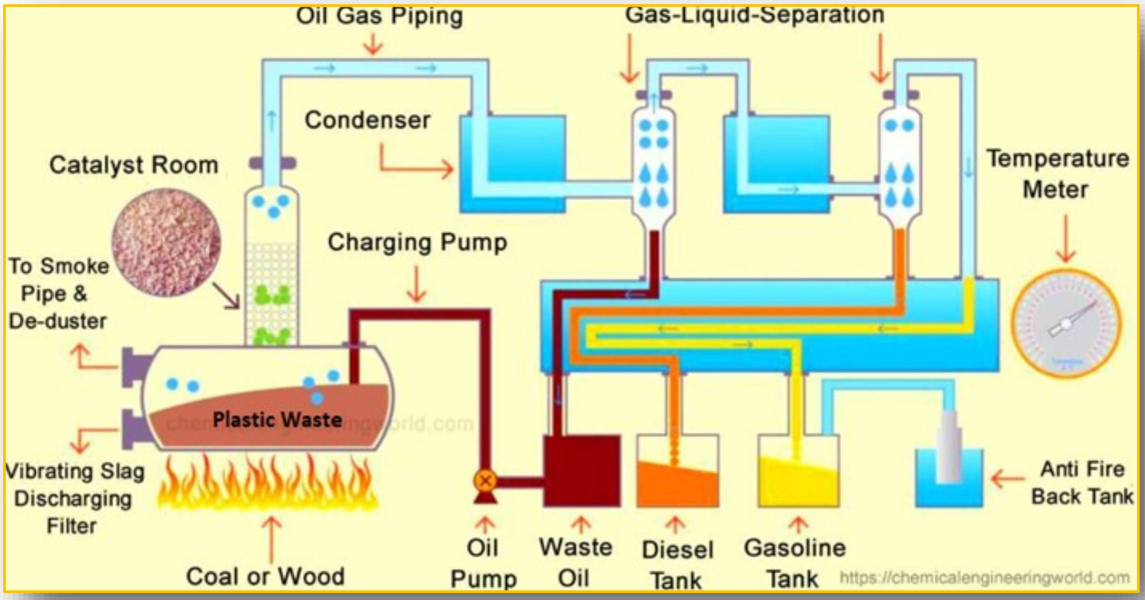


*Figure 4- pyrolysis of polymer waste, (Chemical Engineering World, 2024.)*

## D. Upgrading to SAF

- **Catalytic Cracking:**
  - ◊ Uses zeolites, alumina, or metal-based catalysts to refine the pyrolysis oil.
  - ◊ Converts heavier hydrocarbons into lighter jet-fuel-range hydrocarbons.
- Distillation & Hydroprocessing:
  - ◊ Further separates and hydro-processing upgrading the fuel to meet aviation fuel specifications ,as shown in figure-5.
  - ◊ Removes impurities like sulfur, nitrogen, and oxygen to improve fuel stability.

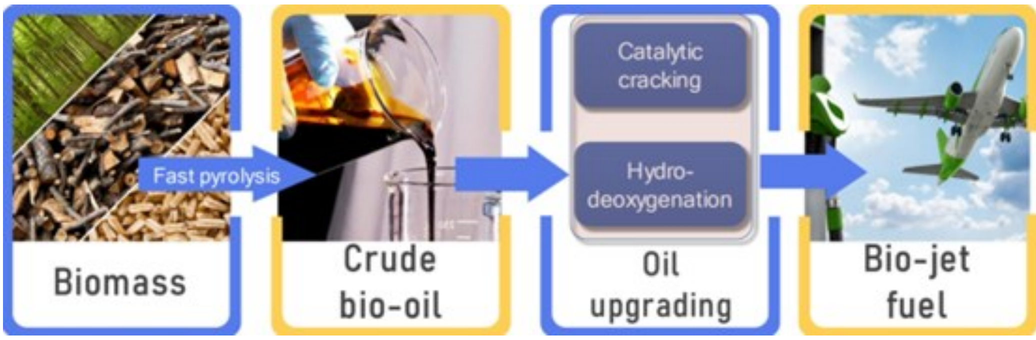


*Figure 5- Upgrading of pyrolysis oil to SAF, (Energy Conversion and Management, 2022)*

**E. Final SAF Production**

- After processing, the final product is a drop-in SAF with low sulfur content, high energy density, and reduced carbon emissions compared to conventional aviation fuels.

Advantages of Co-Pyrolysis (Plastic + SCG) could be following as:

- Higher Liquid Fuel Yield – The presence of plastic increases the hydrocarbon content in the final oil, while coffee grounds improve the quality by reducing heavy waxes.
- Improved Fuel Properties – Blending biomass with plastics balances oxygen and hydrogen content, leading to better-quality jet fuel.
- Sustainable Waste Utilization – Repurposes plastic waste and spent coffee grounds, reducing landfill waste and promoting a circular economy. (Fuel, 2023) (Fan et al., 2023)

## 6. Global Plastic Waste and Spent Coffee Ground Statistics

### Plastic Waste Statistics

- The world produces over 300 million tons of plastic annually.
- About 8 million tons of plastic end up in the oceans every year.
- Over 90% of plastic waste is not recycled, and the vast majority ends up in landfills.

### Coffee Ground Waste Statistics

- 9 million tons of coffee grounds are produced globally every year.
- Coffee waste is typically discarded or composted, with very little being utilized for energy production.

### Potential SAF Yield

Studies suggest that converting plastic waste and coffee grounds into SAF through pyrolysis could result in the production of significant amounts of fuel. For example, 1 ton of plastic waste could potentially produce around 300-400 liters of bio-oil, which could then be refined into SAF as shown in table-1. Similarly, spent coffee grounds could produce approximately 200-300 liters of bio-oil per ton. (Fuel, 2023), (Sharuddin et al., 2016) (PlasticsEurope ,2023)

| Plastic Type | Pyrolysis Oil Yield (wt.%) | SAF Fraction (wt.% of oil) | SAF Fraction (wt.%) | References |
|---|---|---|---|---|
| Polyethylene (PE) | 75–85% | 25–35% | 25–30% | Suriapparao et al. (2023); Fan et al. (2023) |
| Polypropylene (PP) | 70–80% | 20–30% | 20–25% | Mahari et al. (2021); Khandelwal et al. (2022) |
| Polystyrene (PS) | 60–75% | 10–20% | 10–15% | Fan et al. (2023) |
| Mixed Plastics | 50–70% | 10–25% | 15–20% | Sharuddin et al. (2016); PlasticsEurope (2023) |

*Table 1- Different types of plastic polymer efficiency in pyrolysis operation*

## 7. Some main players at world Involved in SAF production from plastic waste and biomass and the few main projects for SAF

A. Leading and main player Countries in SAF Research and Development:

Several countries are actively investing in SAF production from waste materials like plastic and coffee grounds. Here are a few examples:

- **United States:** The U.S. has several companies and research institutions focused on converting plastic waste into biojet fuel, including those involved in pyrolysis technologies. Companies like Agilyx and Vertimass are leading efforts in waste-to-fuel innovations.
- **Germany:** Germany is a leader in biofuel research, with significant investments in sustainable aviation fuel production from waste materials. The country is home to numerous projects investigating the use of pyrolysis for biojet production.
- **India:** India is exploring alternative uses for waste materials, including plastic waste and coffee grounds, to create biofuels. Indian companies like Plastic Energy are developing pyrolysis plants for converting waste into fuel.

B. Few projects and Innovations Around the World:

Several countries are actively developing pyrolysis projects to produce SAF from waste materials such as plastic waste and coffee grounds. While these projects demonstrate significant progress in the field of SAF production from waste materials, comprehensive data on production capacities, especially those utilizing specific feedstocks like plastic waste and coffee grounds, remain limited. The industry continues to evolve, with ongoing research and development aimed at enhancing the efficiency and scalability of these technologies. (ICAO, 2024)

Below is an overview of notable initiatives to produce sustainable aviation fuel (SAF) by different types of wastes (plastic waste, municipal solid waste and agricultural residues, Hydrogenated Vegetable Oil, recycled paper from paper manufacturers) and different technologies (pyrolysis, gasification and Fischer-Tropsch) in the United States, Japan, Portugal, United Kingdom (Energy Conversion and Management, 2024).

