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Maximization of Egyptian Gas Oil Production Through the Optimal Use of the Operating Parameters

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Abstract

Gas oil is the major fossil fuel consumed around the world. Global gas oil consumption is rising at a steadily fast pace because of its higher combustion efficiency (versus gasoline).

The annual increase rate of gas oil consumption in Egypt is 7 % whereas, the world increase rates range from 1.5 % to 2 % .

The main sources for producing gas oil in Egypt refiners is the direct production from the atmospheric distillation process units or it may be produced as a side product from vacuum distillation units .

Gas oil is produced through hydrocracking process of vacuum distillation side streams and heavy cocked gas oil.

Gas oil production yield can be increased through the existing operation process units. Modifications of the current atmospheric and vacuum tower operations will increase gas oil yield rates to 20 % more than the existing production rates.

The modification of the operating conditions and adoption of the optimum catalyst of the existing hydrocracking and mild hydro cracking process units improve gas oil production yield. Operating delayed cocker at high temperatures, low pressure and low cycle ratio also support achieving the maximization of gas oil yield.

General Introduction

The refining industry faces many challenges which affect refinery processing schemes over the next years.

The global demand for gas oil fuel shows an increase of at least 18 % over the next 10 years, while the market share is expected to slightly decline (**Table 1**).

Jet fuel demand climbs rapidly due to the expansion of air transport, while fuel oil demand continues its long – term decline. Refiners must therefore adapt products slate significantly.

Table (1) : World Market Share for Middle Distillates

Years / Items	2000	2005 (expected)	2010 (expected)	Ten years deference
Gasoline	38 %	37 %	37 %	- 1.6
Jet fuel	9 %	11 %	12 %	+ 2.4
Gas oil	33 %	34 %	35 %	+ 2.1
Fuel oil	20 %	18 %	16 %	- 2.9
Total	**100 %**	**100 %**	**100 %**	

Local Market Development

Gas oil production in local market was developed from 4139MT at 1991 to 8276MT in 2003. This difference represents an increase in the production yield reached to 100 %.

The development is referred to the improvement of the performance of the existing atmospheric distillation units and also the installation of a new refiner (MIDOR) which started in operation in 2001. Such new refiner contains a hydrocracker unit and produces about 25 % of the total gas oil production in Egypt .

A mild Hydrocracker is considered an additional operating unit to local gas oil production. It was installed at (AMOC) company and designed to produce 460MT/Y of high gas oil quality.

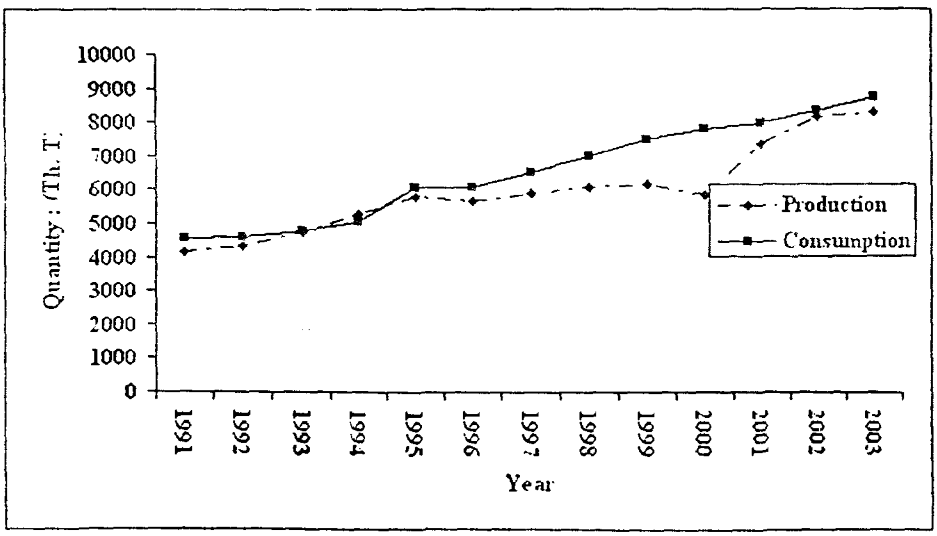
In spite of increasing the production of gas oil and the addition of a new operating unit to maximize gas oil production rate, the consumption was increased by rates over the production rates .

Table (2) represents the development of production and consumption of gas oil in local market started from 1991 – 2003 . This figure explain that the consumption gas oil at 2003 was increased by 90 % over the consumption at 1991 . This can be refer to the new accessional use of diesel engine especially at irrigation sector .

