KNPC TECH

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Contents

EDITOR'S NOTE

In this issue, we publish several work papers presented by our engineers at various local and international events.

Some articles talk about the quality leap in our products following the commissioning of Clean Fuels Project (CFP) in 2022. A major achievement was the production of European winter-grade Diesel after the EU had mandated diesel fuel with a maximum sulfur content of 10 mg/kg (ppm).

One article talks about the refineries expanded flexibility to modify production slates based on market demand. Another outlines the difficulties and challenges that our Company experienced in facing the decline in gasoline demand, as well as the adaptations made to operating units to meet the COVID19 pandemic's impact and improve gasoline production with the Clean Fuels Project commissioning, and how KNPC turned from gasoline importer to an exporter of this important commodity through the new gasoline producing units.

We also cover the decommissioning of certain units to make room for new ones in MAA and MAB refineries to solve various problems, such as the storage setback which was negatively affecting the refineries' ability to meet various specifications for feedstock and treated products.

It is worth noting that some of those articles have already been published in international magazines; a something we are proud of as it reflects the high level of our technical staff.

Rakan Al-Fadala
Manager Corporate Communication

HYDROTREATER CONVERSIONS BOOST REFINERIES' FLEXIBILITY

Mina Al-Ahmadi (MAA) and Mina Abdullah (MAB) refineries recently converted an old Gas Oil Desulfurization (OGOD) Unit and obsolete Diesel Hydrotreater (DHT) to kerosene desulfurization service.
Alongside increasing the units' operational flexibility during planned maintenance shutdowns, the in-house unit conversions have also helped control sulfur specifications of the refineries' aviation turbine kerosene (ATK) production and managed kerosene inventories during shutdown. The unit revamps also have equipped the refineries with greater flexibility to modify production slates based on market demand.

Mohammad B. Matar Coordinator, TSD - MAA
Abdullah Mandani Team Leader - Process Eng., TSD - MAA
Ahmed Al-Motawa Team Leader - Process Eng., TSD - MAB
Ali Al-Mane Sen. Eng, Op. Planning, TSD - MAA
Maryam Al-Shatti Process Eng., TSD - MAB

MAA, MAB refineries

KNPC has launched the Clean Fuels Project (CFP) initiative to meet growing global demand for cleaner and more sustainable fuels conforming to international standards.

Because of more stringent sulfur specifications in line with the (CFP), MAA's OGOD and MAB's DHT were scheduled for decommissioning to make room for new units designed to ensure production of lower-sulfur domestic and export aviation fuels, including ATK, JP-8, Jet A-1, and dual-purpose kerosene (DPK).

CFP is designed for MAA and MAB treatment units to operate in different blocks. This coordinated approach ensures process units within each block are tightly integrated and coordinated across blocks to meet required production capacities as well as accommodate the refineries' feedstock and product storage availability under different operating scenarios.

The CFP specifically outlined a revised configuration for the refineries' ATK production blocks. This entailed tie-ins (TI) of existing processing units with grassroots units added as part of the CFP, which at MAA includes the 37,000 -b/d Delayed Coking Unit (DCU), and at MAB, the new 264,000-b/d Crude Distillation Unit (CDU) and 39,000-b/d Kerosene Hydrotreater (KHT).

Product specification, storage challenges

Ahead of decommissioning OGOD and DHT, MAA and MAB refineries were experiencing problems with storing untreated kerosene during turnaround shutdowns resulting from limited kerosene treatment capacity. The storage problem, in turn, was negatively affecting the refineries' ability to meet various specifications for kerosene feedstock and treated product.

The sulfur content of raw kerosene normally falls within a range of 2,300-2,500 ppm, and historically, MAA and MAB easily managed meeting this specification.

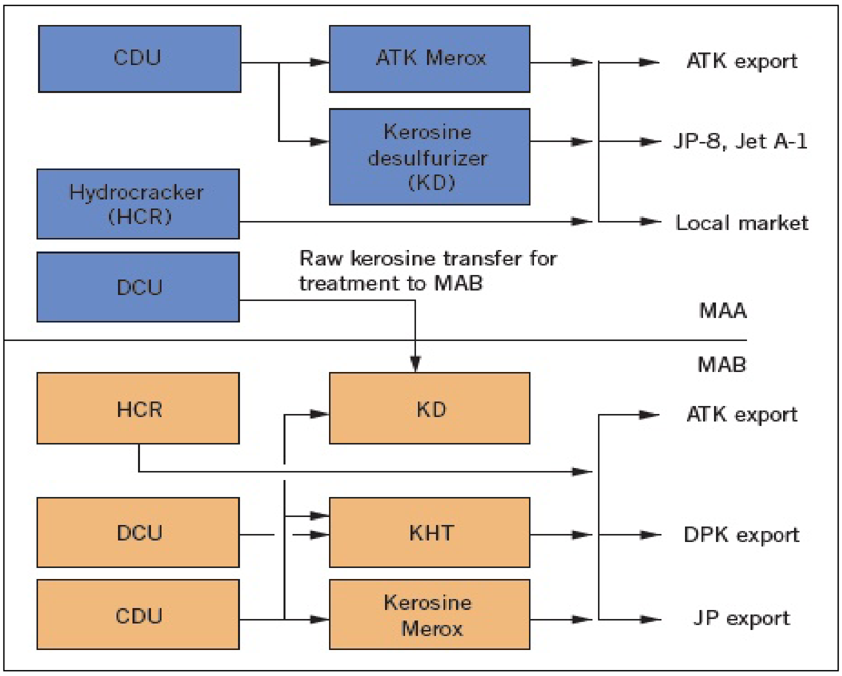


Fig. 1


Mohammad B. Matar


Abdullah Mandani

Before changes to specifications for gas oil volumes sent to Kuwaiti power stations, MAA and MAB could comfortably blend excess raw kerosene into gas oil at various proportions to achieve on-spec product. The refineries' option to blend excess amounts of either treated or raw kerosene streams gradually disappeared, however, as:

- Maximum sulfur content of gas oil supplied to power stations declined to 0.5 wt % from 1.0 wt % before further constriction to 500 ppm.

- Most power stations switched to gas-fired operations, reducing gas oil demand.

- Maximum sulfur content of Kuwait gas oil road diesel (gas oil) fell to 500 ppm from 2,000 ppm, and finally, to 10 ppm.

- Gas oil export specifications became more stringent, progressively narrowing to 2,000 ppm from 0.5% wt, then to 500 ppm, and ultimately to 10 ppm.

Even during scheduled shutdowns of MAA and MAB processing units, the combination of increasingly strict sulfur limits on and reduced demand for gas oil was resulting in large accumulations of surplus untreated kerosene that refineries would have to either transfer, store, or find ways to additionally treat to meet various domestic and export specifications.

Following the CFP's full commissioning and the accompanying planned idling of OGOD and DHT, KNPC understood that sulfur levels in MAA's ATK pool during normal operating conditions would be high, with any unscheduled upset or emergency shutdown of the refinery's Kerosene Desulphurization (KD) or Hydro Cracker (HCR) anticipated to further elevating these levels.

Any major planned or unplanned shutdowns of HCR or desulfurization units at MAA or MAB would also compound the buildup of raw kerosene that could not be stored or treated, creating logistical issues between the two refineries.

While KNPC schedules planned maintenance shutdowns of MAA and MAB's integrated operating blocks to minimize buildup of intermediate kerosene stream inventories, the refineries' operations remain vulnerable to unplanned disruptions during various unit shutdown scenarios.

MAA RUNDOWN STREAM SULFUR PARAMETERS — Table 1

Unit	Sulfur, mg/kg
KD	250
ATK Merox	2,200
HCR	< 10

Table.: 1

MAA SHUTDOWN SCHEDULE, INVENTORY BUILD THREAT — Table 2

Unit	Shutdown, duration days	Shutdown cycle , years	Threat to inventory build
KD	24-28	4	High
ATK Merox	24-28	4	Moderate
	14-18	4 (Salt, clay filter replacement)	High
HCR	42-46	4	Very high

Table 2

Ahmed Al-Motawa

With raw kerosene ullage in intermediate storage tanks not meeting required specifications, inventory management can easily become compromised.

Other potential factors contributing to constraints in managing kerosene inventories include:

- Unscheduled shutdowns of MAA's HCR and KD.
- DPK blending issues at MAB.
- Additional raw kerosene volumes to be generated in the future from KNPC's proposed new CDU and bitumen production unit.
- Interruptions during transfer of raw kerosene from MAA's DCU to MAB for further treatment.

Given the range of existing and foreseeable threats to kerosene inventory management at MAA and MAB during unit shutdowns—whether planned or unplanned—KNPC began exploring options to prevent and control both current and future inventory builds.

Project development, evolution

Initial studies conducted by external agencies to evaluate possible solutions to the refineries' kerosene inventory issues concluded that KNPC would need to make sizeable investments in either adding new desulfurization capacity and storage tanks or executing major modifications to processing units at MAA and MAB scheduled to remain in operation following the full CFP commissioning.

In line with KNPC's increased focus on reducing its overall capital and operating expenditures as well as maximizing use of its existing assets, MAA and MAB investigated the possibility of converting OGOD and DHT to kerosene service to help mitigate inventory constraints in lieu of their original plan to demolish the idled units.

Results of a FEED study on the potential unit modifications suggested the overall conversion project would take more than 4 years to complete and require major, costly changes to the units' process configurations and equipment. Given previous experience with unforeseen work delays and unexpected costs typically associated with these types of externally executed projects, the refineries instead decided to conduct internal feasibility studies and on-site trial testing to determine in-house modifications that could help reduce the overall cost and duration of the unit revamps.

Revamp approach

Hydrotreating processes-which include desulfurization-are widely used in petroleum refining to meet increasingly stringent product specifications and environmental regulations. The aim of any hydrotreating process is to eliminate impurities that diminish production quality via catalytic hydrogenation.

An HTU's reaction section removes impurities such as sulfur from gas oil, with H2S and other dissolved

OGOD OPERATING PARAMETERS Table 3

Reactor start-of-run inlet temperature, °C.	333
Reactor start-of-run outlet temperature, °C.	349
Reactor end-of-run inlet temperature, °C.	378
Reactor end-of-run outlet temperature, °C.	393
Hydrogen (H_2):feed recycle ratio at 60° F., cu ft	100:1
H_2:feed consumption ratio at 60° F., cu ft	26.6:1

Table 3

OGOD PRODUCTION YIELDS Table 4

Product	Yield, wt%
H_2	0.27
Hydrogen sulfide (H_2S)	1.10
C_1-C_2	0.74
C_3-C_5	0.25
Fraction at 176-302° F.	0.60
Gas oil	97.01

Table 4

gases subsequently removed in the unit's stripping section.

OGOD Operating parameters

Most hydrotreating units involve similar configurations, operations, and sometimes, even the same process technology. These similarities often allow operators flexibility to adapt a single unit to treat and upgrade an array of distillate feedstocks into on-spec fuels based on market demand.

MAA's OGOD reactor includes a single catalyst bed and feed distributor that uses about 90 cu m of Haldor Topsoe AS' TK-569 catalyst specifically developed for ultralow sulfur diesel and kerosine applications, providing high hydrodesulfurization activity and stability when processing straight-run and cracked distillate fractions (Fig. 1).

Tables 3 and 4, respectively, show OGOD's operating parameters and product yields.

MAB's DHT already was equipped to use a similar catalyst to that frequently employed in kerosene treating, allowing the unit to process kerosene feed without extensive modifications.

In-house modifications

Based on internal feasibility studies for the project, MAB and MAA identified and executed a mix of outside battery limits (OSBL) TIs to accomplish the unit conversions.

Ali Al-Mane

Notable OSBL TIs completed included the following:

- Installing new lines providing raw kerosene feed from CDUs and feed storage tanks to ensure units maintained adequate processing capacity.
- Installing feed TIs to connect hot coker kerosene and cold kerosene feeds to DHT's feed headers.
- Rerouting gas oil lines from OGOD and DHT, directing ATK to finished-product storage tanks, DPK tanks, or the ATK blender.
- Rerouting DHT's diesel rundown line to direct the unit's finished product to the dehazing unit's feed lines.

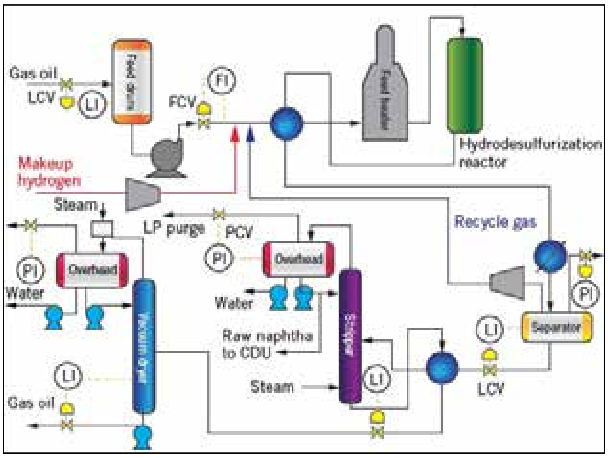


Fig. 2

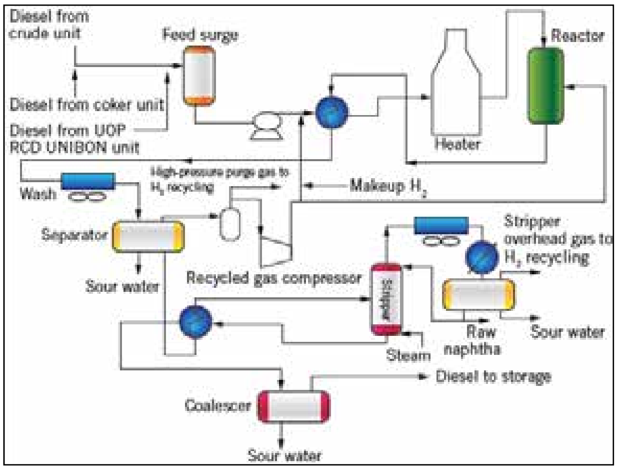


Fig. 3

Unit trial runs, startups

After completing required modifications and necessary test runs, MAB put its revamped DHT into kerosene service Dec. 8, 2019. It has remained in continuous service since.

Tables 6 and 7 show conditions under which DHT continues to operate.

After completing a successful trial run of OGOD in kerosene service during February 2021, MAA continues to flexibly operate the unit in treating kerosene volumes whenever necessary.

Lessons learned

While OGOD continues to treat kerosene on a discretionary basis, MAA observed during the revamped unit's trial run that the percentage yield of lighter ends increases,

which in turn limits the overhead vacuum drier and causes a higher moisture content in the final product. To remedy the issue, KNPC plans to further debottleneck the vacuum drier.

Separately, MAB also observed bottlenecks during DHT's trial run in kerosene service. While the refinery anticipated raw naphtha production would be higher with the unit treating kerosene, the raw naphtha yield notably increased at the unit's chosen operating rate of 25,000 b/d. MAB set that rate because operation at its full 40,000 b/d would increase vapor load and cause flooding due to the smaller diameter of the unit's stripper. The operator additionally observed DHT's recycle-gas compressor encountered limitations caused by varying molecular weights as yield patterns changed.

Since DHT's conversion to kerosene service, however, MAB's diesel dehazing unit—which was previously deemed obsolete and was scheduled for idling following CFP's completion—is successfully meeting Kuwait's required bright-and-clear quality specification when running in kerosene mode.

Results, path forward

In addition to equipping MAA and MAB with increased flexibility to modify production slates, control sulfur levels of ATK product, and manage kerosene inventories, KNPC's conversion of OGOD and DHT to kerosene operations allows

the refineries flexibility to operate the units during regularly scheduled biennial turnarounds without requiring further unit revamps.

Despite minor bottlenecks, desulfurized ATK production from the revamped OGOD and DHT units is on pace to meet KPC's post-CFP stringent specification for ATK of 1,000 ppm maximum sulfur.

Alongside low-cost, minor projects planned to debottleneck issues observed during the units' trial runs in kerosene service, the refineries will also conduct further analyses of OGOD and DHT equipment to ensure long-term integrity and reliability.

Acknowledgment

The authors would like to thank Samir K. Podder, Enni Kuman, and Mayank Garg of KNPC for their respective contributions of engineering, project planning, LP modelling, process simulation, and economic expertise in helping complete the OGOD and DHT unit conversions that led to this article.

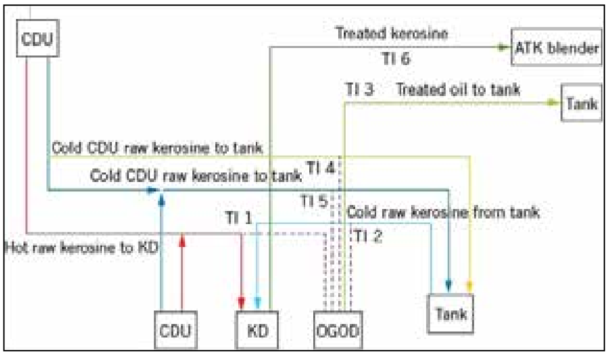


Fig. 4

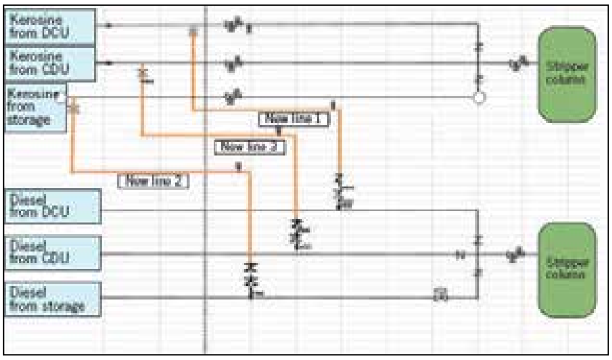


Fig. 5

DHT OPERATING, REACTION CONDITIONS Table 6

Feed	Volume, b/sd
Coker diesel	4,000-8,000
CDU raw diesel	4,000-8,000
Raw diesel from storage (coker unsaturated feed)	13,000-15,000
Reaction temperatures	°F.
Bed 1 inlet	569.3
Bed 1 outlet	590.0
Bed 2 inlet	572.7
Bed 2 outlet	579.8
Weighted-average bed	579.2

Fig. 6

DHT ATK PRODUCT CHARACTERISTICS Table 7

API gravity	45-47
Sulfur, wt %	0.05
Color, ASTM	30
Flash point, °F.	104-110
Freeze point, °C.	-48
Copper strip classification, ASTM D-130	1B
Water, ppm	100-200
Simulated distillation (SIM DIST) boiling point (0.5% off), °F.	307
SIM DIST initial boiling point (10% off), °F.	345
SIM DIST boiling point (50% off), °F.	399
SIM DIST boiling point (90% off), °F.	458
SIM DIST final boiling point (99.5% off), °F.	515
H_2S, ppm	0
Mercaptan sulfur, ppm	0.2
Particulate contamination, mg/l.	0.43

Fig. 7

Kerosene Hydrotreating Unit (U-115) - MAB

QUANTITATION OF DIESEL IN A MIXTURE OF REFINERY DIESEL AND BIO DIESEL BY EXPLORING AROMATICITY

Biodiesel is a liquid fuel often referred to as B100 or neat biodiesel in its pure, unblended form. This eco-friendly fuel is produced from straight vegetable oil, animal oil/fats, tallow and waste cooking oil, which are generally expensive to produce commercially.

Bader K. Al-Ajmi

Senior Chemist, Tech. Services Dept., MAA

Abstract

The present work is aimed at identifying the presence of refinery diesel in bio diesel products and thus addressing the issue of fuel fraud and revenue losses. Mono and di aromatic hydrocarbon components were determined using a High Performance Liquid Chromatography system (HPLC) with Refractive Index detector using an amino column. The percentage of refinery diesel is calculated based on mono aromatic hydrocarbon percentage in blend sample of refinery diesel and bio diesel. This value is further substantiated by correlating with di-aromatic hydrocarbon percentage in the blend. The quantitation is possible with a diesel product of known aromatic content when used for blending where as a qualitative determination is feasible by analyzing the sample for aromaticity for tracing any refinery diesel blends.

Introduction

With the increasing use of biodiesel worldwide, many small as well as large-scale groups are showing considerable interest in the production and distribution of this green energy fuel. Fuel grade biodiesel must be produced to strict industry specifications in order to ensure proper performance it offers.

Biodiesel is a liquid fuel often referred to as B100 or neat biodiesel in its pure, unblended form. This eco-friendly fuel is produced from straight vegetable oil, animal oil/fats, tallow and waste cooking oil, which are generally expensive to produce commercially. Considering the additional cost of converting them into bio-fuel, the commercial production of biodiesel is too expensive to compete with that of fossil diesel. As a result, mixing of fossil fuel with biodiesel could possibly be done by companies and might brand as B100. Such fuel fraud issues must be identified and prevented with regard to revenue loss and national reputation.