- **United States:** LanzaJet's Freedom Pines Fuels Facility: Located in Soperton, Georgia, this facility utilizes ethanol derived from various waste sources, including municipal solid waste and agricultural residues, to produce sustainable aviation fuel (SAF). At full capacity, it is expected to produce approximately 9 million gallons of SAF and 1 million gallons of renewable diesel annually. (GreenAir News, 2024)
- **Japan:** Sanyu Plant Service: This company is constructing a production plant in Ehime Prefecture aimed at mass-producing biojet fuel. The facility plans to use recycled paper from paper manufacturers as its primary feedstock, with operations scheduled to commence in 2025.( JAPAN Forward ,2024)
- **United Kingdom:** LanzaJet and British Airways Partnership: LanzaJet is collaborating with British Airways on a SAF production facility project in the UK.
- While specific production capacities have not been detailed, this initiative underscores the UK's commitment to advancing sustainable aviation fuel technologies. (GreenAir News, 2024)
- **Portugal:** Galp Energia and Mitsui Joint Venture: Portuguese energy company Galp Energia, in partnership with Japan's Mitsui, is investing €400 million to build a Hydrogenated Vegetable Oil (HVO) plant at its Sines refinery. Scheduled to commence production in 2026, the facility will have an annual capacity of 270,000 metric tons, converting waste materials like used cooking oils into renewable biodiesel and sustainable aviation fuel (SAF).
- **United Kingdom:** Velocys' Altalto Project: In collaboration with British Airways and Shell, Velocys is developing the Altalto facility in Immingham. The plant aims to convert household and commercial waste into sustainable aviation fuel using advanced gasification and Fischer-Tropsch technology. Once operational, it is expected to produce approximately 20 million gallons of SAF annually. (Energy Conversion and Management, 2024)
- **Germany:** BioTfueL Project: This collaborative initiative focuses on converting biomass and waste materials into biofuels, including sustainable aviation fuel (SAF). The project integrates thermochemical processes such as pyrolysis and gasification to produce biojet fuel. While specific capacity details are not publicly disclosed, (Energy Conversion and Management, 2024) the project represents a significant step in Germany's commitment to biofuel research and production.
- **India:** Indian Oil Corporation's Plastics-to-Fuel Plant: Located in Dehradun, this facility employs pyrolysis technology to convert various types of plastic waste into fuel. The plant has the capacity to process approximately 10 kilograms of plastic waste per hour, producing around 7 liters of fuel. This initiative aligns with India's broader efforts to manage plastic waste and develop alternative energy sources.
- **Mexico:** Petgas: Based in Boca del Río, Petgas has developed a machine that uses pyrolysis to convert plastic waste into gasoline, diesel, kerosene, paraffin, and coke. The system can process 1.5 tons of plastic per week, yielding approximately 1,350 liters of fuel. Initially, the process requires propane to start but becomes self-sustaining using the gas produced during pyrolysis. Currently, Petgas donates the fuel to local services, including the fire department and food delivery operations.

## 8. Case Studies: Pyrolysis of plastic waste in Action to SAF

**Example 1: Agilyx Corporation (USA)**

Agilyx uses pyrolysis technology to convert plastic waste into high-quality synthetic crude oil, which can be refined into SAF. The company's system processes more than 10 tons of plastic waste per day, contributing to both waste reduction and fuel production.

**Example 2: Plastic Energy (UK)**

Plastic Energy has developed a pyrolysis-based technology to convert mixed plastic waste into a high-quality liquid oil that can be refined into SAF fuel. Their systems are operational in Spain, and they are scaling up to other European countries.

## 9. Challenges and Opportunities

A. The key challenges in producing SAF from waste materials include:

- **Economic viability:** The cost of producing SAF from waste materials plastic waste and coffee grounds) may be higher than producing conventional fuels or other than SAF technologies.
- **Feedstock availability:** Consistent access to quality feedstock (plastic waste and coffee grounds) is crucial for scaling up production.
- **Technological limitations:** While using pyrolysis technology has shown promise for feedstock (plastic waste and coffee grounds), further advancements in efficiency and scalability are really needed and also because the level of oxygen wt.% content of a jet fuel is limited to 5 ppm, especially due to reason of SCG, ,by using only pyrolysis ,we could not produce SAF in 2 stages for example pyrolysis and distillation and it is needed an high expensive hydro-deoxygenation (HDO) unit to reduce the oxygen content of SAF from 40000-50000 ppm of bio-oil to less than 5 ppm in high temperature ( about 400 degree centigrade) and high pressure (about 50 bar) and these processing limitations make it non-feasible in industry.

B. Opportunities include:

- **Waste reduction:** Using waste materials for biofuel production reduces landfill use and environmental pollution.
- **Circular economy:** The process supports a circular economy model, where waste is continuously recycled into valuable products.
- **The Road** to Net-Zero Emissions and decarbonization with SAF.

# 10. The Future of SAF Consumption

From January 2026, airlines and aviation fuel suppliers in several jurisdictions aligned with the EU ReFuelEU Aviation Regulation are required to ensure that at least 2% of aviation fuel is Sustainable Aviation Fuel (SAF)**,** with the mandated share increasing gradually over time (European Commission, 2023; EASA, 2024). This regulation is part of a broader international effort to decarbonize the aviation sector, and similar SAF blending mandates have been introduced or announced in other regions such as the United Kingdom, indicating that SAF obligations are becoming a global policy trend rather than a regional initiative (Climate Catalyst, 2024; IATA, 2023).

As these mandatory SAF blending requirements increase progressively in the coming decades, global demand for SAF is expected to rise steadily, creating long-term market certainty for producers (IATA, 2023). This growing and predictable demand improves the economic viability of SAF production**,** encouraging investment in domestic production capacity and related supply chains. Consequently, countries that invest early in SAF production can benefit economically over time, as scaling effects, technological learning, and stable policy signals gradually reduce production costs and enhance competitiveness (IEA, 2023).

## 11. Conclusion

SAF production from plastic waste and biomass using pyrolysis technology could offer a sustainable solution to both waste management and clean energy production but not by a single and easy step and having some economic technical considerations. As technological advancements continued and more countries invest in these technologies, SAF could become more attractive and a mainstream renewable and sustainable feasible energy source for the aviation industry, contributing significantly to a greener future with considerations of economic & operational in commercial scale of combined processes and technologies especially regarding. Therefore, in response to these expanding global regulations, all countries must move toward this new international climate-change obligation by either developing domestic sustainable aviation fuel (SAF) production facilities or securing SAF supplies through international trade to ensure the future aviation fuel needs of the world are met.

## 12. References

International Civil Aviation Organization (ICAO). (2018). ICAO Environmental Report 2019: Aviation and Climate Change. Montreal, Canada.

International Air Transport Association (IATA). (2023). Sustainable Aviation Fuel: Policy, Technology and Deployment. Geneva, Switzerland.

Suriapparao et al., 2023. Fuel, Volume 337, 127061.

Mahari et al., 2021. Renewable & Sustainable Energy Reviews, Volume 144, 111033.

Fan et al., 2023. Energy Conversion and Management, Volume 302, 118110.

Sharuddin et al., 2016. Energy Conversion and Management, Volume 115.

Climate Catalyst. (2024). Sustainable Aviation Fuel Policy in the UK.

European Commission. (2023). ReFuelEU Aviation Regulation (EU) 2023/2405.

EASA (European Union Aviation Safety Agency). (2024). Sustainable Aviation Fuels.

IATA (International Air Transport Association). (2023). Net Zero Roadmap for Aviation.

IEA (International Energy Agency). (2023). Sustainable Aviation Fuels.