Maximizing methods of gas oil production yield through the existing process units are to be carefully studied to improve the performance of these units to come over the up-normal gas oil consumption.

Table 2 Local Market Production and Consumption of Gas Oil/Diesel from 1991 to 2003

Year / Quantity MT	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003
Production	4139	4319	4747	5255	5790	5652	5867	6078	6147	5857	7325	8130	8276
Consumption	4564	4613	4779	5052	6051	6051	6485	6972	7495	7794	7970	8315	8711



Improving the Performance of Crude Distillation Unit to Maximize Gas Oil Product Yield

Improving crude unit profitability with low revamping capital cost requires modifying only the necessitated systems and equipment. The revamp of a unit should include the practical process criterion which determines the reliability, the operability and the flexibility [1].

Crude-unit product yields depend on heat input, heat removal, operating pressure and fractionation. Heat input into crude charge consists of crude – exchanger duty [2]. Pump around and condenser equipment design and operation determine heat removal. Crude-unit operating pressure is largely determined by the condensing system capacity, fouling and corrosion[3,4,5].

Fractionation depends on the process design and equipment performance. Problems in any of these areas affect unit capacity and product yields and quality. The improvement of the crude-unit profitability requires reliable process and equipment design [6].

The degree of fraction in a crude distillation unit is measured by the temperature difference between the 95% vol. ASTM of the lighter product and the 5% vol. ASTM temperature of the adjacent heavier one.

When the 95% point of the lighter product is less than the 5% point of the heavier one, the difference is called an ASTM gap; and when the 95% point is higher than the adjacent product 5% point, the difference is called an ASTM over lap. The best fractionation performance is when there is an ASTM gap between the products. As fractionation deteriorates and gaps become over laps more and more of the components of the adjacent products are not separated.

Practical Process Considerations [4]:

Revamps, solely based on computer models, often fail to meet their profitable objectives, and in case they do not include practical process criteria, they will be unsuccessful and unreliable. Examples of the practical process criteria are:

- Startup and normal unit upsets .
- Corrosion and fouling[6] .
- Operability .

Crude unit start-ups and normal unit upsets can cause severe mechanical stresses on the column internals. Start-ups can particularly damage column internals if they are not properly designed. These stresses are due to either normal startups conditions or inadequate operating procedures. Whatever the case is, the column areas that are susceptible to be damaged, must be mechanically designed to handle the start-up, shut down and abnormal conditions.

Wet stripping steam and high liquid level can damage trays. Wet stripping steam introduces water to the bottom of the crude column. The water contacts hot oil and expands rapidly. As the water vaporizes into steam, a pressure surge is created to exceed the internal mechanical strength of the column. High liquid level also damages crude column internals due to the energy of the liquid caused by steam or transfer-line vapor velocity. During normal operation, wet stripping steam and high liquid level knock out trays, causing damage during start-up[4]. The stripping-section and wash-section tray performance affects light gas oil (LGO) and heavy gas oil (HGO) product yields and unit profitability **(Fig. 1)** .

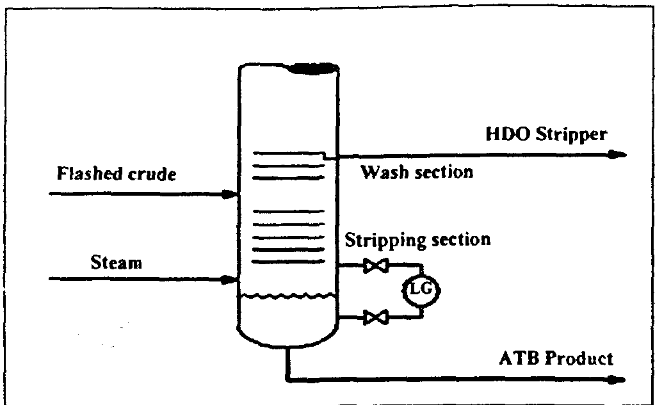


Fig. 1 Crude – Column Stripping and Wash Section

Mechanical stress encountered during start up, shutdown and upset conditions cause lower quality than that expected of light gas oil (LGO) and heavy gas oil (HGO) product yield.

Damaged stripping trays alone resulted in a loss of 10 vol . % of heavy and light gas oil product to fuel oil. Stripping and wash-section tray damage is very common (Fig. 2). Standard mechanical design trays (0.2 psi uplift) had been installed in the wash and stripping sections. Standard strength trays work fine for standard conditions. However, stripping and wash-section trays are not always exposed to standard conditions. They must be designed to stay intact during start-up and normal unit upsets. The stripping-section trays were replaced with a heavy-duty tray design .