Fossil fuel is totally miscible with biodiesel, and their physical separation is not successful.

There is some difference in the distillation characteristics of bio diesel (210-250°C) from refinery diesel which is in the range of 190-350°C (Alptekin et al., 2008). However, confirmation on the identification and quantitation of refinery diesel as a blend component can not be arrived by distillation technique as their distillation range overlaps. Biodiesel fuels are generally more viscous than diesel fuels. The effect on viscosity of blending biodiesel and conventional diesel fuel was investigated by many and the results generally came out with poor accuracy.

Therefore, a rapid but reliable analytical methodology is inevitable to identify the presence of fossil diesel in the biodiesel.

Chemically, biodiesel comprises a mix of mono-alkyl esters of long chain fatty acids derived from vegetable oils and animal fats (Lois, 2007). The most common form uses methanol (converted to sodium methoxide) to produce methyl esters (commonly referred to as Fatty Acid Methyl Ester - FAME) as it is the cheapest alcohol available, though ethanol can be used to produce an ethyl ester (commonly referred to as Fatty Acid Ethyl Ester - FAEE) biodiesel.

Higher alcohols such as isopropanol and butanol have also been used. On the other hand, diesel fuel derived from crude oil is composed of approximately saturated hydrocarbon derivatives (75% v/v, primarily paraffin hydrocarbons including n-paraffins, iso-paraffins, and cycloparaffins), and aromatic hydrocarbon derivatives (25%

v/v), including alkyl benzenes and naphthalene derivatives.

There are certain spectroscopic/chromatographic methods, which could be used to identify the co-existence of both fossil diesel and biodiesel in a blend mixture. For example, an ASTM D7371 method quantifies the Bio Diesel content in diesel fuel oil at a range of 0.1-20% by evaluating its FAME content using FT-IR spectroscopy (ASTM D7371- 14, 2022). However, this procedure is applicable only to FAME and biodiesel in the form of FAEE, if present, will cause a negative bias. Trace (PPM) level of FAME in ATK (Aviation Turbine Kerosene) by gas chromatographic separation followed by mass selective detection of the components is another technique (McCurry, 2011). But refinery diesel blended with higher concentration (% levels) of FAME cannot be quantitated by this technique. There is another GC technique which determines the quality of biodiesel and could be employed to determine its purity. But this technique cannot predict the refinery diesel quantity in a blend of bio diesel and refinery diesel.

A high performance liquid chromatography is an efficient method for the qualitative as well as quantitative analysis of various organic components in liquid/dissolved compounds.

Coupled with a refractive index detector, HPLC could efficiently separate and quantitatively detect various mono aromatic hydrocarbons (MAHs), di aromatic hydrocarbons (DAHs), and tri+aromatic hydrocarbons (T+AHs) present in refinery diesel (ASTM D6591-19, 2019). Since biodiesel generated from vegetable oil doesn't contain any significant aromatic components, this method is expected to be efficient to find out the presence of refinery diesel in biodiesel by monitoring relevant aromatic peaks in the HPLC chromatograms.

In this paper we report our investigation on the HPLC behavior of a batch of biodiesel synthesizes and its blends with different known composition of petro-diesel refined in KNPC.

From the characteristics of resultant HPLC chromatograms it is possible to determine quantitatively the fraction of refinery diesel blended with the biodiesel with good accuracy. The experimental details and results obtained are presented and discussed.

Experimental

HPLC analysis were carried out on an Agilent Technologies 1260 Infinity model HPLC with Refractive Index detector. KNPC gasoil-TK 668 and 603 were used as the reference refinery diesel. Bio diesel used in this study is obtained from a Kuwait group intended to establish its Bio diesel production plant in Kuwait. n-Heptane is the solvent used for HPLC run which is purchased from BDH chemicals (HPLC grade with minimum assay of 99%, absorbance at 210nm 0.3 (max), absorbance at 220nm 0.1 (max.) absorbance at 245 nm 0.01 (max) and with a minimum transmittance of 50, 80 and 98 at 210, 220 and 245 nm respectively). All other reagents and solvents were of reagent grade purity and used without further purification. The HPLC analytical aromatic, di aromatic and tri+aromatic hydrocarbons in diesel fuel (ASTM D6591-19, 2019). Lab blends with different ratios of biodiesel and KNPC diesel product were used for the study.

Results and Discussion

Our literature survey to find out a reliable procedure for the quantitative analysis of fossil diesel

in a blend mixture containing major portion of bio diesel is not successful. Therefore, we investigated the possibility of developing a simple but reliable method to check the percentage of refinery diesel in bio diesel and we chose HPLC technique for this purpose. The HPLC test method we employed for this study was adopted from ASTM D6591 protocol which is the standard method

for identifying mono, di and tri+aromatic components in diesel fuel. However, the Section 1.7 of this standard protocol refers that this test method is developed for diesel samples not containing bio diesel (ASTM D6591-19, 2019). This is due to the possible interference of peak positions corresponds to Tri+aromatic hydrocarbons in diesel fuel and the peak from fatty acid methyl ester of bio-diesel. Nevertheless, we used this method for an entirely different application - to check the percentage of refinery diesel in bio diesel - with an assumption that that the HPLC peaks of mono aromatic hydrocarbons (MAHs) and di aromatic hydrocarbons (DAHs) of fossil diesel would not be interfered by any peaks derived from bio diesel, even if there are much higher concentration of bio diesel is present in the mixture compared to that of refinery diesel.

A high performance liquid chromatography instrument, fitted with a polar amino column is used for the analysis. This column has very little absorption towards non- aromatic hydrocarbon while selectively separates aromatic hydrocarbon types. As a result of the selectivity, the hydrocarbon types of different ring sizes namely MAHs, DAHs, and T+AHs will be selectively separated and eluted out from the column at a predetermined time. In our procedure, after the elution of the DAHs, the column is back flushed to elute fatty acid methyl esters as a single band.

The equipment uses a refractive index detector that detects the components as they elute from the column. The peak areas of mono and di aromatic hydrocarbons with those obtained from previously measured calibration standards as mentioned in ASTM D6591 in order to calculate percent mass of MAHs, DAHs in the sample.

The analysis started with determination of the on MAHs and DAHs values in KNPC Diesel product.

The area corresponds to both MAHs and DAHs were then calculated and this values were assigned to be 100% each (Fig.1; fourth entry). The second analysis was the HPLC run of pure biodiesel having the same concentration of fossil diesel obtained by giving a 5 times dilution with n-Heptane solvent. In this case the peaks corresponds to MAHs and DAHs were absent in the chromatogram. As expected there were intense peak belongs to fatty acid methyl ester which appeared almost in the same elution time as that of T+AHs of KNPC Diesel product (Fig.1; first entry).

Samples of biodiesel blended with different concentration of refinery diesel were investigated next. For this purpose two different blends were prepared- containing 10% and 20% refinery diesel respectively. The chromatographic analysis of both these mixtures were done by taking exactly same quantity of these blends as that used for the KNPC Diesel in the first analysis.

In the case of blend sample (10% refinery diesel + 90% Bio Diesel), the chromatograms showed smaller peaks corresponds to MAHs and DAHs and an intense peak which belongs to both T+Ahs and FAME (Fig.1; second entry). The peak area covering both MAHs and DAHs were then calculated and compared to that of pure refinery diesel. The result showed that the peak area of this 10% blend

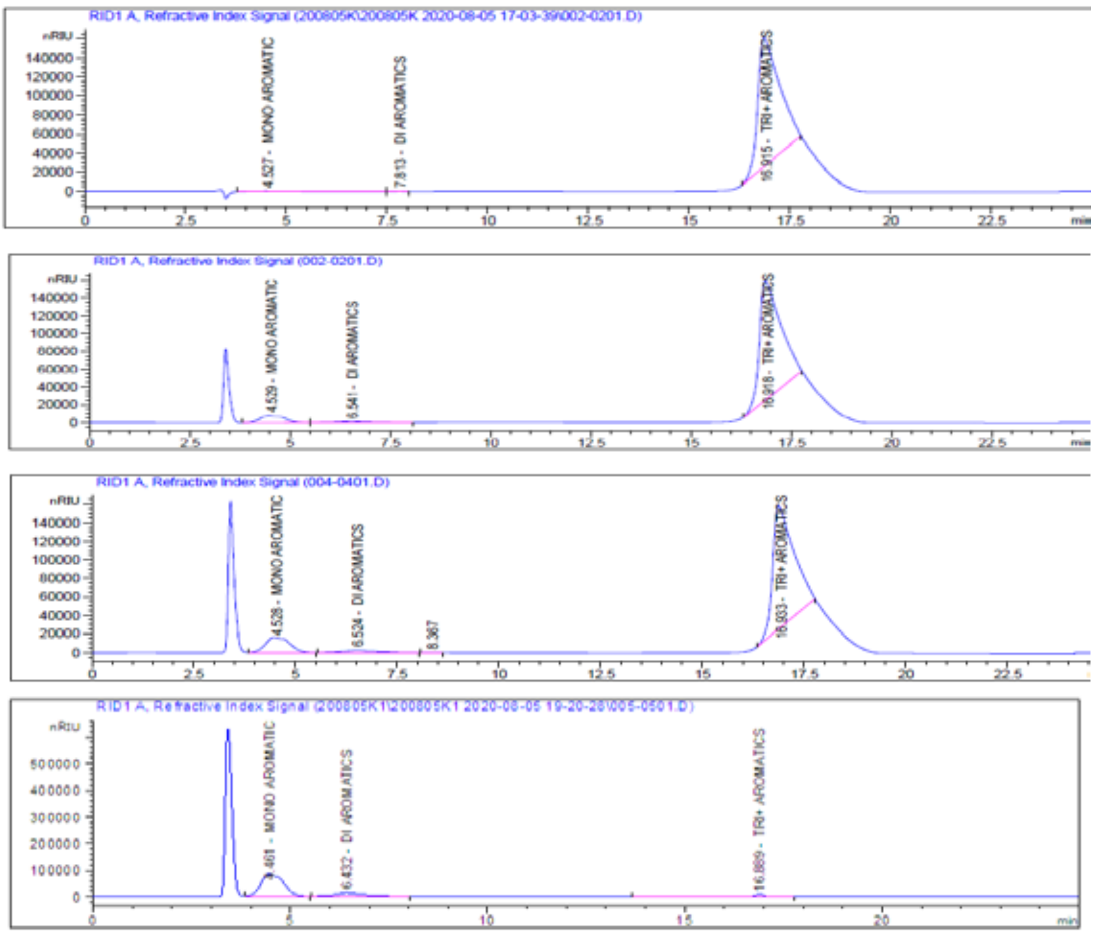


Fig. 1. HPLC chromatograms of different combinations of Bio diesel-refinery diesel blend. The values given in parentheses are the calculated area in nano Refractive Index Units (nRIU) covering MAHs and DAHs.

of refinery diesel are almost 10% (9.90% for MAHs and 10.59% for DAHs) upon comparison with that of refinery diesel, which is perfectly in agreement with the expected values The chromatogram of the HPLC run of second blend sample (20% refinery. diesel + 80% Bio Diesel), showed comparatively bigger peaks corresponds to MAHs and DAHs and as expected intense peak belongs to both T+AHs and FAME and are co-eluted masking each other (Fig.1; third entry). In this case the peak area covering both MAHs and DAHs are almost 20% (18.84% for MAHs and 17.06% for DAHs) upon comparison with that of refinery diesel, which too is in good agreement with the expected values (Table 1 & 2).

We carried out all these analysis in triplicate and confirmed that the component concentrations obtained from HPLC are in agreement with the actual blend values. Even though, there are interference between T+AHs and FAME peaks, the peak area under the MAHs and DAHs are good indicators for the quantitative determination of the fossil diesel component in the biodiesel blend. The quantitation is done based on the aromatic content in KNPC diesel fuel and can vary slightly with different lots as the aromatic concentration varies slightly between samples.

In addition, the repeatability of the test results were carried out on the same equipment by the same analyst under same experimental conditions for a 10% blend of bio diesel with another lot of refinery diesel and the results are reproduced in Table 3. It is found to give very repeatable results. Furthermore, a retention time drift check to assess the impact of bio diesel on elution characteristics was studied. This data is also given in Table 3 which showed very good consistency of elution time for both MAHs and DAHs in the repeated HPLC runs.

Conclusions

Mono and Di aromatic hydrocarbon components in refinery diesel were determined using HPLC technique, and this method has been demonstrated to be highly effective for the qualitative as well as quantitative determination of the refinery diesel in bio diesel blend. Data evaluation showed that the component concentrations obtained from HPLC chromatograms were in agreement with the actual blend values.

Repeatability studies conducted on a 10% blend of bio-diesel in refinery diesel gave very repeatable values with a standard deviation as low as 0.067 for MAHs and 0.006 for DAHs. Retention time drift for the elution characteristics of both MAHs and DAHs were also observed to be negligible in these repeated analysis. All these results clearly indicate that the HPLC method offers an easy and reliable technique for analyzing the presence of refinery diesel in bio diesel and thereby, addressing and preventing the issue of diesel fraud and blend manipulations.

Gas Oil Desulphurization Unit (U-144) - MAA

OIL MOVEMENT AREA-8 - MAA CONTRIBUTION IN KUWAIT ECONOMY AND INFRASTRUCTURE GROWTH

Oil has become the world's most important source of energy since the mid-1950s. It's not only exploration, extraction, refining processes but well managed oil storages and efficient distribution channels are equally important to reach crude and its products to end users. Operations Oil Movement (Area-8-MAA) underpins Kuwait's growth, mainly by supplying fuel to Kuwait power stations, local market, KAFCO (Kuwait Aviation Fuel Company), other Kuwait refineries and international customers through ships.

Khaled Al-Ajmi
Coordinator Operations - CFP - MAA

Introduction

Operations Oil Movement (Area-8-MAA) plays a vital role in making petro-products available at customer's destinations (internal and external) with required quantity and quality standards in a cost-effective manner. Electricity, water and transportation kind of basic infrastructure needs of Kuwait where an uninterrupted fuels supply is crucial being satisfied by Operations Oil Movement (Area-8-MAA). Fuel supply mainly includes heavy fuel oil (26,000 T/day, 3-4% Sulfur) and gasoil (3,000 T/day, 10 ppm Sulfur).

Operations Oil Movement (Area-8-MAA) not only ensuring energy needs of power stations besides, supplying fuel (500 - 1,000 T/day of Kerosene -1000 ppm S, 10,000 T/day of mogas 91/95/98 RON, and 3,000 T/day of Gasoil 10 ppm) to local markets, aviation fuel (namely, JP8 40-60 T/Day with addition of icing and corrosion inhibitor, antistatic additives) for military air crafts through road tankers, civil aviation fuel (namely Jet A1 , 2,000 T/day 3000 ppm S with antistatic additives) to KAFCO and performing inter refinery oil transfers to satisfy Mina Abdullah (MAB) refinery requirements.

In addition to that, most of petro product exports (viz. aviation fuel called Jet A1 used for civil aircrafts with antistatic additives cans, gasoil, mogas, naphtha, heavy fuel, propane, butane) also being performed by Operations Oil Movement (Area-8-MAA) via oil terminals (New Oil Pier & North Pier). Monthly almost 25 export ships being loaded, which mainly includes 6 no's of naphtha ships, 10 LPG, 6 gasoil and 3 for aviation fuel ships per month.

Dedicated team

Supply of fuels at different destinations as per schematic, is being performed by complex networks of cross-country pipelines, pumps, storage tanks, efficient metering and quality checks. A dedicated team of professionals/engineers involved behind the scenes, is divided into different departments to take care each segment of fuel distribution and storage.

Mainly there are 3 segments of Oil movement & storage dept. One dedicated for Petroleum products & crude loading to ships via New Oil Pier, North Pier, and Sea Island. MAA owns total 12 berths (8-New Oil Pier + 4- North Pier) for ship loading which includes 70 loading arms dedicated for specific product ranges. For north pier, there are 6 submarine line from MAA refinery to berths carrying dedicated products. All these 6 lines are passing through sea for approx. 6 Kms. Moreover, this division also takes care about inter refinery oil transfer operations through 7 dedicated IRT (inter refinery transfer) headers for specific products. (1 for Mogas, 2 for Black oil, 1 for Naphtha, 3 for Gasoil different grades).

MEW division of Operations Oil Movement (Area-8-MAA) is involved in supplying fuel oil to different power stations in Kuwait. These are Doha, Alzour, Sabiya, Kadma

and Shuiba power stations. Around 10 headers are supplying fuel to these power stations from Mina Al-Ahmadi Refinery (MAA). Line size ranges from 10" to 24". Almost 25 pumps being utilized to ensure continuous , uninterrupted supply of heavy fuel Oil (3-4% S) and diesel (10 ppm S) to these power stations along with crude supply to Al-Zour Refinery.

Blending process

Crude oil refining processes do not generally produce commercially usable products directly, but rather semi-finished products, which must be blended to meet the specifications of the demanded products. Such kind of product blending process also being performed by Operations Oil Movement (Area-8-MAA). It is a last step in the refining process that mixes the optimal combination of components (among various petroleum streams) to produce the final finished product. Blending is much more complicated than a simple mixing of components.

Product blending section encompasses 6 product blenders for this process (Gasoil-2, Mogas-2, Naptha-1, Fuel Oil-1).

Oil storage and movement section at MAA, as mentioned above, plays an important role in fuelling Kuwait' s economic and infrastructure growth by back boning power stations, aviation industry (KAFCO), local market and exporting oil & gas to international customers. Successful supply of crude oil (approx. 1,000,000 barrels) to KIPIC for their CDU plant commissioning is just another feather in cap of Operations Area-08 Team.

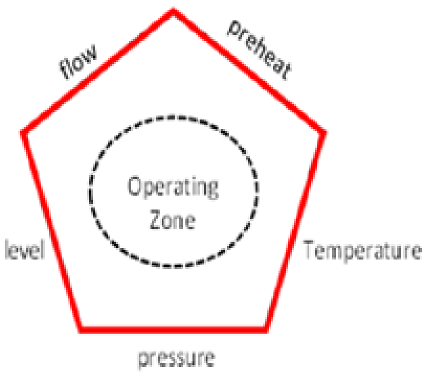
USING ADVANCED PROCESS CONTROL TO SOLVE LPG QUALITY ISSUES

Advance Process Control (APC) is widely used, not only in oil and gas industry but also in other industries like petrochemical, paper production, industrial power...etc. APC enhances the yield and quality of the product. Moreover, APC helps decreasing energy usage and operational cost. Mainly in oil and gas industry the benefit of APC implementation will payback within 1 to 2 months. A study was conducted by Honeywell shows that 1-5% increase in upstream production prior to APC execution.

Hussain Al-Salman
Process Control Engineer, Tech. Serv. – MAA

APC drives the process out of the comfort zone of individual operator towards the optimum performance while keeping the plant operating constraints:

- High and low limits (set points) for control variables.
- Product composition specifications & valve output.
- Highlight Safety and Operational Constraints.



- All constraints are considered and accounted for in the overall control and optimization strategy.

Further to operation process of APC, below are the features related to APC:

- Multi-Variable: APC helps manage the effects of multiple process variable interactions.
- Model-Predictive: APC uses dynamic models to predict process behavior into the future.
- Constraint Aware: APC monitors and maintains MVs and CVs within limits while it is controlling the process.
- Optimized Control: APC has integrated optimization capabilities to push applications toward specified design intentions.

APC implementation in Mina Al-Ahmadi

The Gas Plant LPG Fractionation Unit (Unit 30,31,32,33) consists of three trains having associated gas and liquid condensates from Kuwait Oil Company (KOC) gathering centers and gas booster stations as feed. The gases are processed by cooling, condensing and fractionation to produce C2, C3, C4 and KNG products.