Fuel, 2023. Volume 337, 127061.

ASTM International. (2022). ASTM D7566: Standard Specification for Aviation Turbine Fuel Containing Synthesized Hydrocarbons. West Conshohocken, PA, USA.

Energy Conversion and Management, 2022. Volume 268, 115956.

Energy Conversion and Management, 2024. Volume 302, 118110.

Chemical Engineering World, 2024. http://chemicalengineeringworld.com.

Pearlson, M., Wollersheim, C., & Hileman, J. (2019). A techno-economic review of hydro-processed renewable jet fuel. Renewable and Sustainable Energy Reviews, 109, 542–557.

Dry, M. E. (2002). The Fischer–Tropsch process: 1950–2000. Catalysis Today, 71(3–4), 227–241.

JAPAN Forward ,2024

Energy Conversion and Management, 2025. Volume 268, 115956.

International Energy Agency (IEA). (2020). Sustainable Aviation Fuels. Paris, France.

Schmidt, P., et al. (2018). Power-to-Liquids as renewable fuel option for aviation. Energy & Environmental Science, 11, 327–343.

GreenAir News, 2024. https://greenairnews.com.

ICAO, 2024. https://www.icao.int.

Staples, M. D., et al. (2018). Aviation $CO_2$ emissions reductions from the use of alternative jet fuels. Energy Policy, 114, 342–354.

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# Decarbonization Strategies In Refinery Distillation

Dr. Shahzeb H. M. Ismail
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**Introduction**

Refineries play a pivotal role in the global energy supply chain, but they are also among the major contributors to greenhouse gas (GHG) emissions. In response to growing environmental awareness and international sustainability commitments, many refineries are adopting structured decarbonization roadmaps to minimize their carbon footprints. This article presents an in-depth overview of the decarbonization roadmap and energy optimization initiatives undertaken in a large-scale refinery. The initiatives aim to reduce carbon emissions, enhance energy efficiency, and align with the global sustainability commitments of the refining industry. There are seven major energy optimization projects that collectively contribute to significant $CO_2$ reductions, energy savings, and improved operational reliability.

## Decarbonization Strategies

### 1. Waste Heat Boiler (WHB) Installation

The Distillation Units (CDU/VDU) is the heart of any refinery, where significant quantities of fuel are consumed to heat crude oil to the desired fractionation temperatures. Typically, flue gases from the large process heaters in this unit exit at high temperatures often between 350°C and 490°C representing a substantial source of recoverable thermal energy.

To capture this lost energy, a Waste Heat Boiler (WHB) is required to install downstream of the CDU/VDU heater stacks. The system should be designed to recover waste heat from the flue gas and generate at least 50 tons per hour (TPH) of Low pressure LP steam, which can be exported to the refinery's steam header. The WHB design should optimize to minimize pressure drop across the flue gas side and prevent fouling through soot-blowing and on-line cleaning provisions. The integration also required careful assessment of stack draft, combustion air flow, and heater firing control to maintain combustion stability and safety margins.

By generating 40-50 TPH or more of LP steam from waste heat, the refinery effectively reduced its dependence on utility generated LP steam. The displaced steam production corresponded to a daily fuel saving of approximately 2200-3,360 MMBTU, translating to significant OPEX savings.

The project can contribute to an overall 1.5-2% improvement in Energy intensity index, a key metric for refinery energy performance benchmarking. This can be achieved through reduced specific energy consumption per barrel of crude processed.

The reduction in fuel firing within utility boilers resulted in an estimated annual $CO_2$ emissions reduction of about 50-65 kilotons. This outcome supports both corporate and national sustainability targets, contributing to the refinery's long-term decarbonization roadmap.

### 2. Maximizing LPG Recovery From Refinery Off-Gases

The Distillation Unit (CDU/VDU) is the primary separation section of the refinery and a major source of off-gases generated during crude heating, flash vaporization, and stabilization processes. These gases typically contain light hydrocarbons ($C_1$–$C_4$), hydrogen, and small amounts of $H_2S$ and $CO_2$. While traditionally routed to the refinery fuel gas network, these streams often contain recoverable LPG-range hydrocarbons ($C_3^+$) that can be economically extracted and utilized for higher-value applications.

To capitalize on this potential, the refinery can implement an Off-Gas LPG Recovery Project, aimed at recovering propane and butane from the light ends produced in the atmospheric and vacuum distillation sections. This initiative enhanced LPG yield, improved energy utilization, and reduced $CO_2$ emissions by decreasing the need for higher-carbon fuel firing in the heaters. For example;

- Integrating a light end absorber column
- Revamping of the overhead vapor recovery system.
- Optimization of reflux ratio and reboiler duty of the debutanizer column
- Enhancing off-gas compression and cooling stages to minimize hydrocarbon loss to the flare system.

By diverting LPG from the fuel gas system and replacing it with lower-carbon hydrogen-rich gas, the refinery can achieve an annual $CO_2$ reduction of 18-24 kilotons. This contributed meaningfully to the refinery's decarbonization roadmap.

### 3. Heater Stack Oxygen Optimization

The Distillation Unit (CDU/VDU) contains some of the largest fired heaters in the refinery, responsible for heating crude oil to its flash temperature prior to fractionation. These furnaces are among the most energy-intensive assets in refining operations, consuming a substantial portion of total fuel gas and contributing significantly to overall $CO_2$ emissions.

To enhance thermal efficiency and minimize excess energy losses, the refinery can implement a Heater Stack Oxygen Optimization Project, focusing on precise control of combustion air and improved burner management. The objective is to maintain optimal air-to-fuel ratios under all operating loads, thereby reducing fuel consumption, improving combustion quality, and lowering stack emissions. As manual damper control and non-linear draft response often resulted in fluctuating $O_2$ levels, typically between 4.5% and 6% at the stack. Engineering studies revealed that every 1% reduction in excess $O_2$ could save approximately 0.5–0.6% of fired fuel, providing a clear incentive for automation and optimization. Key actions under this project could include;

- Installation of electronic damper actuators with precise modulating capability to replace manual control.
- Integration of $O_2$ analyzers and smart PID control loops in the DCS to maintain optimal $O_2$ levels in real time.
- Sealing of air ingress points at heater doors, viewports, and inspection openings to eliminate false air entry.
- Periodic calibration and maintenance of flue gas analyzers to ensure measurement accuracy.
- Training of operations personnel on optimized firing practices and interpretation of $O_2$ trends.

These modifications can enable continuous control of excess oxygen around 2.5–3.0%, a significant improvement from baseline operation. The tighter control of combustion air resulted in daily fuel savings of approximately 4,300 MMBTU/day across the CDU and VDU heater network. This equates to an annual saving of around USD 0.8 million, depending on fuel gas pricing.

Lower fuel consumption directly translated to an annual $CO_2$ reduction of 0.4 kilotons, contributing to the refinery's decarbonization targets and improved Energy Intensity Index (EII).

### 4. Boiler Feed Water Chemical Treatment Enhancement

To improve long-term reliability and reduce chemical consumption, a comprehensive Boiler Feed Water Chemical Treatment Enhancement Program is required to be implemented. The project to optimize chemical dosing practices, introduce online monitoring systems, and improve feed water conditioning across the high or low pressure steam system.

Notably, chemical injection systems is usually operate in manual dosing mode, relying on periodic laboratory analysis to adjust treatment rates. This approach leads to fluctuations in phosphate and oxygen scavenger residuals, inconsistent pH control, and occasional episodes of high dissolved oxygen (DO), conditions that can accelerate internal corrosion and boiler tube scaling.

One option is to introduce automated and feedback-controlled chemical dosing, driven by real-time field instrumentation and integrated with the DCS. The optimization focused on three key treatment areas:

Oxygen Scavenging:

- Transition from batch dosing to continuous, flow-proportional injection based on de-aerator outlet flow.
- Install an online DO analyzers (range: 0–50 ppb) to continuously monitor oxygen levels and fine-tune scavenger dosage.
- Achieve a stable DO level below 5 ppb, compared to baseline values fluctuating up to 30 ppb.