Fig. 2. Stripping and wash section tray damage .

Crude-column over flash is the vaporized oil that returns to the flash zone as liquid. Minimum column over flash maximizes the HGO-product yield. Controlling the wash – oil rate by changing the HGO – product yield (As the existing situation of the major crude distillation units) causes difficulty to operate and either lower or poor HGO quality. The revamp used a total-draw tray with flow-controlled reflux to a packed bed. This design permitted minimum over flash operation and maximum LGO and HGO yield. The wash-section mechanical design can be upgraded to maximize the HGO by rate reached to 7 vol . %[7-9] .

Corrosion and Fouling[6,7]

Crude-column top trays and overhead condensers are susceptible to high levels of general corrosion and under-deposit corrosion and fouling. Neutralizing and filming chemicals are used to reduce corrosion rates and extend condenser life. Corrosion of overhead condensers affects the condenser system performance, which in turn affects the column operating pressure. Column top-tray corrosion decreases fractionation between naphtha and jet fuel. Top-tray corrosion is a function of reflux rate, temperature and water content .

Reduced condenser heat removal increases column operating pressure and decreases light and heavy-gas oil product yields. When light and heavy-gas oil yields decrease, profits are lowered. Over head condenser system design causes an increasing of column operating pressure which decreases LGO and HGO product yields.

The crude-column operating pressure increases as the condensers foul. Large increases in the column operating pressure from start-of-run to end-of-run are a common cause of distillate-yield loss. The column operating pressure determines the amount of vapor generated at the heater outlet temperature for any given crude oil mix and temperature. The heater outlet oil vaporization strongly affects product yields and economics[10] .

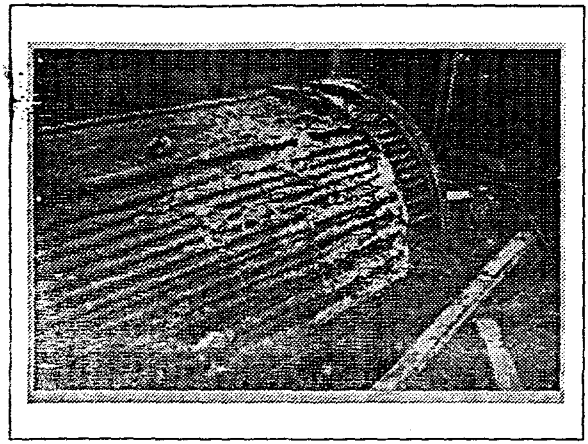


Fig.3. Fouled overhead exchanger bundle.

Crude – unit over head systems that exchange heat against crude, commonly experience corrosion and fouling (fig .3). The root cause is insufficient water at the point of salt deposition to dissolve the salt. The only effective way to remove these salts is by injecting water in front of the exchangers. Whenever column over head vapor is experiencing against crude oil, the purpose is heat recovery. However, injecting sufficient wash water to remove these salts lowers the temperature to the first exchanger by 20 $^{\circ}$c, which in turn reduces the driving force for heat exchange. Proper water washing of crud – column over head systems is difficult.

The parallel exchangers can be periodically isolated for cleaning. Cleaning overcomes the inherent reliability problems with these condenser-system designs. Living with the consequences of this system design is purely a business decision. The benefit of cleaning the exchangers is lower with average operating pressure. Reliable operation at lower pressure increases distillate yield and improves unit profitability.

Operability (Normal Operation and Start – Up)(11) :

Unit operability is a measure of the unit capability to function during start-up through normal unit disturbances, and the inevitable bounces associated with crude switches.

With the gas oil pump around located between light and heavy gas oil products, it is possible to dry out the LGO-stripper draw tray. Varying crude slates and charge rates require constantly changing heat removal in the gas oil pump-around, otherwise; the LGO stripper draw can dry out. Once this occurs, the LGO stripper losses level, and the stripper-bottoms pump suffer from cavitations. The operators had to constantly shift the heat balance to maintain stable operation. Therefore, maximum LGO yield could never be achieved. Fig. (4) shows a previous modification to the LGO stripper and product draw system. A flow controller was installed to feed the stripper . This design was used to overcome the dry out problem, and the LGO product was drawn to storage on level control.