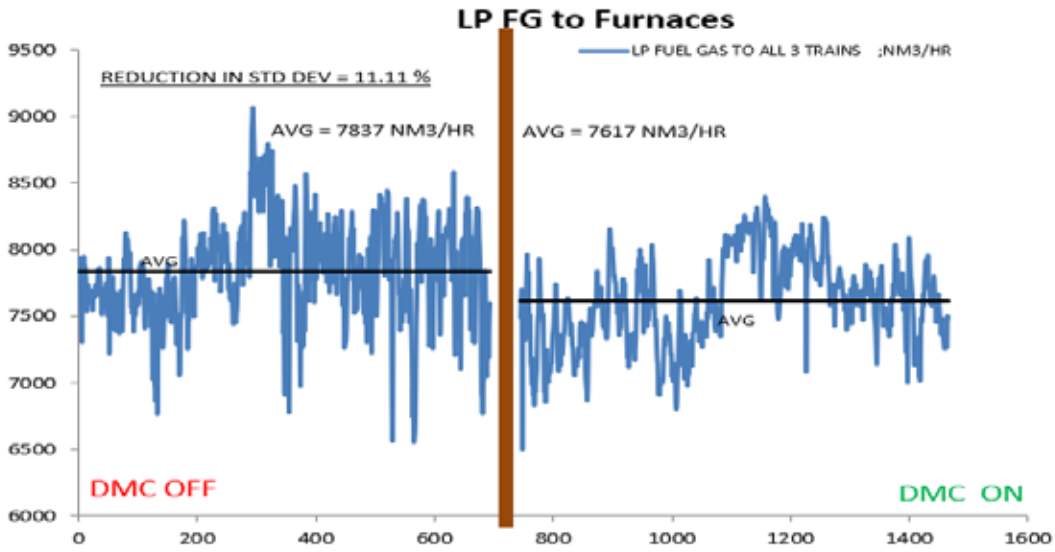
The main process variables that are monitored and optimized in the operation of the fractionators of LPG Trains are the C2, C3, C4 and KNG properties. In order to have better control on these properties, fractionators temperatures were considered in the new design.

APC Objective in Mina Al-Ahmadi

- Maximize the Ethane product from the De-Ethanizer section while maintaining the desired product specifications.
- Maximize the Propane product from the De-Propanizer section while maintaining the desired product specifications.
- Maximize the Butane product from the De-Butanizer section while maintaining the desired product specifications.
- Provide stable unit operations in the wake of disturbances, such as feed fluctuations, ambient temp, product injection to LP fuel gas header
- Maintain stable reboiler furnace operations

Achieved benefits

Post APC application several studies have been done with some benefit studies are still going on. As far as today, the recent benefit achieved for the three trains is shown the following graph:



The achieved benefit for train-2 in the 2018 was 477,000 $/yr. Currently, studies are going on to further enhance the APC performance as well as increase the benefit.

Deisopentanizer (Unit-137) - MAA

IMPLEMENTATION OF AN IN-HOUSE-BUILT APC BY USE OF DCS

Advanced Process Control (APC) enhances unit optimization through improved control, maximizes profit by increasing valuable product yield and minimizes utility consumption. However, its implementation on a third-party system with separate hardware and software incurs significant costs. Therefore, implementing an in-house-built APC system using readily available tools in the refinery was deemed necessary. The implemented APC demonstrated substantial cost savings and enhanced process reliability.

Jumanah O. Ghanim, *Process Engineering (Gas), Tech. Servs. Dept. – MAA*
K. S. Sabapathi, *Process Engineering (Gas), Sen. TPL Specialist, Tech. Servs. Dept. - MAA*

Introduction

Configuring APC typically involves a third-party system with its own hardware and separate software applications, requiring an intermediary application to bridge the gap between APC and Dynamic Control Systems (DCS). Consequently, implementing a standard APC system often results in high costs and long implementation durations.

In some cases, within the natural gas processing complex, APC applications involve only two or three manipulated variables, such as those found in De-Ethanizer, De-Propanizer, and De-Butanizers fractionation towers. Therefore, establishing proper automated and optimized control does not necessarily require a complex APC system. The main economic objective of a fractionation comes from the following two areas:

- Optimizing the Overhead and Bottom Product Purities The prices of products vary from month to month, thus increasing the production of one product over the other while allowing slippage of the less valuable product. This is to be done in a controlled manner as to optimize the production of more valuable products without crossing the allowable impurity limit for each product.
- Energy Savings Achieving its objective, an APC system will push the operation towards the borders of the operational envelope prioritizing the products qualities while maintaining the lowest possible utility consumption (steam or fuel gas). The APC achieves this by reducing the reflux ratio and reducing the separator's pressure within the allowable design constraints of the unit.

Technical content

Problem statement

For years, the LPG Train-4 operated without an APC system, with the expectation of several more years until the APC contract is awarded and a standard APC system is implemented. Therefore, to introduce the benefits of an APC, such as reducing operator workload, improving product quality, and minimizing utility consumption without incurring additional costs, implementing an in-house-built APC system with readily available tools in the refinery was deemed necessary.

Methodology and Implementation

To introduce an APC in the fractionation section of Train-4,

Jumanah Ghanim

the manipulated and controlled variables of both the De-Propanizer and De-Butanizer were identified. The manipulated variables considered were the tower's reflux flowrate, the tower's bottom temperature controller, and the tower's pressure. These manipulated variables were used to achieve the objective of controlling the two controlled variables, which were the concentration of the top product in the bottom product and the concentration of the bottom product in the top product. Once both the manipulated (MV) and controlled (CV) variables have been identified, the impact of changes in the manipulated variables on the controlled variables needed to be studied.

To develop and create a correlation between the three main manipulated variables and two main controlled variables. The tower was simulated for all the possible different cases within the design limit in terms of the CV and MV changes. The simulation results were then listed in an Excel Workbook, where the data for all cases was correlated into one formula for each CV by data regression. The two resulting equations served the purpose of calculating the MVs for the desired CVs target to be set by the operator. To avoid significant fluctuations in the parameters and possible errors, it was decided to calculate the difference between the current and actual MV values and consider the change as a delta change to move the MV from current to target value.

After setting the action plan and desired calculations and correlations, the two equations were embedded into the DCS using CALCA blocks, where current MV, CV, and target CV values were given as input to the block. Configured CALCA blocks then calculated the desired delta change of MVs as output for each MV. Multiple other blocks were configured, such as delay, tracking, clamp, operator setpoint input, and limit blocks to ensure proper and smooth operation of the APC being built with proper tuning and desired target with respect to constraint limits. The graphic display screen provided to operators for target CV and MV limits adjustment is shown below.

The APC was commissioned on stepwise basis for one tower at a time to allow time for the unit to stabilize and avoid any sudden upsets as well as allow time for monitoring. As highlighted earlier, the program was implemented in the DCS, without any additional hardware or software requirements. It was installed and commissioned on 15/04/2019. This APC implementation methodology was achieved in KNPC marking the first time for such implementation in the entire GCC.

K. S. Sabapathi

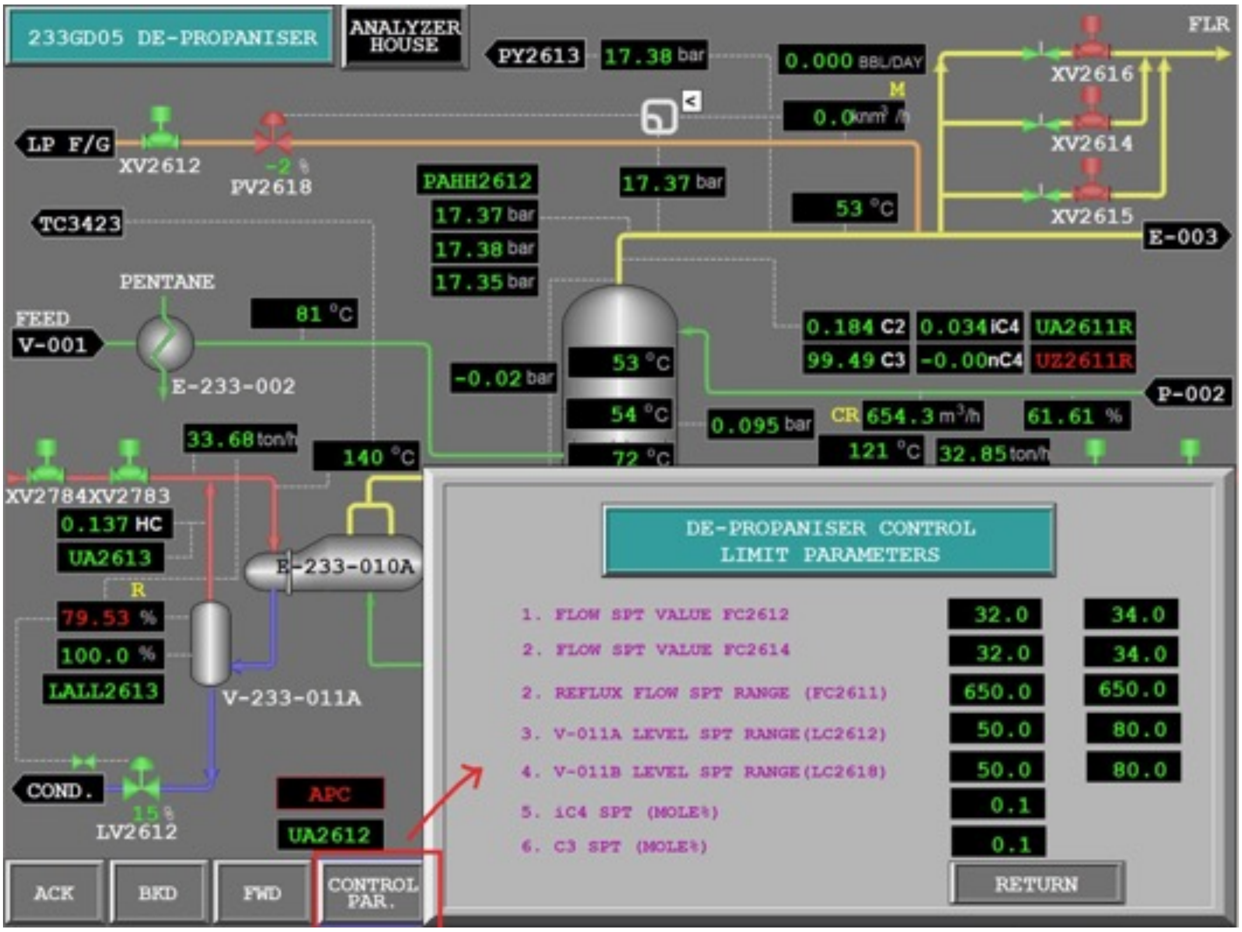
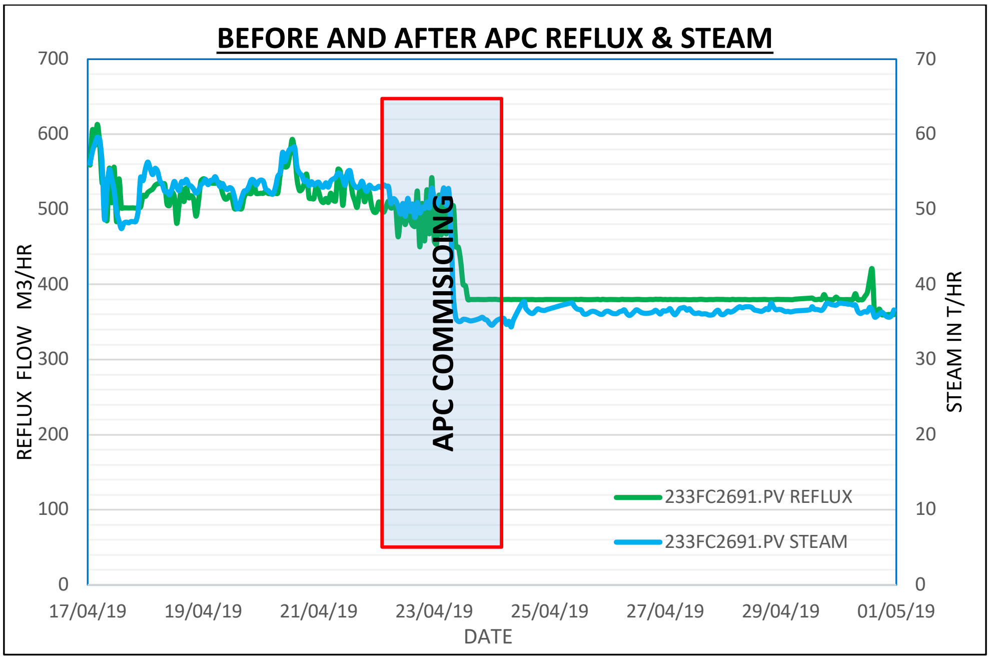


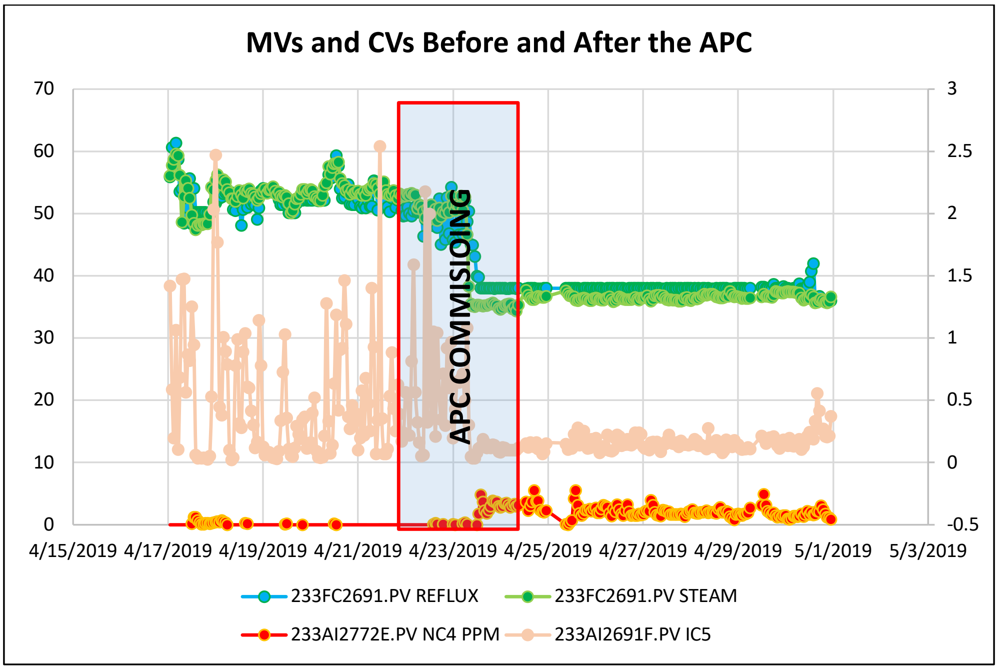
Figure 1 In-House Built APC DCS Graphic Display

Result analysis and discussion

The introduced APC improved the unit's operation and reduced the operator's intervention for the columns in which it was implemented, resulting in positive feedback from the Operations Team. Moreover, the in-house-built APC resulted in a significant reduction in the steam consumption of the towers' reboilers, as observed in the trend below.

Calculated steam savings were approximately 16 - 18 MT/hr, leading to an estimated annual saving of US$ 5 million based on actual energy prices and US$ 1 million by KNPC energy prices. The dollar value savings were calculated as per the breakdown shown in Table 1. Furthermore, consistent overhead and bottom product qualities were achieved with less undesired slippage, leading to quality improvements and greater unit reliability.





Parameter	Units	Actual Energy Price	KNPC Energy Price
FG cost	$/MMBTU	9.47	1.88
Steam cost	$/MMBTU	13.76938	2.82
Steam saving	Ton/hr	18	18
LP Steam CV	MMBTU/lb	1180	1180
LP Steam CV	MMBTU/Ton	2.6	2.6
Energy saving	MMBTU/hr	46.8	46.8
Savings/ hr	$/hr	644.40698	131.976
Savings/ Annum	$/yr	5645005.2	1156110
Savings	Million$ /Yr	5.6450052	1.15611

Table 1 Energy Costs and Savings

Conclusion

For simple systems where there are only two or three manipulated and/or controlled variables, introducing an in-house-built APC system could be more feasible than introducing a standard APC system. An in-house-built APC provides all the advantages and savings of a standard APC, without the added cost of new hardware and software. Implementing this APC in KNPC MAA Train-4 has proven successful and has provided an annual saving of US$ 1 million based on KNPC energy prices.

Deisopentanizer Unit (U-137) - MAA

ASPEN PIMS LP MODEL FOR LPG TRAINS LONG TERM PLANNING

Advance Process Control (APC) is widely used, not only in oil and gas industry but also in other industries like petrochemical, paper production, industrial power...etc. APC enhances the yield and quality of the product. Moreover, APC helps decreasing energy usage and operational cost. Mainly in oil and gas industry the benefit of APC implementation will payback within 1 to 2 months. A study was conducted by Honeywell shows that 1-5% increase in upstream production prior to APC execution.

***Muthusamy Rajendran**, Eng. Operational Planning, Tech. Serv. Dept. - MAA*
***Arbeea Al Ajmi**, Sen. Eng., Operational Planning, Tech. Serv. Dept. - MAA*

Abstract

The objective of this paper is to demonstrate how Aspen PIMS LP model can be utilized for Gas Processing facility and the benefits derived while using the same for long term production planning.

Gas processing facility at KNPC is capable of processing associated gas, free gas and condensate generated from upstream Kuwait energy companies such as Kuwait Oil Company (KOC) and Kuwait Gulf Oil Company (KGOC). In addition, KNPC refinery generated LPG also processed. Gas processing facility products are high pressure (HP) & low pressure (LP) fuel gas, ethane rich gas, propane, butane, Kuwait Natural Gasoline (KNG) and sulphur. HP & LP fuel gases are consumed internally for Kuwait energy requirements. The valuable products propane, butane and KNG are marketed through exports in international markets.

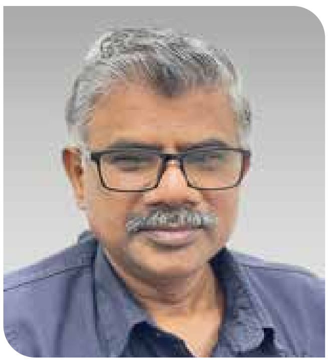
To achieve maximum profitability, planners must be able to make fast, accurate and optimal decisions about feedstock selection, modes of process operations and product mix while considering numerous other factors, such as changing market demands and price volatility and more. To mitigate the above challenges, we have developed LP Optimizer using Aspen PIMS LP optimization software for LPG Trains. That enabled KNPC to solve challenging business problems by clearly and logically capturing relevant data such as unit operation yields, stream compositions and properties, blending plans, feed and utilities requirements, process limits (constraints) and potential flexibility of cargo sizes etc. This helped to:

- Determine optimal operating conditions for LPG trains
- Seize spot sales opportunities
- Secure long-term product sales
- Optimize supply chains and value chains
- Understand and mitigate business risk due to volatility in price or reliability
- Integrate Refinery Operation to Gas Plant operation for stream exchange

Arbeea Al Ajmi

Introduction

The feedstock to gas processing facility is pool of recovered gases and un-stablished liquid generated while crude oil production and free gas generated from exclusive gas wells [1]. Hence, controlling the quality of the feedstock pool will be very difficult resulting into frequent fluctuation in quantity and quality of the feedstock available for Gas Processing facility. Accordingly, the output from Gas Processing facility is also highly fluctuating. Managing the fluctuation in the feed quality is really challenging task for production planner while forecasting production numbers to marketing team to meet the commitments. In addition to that selecting proper tools for forecasting the production numbers also tremendously influences the quality of production forecast.



Muthusamy Rajendran

Gas processing is separation process using various separation technique. Hence, excel can be helpful for calculating the yield across gas processing facility. However, it will not include process constraints while forecasting the yield. In order to utilize the process constraints while forecasting the gas processing facility production, usage of optimization tool is essential.

Excel solver will be helpful tool as long as Gas processing facility is small and no complex configuration. If complexity increases than representing the configuration in excel solver is difficult. Increased complexity will demand more process constraints consideration. Solver has limitations in the number of constraints that can be considered.

Hence, robust LP model such as Aspen PIMS are essential to represent very complex configuration, unlimited process constraints, feed stock availability, product demands, limitations on unit availability, price impact, etc.. Accordingly, KNPC started developing LP model for Gas processing facility and implemented the same for KNPC's long term planning requirement.

In this paper, we will discuss the details of approach considered while developing in house Gas processing Aspen PIMS LP model[2] and its benefit while applying the same in various planning process.