Phosphate Treatment for Scale Control:

- Implement tri-sodium phosphate (TSP) feed control using automated dosing pumps tied to drum blow down conductivity.
- Maintain phosphate residuals in the optimum range (2.5–4.0 ppm) to prevent both under- and over-phosphate, ensuring proper scale inhibition without sludge formation.

PH and Condensate Return Conditioning:

- Install online pH analyzers at the BFW outlet and condensate header.
- Adjust amine feed rates dynamically based on condensate return quality, maintaining system pH between 8.5 and 9.5, ideal for carbon steel protection.

Tighter blow down control and reduce sludge formation can improve water use efficiency, saving an estimated 5,000 $m^3$/year of demineralized water.

Improving heat transfer and lower blow down requirements may result in an annual $CO_2$ reduction of approximately 8 kilotons, complementing refinery decarbonization initiatives.

### 5. Condensate Recovery System Revamp

The Condensate Recovery System (CRS) plays a critical role in refinery steam network efficiency, particularly in energy-intensive units such as the Crude and Vacuum Distillation Unit (CDU/VDU). Condensate generated from process heaters, exchangers, and column reboilers carries substantial thermal energy and high-purity water, making its effective recovery vital for overall refinery sustainability and cost optimization. Following items can be included as a part of revamp;

Header Pressure Optimization:

- Back-pressure control valves is to be installed to maintain steady return pressures (1.5–2.0 barg), minimizing flash losses and cavitation in return pumps.

Installation of Flash Steam Recovery Unit:

- A flash vessel and steam ejector system is to be installed to capture low-pressure flash steam from high-temperature condensate streams.
- Recover steam to be routed to the LP steam header, may provide an additional 2-5 TPH of reusable steam and reducing venting to atmosphere.

Condensate Polishing and Monitoring:

- A new conductivity-based contamination monitoring system is to be implemented at key collection points.
- Automatic diversion valves to ensure that only clean condensate is routed back to the Boiler Feed Water (BFW) system, while contaminated streams require to isolate for treatment.

Pump and Control Upgrade:

- Energy-efficient pumps with variable frequency drives (VFDs) to be replaced with the old fixed-speed units, reducing power consumption and improving flow stability.
- Condensate return rates require continuous monitoring via the DCS to ensure real-time performance tracking.

The improved heat recovery may lead to an estimated annual $CO_2$ reduction of 15 kilotons, supporting refinery decarbonization targets.

## 6. Heat Integration And Crude Preheat Train Optimization

The Crude Preheat Train (CPT) is the back-bone of energy efficiency in any refinery's Distillation Unit (CDU/VDU). It recovers heat from hot process streams mainly product draws and pump around circuits to preheat incoming crude before it enters the fired heaters. Over time, fouling, suboptimal exchanger configuration, and bypassing significantly degrade heat recovery, leading to increased fuel consumption and $CO_2$ emissions.

Recognizing this, a comprehensive Heat Integration and Optimization Project is required to undertake at Heat integrated train distillation units to enhance thermal efficiency, reduce energy intensity, and align with the refinery's decarbonization roadmap. Key initiatives under this project could include:

Pinch Analysis and Energy Mapping:

A refinery-wide pinch analysis is to be performed using Aspen Energy Analyzer to identify energy recovery bottlenecks and quantify heat losses across the preheat train.

Exchanger Cleaning and Bundle Replacement:

- Fouled exchangers particularly those handling HVGO and AGO pump around stream is required to be mechanically and chemically cleaned based on the temperature and pressure drop surveys.
- Severely fouled bundles may replace with high-efficiency corrugated plate designs, improving heat transfer coefficients by up to 25%.

Network Reconfiguration:

- Low-duty exchangers is to re-sequence to recover waste heat from product streams that previously bypass the preheat section.
- New bypass control valves to be install to enable dynamic flow switching based on feed composition and exchanger outlet temperatures.

Fouling Monitoring and Predictive Maintenance:

- Differential temperature ($\Delta T$) monitoring and pressure drop trends to be integrated into the DCS for early fouling detection.
- An AI-based predictive fouling model to be developed, correlating fouling rate with crude type, flow velocity, and operating severity.

Integration with Heaters Flue Gas:

A portion of waste heat from the CDU/VDU heater stack can be recovered via a new air preheater considering the acid dew point temperature margin and waste heat exchanger existing design targets. This has a potential of improving the preheat of the crude train about 15-25 C.

This fuel reduction against the improved preheat can translate into an estimated **annual $CO_2$ emission reduction of ~40 kilotons**, contributing meaningfully to the refinery's decarbonization objectives.

## 7. Stripping Steam Optimization

Stripping steam is essential in CDU side strippers and VDU vacuum columns for hydrocarbon vaporization and product quality control. However, overuse of stripping steam leads to energy wastage and off-gas handling issues.

A detailed operational study is required using Aspen HYSYS and Data Historian analytics to correlate stripping steam rates with product flash points and flash zone temperature. The study may reveal that steam rates is running 15–20% higher than required, mainly due to conservative operation. Following items can be considered for steam optimization;

- Install flow control valves with tighter cascade control to maintain precise steam-to-feed ratios.
- Develop an inferential models to predict optimum steam rates based on real-time product specifications.
- Implement advanced process control (APC) logic for CDU and VDU strippers to auto-adjust steam based on composition data or change in plant loads.

This can leads to significant reduction in Steam consumption refinery-wide. Expected $CO_2$ reduction of 10-13.5 kilotons/year fulfilling the decarbonization target.

**References**


I. Smith, R. (2016). Chemical Process Design and Integration (2nd ed.). John Wiley & Sons.

II. Towler, G., & Sinnott, R. K. (2022). Chemical Engineering Design: Principles, Practice and Economics of Plant and Process Design (3rd ed.). Butterworth-Heinemann.

III. Bejan, A., Tsatsaronis, G., & Moran, M. J. (1996). Thermal Design and Optimization. John Wiley & Sons.

IV. El-Halwagi, M. M. (2017). Sustainable Design Through Process Integration: Fundamentals and Applications to Industrial Pollution Prevention, Resource Conservation, and Profitability Enhancement (2nd ed.). Butterworth-Heinemann.

V. Agarwal, R., & Gupta, J. P. (2018). “Energy Optimization in Crude Distillation Units Using Pinch Analysis.” Energy Procedia, 153, 23–29.

VI. Ochoa, J. F., & Jiménez, L. (2020). “Integration of Waste Heat Recovery Systems in Refinery Operations for $CO_2$ Emission Reduction.” Journal of Cleaner Production, 256, 120498.

VII. Rahimpour, M. R., & Jokar, S. M. (2012). “Feasibility of Energy Conservation in Refining Systems through Process Revamp and Heat Integration.” Applied Energy, 93, 651–663.

VIII. Energy Agency (IEA). (2021). Energy Efficiency in Refineries: Best Practice Guidelines for Energy Management and $CO_2$ Reduction. Paris: OECD/IEA.

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# Sensorless Pump Performance & Condition Monitoring: A 50% Energy Savings Case Study

Tümay Karaver
Cumali Özel

## Introduction

As energy costs continue to rise and environmental concerns drive sustainability efforts, optimizing industrial pump performance has become critical. This case study details how a comprehensive pump performance monitoring program using sensorless technology not only diagnosed key inefficiencies in a pump system but also identified a cost-effective solution with significant energy savings. By analyzing data from a 110 kW pump operating far from its Best Efficiency Point (BEP), the study demonstrates the benefits of an integrated monitoring approach that combines Electrical Signature Analysis (ESA) with AI-based predictive maintenance algorithms.

By installing hardware solely at the Motor Control Center (MCC) panel, this method eliminated the need for traditional flow and pressure sensors while delivering real-time insights into efficiency, mechanical health, and operational risks like cavitation. The findings guided the selection of a more suitable pump, showcasing how data-driven decisions can enhance sustainability and reduce costs.