Data gathering

The following information are essential for developing LP model for gas processing facilities:

- Configuration details of gas processing facilities
- Feed streams and its qualities
- Product streams and its specification requirement
- Various utility consumptions details
- Process limitations and constraints
- Capacities of each facility

A. Configuration:

Let us explain that how we have to model Aspen PIMS LP for the following configuration. The gas processing facility has two LPG

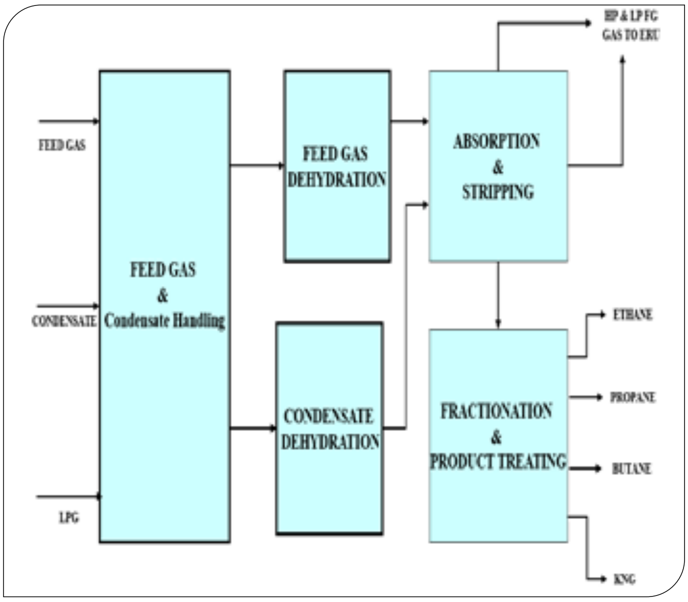


Figure -1 Train-1 Block Flow Diagram

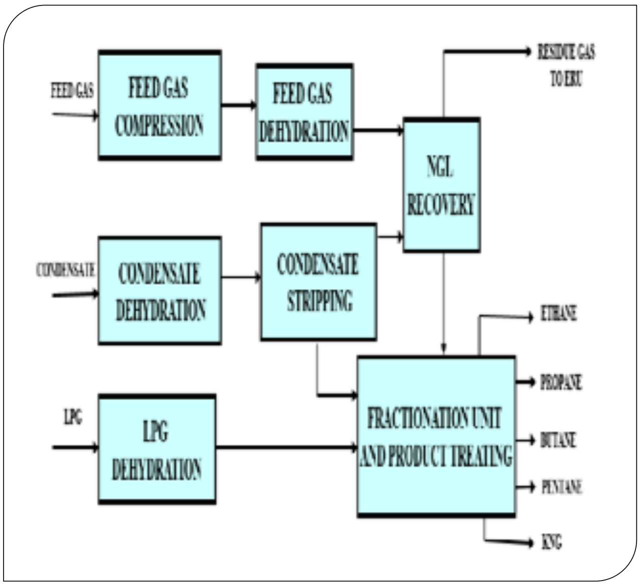


Figure -2 Train-2 Block Flow Diagram

trains with differential processing capabilities. Their Process flow diagram is as below.

B. Feed streams

- Gas
- Condensate
- Refinery LPG

Table 1 lists the feed stream qualities. Main quality is composition in percentage mol and impurity content.

C. Product streams

- HP/LP fuel gas
- Ethane rich gas
- Propane
- Butane
- Natural Gasoline

HP/LP fuel gas will contain mainly C1 and excess C2. Ethane rich gas as name indicates major portion is C2 content. The following Tables 2&3 are the qualities requirement for the export grade Propane and Butane.

Natural Gasoline is residue product. Its quality cannot be controlled. However, the quality requirement will depends on the final destination and available infrastructure for storage. For study purpose, we will assume that it will be blended in the refinery Naphtha pool.

D. Utility consumption

Table 4 lists utility consumption of each unit for one MT of respective feed.

E. Process limitations

Range of each constituent component distribution across trains products in mol %. Table 5 indicate the same:

In addition, the following process limit also exists.

- Train-1 has max C2 recovery limit of 58 %.
- Train-2 has max C2 recovery limit of 75 %.
- Train-1 absorber and stripper feed distribution should be 4:1.

F. Capacities

Table 6, represents the max

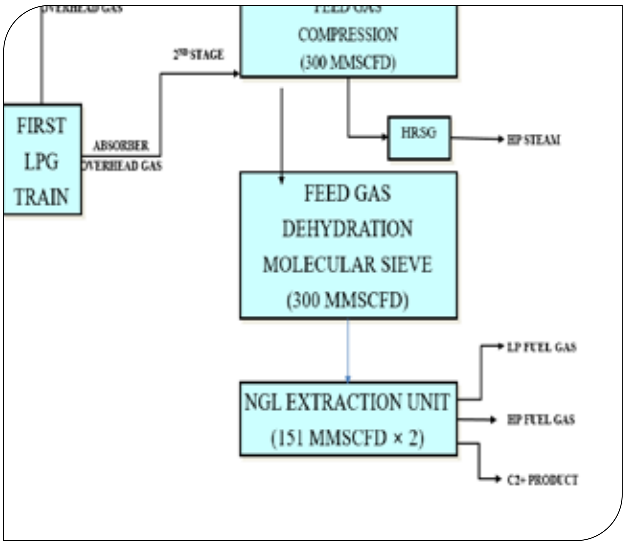


Figure -3 ERU Block Flow Diagram

Composition	Gas	Cond	Refinery LPG
Units	Mol %		
C_1	70.0	5.1	0
C_2	14.8	12.6	3.5
C_3	8.2	29.1	35.8
iC_4	1.2	7.7	19.9
nC_4	2.5	21.5	39.7
iC_5	0.5	8.2	0.7
nC_5	0.5	8.4	0.4
C_6	0.1	6.7	
C_7^+	0	0.4	
H_2S	0.1	0.1	
O_2	0	0	
N_2	0.7	0.1	
CO_2	1.4	0.1	
TOTAL	100	100	100

Table 1. Feed stock qualities

Characteristic	Specifications
C2 & Lighter (% Mole)	2 Max
C3 (Propane) (% Mole)	96 Min
C4 & Heavier (% Mole)	2.5 Max

Table 2. Propane

Characteristic	Specifications
C4 (Butane) (% Mole)	95 Min
C5 & Heavier (% Mole)	2.0 Max

Table 3. Butane

Utility	Unit	Train(1)	Train (2)	ERU
Fuel gas	mmbtu	0.708	0.788	1.03
Electricity	kwh	6.020	7.865	3.87
Steam	mmbtu	0.642	-0.250	-0.31
Dist.water	mig		0.03	0.04
Cool.water	mig			
Sea water	mig	7.260	2.061	0.57
Cat. Cost	$	0.026	0.026	0.07

Table 4. Utility consumption details.

capacities of each facility.

LP Modelling

Gas processing is simple separation process; there is no complicated conversion process involved. In gas processing C1, C2, propane, butane and natural gasoline are separated. C1&C2 is used for fuel gas requirement or as feedstock to petrochemical complex. Propane, butane and natural gasoline are exported after treatment.

Hence, we need to separate the input stream as per its composition and pool it in the respective product streams with required constraints, process limit and limitations. In addition to that we need to assign feed stock availability, product demand and price information.

Aspen PIMS LP solver will read the input from excel format. Hence, this inputs to be converted into excel format which Aspen PIMS will be able to read. The following table indicates the excel converted versions of each trains along with constraints. Aspen LP model will understand each feed, product and intermediate streams through unique three letter code. Thus, in the above example feed streams feed gas is captured as “FGs”, feed liquid is as “FLQ” and refinery LPG as “C34”. As per each train configuration, the different output as indicated process flow diagram is captured.

Control rows is required to distribute the feed constituent to the product and control the quality of the product stream. Recursion rows are used to transfer the feed stream qualities to product.

ERU receives feedstock from Train-1 stripper & absorber overhead and from Train-2 Demethanizer overhead. The two products are C2 rich gas and HP fuel gas. Steam product is represented in the utility section.

The feed stock availability is indicated through buy table. Table 7 is a sample.

The demand for product from gas processing are captured in sell table as indicated Table 8.

Process limit row will control the product quality and units process limits. In addition, we need to include some specific details for pooled streams properties either through blnprop or pguss or pcal tables. Capacities limits should be mentioned in the capacity table.

The above is the major Aspen LP model requirement to represent the gas processing facility.

The above information are needs to be properly captured in the excel as indicated. The completed excel sheet needs to be attached to Aspen PIMS LP model tree with respective tables.

Upon attaching all the required table model gas to run for data validation. It will help to identify the missing data requirements. Once validation is completed without any error than we can say LP model is ready for planning purpose.

That is Gas processing model is ready to run optimization.

Execution of run

Running the model requires Aspen

	ABS OH		ST OH		DE-C2		C3 SEP.		C4 SEP.		NG	
	Min	Max	Min	Max	Min	Max	Min	Max	Min	Max	Min	Max
C1	83.0	97.3	70.0	100.0	3.0	7.0						
C2	1.0	10.0	1.0	14.0	86.0	92.0		1.0				
C3	0.2	1.0	0.5	1.0	0.1	1.5	96.0	99.8	0.3	1.0		
IC4		0.4		0.5				0.8	20.9	38.0	0.1	0.2
NC4		0.2		0.5			0.2	0.4	60.0	79.0	0.1	1.0
IC5		0.4		0.6						0.6	27.0	40.0
NC5		0.2		0.2					0.1	0.2	29.0	42.0
C6	0.0	0.0	0.0	0.0							12.0	30.0
C7	0.0	0.0	0.0	0.0							2.0	10.0
CO2	1.0	2.0	1.0	2.0	2.0	3.6						
H2S		0.0		0.0	0.1	0.5		0.0				
N2		1.0		0.6	0.2	0.4						

Table 5. Train-1 component distribution limit

Feed Streams	Unit	Train(1)	Train (2)	ERU
Gas	MMSCFD	400	700	300
Liquid	MBPD	20	60	NA

Table 6. Capacities of each facility.

* TABL	BUY	FEEDSTOCK PURCHASES		
*		Rate/Day	Rate/Day	
	TEXT	MIN	MAX	COST
*	Imported Gases/Condensates			
FGs	GAS	-----	100.000	521.14
FLQ	Condensate	-----	100.000	506.58
BLG	Refinery LPG	-----	100.000	608.00

Table 7. Sample buy table

* TABLE	SELL	PRODUCT SALES		
*		Rate/Day	Rate/Day	
	TEXT	MIN	MAX	PRICE
MLp	Low Pressure Gas	-----	100.000	$533.63
MHp	High Pressure Gas	-----	100.000	$533.63
C2E	C2 ex TR1	-----	100.000	$537.58
C2R	C2 ex ERU	-----	100.000	$537.58
C4E	C2 ex TR2	-----	100.000	$537.58
BC3	Propane Blend	-----	100.000	$601.60
BC4	Butane Blend	-----	100.000	$625.00
KNG	Natural Gasoline	-----	100.000	$677.00
*				

Table 8. Sample sell table

PIMS LP training. Once understand the concept of LP model, running the model will be less time consuming. Trained people can complete the run within few minutes.

Output of run

LP reports are robust and will contain all the details that are required for planning purpose. The following will be the content of LP model output report.

The following tables indicate some of the above topics, which will be most relevant for planning purpose.

Analysis of LP output

LP outputs can be used to understand the detailed units operation, limitations in the unit operations, stream disposition, utility consumption details. Further, it will indicate very clearly the limiting factors for further study. The marginal value reports will indicate the economic potential of the streams. Example: For the given feed condition and the available configuration, processing of refinery LPG is not economical. It was indicated as negative margin. Model wants to increase processing of other two feed streams. None of the product sales are at limit, hence model suggests to increase the production of all products for the given prices.

Economic summary analysis will indicate how much gross margin, operating expenses and net margin in order to feel the net benefit.

Negative marginal value in capacity utilization and process limit table will indicate that particular unit has potential for increase further. Positive value indicate we are losing per unit.

Feed Stock		Units/Day	$/Unit	$/Day	Marg Val
FGs	KOC Gas	34,414	521.14	17,934,513	13.339
FLQ	KOC Condensate	2,938	506.58	1,488,332	99.359
BLG	Refinery LPG	1,000	608	608,000	-16.487
Total Purchases		**38,352**		**20,030,845**	

Table 9. Purchase Table

Utility Purchase		Units/Day	$/Unit	$/Day
CHM	Chemicals $	3	1	3
CAT	Catalyst $	1,361	1	1,361
SWT	Sea Water Mi G	156,746	0.716	112,165
CWT	Cooling Water Mi G	4	0.723	3
DWT	Distilled Water Mi G	207	8.609	1,785
KWH	Electricity kWH	295,617	0.111	32,753
GLP	LP Fuel Gas MMBTU	0	7.909	
STM	Steam Pool MMBTU	0	11.485	
	Total Utilities			**148,070**

Table 10. Utility Purchase

	$/Day
Product Sales	**21,345,798**
Feedstock Purchases	**20,030,845**
Gross Margin	**1,314,952**
Utility Purchases	**148,070**
Net Operating Margin	**1,166,882**
Penalty Cost	**0**
Net Variable Margin	**1,166,882**
Non Discounted Objective	**1,166,882**

Table 11. Economic Summary

Table of Contents

Report Content

Specification blend table will indicate the component blended in volume and weight percent. Further, it will have each property min, max and actual blend values. Marginal value will indicate the limiting property potentials.

Each unit sub-model report will contain the feed/product stream quantity and its qualities. In addition, utility consumption details and capacity consumption details.

Thus, we can understand that using the Aspen PIMS LP model for Gas Processing facility planning purpose is most appropriate tool. It will include all the constraints and will be the most optimized plan for the given inputs. It is the only tool that can capture very complicated processing facility configuration and will optimize economically.

Long term plan of gas processing facility

Long Term plan consists of bounded inputs. Gas processing facility has to predict the accurate production forecast for marketing departments in order to identify the reliable customer. Identifying the reliable customer is essential to continue the gas processing facility continuous operation.

Hence, commitment to customer and meet the committed is essential part in whole process. For that accurate forecast is essential. That can be achieved using Aspen PIMS LP model provided that forecasted inputs such as feed quantity and quality are reliable.

Units		Activity	Minimum	Maximum	Marg Val
GTG	GAS: Total Feed	14,352	0	48,000	
GT4	LPG 4: Total Feed	24,000	0	24,000	-18.046
ERU	C2 Recovery Unit	5,180	0	1,000,000	
ERX	C2 Recovery Unit Byp	0	0	0	-6.788
C2B	C2 to Equate Billion	244	0	1,000,000	
GLD	HP Gas let down to L	2,000	2,000	1,000,000	0.355

Capacity Utilized

Component to Blend		Bbls/Day	Vol%	MeTons/Day	Wt%
KNG	Comb Kng	18,434	65.36	1,897	66.13
PEN	Pentane TR4	9,771	34.64	972	33.87
Total		28,205	100	2,869	100

Product Qualities	Minimum	Product	Maximum	Margval ($/unit/bbls)
Specific Gravity	0.6000	0.6411	0.7250	
Specific Volume	0.50	1.5602		
Sulfur, wppm		463.2	950.0	
Olefins, vol %		0.0	1.0	
Aromatics, vol %		0.001		
Paraffins, vol %	76.00	100.0		
Vapor Press. Index		30.8	35.8	
Reid Vapor Pres.		15.5	17.5	

Specification Blend

Train 1 Submodel Report		Units	Units / Day	Vol%	Units	Units / Day	Wt%	Avg SPG	Avg SPV	Avg SUL
Feed										
BLP	MAB LPG	Bbls	11,238	4.72	MeTons	1,000	6.97	0.5608	1.78	0.00
FGs	KOC Gas	Bbls	199,508	83.86	MeTons	11,049	76.98	0.349	2.87	0.08
FLQ	KOC Condensate	Bbls	27,148	11.41	MeTons	2,303	16.05	0.5347	1.87	0.06
	Total		237,894	100		14,352	100	0.38	2.63	0.07
Product										
ghp	Absorber OH	Bbls	96,121	40.4	MeTons	5,180	36.09	0.3396	3.11	0.00
glp	Stripper OH	Bbls	33,574	14.11	MeTons	1,682	11.72	0.3157	3.22	0.00
C2E	C2 to Equate	Bbls	19,119	8.04	MeTons	1,148	8	0.3784	2.74	0.00
GC3	Propane	Bbls	16,928	7.12	MeTons	2,686	18.72	0	0.00	0.00
GC4	Gas Plant C4's	Bbls	24,759	10.41	MeTons	2,266	15.79	0.5771	1.73	0.00
kng	Kng	Bbls	13,481	5.67	MeTons	1,391	9.69	0.6502	1.54	0.07
	Total		203,983	85.8		14,352	100	0.361	2.14	0.01

Utility Consumption			Units/Day	$/Units	$/Day
GLP	LP Fuel Gas	MMBTU	10,161	7.909	80,368
KWH	Electricity	kWH	86,399	0.111	9,572
STM	Steam Pool	MMBTU	9,214	11.485	105,819
SWT	Sea Water	Mi G	104,196	0.716	74,561
CAT	Catalyst	$	373	1	373
Total					270,694

Capacity Utilized			Minimum	Units/day	Maximum
GTG	Tr-1 Feed	MT/Day	0	14,352	48,000

		Minimum	Units/Day	Maximum
ERU	C2 Recovery Unit MT/day	0	5,180	1,000,000

ERU detail

KNPC INITIATIVES TOWARDS CLIMATE CHANGE TRAINS LONG TERM PLANNING

The environment and climate change have become a serious concern in our modern world. As we witness significant shifts in weather patterns, rising temperatures, and depletion of natural resources, it is evident that our planet is undergoing significant transformations. The delicate balance of ecosystems is being disrupted, leading to great consequences for both nature and humanity.

Eng. Fatmah Abdulhadi
Environment Engineer, Field – MAB

Introduction

Climate change, primarily driven by human activities, causes threat to our planet's stability. The burning of fossil fuels, deforestation, and industrial emissions have intensified the greenhouse gas (GHG) effect, resulting in global warming. This warming trend has triggered environmental issues, such as melting ice caps, rising sea levels, and increased frequency and intensity of extreme weather events like hurricanes, droughts, and heatwaves. These changes, not only endanger vulnerable ecosystems, but also have severe implications for human health, food security, and economic stability. Addressing these environmental and climate challenges requires collective actions and a shift towards sustainable practices. Governments, businesses, communities, and individuals must come together to reduce GHG emissions, promote renewable energy sources, protect, and restore ecosystems, and adopt sustainable land and resource management practices.

Refineries play a significant role in the production and processing of fossil fuels, which are major contributors to GHG emissions and climate change. The refining industry is crucial for meeting the global demand for energy, but it also has a responsibility to minimize its environmental impact. As it emits Carbon Dioxide (CO_2), volatile organic compounds (VOCs), and other pollutants that contribute to air pollution and climate change. However, there are opportunities within the refining industry to address these challenges. Embracing cleaner and more efficient technologies, such as advanced emissions control systems and energy-efficient processes, can significantly reduce the environmental footprint of refineries.

In today's era of environmental consciousness, organizations across various industries are recognizing the urgent need to address climate change. Kuwait National Petroleum Company (KNPC), a prominent player in the refining sector, is taking significant steps towards mitigating its environmental impact and embracing sustainable practices. As the State of Kuwait commitment to 7.4% CO_2 reduction, KNPC has taken various initiatives to support country's carbon reduction target through many programs such as: GHG Inventory, Leak Detection And Repair (LDAR), Relief Gas Management System (RGMS), Flare Gas Recovery Units (FGRUs), Energy Efficiency, and Renewable Energy.

GHG inventory

KNPC has been a leader in the State of Kuwait in accurate accounting of Scope 1 & 2 GHG to support Kuwait's oil sector strategic initiatives. Through comprehensive GHG inventories, the company accurately measures and tracks its carbon footprint. This data enables KNPC to identify emissions, set reduction targets, and implement effective strategies to minimize its environmental impact by adopting the methodology of Intergovernmental Panel on Climate Change (IPCC) reporting guidelines since 2015 - 2016. The inventory involves identifying and quantifying the sources of GHG such as: flaring, electricity consumption, leaks, HFCs, and hydrogen production. Emissions of different gases are converted into CO_2 equivalent based on their global warming potential over a specific time.

Based on figure 1, only 9% of the total amount of GHG emissions in the State of Kuwait is caused by KNPC.

RGMS & LDAR

To address and minimize fugitive emissions, the company has implemented RGMS and LDAR programs. These initiatives involve regular inspections, maintenance, and prompt repairs of equipment and pipelines to prevent leaks and reduce emissions of VOCs. KNPC's voluntarily implemented these programs in its refineries and local marketing operations with a six-month frequency. Example of significant reduction of emission in 2022 in MAB Refinery using LDAR program is shown in figure 2.

FGRU

Flaring, a common practice in the refining industry, releases significant amounts of GHG into the atmosphere. By capturing and recovering these gases, KNPC not only reduces emissions, but also maximizes the utilization of valuable resources. The FGRU minimize the flaring by recovering the gases.

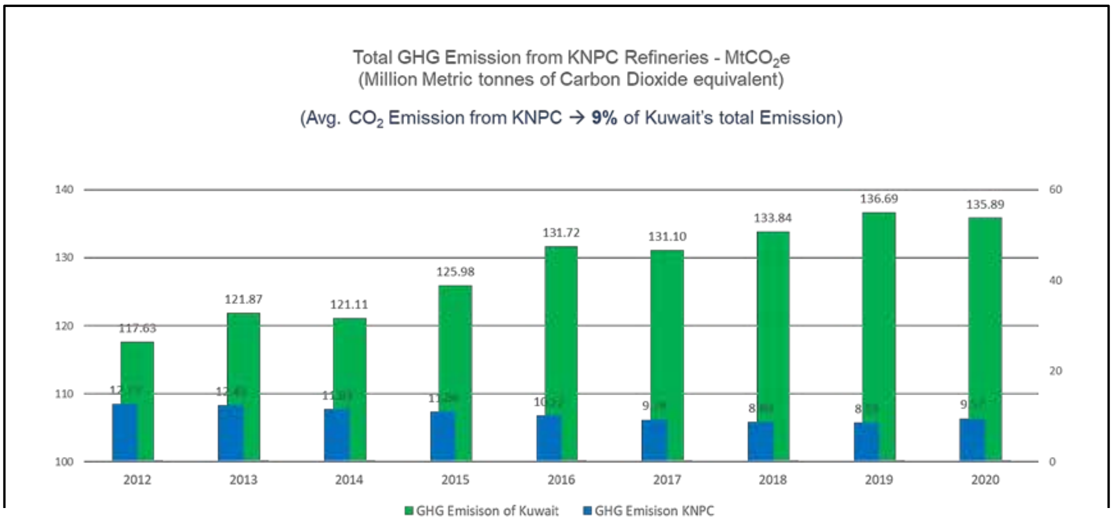


Figure 1: Total GHG Emission from KNPC Refineries

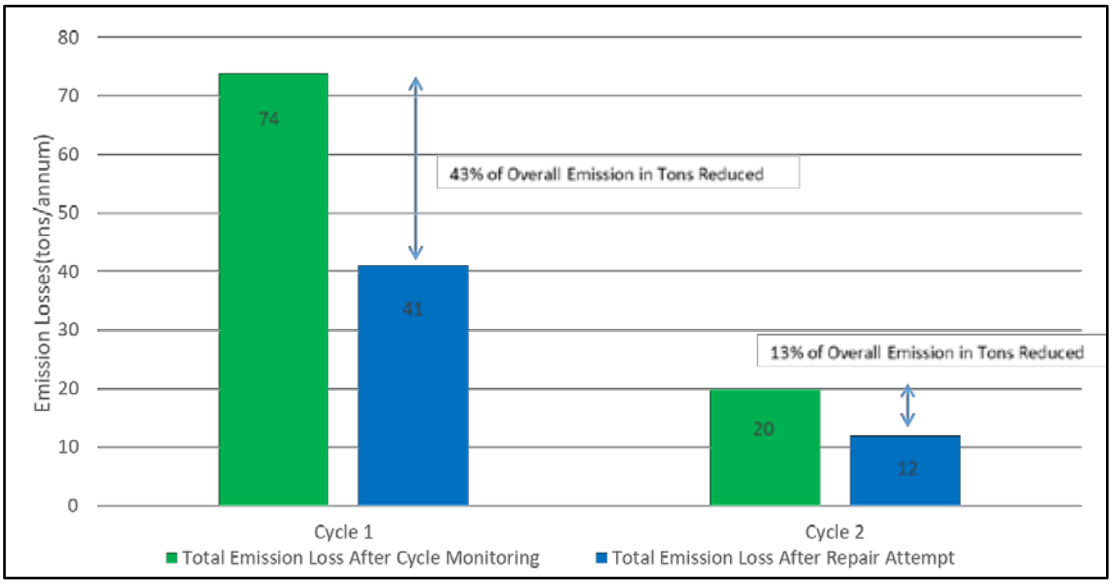


Figure 2: Reduction of emission using LDAR in 2022 in MAB Refinery

This project considered as the first Clean Development Mechanism (CDM) project registered in the state of Kuwait in both MAA in 2012 and MAB in 2014. The potential reduction of CO2 emission from the project is approximately 54,000 tons in MAA and 91,000 tons in MAB as shown in figure 3.

Energy efficiency

Energy efficiency initiatives in refineries can lead to substantial cost savings and directly translate into reduced energy consumption. KNPC has implemented an air preheated system for heaters. This innovative technology preheats the combustion air before it enters the heaters, resulting in improved energy efficiency and reduced fuel consumption. By maximizing the efficiency of processes, equipment, and systems the refinery can decrease its carbon footprint and mitigate the environmental impact associated with the operations. This improvement in energy efficiency can be seen clearly in figures 4 & 5.

Renewable energy

Solar panels utilize the power of the sun to generate clean energy, reduce dependence on conventional sources and mitigate carbon emissions. KNPC has taken a significant step towards sustainable energy by integrating renewable sources into its operations. One notable initiative is the installation of solar panels in KNPC petrol filling stations and in refineries streetlights.

Summary and path forward

In conclusion, to be in line with Kuwait Petroleum Company (KPC) 2050 Strategy, KNPC has a future plan toward sustainability in energy transition through reaching zero non-emergency flaring, commissioning FGRU in MAB CFP, placing charging points for electrical vehicles in KNPC's

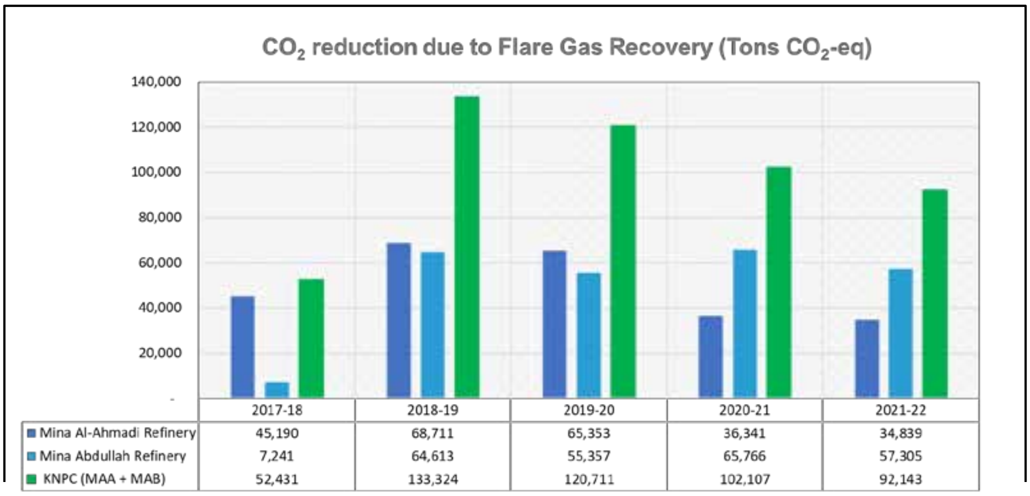


Figure 3: CO2 reduction due to Flare Gas Recovery (FGRU)

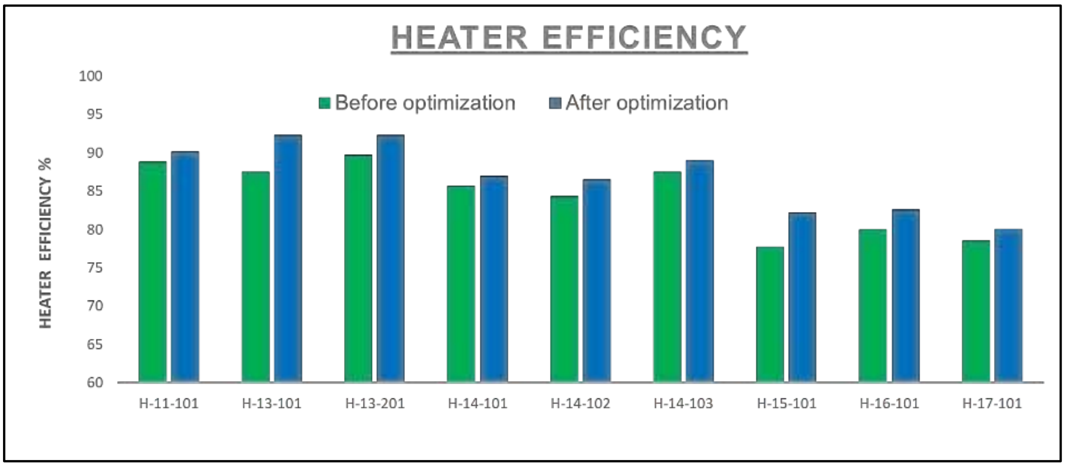


Figure 4: Heaters Efficiency

petrol filling stations, starting biofuels business, and coordinating with KEPA for the plantation of Mangrove. As climate change becomes an increasingly urgent global challenge, KNPC is at the forefront of driving change in the refining industry. Our QHSSE Policy stands firmly on the principle of environmental preservation, reflecting our strong commitment to protecting our planet. Thus, with initiatives mentioned above, KNPC is actively working towards reducing its environmental impact, contributing to global climate change mitigation efforts, and securing a greener future.

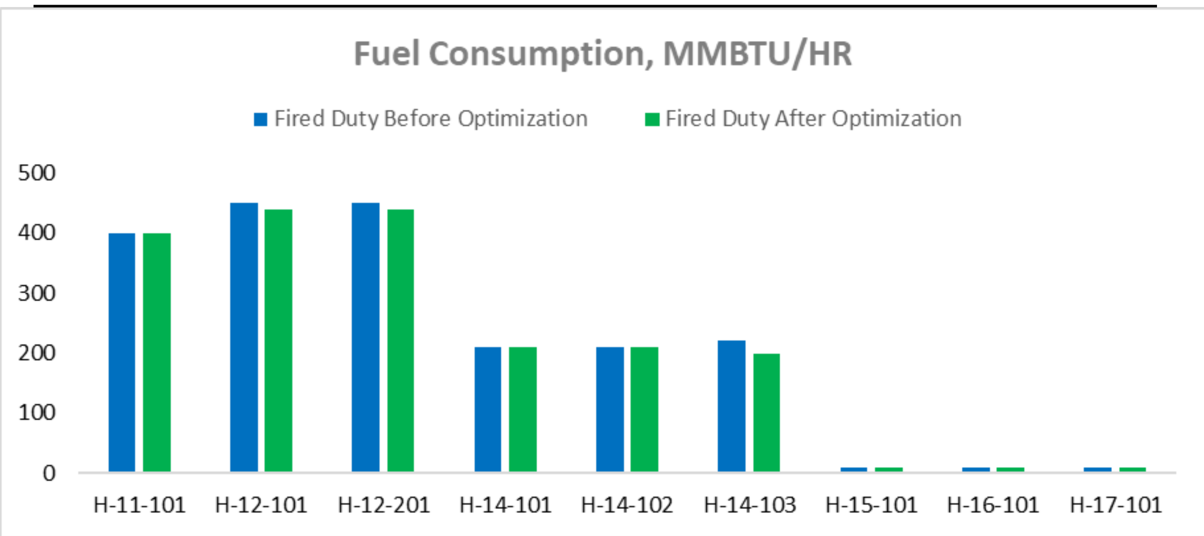


Figure 5: Fuel Consumption

FGRU Unit - MAA

REFINERIES' PANDEMIC TURMOIL–KNPC'S GASOLINE DEMAND: OPPORTUNITIES AND CHALLENGES

The COVID-19 pandemic is regarded as one of the most exceedingly difficult challenges faced in the oil and gas industry's history. Refineries around the world encountered daunting periods in operating their units at their turndown capacities and were forced to temporarily shut down units at their facilities.

Mohammad Matar*, Coordinator, Tech. Srvs. Dept. MAA*
Dalal Alqallaf*, Sen. Eng., Operational Planning, Tech. Srvs. Dept. - MAA refinery*
Ibrahim Alesmail*, Chemical Engineer, Operational Planning, Tech. Srvs. Dept. - MAA*
Faisal Al-Mutairi*, Chemical Engineer, Operational Planning, Tech. Srvs. Dept. - MAA*

Introduction

The unexpected conditions added substantial pressure on refinery production volumes and profit margins to survive in this challenging market situation, compelling refiners to adapt and improvise. The unparalleled quarantines and lockdowns imposed due to COVID-19 had serious consequences on fuel demand and oil prices.

Gasoline demand declined drastically during the pandemic, forcing refineries to adjust units' operations. However, some units had flexibility such as fluid catalytic cracking (FCC) and naphtha reforming units, as they play a vital role in optimizing refinery economics. The FCC unit converts vacuum gasoil/coker gasoil (VGO/CGO), upgrades feedstocks into high-value fuels and chemicals feedstock, which boost the value of refinery product slates. Moreover, naphtha reforming, isomerization and alkylation have a significant contribution to gasoline production.

This article features a comprehensive analysis of the difficulties and challenges that Kuwait National Petroleum Co. (KNPC) experienced in facing the decline in gasoline demand, as well as the adaptations made to operating units to meet the COVID19 pandemic's impact and improve gasoline production.
The decline in gasoline demand coincided with KNPC's Clean Fuels Project (CFP) commissioning, which included new gasoline producing units. The following is a detailed description of the impact the epidemic had on gasoline demand in Kuwait, and how KNPC turned from gasoline importer to an exporter of this commodity.

KNPC: An Introduction on gasoline

KNPC two refineries, Mina Abdullah (MAB) and Mina Al-Ahmadi (MAA), have modern units to produce clean-burning fuels conforming to Euro-V standards.

Prior to the completion of the CFP, the MAA was the only gasoline producer in Kuwait, with three different octane grades: UL-91, UL-95 and UL-98. Various refinery streams are blended to produce gasoline in different grades. The primary streams are:

- Reformate produced in catalytic reforming Units 1 and 2 of the MAA refinery. These units produce a high-octane rating and high aromatic content.
- Catalytic-cracked light naphtha and heavy naphtha produced in an FCCU, with a moderate octane rating (with a sweetening treatment unit), high olefins (alkene) content and moderate aromatics levels.
- Hydrocracker light naphtha produced in a hydrocracking unit, with a medium-to-low octane rating and low aromatic levels.
- Coker light naphtha that is imported from the MAB refinery. It has a lowoctane rating, high olefin content and a low aromatics level.
- Straight-run naphtha directly from crude oil Units 1, 2 and


Dalal Alqallaf

- 3. It has a low octane rating, low aromatic content and no olefins (alkenes), which feed to the reforming units.
- Alkylate produced in an alkylation unit, with a high octane rating, which is pure paraffin.
- Methyl tert-butyl ether (MTBE) is added to gasoline to increase its octane rating and help prevent engine knocking. In addition, oxygen present in MTBE helps gasoline burn more completely—the MTBE volume limit is 10%.

High-octane reformate is the main gasoline component, which is produced in two identical platforming processes. Each process train consists of a naphtha hydrotreater (NHT), a naphtha splitter and a continuous catalyst regenerator (CCR) platformer. The aim of the complex is to produce reformate with a research octane number (RON) of 102, a motor octane number (MON) of 90.06 and a maximum benzene content of 1 vol%. The primary gasoline stream qualities are shown in TABLES 1 and 2.

Each platformer train has a capacity of 18,000 bpd. The CCR's throughputs can be varied depending on gasoline demand. Prior to 2001, gasoline requirements were normally met by one reformer; however, the second reformer's operations may be needed as per KPC's reformate export requirements directive or in the case of FCCU shutdown (FIG. 1). Post-2003, domestic demand increased steadily until it surpassed the nation's refineries' production rate, leading to continuous gasoline and MTBE imports until the commissioning of the CFP in 2020.

FCC light and heavy gasoline are products from the FCCU—also referred to as FCC light and heavy naphtha. The 40,000-bpd FCCU has two streams that are treated separately in merox treatment units to remove mercaptans. Both streams have a moderate octane rating of 92 RON and 93.1 RON, respectively, with high olefins (alkene) content and moderate aromatics levels. Although both components are excellent in gasoline blending, FCC heavy naphtha blending is limited due to its high density and exceptionally low distillation recovery.

Alkylate is produced in the alkylation unit, which is part of

TABLE 1. Gasoline stream qualities

UNIT	STREAM	Specific gravity	Research octane number (RON)	Motor octane number (MON)	Reid vapor pressure (RVP)	70°C	100°C	120°C
Continuous catalytic reformer (CCR) Units 1/2	Reformate	0.8	101	85	7.4	7%	23%	47%
FCCU	FCC light naphtha	0.705	92	80	9.8	48%	75%	87%
	FCC heavy naphtha	0.846	93.1	82	1.3	0%	0%	0%
Alkylate	Alkylate	0.695	93.3	93	9	8%	30%	100%
Methyl tert-butyl ether (MTBE)	MTBE	0.745	111.4	99	9.9	96%	100%	100%
Coker (MAB refinery)	Light naphtha	0.673	71	66	10.9	20%	62%	80%
Hydrocracker	Hydrocracker light naphtha	0.6758	74.2	70.4	-	79	100	-
Naphtha complex	Light straight-run naphtha	0.665	66	60	-	85	100	-

TABLE 2. Gasoline stream qualities

Gasoline components		Properties	
Unit	**Stream**	**Aromatics**	**Olefins**
CCR Units 1/2	Reformate	80	0
FCCU	FCC light naphtha	12.9	35.8
	FCC heavy naphtha	70	14.5
Alkylation	Alkylate	0	0
MTBE	MTBE	0	0
Coker (from the MAB refinery)	Light naphtha	6	20
Hydrocracker	Hydrocracker light naphtha	1.3	0
Crude distillation Units 3, 4, 5	Light, straight-run naphtha	1.3	0

the MAFP (MTBE, alkylation, FCC) block. The unreacted C4 raffinate from the MTBE unit is fed to the alkylation unit to produce alkylate–the production rate is 3,900 bpd. Alkylate is produced by reacting isobutene and light olefins in the presence of sulfuric acid catalysts. The alkylate product has a boiling range of gasoline, a high-octane rating of up to 97 RON and is an excellent blending component in gasoline.

MTBE is produced in the MTBE unit, which is part of the MAFP block. The feed is the total C4s bottom stream of the C3/C4 splitter in the FCC liquefied petroleum gas (LPG) splitter unit. This unit converts most of the isobutene in the C4 stream into MTBE. This is achieved by reacting methanol and isobutene in the presence of an ion exchange resin-type catalyst. MTBE is used in gasoline as an additive to enhance octane rating and improve combustion.

Gasoline components are blended to meet product specification requirements, which have winter and summer specs. Typical product specifications are summarized in TABLE 3.

Post-CFP gasoline production

KNPC's CFP involved the upgrade and integration of the MAB and MAA refineries, and the closure of the SHU refinery. The project increased the combined capacity of the refineries from 736,000 bpd to 800,000 bpd and lowered the sulfur content of petroleum products to 10 parts per million (ppm). Additionally, certain facilities at the neighboring SHU refinery were renovated as part of the project. Selected offsite facilities of the refinery-including storage, blending and shipping/logistics-were integrated with MAA and MAB refinery operations.

After the CFP's completion, the country was able to produce ultra-low sulfur gasoline. The existing gasoline blending facilities were upgraded to adhere to KPC's Euro-V specification and production was increased. Different streams were included in the project to reduce aromatic content and limit the presence of benzene. The following blend components were added to produce different grades of gasoline:

CCR Units 1 and 2 and an LPG treating unit to remove chlorides and hydrogen saturation of olefins in the combined LPG stream from the debutanizer of the platforming unit. Its main purpose is to reduce reformate Reid vapor pressure (RVP) from 6 psi to 2.5 psi to allow high-octane isomerate and isopentane blending, which is essential to meet the gasoline RVP specs, especially in the summer.

An FCC NHT unit to selectively desulfurize cracked naphtha to a maximum of 10 ppm sulfur in each of the two FCCU streams.

A deisopentanizer unit to separate isopentane from the natural gasoline stream via a fractionation column and provide sufficient isopentane to meet gasoline-blending requirements. The unit's primary aim is to upgrade the isopentane to gasoline rather

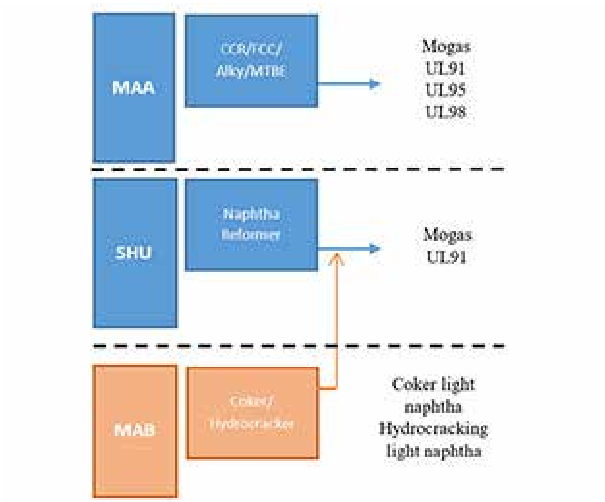


FIG. 1. Gasoline production among KNPC's refineries.