*Figure 1- Process Water Pump*

## The Existing System: Identifying Inefficiencies

The pump under review, a process water unit with a 110 kW motor, was designed for a nominal flow of 421.5 m³/h at its BEP. The installed continuous monitoring ESA system, however, revealed it frequently operated at 90-115 m³/h—well below its optimal range. This mismatch, common in oversized legacy systems, resulted in efficiencies between 40% and 60%, with annual energy consumption reaching 602,880 kWh.

Several issues compounded the inefficiency. A belt-and-pulley drive system, increasing shaft speed from 1,480 rpm to 1,630 rpm, added a 2% energy loss. More critically, operating outside the BEP heightened cavitation risk, as shown by broadband noise in the power spectral density (PSD) analysis. This noise signaled flow irregularities, leading to mechanical stress, vibration, and potential damage to bearings and seals.

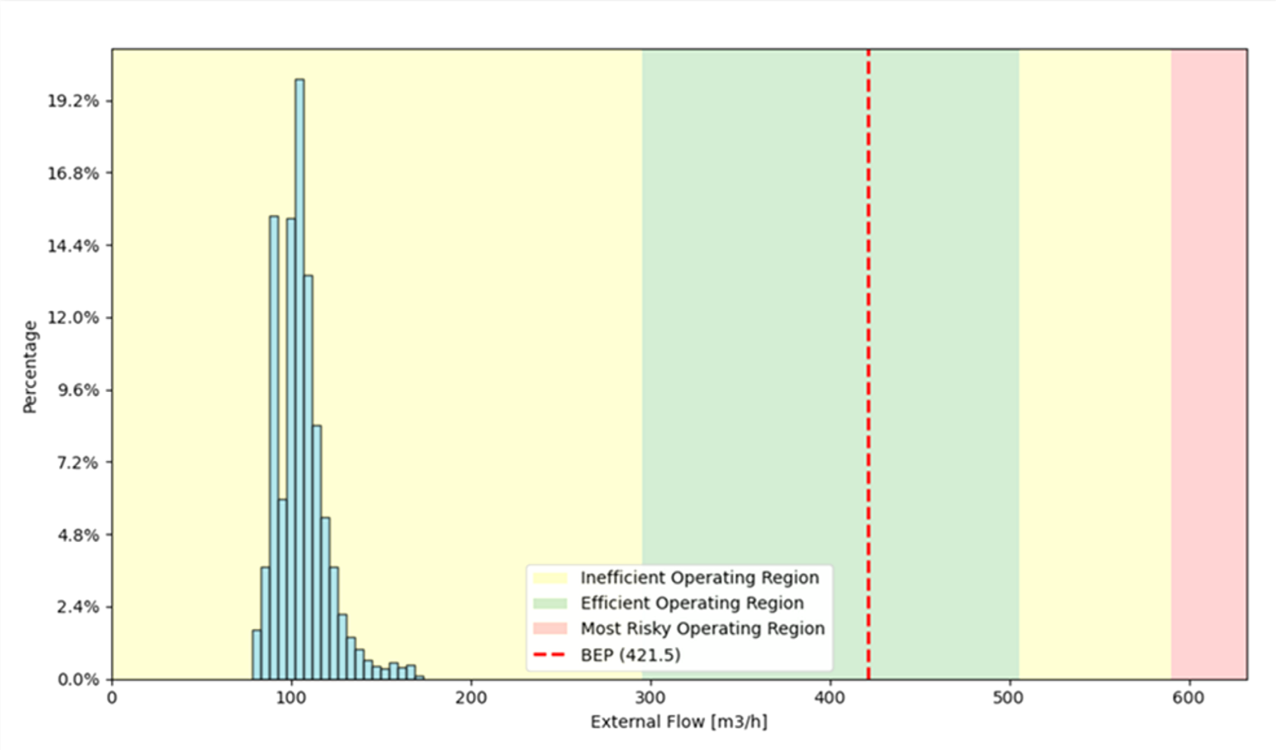


*Figure 2-Flow distribution illustrating operation below design capacity.*

**Sensorless Monitoring Methodology**

To evaluate the system, ESA was implemented using a device installed at the MCC panel. This system measured motor electrical signals—voltage and current—to calculate flow, pressure, and power without requiring external sensors on the pump or piping. The hardware interfaced directly with the motor's power supply, simplifying deployment by avoiding additional instrumentation. This approach provides several advantages:

- **Unified Data Collection:** By gathering data directly from the motor control panel, the system eliminates the need for separate flow or pressure sensors.
- **Real-Time and Historical Analysis:** The Artesis Pump Performance Analysis Software consolidates real-time data with historical records, enabling operators to quickly identify deviations from optimal operating conditions.
- **Predictive Maintenance:** Through the use of advanced AI algorithms, the software delivers early warnings for issues such as low efficiency, low flow, temperature risk, and cavitation, thus allowing for proactive maintenance.

The combination of sensor data (when available) with the estimation algorithm provides a robust solution for continuous performance monitoring. This methodology simplifies the monitoring process, reduces the reliance on costly physical sensors, and delivers accurate performance assessments even in challenging operational conditions.

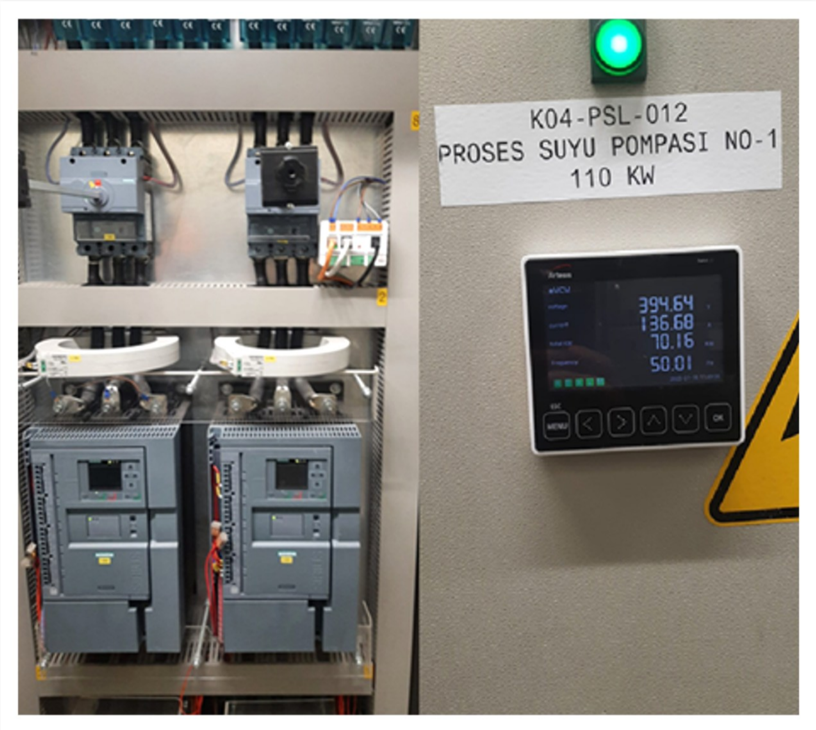


*Figure 3 Hardware Installation at the MCC*

## Analysis of Current Pump Performance

The ESA system incorporated AI algorithms to process electrical signals and estimate operational parameters. These algorithms correlated signal patterns with hydraulic performance, enabling flow and pressure predictions with a deviation of ±3% compared to on-site sensor data during validation. Beyond performance metrics, the AI analyzed signal anomalies to identify mechanical conditions, such as cavitation or bearing stress, by detecting irregularities like sideband noise in the PSD. The software presented data through graphical outputs: flow-pressure curves plotted actual operation against the BEP, with zones delineating efficient and inefficient regions; power curves showed energy consumption across flow ranges; and time-series trends tracked 24-hour and annual performance. These tools provided a detailed diagnostic framework for assessing pump behavior.

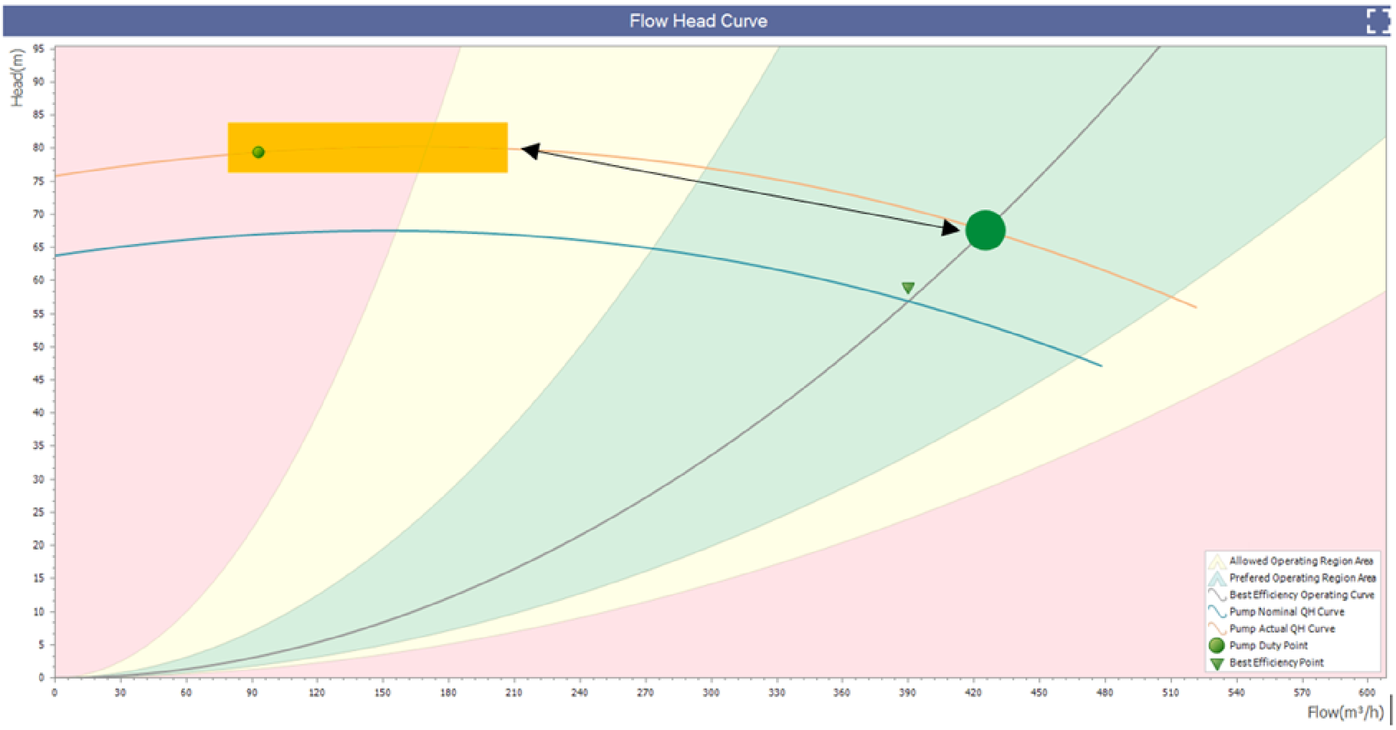


*Figure 4-Comparison of measured performance against design specifications*

### Analyzing Energy Consumption and Mechanical Health

Operational data over one year revealed energy consumption patterns at various flow rates. At 83 $m^3/h$, motor input power was 68.58 kW, rising to 83.14 kW at 168 $m^3/h$. The low-flow bias—concentrated at 90-115 $m^3/h$—drove inefficiency, with total consumption reaching 602,880 kWh annually.

Mechanical analysis identified multiple risks:

- **Low Efficiency:** Operating away from the design curve leads to increased energy usage and reduces overall system efficiency.
- **Cavitation Risk:** The low-flow, high-head conditions increase the risk of cavitation—a phenomenon where vapor bubbles form and collapse in the pump, potentially causing significant mechanical damage.
- **Component Wear:** Elevated operating stress due to improper flow conditions accelerates wear on critical components such as bearings and seals, which can lead to more frequent maintenance and potential downtime.

These findings indicated a mismatch between the pump's capacity and system demand, necessitating a reassessment of equipment sizing.

| Flow ($m^3/h$) | Motor Input Power (kW) | Annual Operating Hours | Annual Consumption (kWh) |
|---|---|---|---|
| 83.06 | 68.58 | 446.04 | 30,603 |
| 92.60 | 68.72 | 1791.72 | 123,064 |
| 102.13 | 71.26 | 2949.24 | 210,008 |
| 111.66 | 72.83 | 1827.84 | 133,091 |
| 121.19 | 74.35 | 761.88 | 56,669 |
| 130.72 | 76.03 | 295.68 | 22,472 |
| 140.25 | 77.56 | 136.08 | 10,546 |
| 149.78 | 79.02 | 63.84 | 5,047 |
| 158.31 | 81.08 | 78.96 | 6,400 |
| 168.85 | 83.14 | 47.88 | 3,980 |

*Figure 5-Energy use distribution highlighting low-flow inefficiency*

**Alternative Pump Selection and Performance Optimization**

To address these inefficiencies, an alternative pump model was evaluated. This alternative solution was selected based on its ability to meet the required operational parameters while delivering higher efficiency. Key technical improvements include:

- **Improved Flow and Efficiency:** The alternative pump has a nominal flow rate of approximately 215 $m^3/h$ and a nominal power consumption of 55 kW. At its BEP, it operates with an efficiency of around 81.3%, a significant improvement over the current pump.
- **Energy Savings:** With the alternative pump, the projected annual energy consumption drops to approximately 264,707 kWh. This represents an energy savings of roughly 338,173 kWh per year—a reduction of over 56%.
- **Enhanced Operational Flexibility:** Integration with a Variable Frequency Drive (VFD) allows for speed control based on operating conditions. Adjusting the pump's speed to match the required flow further optimizes efficiency and reduces energy consumption across varying load conditions.

The performance data clearly indicate that a switch to the alternative pump would not only resolve the issues related to low efficiency and energy waste but also extend the equipment's operational life by reducing mechanical stress and wear.

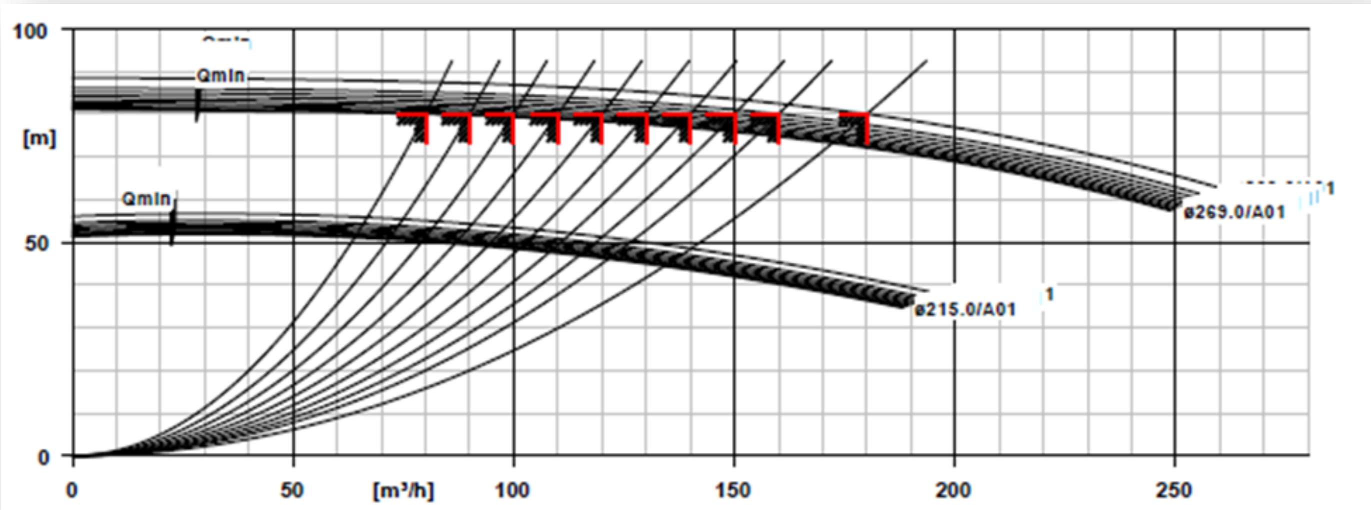


*Figure 6-Performance curves illustrating improved alignment with system demand*

**Economic and Technical Outcomes**

The alternative pump and VFD required an estimated investment of $39,000. With energy savings of 338,173 kWh annually—equivalent to $30,435 at $0.09/kWh—the payback period was 1.28 years, with a simple ROI of 78%.