TABLE 3. Gasoline specifications

Motor gasoline Unleaded 91, 95, 98 RON	Limits: Pre-CFP	Limits: Post-CFP
Composition		
Doctor test	-ve	-
Aromatics, vol%	-	35
Benzene, vol%	4	1
Total sulfur, mass%	0.05	0.001
Volatility		
Vapor pressure at 37.8, kpa (psi)	70 (10.1) W 62 (9) S	70 (10.1) W 62 (9) S

than LPG, thus increasing gasoline production and adhering to aromatics limits.

An isopentanizer merox unit to remove mercaptans from the isopentane stream to meet the CFP sulfur specification of 10 ppm.

A deisobutanizer unit to provide incremental isobutane required for the alkylation revamp for a higher capacity. Feed to the deisobutanizer unit is field butane produced by the LPG trains in the MAA refinery.

An isomerization unit to produce isomerate (RON 87.5) from light naphtha for mogas blending to increase gasoline production and meet the aromatics limit.

Light naphtha from the MAB's hydrocracker is transferred to the MAA refinery's isomerization unit to increase gasoline production.

The MAB refinery's CCR unit was designed to produce reformate with a RON of 102 to upgrade naphtha and increase gasoline production.

A new 230,000-bpd mogas blender was built to meet the higher demand. The existing 50,000-bpd gasoline blender was modified to provide new components required for the blend recipes. Both blenders can be operated simultaneously.

The gasoline streams' properties post-CFP are detailed in TABLE 4.

COVID-19 and KNPC's refineries shift in demand and prices

The COVID-19 pandemic was

TABLE 4. Post-CFP gasoline streams' properties

Unit	Stream	Specific gravity	RON	Sulfur	RVP
MAB refinery's CCR	Reformate	0.8	102	0	2.5
FCCU	FCC light naphtha	0.705	91	10	10.5
	FCC heavy naphtha	0.846	92.1	10	3.5
Deisopentanizer	Isopentane	-	81	0	20
Isomerization	Isomerate	-	87	0	12.5

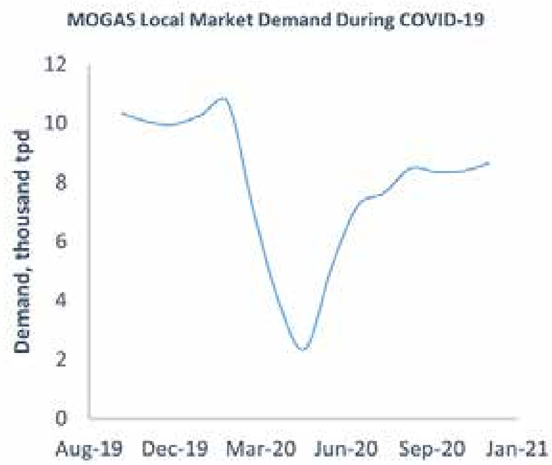


FIG. 2. Kuwait domestic mogas (gasoline) demand during the COVID-19 pandemic.

Faisal Al-Mutairi

Mohammad Matar

Ibrahim Al-Esmail

a global crisis, with nations around the world experiencing unprecedented levels of disruptions. The oil and gas industry was significantly affected by the pandemic. Operations were interrupted, and production reduced significantly due to workforce movement curfews, health regulations, and decreased fuel and petrochemicals demand. Several companies set up systems to protect their workers and perform their operations safely to mitigate the crisis. Efforts to contain the virus disrupted the global workforce and material supply chains.

KNPC, like other businesses, was not immune to the challenges brought on by the effects of COVID-19. Travel restrictions and lockdowns profoundly affected Kuwait's domestic gasoline demand, which constrained refinery operations to process crude oil at maximum capacities.

Oil prices dropped drastically in March and April 2020 due to plummeting demand, rising crude oil supplies and diminished storage capacities. It caused such a pronounced crude petroleum price drop that, on April 20th, crude petroleum traded at a negative price in the intraday futures market. Producer prices for crude petroleum declined 34% and 48.8% in March 2020 and April 2020, respectively.

During the COVID-19 pandemic, the global gasoline demand experienced a significant decline, as well. As a result, gasoline production also declined in most countries. When partial curfews were implemented in Kuwait, gasoline consumption fell to record lows (FIG. 2). In turn, this affected the gasoline production units at KNPC's refineries. Domestic gasoline demand declined from pre-pandemic levels of 10,000 tpd

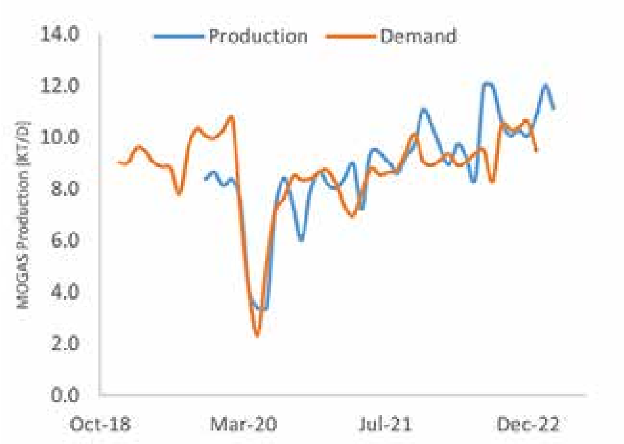


FIG. 3. Kuwaiti gasoline demand and production, October 2018-December 2022.

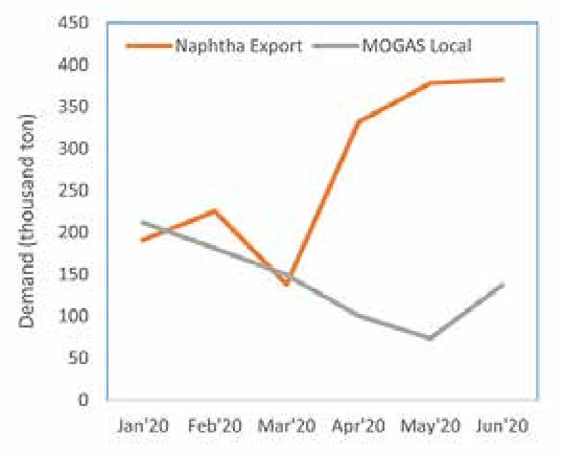


FIG. 4. KNPC's naphtha and gasoline production, January 2020-June 2020.

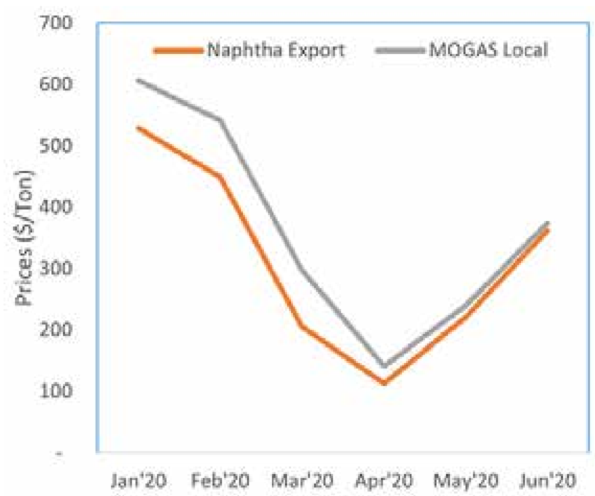


FIG. 5. KNPC naphtha and gasoline prices, June 2020-June 2020.

to 4,000 tpd and reached a low of 2,400 tpd in May 2020 (FIG. 3). From May 2020-August 2020, all gasoline imports ceased. FCCUs operated at turndown capacity, and one of the naphtha reformers was kept idle. Consequently, naphtha exports increased (FIG. 4). Furthermore, naphtha and gasoline prices fell dramatically, with margins reaching a record low in April 2020 (FIG. 5). In view of decreased domestic demand and no export options available, the MAA refinery was forced to lower utilization or shut down their units to match the diminished demand.

Unit operations have played a significant role in managing demand crises. Units such as the naphtha reformer and FCCU have operational flexibility in terms of unit throughput and were kept at turndown capacity. However, after the lifting of COVID-19 restrictions, Kuwait's domestic gasoline demand rose steadily and KNPC increased production to satisfy this demand. Although gasoline imports were needed to bridge the nation's supply and demand gap, the completion of the CFP enabled KNPC to increase production capacity by 40%. This increase—from the addition of a CCR unit at the MAB refinery and the utilization of isomerzation, alkylation and de-isopentanizer units—played a significant role in providing the flexibility to meet local demand and export excess production.

However, gasoline exports were not attractive due to a low naphtha-gasoline margin; therefore, post-pandemic plans focused solely on meeting domestic demand. These economics changed in 2022 due to worldwide political events and sanctions on various nations, which boosted refiners' margins to record highs.

Throughout 2022, KNPC managed to satisfy local market demands and export excess quantities of gasoline. The exported cargoes—un-oxygenated grades and 92 octane ratings—were sent to European markets. Since August 2022, several gasoline cargoes have been successfully exported. The primary driver was the substantial difference in the naphtha-gasoline margin, which was an attractive market fundamental.

Takeaway

The COVID-19 pandemic had a tremendous impact on the world economy, especially in the disruption of supply chains and global fuels demand. Refineries suffered challenges in maintaining operations due to movement restrictions and drastic changes in demand patterns. The reduction in demand necessitated more flexibility in refinery operations.

The drastic change in the world's economic environment led to financial losses and several refineries were forced to shut operations. The lesson learned is this: refineries must have plans, strategies and built-in flexibility in place to hedge against unfavorable circumstances and to capture favorable opportunities. HP

Acknowledgment

The authors would like to recognize the dedication and challenging work performed by experienced KNPC staff for the growth of the company to enhance the overall profitability of KNPC's refineries.

FCC Unit (86) - MAA

HIGH CORROSION RATE IN GAS SWEETENING PLANT – CASE STUDY

The COVID-19 pandemic is regarded as one of the most exceedingly difficult challenges faced in the oil and gas industry's history. Refineries around the world encountered daunting periods in operating their units at their turndown capacities and were forced to temporarily shut down units at their facilities.

Abdullateef Alfares, *Eng. Corrosion & Metallorogy*
Laxma Reddy, *Eng. Corrosion & Metallorogy*
Khaled Al-Mutairi, *Snr. Inspection Engineer*
Quality Assurance Dept., MAA

Abstract

Acid gas removal plants commonly use the concept of amine treatment to remove Hydrogen Sulfide (H_2S) and Carbon Dioxide (CO_2) from natural gas. This method is also applied to remove H_2S from refinery off gas.

Depending on feed conditions, for all practical purposes, the most economical material is carbon steel with the exception of stainless steel cladding in areas of high severity of corrosion. Increased throughputs and deterioration of feed quality, which are common over the years across the oil and gas industry, will likely yield abnormal corrosion rates and severe fouling of plant equipment and piping. Such changes not only reduce the plant reliability but also increase operating and maintenance costs, and often cause safety and environmental concerns.

During the past turnarounds of the acid gas removal plant, a high corrosion rate was observed in most of the equipment and piping. The degree of severity was such that some equipment leaked within 10 years of commissioning.

The subject paper demonstrates details of salient observations during the turnarounds, the remedial short-term measures, and the long-term approach with respect to material up-gradation and design changes. The paper also highlights the necessary improvements foreseen in material selection and design for new projects keeping into consideration the life cycle cost of the equipment, in addition to reliable and safe operation of the plant.

Introduction

Sour gas treatment in the oil and gas industries is getting more complex to meet stringent gas specification and emissions requirements established by environmental regulatory agencies. Furthermore, increase in demand for utilizing new wells of varying sources of sour gas composition is directing the industry to fine-tune preconditioning in the upstream and gas treatment in the downstream segments.

The wet chemical absorption (solvent based) acid gas removal remains the most cost-effective for natural gas or refinery off gas treatment. Various types of

Khaled Al-Mutairi

Abdullateef Al-Fares

Laxma Reddy

water-soluble amines have been used for treatment of process streams. The most commonly used amines are aqueous solutions of monoethanolamine (MEA) and diethanolamine (DEA). Other amines, such as methyldiethanolamine (MDEA), di-isopropanolamine (DIPA), and diglycolamine (DGA), are also used in various treatment processes.

Comparison is made between two gas-treating plants of varied acid gas loadings, and highlighted corrosion problems. The focus of this paper is to consider the life cycle cost of the plant taking into account the production losses, maintenance costs, reliability and versatility, instead of economic selection of material in the project inception. Corrosion issues shall be mitigated at the design stage instead of combating while in operation.

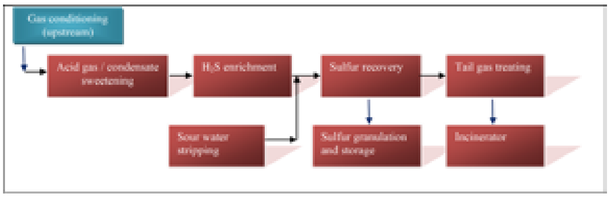


Figure 1. Gas treating facilities

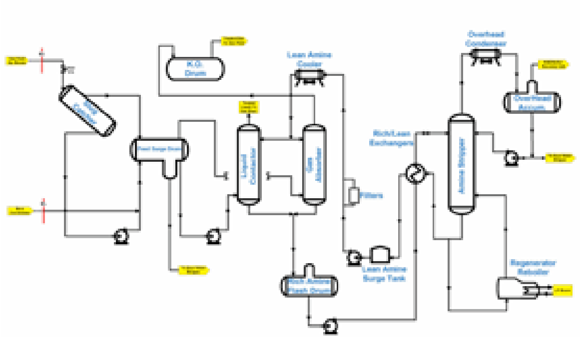


Figure 2. Process flow diagram of amine treating

Composition	Sour gas (Mol%)	Sour Condensate (Mol%)	Refinery off gas (Mol%)
H2O	0.02	0.03	0.02
H2S	2.50	2.50	10.5
CO2	0.80	5.00	1.38
O2	0.02	0.00	0.00
N2	0.51	0.08	0.00
C1	55.32	20.42	66.00
C2	18.27	19.15	6.40
C3	10.75	24.08	2.21
IC4	0.88	3.34	1.64
NC4	2.65	12.79	0.36
IC5	0.43	3.67	0.81
NC5	0.58	6.12	0.00
C6+	0.08	2.83	0.72
H2	-	-	13.3
Total	100.00	100.00	100

Table 1. Feed composition specification

Plant description

The acid gas removal plant consists of a feed handling unit, an acid gas/ condensate sweetening unit, an acid gas enrichment unit, a sulfur recovery unit, a tail gas treating unit, and a thermal incinerator as shown in block diagram Figure 1. [1] In case of refinery off gas, it is limited to acid gas H2S removal from refinery off gas.

Refer figure 2 is a process flow diagram for main amine treating section of the plant. The gas or liquid streams containing one or both of the acidic components are fed to the bottom of a gas-absorber tower or a liquid-contactor vessel, respectively. The lean (regenerated) amine solution flows counter to the feed stream in the tower and absorbs the acidic components during the process. The purified gas or liquid stream passes to the overhead system. The rich amine solution is fed to a regenerator (stripper) tower, where the acidic components are removed by pressure reduction and by the heat supplied from a re-boiler. The acidic components are removed from the overhead and sent to the H2S enrichment unit, the sulfur removal unit, tail gas recovery unit and finally to the incinerator. The lean amine solution that leaves the bottom of the regenerator is returned to the absorber or contactor to be used again for purification of the hydrocarbon streams.

The acid gas removal plants with their capacity, acid gas loading, filtration capacity and material of construction are tabulated in Table2.

The feed compositions of sour gas, condensate and refinery off gas are listed below in Table 1.

Observations & findings

Visual Examination

Salient inspection observations and findings discussed below:

Feed Surge Drum: The drum was internally inspected during the last four major overhauls. Every time, a significant quantity of sludge was found. The sludge was analyzed and the details were reported in Table 3. Severe pitting was observed towards the bottom as shown in figure 3a. Cold repairs were carried out in previous major turnarounds. During recent turnaround, thickness loss was beyond the corrosion allowance and hence it was recommended for weld repair. While carrying out preheat and hydrogen bake-out, a crack was observed at a heavily corroded area due to suspected hydrogen induced cracking (HIC), though HIC resistant shell material was used, as shown in Figure 3b. Wet fluorescent magnetic particle and Phased Array were carried out. The cracks were arrested and repaired by weld buildup. Stress relieving was done at repaired locations.

DEA Stripper tower: The vessel shell was originally cladded with UNS S30403 (SS304L) on the top (above feed nozzle) and no corrosion/ fouling was observed in this section. Below the SS cladding, the carbon steel portion was found to have localized corrosion and roughening up to 5mm metal loss in approx. 10 years of service, against 6.4 mm corrosion allowance as shown in Figure 4. Significant amount of scale and sludge were observed in the vessel bottom which usually takes a few days to gas freeing, cleaning and entry.

Flash corrosion: [2] higher corrosion rate with increase in temperature, particularly in rich amine service, was observed in the shells of the rich amine exchangers. The associated piping up to pressure reducing valve upstream of regenerator was subjected to acid gas flashing. This resulting in severe localized corrosion and leaks commonly found in the piping as shown Figure 5 since the pressure drop is high enough as the process stream reaching the regenerator. Only downstream of the pressure control valve piping is provided with corrosion resistant material, UNS S30403 and such spools were found in good condition

AGE Absorber tower: The vessel shell was originally cladded with SS304L at the bottom section (below the feed nozzle) and no corrosion/ fouling was observed in this section. Above the SS clad, the carbon steel portion was found to have general corrosion

Plant No.	1	2	3
Commissioned	1999	1999	1984
Capacity			
(MMSFD) (1)	145	145	8.6
Type of amine (strength)	DEA (30%)	MDEA (30%)	DIPA (30%)
Acid gas loading	0.4	0.4	0.75
Selective absorption	H2S + CO2	H2S	H2S
Side stream filter capacity	30%	30%	15%
Corrosion inhibitor (Y/N)	No	No	Yes
Absorber MOC(2)	CS + UNS(3) S30403 clad	CS + 3.2 CA(4)	CS + 3.2 CA
Regenerator MOC	CS + UNS S30403 Clad (top)	CS + UNS S30403 Clad	CS + 6.4 CA

(1)MMSFD - Million Metric Standard Cubic Feet per Day
(2)MOC - Material of Construction
(3)UNS - Unified Numbering System for materials
(4)CA - Corrosion Allowance
Table 2. Process Design Conditions

Test	Units	Results
Loss on Drying @ 105oC	wt%	41.4
Loss on Drying @ 600oC	wt%	18.9
pH of 5% solution	-	5
Sulfide as S	wt%	15.5
Iron as Fe	wt%	35.6
Carbon as C	wt%	0.9
Heavy metals (Hg, V)	ppm	0.00

Table 3. Chemical Analysis of Sludge (Feed Surge Drum)

and roughening up to the 4mm metal loss in approx. 10 years of service against 6.4 mm corrosion allowance as shown in Figure 6. To mitigate such heavy corrosion, it was planned to go for UNS S30403 strip lining in the next planned overhaul.

Lean/Rich Amine Exchangers: Corrosion in addition to multiple cracks were observed in rich amine circuit as shown in Figure 7. The

Figure 3b. Crack in the corroded area after Hydrogen bake out and prior to drum repair

Figure 4. DEA Stripper tower interface between SS clad and CS. 4~5 mm metal loss at the interface due to galvanic effect and erosion - corrosion

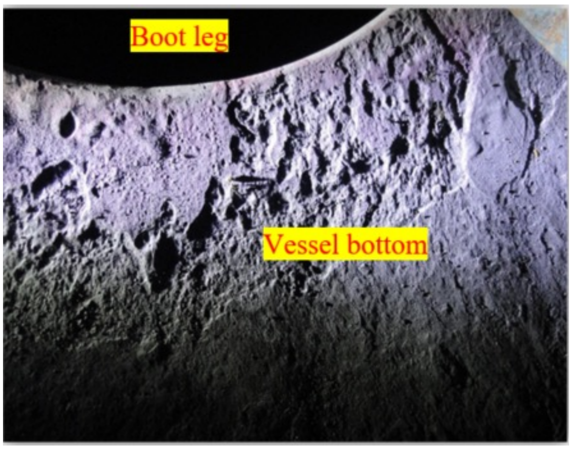


Figure 3a. Feed surge drum (severe corrosion and grooving 4 to 8 O'clock position)

Figure 5. Flash corrosion in the CS piping downstream of the rich amine exchanger to stripper

Figure 6. Acid gas enrich absorber Vessel shell deep pitting

Figure 7. Cracks in shell material (CS) of a Lean/ Rick Amine Exchanger

cracks were found in post weld heat treated pressure vessel quality shell material (carbon steel). Localized corrosion can be seen on tube sheets and tube-tube sheet in rich amine side.