Additional benefits included:

- Reduced Downtime: Early fault detection via MCC-based monitoring.
- Environmental Impact: Lower energy use reduced the carbon footprint.
- Scalability: The sensorless approach could be replicated across other pumps on-site.

| Flow ($m^3/h$) | Current P1 (kW) | Alternative P1 (kW) | Annual Savings (kWh) |
|---|---|---|---|
| 83.06 | 68.58 | 27.17 | 18,482 |
| 92.60 | 68.72 | 28.88 | 71,357 |
| 102.13 | 71.26 | 30.71 | 119,526 |
| 111.66 | 72.83 | 32.60 | 73,482 |
| 121.19 | 74.35 | 34.57 | 30,341 |
| 130.72 | 76.03 | 36.65 | 11,639 |
| 140.25 | 77.56 | 38.80 | 5,267 |
| 149.78 | 79.02 | 41.02 | 2,428 |
| 158.31 | 81.08 | 43.34 | 2,979 |
| 168.85 | 83.14 | 48.19 | 1,672 |

*Figure 7-Projected energy and cost savings with the alternative pump*

**Implementation and Validation**

A key aspect of this case study was the validation of the sensorless monitoring approach. Side-by-side comparisons between traditional flow sensor data and the values estimated by the Pump Performance Analysis Software showed a high degree of accuracy, with deviations of only around ±3%. This close alignment validates the effectiveness of the sensorless approach and underscores its potential as a reliable alternative to conventional monitoring systems.

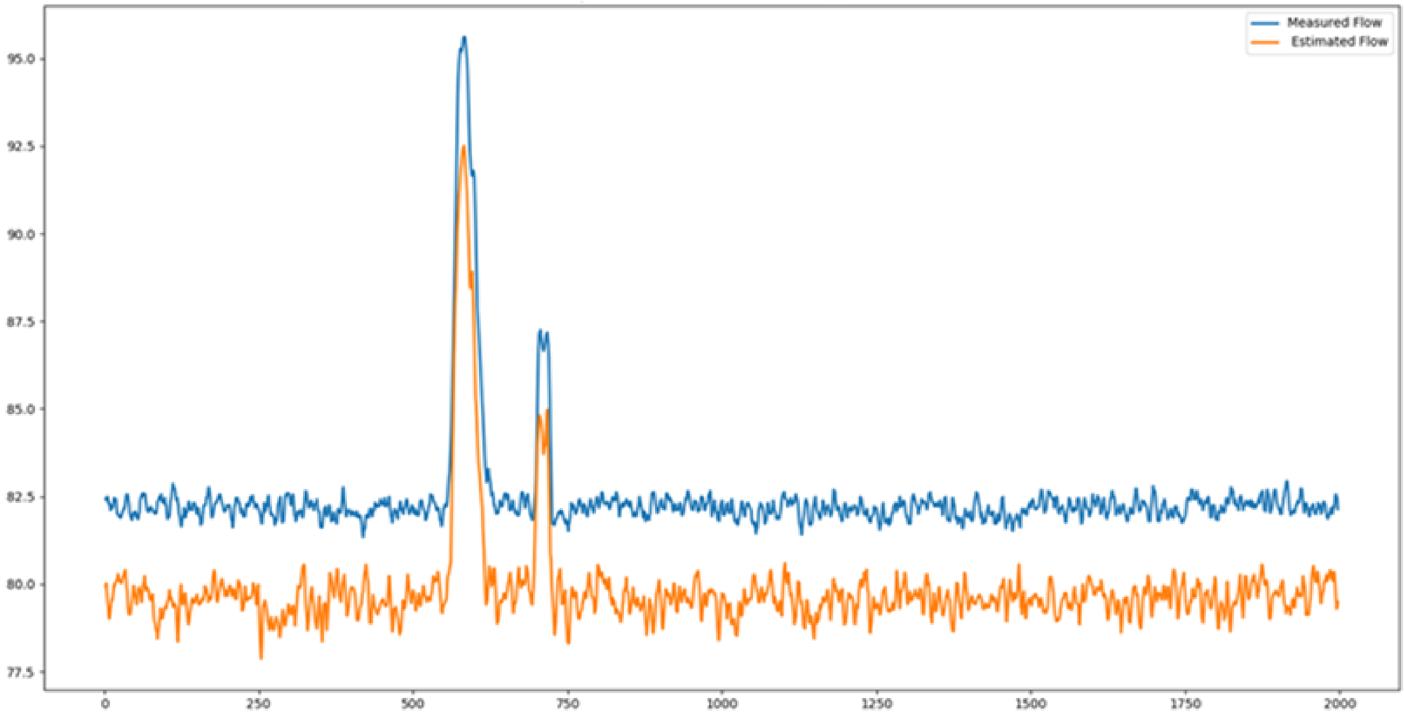


*Figure 8 Flow sensor data vs Estimated Flow*

**Conclusion**

This case study underscores the importance of integrating advanced monitoring techniques into pump operations. The key takeaways include:

- **Comprehensive Monitoring:** The sensorless monitoring system offers a unified and accurate method to track pump performance and condition without the need for expensive physical sensors.
- **Efficiency Improvement:** Operating a pump outside its BEP not only wastes energy but also increases wear and tear. Optimizing pump selection and integrating speed control mechanisms can lead to significant energy savings and operational improvements.
- **Economic Viability:** With an ROI of 78% and a payback period of just over one year, the financial benefits of upgrading to a more efficient pump system are clear.
- **Sustainability:** Reduced energy consumption lowers operational costs and contributes to environmental sustainability by decreasing the carbon footprint.

For operators and engineers seeking to improve pump system performance, the lessons from this study are clear. Regular performance monitoring—especially when combined with sensor-less technologies—can identify inefficiencies early and guide the selection of more appropriate equipment. By addressing issues such as low efficiency, cavitation risks, and component wear, companies can achieve significant cost savings and support long-term sustainability goals.

# Guidelines to Reduce Process Plants Incidents

Karl Kolmetz

**Guidelines to Reduce Process Plants Incidents**

Smarter Than the Average Bear – Yogi Bear Syndrome

As a child I liked to watch cartoons, guess I am still not fully grown as I still like to watch. One of my favorites was Yogi Bear, which started in 1961 so I would have been six years old. Yogi's favorite saying was "I am smarter than the average bear." Any survey of automobile drivers would confirm that the driver felt they were better than the average driver. This is a normal human condition. I am smarter, stronger and better looking than the average person. We could call this - the Yogi Bear Syndrome.

This normal human condition can also filter into our safety system thinking. Our plant and our industry are smarter, stronger and better looking than the average plant. Unfortunately, the same type of human still runs your plant, with the same tendencies as every other plant so you will have the same safety issues.

We get blinded by the day-to-day familiarness of our plants. We have always done it this way so it must be safe. That is why we need a cold eye review of the hazards in our plant. A third-party knowledgeable safety professional should review your plant for safety hazards. The USA EPA now requires third-party audits due to this day-to-day familiarness.

One of the worst cases of "we have always done it this way" is how pipeline pigging systems are designed. Many piping systems have 1500 PSI (10 Bar) pressures with a single isolation valve, and you reach into the confined space without a permit. This would never pass a real HAZOP, but "we have always done it this way". We need to take a cold eye view of the hazards in our plant, which is difficult.

One of the best ways to improve the safety of your plant is to review the incidents that have happened in your industry, even though you are much smarter than the other guys. Amazingly the same types of incidents happen in every industry. Maybe we are not smarter than the average bear. Normal incidents include:-

1. Hot Work Issues
2. Working at Heights
3. Fire Hazards
4. Hot Oil Hazards
5. Pressure Vessel Hazards
6. Boiler Steam Hazards
7. Electrical Hazards
8. Dust Hazards

**Palm Oil Plant Incidents**

In 2003 I began consulting in the Palm Oil Industry with a distillation company. We have published multiple papers on Palm Oil Distillation. Palm Oil, like every other industry – Refining, Ethylene, Ammonia, Natural Gas – all suffer from the Yogi Bear Syndrome. Our plant is not as dangerous as the other bears. But it is not true – any survey of incidents will find the same issues. Here is a list of some recent Palm Oil Incidents. I could find similar incidents for your plant and industry also.

1. Johor Malaysia, Jan 16, 2021 – At a biodiesel plant there was a small explosion when welding works were carried out on a storage tank at a height of 15 meters. The impact caused the welder to fall to his death while another two were injured. Hot work is a common cause of incidents.

2. Selangor, Malaysia, May 1, 2021 - At a fatty alcohol plant, a man doing welding on a storage tank at height of 10 meters fell to the ground when the tank exploded. The man died while three others were injured.

3. Sumatra, Indonesia - June 16, 2021 - At a biodiesel factory that was just commissioned in March, contractors were carry out welding on top of a tank when there was an explosion. The contractors were seriously burnt and two of them died.

4. Malaysia – 06 May 2025 - The accident has killed four workers and caused damage to a sterilizer and building structures. The force of the explosion fractured the welded joint between the sterilizer door and its support.

   This caused the door to fly about 11 meters away, while the 6 cages , which were processed that time containing palm fruit were thrown out and scattered around the sterilizer.

5. During the event , all victims were working near the sterilizer. Preliminary investigation Nilai, Negeri Sembilan, Malaysia – 19 Feb 2011 - Fire caused by hexane leakage with five injuries.

6. Bikit Kepah, Tereengganu, Malaysia – 16 Jan 2013 – Explosion of pressure vessel – four deaths.

7. Prai, Penang, Malaysia – 25 March 2013 – Oleochemical Plant Dust Explosion and fire – two deaths and three injuries.

8. Pulau Indah, Selangor, Malaysia – 07 June 2018 – Palm Oil Refinery fire – no casualties.

9. Chicago Illinois, USA – 09 August 2018 – Hot Oil Burn - An employee was in the process of preparing to transfer 180 degree F (82C) palm oil through a series of tanks and a filter press. One or more of the valves between the tanks and filter press was not in the correct position and caused the oil to overfill a tank and create pressure in the lines.

   When the employee opened the valve, which was located directly below the overfilled tank, the air pressure caused the heated oil to propel up and out the unsecured hatch lid and subsequently spray down onto the employee. The employee sustained second degree burns on his shoulder, chest and face.

   Critical Valves should be located away from the hazards.

10. Malacca Malaysia – Sept 2020 – Worker was killed after he fell into a boiling tank at a palm oil mill

11. Johor, Malaysia - May 2020 - Worker died from electrocuted at a palm oil mill

12. Malaysia - 3 May 2025 - a steam boiler explosion left four workers injured.

**Plant Challenges – Inspite of Yogi**

Today's processing plants are increasingly complex and integrated, making safe and efficient operation challenging. Any survey of process plant failures includes

1. human error
2. equipment malfunctions
3. improper maintenance
4. inadequate safety procedures.

These factors can lead to fires, explosions, leaks, and other accidents that disrupt operations and cause harm.

**Human Error:**

Mistakes made by operators, such as incorrect chemical mixing, improper equipment use, or failure to follow safety protocols, are a significant contributor to accidents.

**Equipment Malfunctions:**

Defective equipment, wear and tear, or electrical problems can lead to failures that trigger accidents.

**Improper Maintenance:**

Lack of regular inspections, repairs, or failure to follow maintenance procedures can result in equipment breakdowns and hazardous situations.

**Inadequate Safety Procedures:**

Insufficient safety protocols, lack of training, and poor safety culture can create an environment where accidents are more likely to occur.

**Plan to Improve**

The two largest causes are human error and instrumentation.

A review of these causes and a plan to reduce the largest causes would be a great start in reducing processing plant incidents

Human Error is typically the highest cause of processing incidents. Types of human error include

1. Incorrect Design
2. Operator Error, especially when starting equipment – almost 40% of equipment damage happens at start up
3. Incorrect Maintenance

Every Engineer, Operator and Maintenance Person should have an equipment fundamentals course that covers the following topics custom build for your systems.

1. Piping Temperate and Pressure Limits
2. Heat Exchanger Design, Commissioning, Maintenance and Troubleshooting
3. Process Vessel Design, Commissioning, Maintenance and Troubleshooting
4. Rotating Equipment Design, Commissioning Maintenance and Troubleshooting
5. Instrumentation Design, Commissioning Maintenance and Troubleshooting – understand instrumentation required for safe operation

6. Crystallizing / Distillation Design, Commissiong Maintenance and Troubleshooting
7. Boilers and Hot Oil Systems Design, Commissiong Maintenance and Troubleshooting

Every Engineer, Operator and Maintenance Person should have a safety fundamentals course that covers the following topics custom build for your systems

1. Hazard Identification
2. Job Safety Analysis
3. Incident Investigation
4. Hot Work Permits
5. Electrical Basics and Hazards
6. Working at Heights and Fall Protection
7. Working with hot systems and hazard identification
8. Pressure Vessel systems and hazard identification
9. Potential Fire Hazards and Mitigation

**Keys to Reduce Incidents**

Very Easy to become Yogi Bear which is the normal human condition. We must try to improve our safety knowledge even thought we are already smarter than others.

1. Review incidents that has happened in your industry and capture that knowledge in your plants training
2. Capture the incidents that has happened in your company and ensure that knowledge is in your plant training standard operating procedures
3. Ensure that your engineers, operators and maintenance understand process equipment fundamentals
4. Ensure that your engineers, operators and maintenance understand safety fundamentals that are relevant to your plant
5. Conduct a Hazard Analysis on your site – cold eye review

**References**


1. https://www.osha.gov/ords/imis/accidentsearch.accident_detail?id=108127.015 – Employee sustains burns from hot oil
2. https://www.icheme.org/knowledge-networks/communities/special-interest-groups/palm-oil-processing-sig/news/triple-tragedies-in-2021/
3. https://intranet.dosh.gov.my/osh-info-2/safety-alert/750-accident-involving-an-explosion-of-a-palm-oil-mill-steriliser
4. K Kolmetz et al, Kolmetz Handbook of Process Equipment Design, Safety in Process Equipment Design, Engineering Design Guidelines, 2014
5. Peter Faessler, Karl Kolmetz, Kek Wan Seang, Siang Hua Lee, Advanced fractionation technology for the oleochemical industry, July 2007, Asia-Pacific Journal of Chemical Engineering
6. Peter Faessler, Karl Kolmetz, Kek Wan Seang, Revamp Strategies for Fatty Acid Distillation Section In Oleo-chemical Plant, Sept 2004
7. K Kolmetz, Review of Bio Diesel Technology, Jan 2025, Engineering Practice Magazine
8. K Kolmetz, Guidelines for Palm Oil Management, Nov 2024, Engineering Practice Magazine
9. K Kolmetz, Design and Revamp Guidelines for Palm Oil Stripper Columns, Sept 2023, Engineering Practice Magazine

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