Piping: Several leaks were observed in associated piping in the unit due to internal corrosion as shown in Figure 9,10 and 11.

Material Design: Economic design and minimizing project capital cost often end up with higher corrosion rate. Here are two examples. First, an exchanger cage was made of CS material and the tubes of SS. There was a galvanic corrosion, which was compounded by high aspect ratio of anode to cathode surface areas as shown in Figure 14. Second, an expansion bellow was made of UNS S31603 downstream of incinerator. Sulfurous flue gases are highly corrosive and often perforations were observed as shown in Figure15.

Figure 8. Tube sheet and tube end corrosion

Figure 9. Internal corrosion of piping

Figure 10. Waste heat boiler tube failure by external acidic gases in Incinerator section

Intermittent Operation: Acid gas plants depend on upstream feed. Any upset or leaks, will compel to have plant shutdown for short periods. During such unplanned shutdowns acid condensation takes place. The plant suffered higher corrosion rate during idling than that of normal operation. Highlighted are two such corrosion problems and leaks as shown in Figure 16a and 16b.

Discussion

Corrosion rates vary from plant to plant since the control parameters are many, those include feed composition, particulate matter, type of amine and its strength, acid gas loading (mole acid gas/mole active amine), heat stable amine salts (HSAS), velocity and material of construction. Estimated amine corrosion rate of carbon steel and stainless steels with different combination of parameters were well documented in API-581(2)[3] - Base Resource Document of Risk Based Inspection (RBI). However, ideal operating conditions and continuous on-stream of plant will follow such corrosion rates. Any parameter variation will have significant impact on corrosion, especially on carbon steel material.

In the subject discussion, the plant was forced to shut down several times due to feed supply related issues, operational upsets, frequent leaks in the plant (from slug catcher up to incinerator). Eventually, all these stoppages had a great impact on the corrosion rates, in addition to loss in production as well as flaring of feed gas in the upstream, which is an environmental concern.

Figure 11. Drain boot in the pit. Internal corrosion (over 50% metal loss) in boot

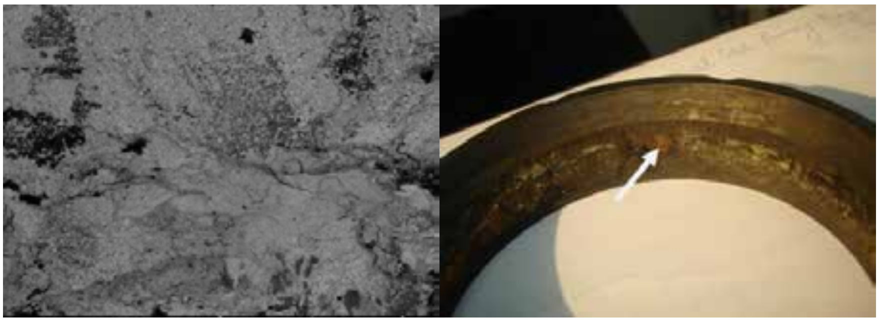


Figure 12. Frequent mechanical seal failure of ADIP pumps. SEM photograph of scoring of FeS particulate on the seal (below)

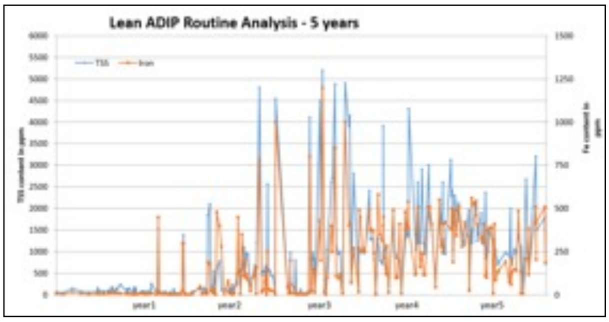


Figure 13. Iron (Fe) and Total suspended solids (TSS) trend of Plant 3

Figure 14 a &b. Perforation of tie-rod and sleeves (galvanic corrosion). Localized corrosion and enlargement of baffle plate holes

Over the years, maintenance costs (manpower, materials and equipment) have been phenomenally increasing, while emphasizes on safe work practices. During the project construction, invariably, impetus made to cost-effective plant design rather than unforeseen difficulties arise in operation and maintenance. [4]

It is evident that every time the gas sweetening plant is shut down for a week, its production losses wipe out all project cost savings. Moreover, environmental concerns arise by flaring the gas.

While designing the plant, it is equally important to consider life cycle cost of plant. A little additional initial capital cost, will have leverage or insure to have leak-free and reliable operation. Oversizing of feed and amine solution filters can effectively handle the plant upset conditions. UNS S30403 clad is another viable option in vessels (feed drum, absorber, and stripper), Lean/Rich amine exchanger circuit including its piping up to stripper. Moreover, good operating practices put corrosion at check.

Conclusions

Based on the observations, findings, analyses, the following are concluded:

1. Particulate matter in feed was reportedly high and getting accumulated in feed surge drum.

2. Corrosion rates are exceptionally high in surge drum, knock out drums, and quench vessels, absorber, stripper towers, lean/rich amine exchangers and its piping.

3. Fe and TSS levels were high in acid loading amine stream

during plant upset condition resulting in seal leaks and thus amine loss.

4. Frequent upset / leaks resulting in low reliability of the plant.

Recommendations

The following are short and long term recommendations in order to improve the plant reliability and safe operation:

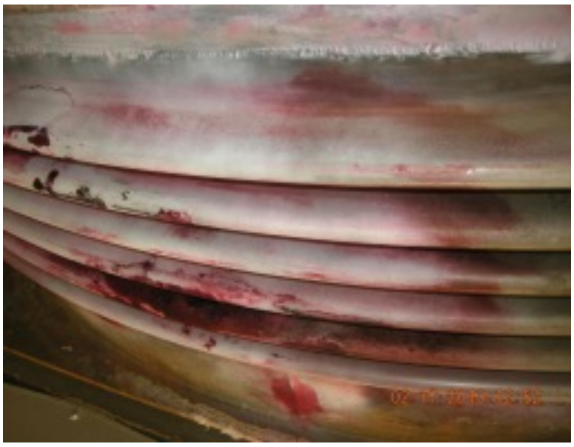


Figure 15. Perforations on expansion bellow (UNS S31603) connected to incinerator.

Figure 16a. Perforation of sour water pump

Figure 16b. Perforation quench column bottom sour water nozzle

Short Term Recommendations

1. Optimize maintenance/ inspection interval to clean the fouled system and make necessary repairs/ replacement.

2. Protective linings and coatings are applied wherever applicable.

3. Oxygen ingress into amine tanks is minimized by nitrogen blanketing.

4. Using mobile amine filters to reduce load during upset conditions to bring down Fe and TSS levels within acceptable range.

Long term recommendations

1. UNS S30403 clad shall be considered for feed surge drum, absorber, and stripper tower and lean/rich exchangers to reduce fouling and under deposit corrosion.

2. Use UNS S30403 in lieu of carbon steel piping where flash corrosion is expected between rich amine exchangers and regenerator tower.

3. Exchanger tube bundle cage shall be of the same metallurgy as the tubes in lean/rich amine exchangers to avoid galvanic corrosion.

4. Physical filtration shall be increased from 30% up to 50% for amine system to take in to consideration of upset operating conditions and unforeseen feed impurity levels.

5. Plant redundancy or two trains will help to minimize

sour gas flaring which has increased concern on environment.

6. Follow operations best practices to minimize amine corrosion problems and plant shutdowns.

Acknowledgements

The authors wish to thank the management of Kuwait National Petroleum Company for their support in presenting this paper.

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LPG Train 4

HIGH PRESSURE DROP OF COLDBOX IN LPG TRAIN-4 CAUSES AND MITIGATION

High pressure drop was detected through the LPG Train - 4 Cold Box. Many attempts have been taken to de-dust the system. However, these efforts have only worked for a short period of time until total plant shutdown occurred

***Sr. Eng. Anwar Behbehani**, Maintenance Department -MAA*
***Eng. Mahmoud Mukamis**, Maintenance Department -MAA*

Abstract

The Cold Box was obtaining pressure because of passing foreign particles through the Dry Gas Filter as a result of the damaged inner sealing ring. Furthermore, the misalignment and deformation of the partition ring caused the Filter to semi-chock. A new partition ring including a sealing Vegetable Fibre Gasket was fabricated by taking actual measurement of the outer diameter of the internal bracket in order to seal the gap between the filter elements. Also, the filter elements were replaced by new ones.

Introduction

The design of the LPG plants in Mina Al Ahmadi are intended for the treatment of gases produced in Kuwait Oil Company (KOC). The collected gases from the gathering centers are processed by cooling, condensation, and fractionation to produce LPG products (Propane, Butane, Pentane, Natural gasoline). The feed streams are received in feed accumulators of common plant from where the gas and liquid are distributed to three parallel identical extraction trains (Train-1, Train-2, and Train-3). Products from the unit are sent to LPG storage for storage and shipping. The newly commissioned LPG Train-4 gas plant is able to produce ethane, propane and butane by treating about 805 MMSCFD (Million standard cubic feet per day) gases and 106.3 MBPD (Thousand barrel per day) condensates in the Mina Al Ahmadi oil refinery plant complex located 40km south of Kuwait City. Feed Gas supplied to Fourth Gas Train facilities consist of mixture of Associated Gas and Condensate from the KOC Gathering Centers Southeast Kuwait and North Kuwait oilfields. In addition, existing KNPC Refinery Gases from the Mina Al-Ahmadi AGRP (Acid Gas Removal Plant) are supplied to the Fourth Gas Plant facilities [1].

During the dehydration process the dehydrator beds are saturated with moisture and hence has to be regenerated. Separate regeneration system with dryer regeneration gas compressor and dryer regeneration gas heater is provided. Separate and dried intermediate products from this unit is the feed for the downstream units, which includes Dried Feed Gas, Dried Condensate and Dried LPG to NGL Recovery Unit. Spent Regeneration Gas generated from the dehydrator regeneration system is sent to Residue Gas header. [2] An important component in gas separation plants are the so-called "Cold Boxes". Cold boxes are (pressure) vessels that hold a gas or liquid at a very low temperature. The distinctive feature of cold boxes is the double-wall construction, which allows the insulation to be fitted between the inner and outer walls. In July 2016, after one-year of commissioning LPG Train-4, high Delta P was recorded across the Cold Box. At the time it was believed that the Cold Box strainers could have been filled with foreign particles which resulted in the high Delta P scenario. The Stationary Maintenance was requested by Operations to de-dust the system. After line-breaking the strainers, dust and particles were indeed spotted. The strainers were cleaned by Nitrogen gas and fixed back for operation. However, the de-dusting process has only worked for a period of three months where the system was running normal until high Delta P was recorded again. Although the performance of the Cold Box with respect to heat transfer is found satisfactory, but it is higher than the designed pressure differential of 0.5 bar. [2] An investigation team was formed to identify the root cause and propose necessary recommendations for the repeating incident.

The process

The Cold Box located in Train-4 consists of a rectangular carbon steel casing that supports and houses heat exchangers, piping, and insulation material in an inert atmosphere. The cold box is sealed after the insulation has been fitted, so the insulation can no longer come into contact with, for example, water, snow, dust and

Anwar Behbehani

contaminants. Figure (1) shows the Cold Box in Train-4.

The process starts out by the direct feed gas coming from the Gas Compressor. The feed gases first go through the Feed Gas Filter where it is screened for any dust or particles before it makes its way through the Gas Dryers. The existing flowing gas contains amount of water vapor. Water vapor must be removed before it undergoes cryogenic fractionation to avoid formation of hydrates and ice. It is not uncommon to see hydrates obstructing flow through the Heat Exchangers of the Cold Box or the strainers that protect the heat exchanger from construction dirt and debris. Hydrates can cause enough pressure drop to rip apart strainers, allowing dirt and debris to enter and damage the downstream heat exchanger. Hydrates may also form on the cold plant fractionation column trays and packing. The result is a decrease in efficiency causing low product recovery and potentially off-specification liquid product. Hydrates are also known to plug control valves and plant instrumentation. The gas is compressed and then cooled to remove as much entrained water vapor as possible before it is sent to molecular sieve dryers. [3] There are four Gas Dryers in LPG Train-4. The dryers remove the moisture content from the Feed Gas Compressor by means of catalyst reaction with the process gas. Multiple molecular sieve dryer vessels are typically operated in parallel with a piping system that allows a saturated adsorbent bed to be taken off line for regeneration with heated gas. After the moisture removal is completed the product is filtered furthermore through the Dry Gas Filter before entering the Mercury Guard Bed. Almost all hydrocarbons contain mercury. In the case of natural gas and natural gas liquids it is likely to be present as elemental mercury. In the case of crude oil, it may also be present as organo-metallic and ionic mercury. [4] The Bed is a very important aspect in the process since it eliminates any Mercury content that might exist in the processed gas. The Cold Box heat exchangers are composed of Aluminum and Mercury is known to damage aluminum heat exchangers to the point of catastrophic failure, therefore, its extraction is important. When mercury reacts with aluminum, it forms an amalgam (a mercury alloy). Although Aluminum is normally protected by a thick oxide layer, but the formation of the amalgam disrupts it. It allows fresh Aluminum to react with air to form white Aluminum oxide. After the Mercury Guard Bed, the processed gas is further screened by the After-Mercury Filter before entering the Cold Box strainers. The strainers filter out any additional particles that might exist in the pipelines before eventually entering the Cold Box. Figure (2) shows a schematic diagram of the process. It shows a simplified path where the feed gas coming from the Dry Feed Compressor goes through four different filtration processes before entering the Cold Box.

Root cause analysis

Upon investigation high Delta P was recorded in July 2016, October 2016, and January 2017. The trend suggests high differential pressure was recorded frequently every three months. The three incidents resulted in the unplanned shutdown of Train-4. Production loss was exceeding $100,000 USD and maintenance cost was nearing $10,000 USD. [2] At the time, investigation revealed that the Mercury Guard Bed was bypassed due to some sort of mechanical issue which have led its catalyst to escape through the pipelines. The internal grid holding tons of catalyst in place was assumed to be damaged. This meant that two important equipment (the Mercury Guard Bed and the After-Mercury Guard Filter) were no longer in-line with the system. In fact, the escaping catalyst have damaged the outlet Orbit Valves connecting the Bed to the pipelines. Huge quantities of catalyst were found choking the After-Mercury Bed Filter. The incident became obvious since the After-Mercury Guard Filter high Delta P alarm was first triggered. Until maintenance activities of the Mercury Guard Bed is completed,

LPG Train 4

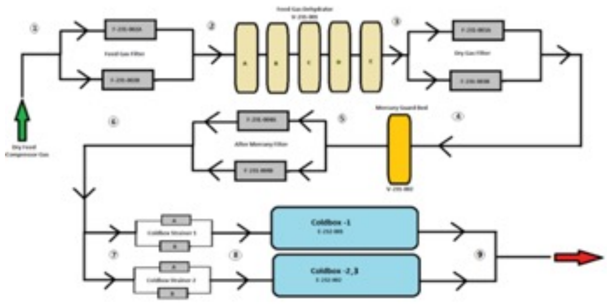

Figure 2: Schematic Diagram

risk was taken to bypass the system and continue operation. This meant that the overall system still has three out of four filtration processes active but could be risking any Mercury content to pass through the Cold Box. The Cold Box strainers were cleaned and put back to work. Everything seemed fine until high Delta P was recorded again through the Cold Box.

Tracing back the pipelines all the way to the Gas dryers, the Dry Gas Filter was the next suspect since high trends of Delta P was also recorded. Anything passing the finer mesh Dry Gas Filter will surely pass the Cold Box larger mesh strainer, therefore, the Maintenance Department was requested to line-break the Dry Gas Filter. Upon line-breaking, large amount of catalyst started to flow out the filter's channel head. The filter was indeed chocked with catalyst but this time it came from the connected Gas Dryers. Figure (3) shows the Dry Gas Filter after opening the channel head. Figure (4) shows the filter elements chocked with catalyst and ceramic balls.

Like the Mercury Guard Bed, one of the Gas Dryers internals was believed to be damaged allowing catalyst to escape. Even though the existing damage has affected the pipelines connecting all the way to the Cold Box, however, the Dry Gas Filter should have stopped any undesired particles to go through. So why were foreign particles passing the Filter? It was best practice to dismantle the Filter and inspect its internals. Conclusions came through when the sealing ring holding the filter elements in place was noticed to be damaged. Furthermore, the partition ring in figure (3) was observed to be misaligned. This made sense as to why particles were passing and suggestions came through to line-break the Cold Box strainers one more time for final inspection. Upon accessing and close examination, it was observed that the strainers' wrapping metal cloth is ripped off in some areas. The damaged metal cloth is definitely an access path for larger particles to pass through. The root cause of this dilemma is starting to become clear.

Although, the reason for such incident was becoming obvious, however, as to why was both of the Mercury Guard bed and Gas Dryers internal grids were damaged is a whole different story. Nevertheless, the Dry Gas filter did not do a good job by eliminating particles to pass.

Recommendations

After the root cause for such incident was addressed, the team came up with various solutions to solve the problem. First, the Dry Gas Filter and the After-Mercury Bed Filter were to be inspected and rectified for the disorientation/ damage of inner partition ring. The Stationary Maintenance Department certainly has followed this recommendation by line-breaking all four filters for final inspection and resolution. Luckily, one filter (Dry Gas Filter) was in fact damaged and repair was followed. The misalignment and deformation of the partition ring caused the Filter to semi-chock. A new partition ring including a sealing Vegetable Fibre Gasket was fabricated by taking actual measurement of the outer diameter of the internal bracket in order to seal the gap between the filter elements. Also, the existing filter elements were clog with catalyst and ceramic balls which meant it was only best practice to replace them with new ones for best efficiency. Second, it was recommended by UOP to perform

Feed Gas Filters & Dry Gas Filters

baroscopic inspection on the Drier Beds to ensure its internal mechanical integrity.

Conclusion

As can be seen it took a while to crack down the puzzle and figure out what was happening. But, all the solutions carried out were found to be successful with no high differential pressure recorded ever since in the Cold Box. After tracing down the pipelines from the Cold Box all the way back to the Gas Dryers, it was clear where the origin of the problem was from. The internal mechanical integrity of the Vessels was in doubt. The grid holding tons of catalyst in place has experienced some sort of impact/fatigue were a fault was present for catalyst to escape through the pipelines. Although it is still not clear if this impact/ fatigue occurred due to running operational error, design issues, or installation inaccuracy. However, it was obvious that the Dry Gas Filter did not do its intended job properly by allowing catalyst particles to escape. This defect is a result of the misalignment and deformation of the filter's inner partition ring. The incorrect position of this partition ring made a passage way for catalyst and particles to get away. It was believed this experience was a consequence of improper installation when the unit was first commissioned. To fully disclose this matter, an additional modification was applied. A sealing Vegetable Gasket was fabricated by acting actual measurements of the outer diameter of the internal bracket in order to seal the gap between the filter elements. The Vegetable Fiber gasket would act as an insulator sealing the minuscule gap between the partition and filter elements. Furthermore, the two filter elements found in the Dry Gas Filter were found to be chocked with catalyst and ceramic balls upon line-breaking and dismantling the filter. The old filter elements were replaced by new ones for better screening efficiency. Since the Mercury Bed was by-passed because of suspected internal mechanical fault, the After-Mercury Filters were also dismantled, inspected, and repaired. The same issue was repeated with the Gas Dryers. Although, the reasoning for such phenomena is still not clear. But as a Stationary Mechanical Maintenance perspective, the job is to repair and maintain the equipment. The $100,000 USD cost of lost production was resolved by a $5 USD Fiber Gasket. After maintenance activities were completed, the system was put back on line and was running normally. The only maintenance recorded ever since this incident were periodically Preventive Maintenance activities.

Mahmoud Mukamis

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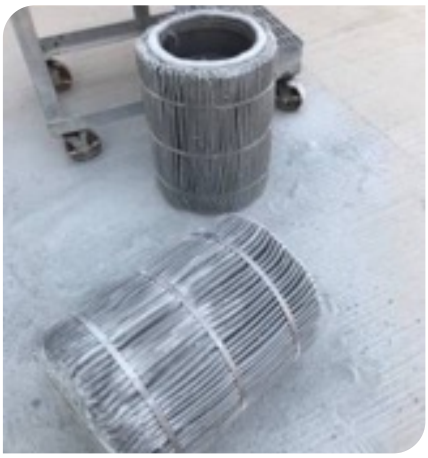

Mercury Guard Bed & After Mercury Filter

GUIDELINES FOR OPTIMIZING ISOMERISATION UNIT TO REDUCE ISOMERATE RVP MAINTAINING HIGH RON

This article explains how a lower RVP product can be achieved while maintaining a high RON and presents guidelines for other optimisers that share the same goal.

***Fatemah Jameel Fraidoun**, Process Eng. – Tech. Svcs. MAA*

Introduction

The ever-growing global demand for cleaner fuels is challenging refineries to increase the production of low-sulfur and aromatic gasoline components.[1] The light naphtha isomerisation process can reduce the aromatics from catalytic reforming products.[2] A crucial barrier is the highisomerate Reid vapour pressure (RVP), which affects the MOGAS pool blend. As a result, extra quantities of other higher research octane number (RON) and lower RVP gasoline streams are sacrificed, or product sales quality is jeopardised. Reducing isomerate RVP is accompanied by a loss in octane barrels. Previous research has focused on enhancing product yield or RON.

This article explains how a lower RVP product can be achieved while maintaining a high RON and presents guidelines for other optimisers that share the same goal. The subject of the case study presented in this article is a C5/C6 isomerisation unit at the Mina Al Ahmadi (MAA) refinery in Kuwait constructed of treating sections for hydrogen and light naphtha feed to remove sulfur, chloride, and moisture. The feed mixture is then heated sequentially by exchange with the effluent of two reactors operating in series. Raw isomerate is then sent to a stabiliser column for C4- components removal.

The stabiliser overhead gas is treated in a net gas scrubber for HCl neutralisation. The isomerate is routed to the deisohexaniser (DIH), in which further optimisation of the column's streams achieves the required product specification. The DIH overhead stream contains the main isomerate composition of normal pentane (nC5), iso-pentane (iC5), and dimethylbutanes (DMBs).

The bottom flow of DIH contains heavier C7+ components that contribute to lowering the final product RVP. Finally, a side cut from the column is recycled to the reactor with fresh feed. This stream contains normal hexane (nC6) and methylpentane (MPs), with RONs that can be further upgraded.

The author's strategy is:

- Firstly, to understand the effect of feed composition on unit operation and desired product quality.
- Secondly, to record and maintain the optimal reactor operating temperature at which desired conversions are achieved.
- Thirdly, study the stabiliser column behaviour and construct an approach to identify the changes in feed composition without sample analysis.
- Finally, understand the DIH operation and its effect on product quality.

It was concluded that the reactor feed composition affects the performance of all columns and, therefore, product quality. Consequently, the author presents an inclusive and systematic optimisation approach for feed composition and column operation.

Plan and Strategy

The trial aims to reduce isomerate RVP from 14.1 to below 12.8 psi while maintaining around 87 RON. The reduction in nC5 to iC5 conversion results in a lower presence of the higher RVP iC5 components in the final product.

Hence, the desired isomerate RVP is expected to be accompanied by a lower RON by at least two figures, i.e., 87 to 85. Therefore, familiarizing oneself with RON and RVP for each component, as shown in Table 1, is the first step towards achieving the desired product quality.

The isomerate composition is affected by several factors, such as operating capacity, fresh and combined feed composition, hydrogen purity, HLSV, reactor temperature, stabiliser and deisohexaniser column operation.

COMPONENT	RON	RVP (PSI)
IC4	100.2	71.9
NC4	95	51.5
IC5	93.5	18.93
NC5	61.7	14.42
CP	102.3	9.18
22DMB	94	9.13
23DMB	105	6.85
2MP	74.4	6.27
3MP	75.5	5.65
NC6	31	4.59
MCP	96	4.17
CH	84	3.02
BZ	230	2.98
C7+	82	2.1

Table 1. Reference for hydrocarbon components RON and RVP[3]

The unit operating capacity during the trial is 75% due to feed unavailability. Despite this, the author was able to achieve similar results at a higher operating capacity. The hydrogen purity and HLSV remained constant throughout the trial and will therefore be excluded from the analysis.

Feed Composition

Maintaining steady feed quantity and quality is the main factor affecting the results' consistency. The fresh feed and DIH recycled stream affect the quality of combined feed to the lead reactor. Therefore, comparing reactor feed composition with remaining factors constructs a representative analysis of feed quality impact on product quality. Generally, lowering C4- and C7+ components reduces isomerate RVP. However, a steep reduction in C7+ may lead to disturbances in the stabiliser, and their absence in the DIH bottom will eventually lead to an increase in isomerate RVP. For higher than 50% operating capacity, it is advisable to maintain 2.7 - 4.1mol% of C7+ components in the feed.

Three feed cases (A - C) with different compositions were selected for the trial, where feed cases A and B have similar C7+ components and case A contains more C5s. On the other hand, cases B and C have similar C5s, and the second is lower in C7+.

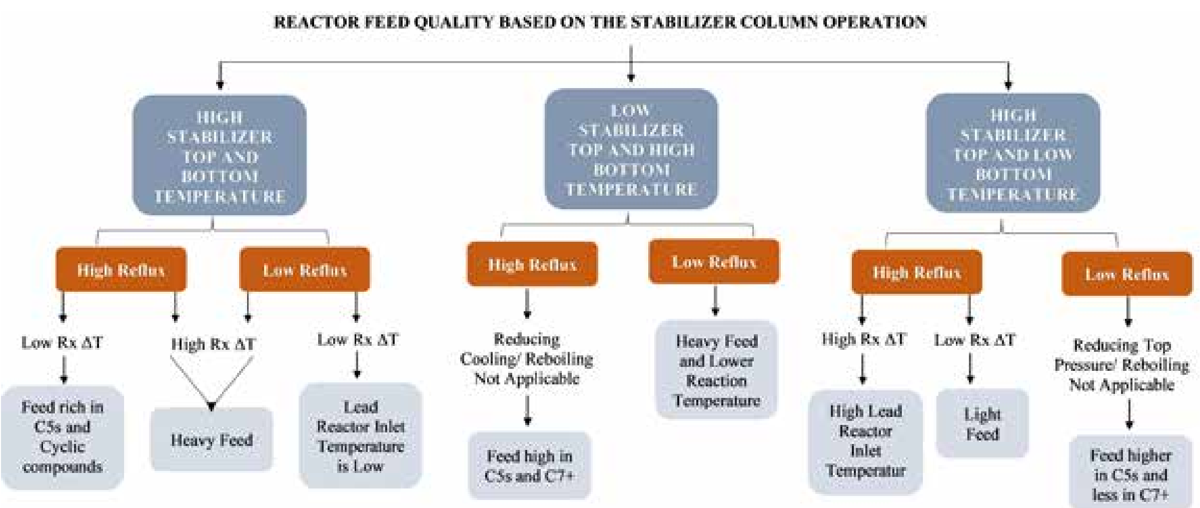

Figure 1. Reactor feed quality based on the stabiliser column operation.

The components were altered in these manners to understand the effect of feed composition on product RON and RVP.

Reactors Performance

The isomerisation reactions, benzene saturation, ring opening and cracking occur in the lead reactor. Hence, the lag reactor inlet temperature is lowered to maintain paraffin to iso-paraffin conversion and avoid reaction reversibility.

The focus is on achieving an optimal inlet temperature for both reactors at which DMBs production is maximised and nC5 to iC5 conversion is lowered without highly impacting product RON. For this study, lead and lag reactor inlet temperatures are reduced below normal operating conditions by 7 and 5°C, respectively. Most importantly, the lead reactor inlet temperature should not result in complete C7+ cracking to avoid eliminating its presence in isomerate and increasing C4- components. Different outcomes are expected for cases with varying numbers of reactors and catalyst types.

Stabilizer Column

Promoting the stabiliser column separation allows for improving product RVP due to minimizing C4- components slippage with the raw isomerate. Lifting a small portion of the isomerate C5s with the stabiliser off-gas will improve RVP with minimal effect on RON, especially at high DMBs conversion.

The stabiliser reflux rate and bottom reboiling remained constant throughout the cases to assess the impact of different feed compositions on product quality. Double membrane technology can be introduced to recover hydrogen and LPG components from the net gas to accommodate losses in isomerate yield. Hydrogen can be routed to a hydrogen recovery unit, while LPG can be sent as a product.

Deisohexanizer Column

The most potent streams that process engineers can artistically optimise to impact isomerate quality is the DIH. The type of adjustment differs based on the reactor and DIH feed components. The guidelines in Figures 1 and 2 assist in identifying the changes in reactor feed quality without a sample and advise on the required DIH stream optimisation. The focus should be on optimizing the column to allow the accumulation of DMBs in the system as this maintains higher product RON and lower RVP. It was found that this is more achievable at lower operating capacities than 100% as there is more flexibility in optimising the unit.

Results and Discussion

Feed Quality

The composition of the reactor's three feed cases (A - C) were analysed to understand the impact of components, RON and RVP on product quality. The results revealed that feed RON and RVP only sometimes directly correlate with product RON and RVP. Therefore, the comparison was dedicated to specific components. As shown in Figures 3 and 4, an increase in higher RVP contributors (C4s, C5s, C7+) and a decrease in lower RVP components (nC6, MPs, DMBs) is directly proportional to high product RVP and lower RON.

The Case A feed contained 31 mol% of C5s and 3.3 mol% of C7+, twice the content of that in Case C. Moreover, the total nC6s, MPs and DMBs in Case A were lower than in Case C by 16.4 mol%. As a result, Case A product RON and RVP were 87.6 and 11.5 psi, respectively, while Case C product RON was 86.3 with a RVP of 10.6 psi. On the other hand, Case C feed contained higher 2.8 mol% nC6, MPs and DMBs, 0.9 and 1.67 mol% lower C5s and C7+, respectively, than Case B. The differences were minor however, with different stabiliser and DIH column operating parameters, the resulting Case B product RVP is 11 psi with a RON of 87. Although the isomerate RON for Case C was lower than the minimum requirement of 87, the reduction was considered minimal for the RVP difference of 2.2 psi.

In conclusion, the results of the three presented cases met the trial's target of RVP below 12.8 psi while maintaining a higher RON. To illustrate, the original product RON was 87, with an RVP above 13.0 psi. From the author's experience, the results shown in Case B are more achievable and optimal.

Reactors Performance

The author prefers to keep the reactor temperature adjustment as the last optimisation option to preserve the catalyst lifecycle.

The key is that the selected temperature does not result in lower RVP by a drastic reduction in nC5 to iC5 conversion, as presented in Figure 5. Therefore, the optimal lead reactor inlet temperature was maintained throughout the trial, while the lag reactor's temperature was slightly increased to be aligned with the increase in nC6/MPs components. The temperatures were deliberately maintained constant to observe the stabiliser behaviour.

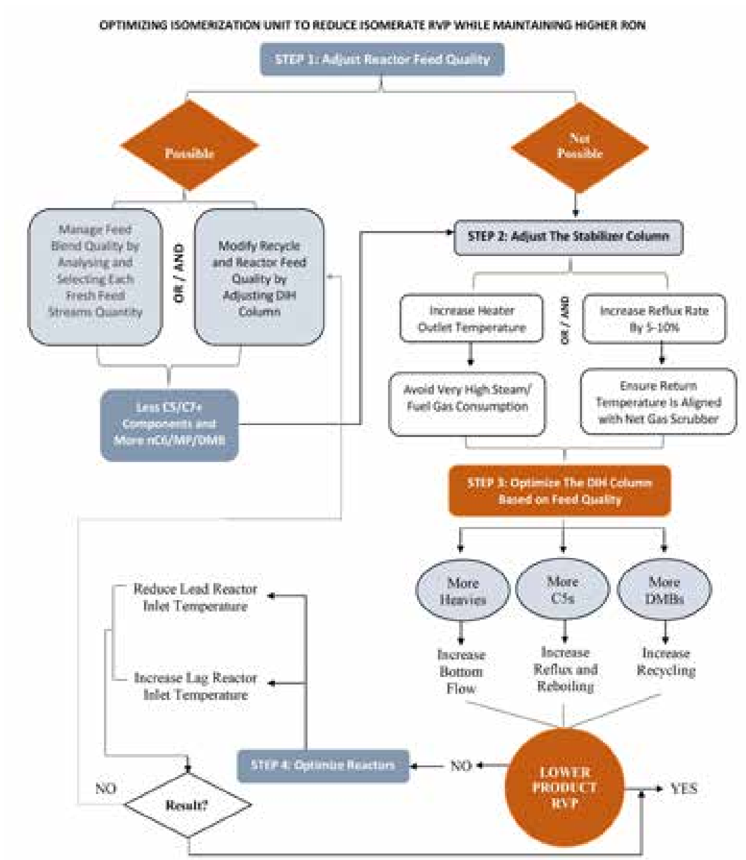

Figure 2. Optimising isomerisation unit to reduce isomerate RVP whilst maintaining higher RON.

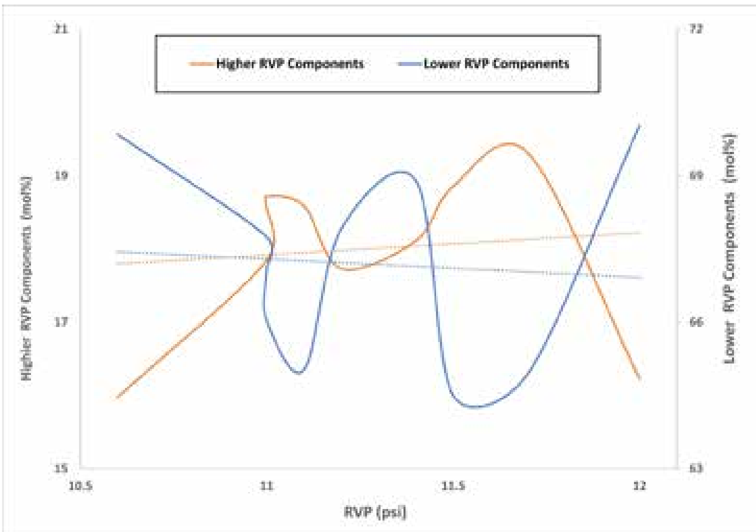

Figure 3. The impact of feeding higher and lower RVP components on isomerate RVP.

Stabilizer Column

The stabiliser bottom temperature was maintained at a standard operating value with a reflux flowrate higher by 8% throughout the three cases.

The higher C5 and C7+ components in the Case A feed resulted in a top temperature of 98.5°C against 96.9°C and 95.5°C for Cases B and C, respectively. At the same operating conditions, the higher the C4s, C5s and C7+ components in the reactor feed, the higher the stabilizer overhead and return reflux temperatures. Lifting higher RVP components with the off-gases is more achievable for Case C due to the feed's lower presence of C4s and C5s. Consequently, reducing the stabiliser bottom temperature would increase Case C product RVP, and a higher temperature would decrease Case A RVP.

Higher return temperature and overhead cooling were required for Case A due to the feed's higher C4 and C5 content.

Subsequently, cooling requirements dropped for the other cases. Reducing the reflux rate would affect the separation efficiency, leading to more C5s in the raw isomerate.

Therefore, slightly increasing the return temperature of Cases B and C would result in more effective separation. Although the top temperature may drop with the higher reflux rate, unnecessary energy losses due to heating the additional reflux shall be avoided.

Moreover, the reflux return temperature must be aligned with the downstream net gas scrubber operating condition to avoid foaming. In conclusion, based on the optimiser requirements, a decision can be made on what elements to sacrifice while operating the unit safely.

Deisohexanizer Column

The DIH top and bottom temperatures, reflux rate, and overhead flowrate decreased along Cases A - C. On the other hand, DIH bottom flow and recycling increased.

The column behaviour is associated more with the quality of DIH feed, which changed due to the different reactor feed quality and maintained stabiliser bottom temperature.

A higher presence of C4 and C5 components in the DIH is expected when the reactor feed is lighter. Hence, an increase in the DIH overhead flow and reflux rate was observed in Case A. Subsequently, the bottom and recycle were reduced to maintain the overall column balance, resulting in higher product RVP and RON.

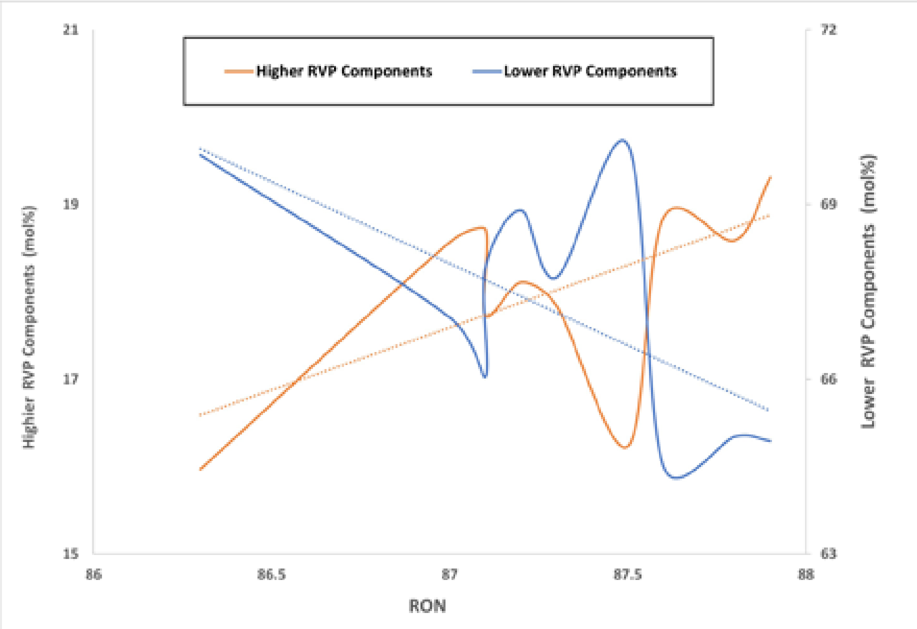

Figure 4. The impact of feeding higher and lower RVP components on isomerate RON.

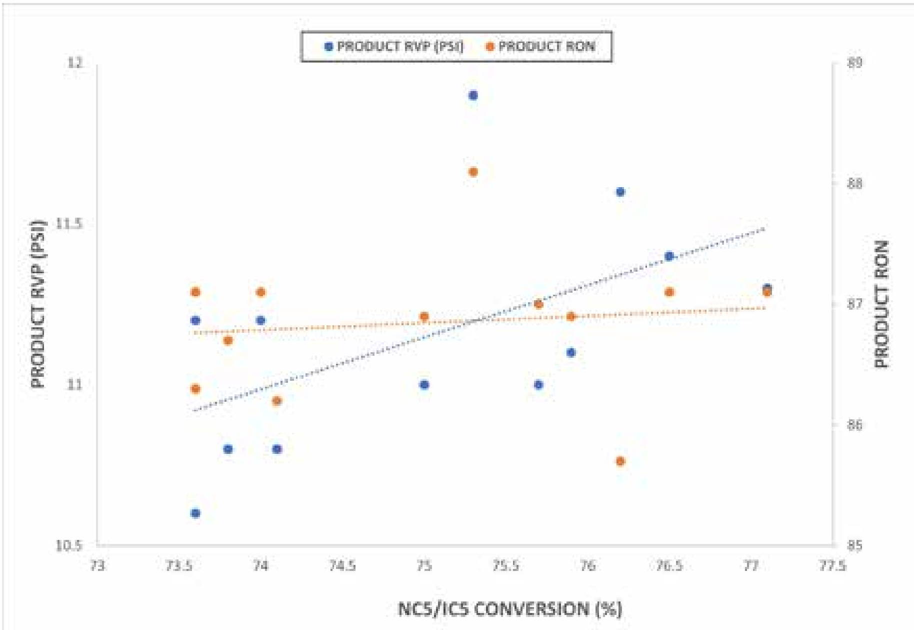

Figure 5. The effect of lower iC5 to nC5 conversion on product RON and RVP.

The Case C feed was lower in C4 and C5 and higher in lower RVP components, which resulted in lower DIH overhead flow and reflux rate.

Consequently, the bottom and recycle flows increased to maintain material balance, contributing to lower product RVP with minimal reduction in RON. In conclusion, minimising product RVP while boosting RON requires lower iC5 and a higher DMBs presence in the overhead with maximising nC6 and MPs in recycle to increase DMBs conversion.

Conclusion

Isomerisation is a leading process for refineries that aim to compete in global oil markets with low aromatic and sulfur gasoline.

Enhancing isomerate quality depends on operational planning requirements of higher yield or RON. Despite this, high gasoline RVP can be a challenge for refineries, and reducing isomerate RVP is expected to result in lower RON. An in-house exercise was conducted at the MAA refinery to minimise the isomerate RVP while maintaining a high RON.

Isomerate RVP was reduced from a maximum of 14.1 to 11 psi with a RON of 87. It is worth mentioning that an isomerate RON of 89 was achievable at an RVP lower than 12.8 psi. The conclusion of this trial is constructed in Figure 2 as a guideline to assist other optimisers in their journey to achieve similar objectives.

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