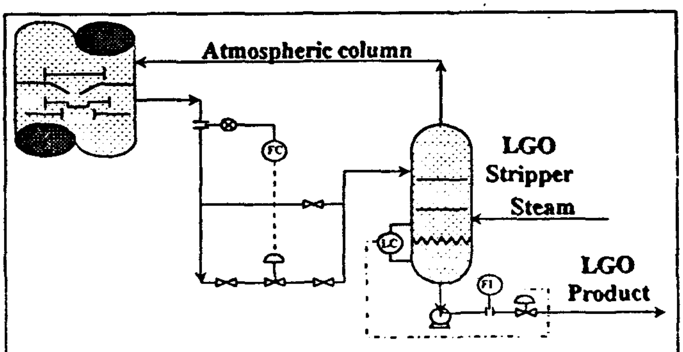


Fig.4. LGO Side – stripper draw.

Crude – Column Fractionation:

The LGO-product end point control was poor because reflux below the LGO draw could not be maintained through crude switches, charge rate changes or normal unit upsets. The gas oil –pump around and exchangers surface area sizes were large, where the gas oil –pump around heat removal reduced the reflux from the LGO product draw tray. Reduced reflux downgraded light gas oil to heavy gas oil. Ultimately, the gas oil pump around was taken out of services because it reduced LGO yield. This reduced the total column heat removal capacity, which limited the fired heater duty and lowered the light and heavy gas oil product yield.

Pump around location has significant impact on column internal liquid and vapor rates. Increasing gas oil pump around heat removal at constant heat input lowers the vapor rate leaving the pump around section. This, in turn, reduces the internal reflux flowing from the LGO product draw tray. Reduced reflux increases the LGO product endpoint. When the internal reflux from the LGO draw tray reaches zero, there will be insufficient liquid flowing to the LGO stripper maintain stable operation.

Fractionation Efficiency – LGO / HGO Section

LGO product yield and fractionation between LGO and HGO product should be maximized. When LGO/HGO section had poor fractionating efficiency which indicated that poor LGO quality and yield control.

LGO / HGO fractionation section experienced large changes in vapor and liquid rates from the top to bottom trays. The liquid rate decreased to 50 % from the top trays to the bottom. The vapor rate also increases by 30 % from the bottom tray to the top. A single tray design used throughout the LGO / HGO section will cause low efficiency or reduced capacity in the column. The revamped LGO / HGO section trays used two different tray designs to handle the process requirements. Often, small equipment design changes result in significant profit improvements.

Improve the performance of Hydrocracking Process Aiming to maximize gas oil yield

The world's refineries have accepted hydrocracking as the process for giving superior products from a wide variety of feed stocks. Its success is due to its ability to handle and convert heavy aromatic compounds into lighter material, rich in naphthenes and isoparaffins. Process improvements, in optimizing configurations and taking full advantage of the latest zeolities, will reduce capital expenditures and operating expenses.

Because of similarities in designing conditions and equipment size, the mechanical design and operating features developed for the hydrocracking process, have been incorporated into the design of residuum hydrotreating plants.

The decline in usage of heavy fuel oil is expected to continue for the foreseeable future. Careful selection and implementation of hydroprocessing schemes will allow refiners to produce high-quality products from low-quality feeds.

The demand for middle distillates is expected to continue to increase in the coming years. Investment in options reduce refiners output of gasoline and heavy fuel oils and increase the production of high-quality distillates (gas oil) is consider a new trend in refinery process . Hydrocracking will be important. This is prompting hydrocracking catalyst suppliers and process licensors [12,13] (such as **Akzo Nobel, Axens, Chevron Lummus Global, Criterion Catalysts, Exxon Mobil, Haldor Topsoe, PDVSA's Intevep, Shell Global Solutions, Sinopec's RIPP, Technip-Coflexip, and UOP**) to improve upon their existing technology and expand their product portfolios .

Optimization of the composition of amorphous catalysts can maximize gas oil yield and meet aromatics and cetane specifications. Some catalyst suppliers have chosen to incorporate modified zeolites into catalysts to obtain product slates close to that of amorphous catalysts without greatly sacrificing activity. Several of these catalysts contain novel base metals, which offer higher aromatics hydrogenation ability than other base metal components at a cheaper cost than Pt or Pd. Lower-pressure hydrocrackers can be integrated with downstream hydrotreaters for the production of high-cetane diesel with a maximum sulfur level of 10 or 15 ppm .

More costly two-stage configurations minimize the aromatics content of the gas oil product [14]. High-pressure units currently on-stream are able to produce gas oil with a cetane level above 60 and a sulfur content below 10 ppm .

Operations at lower overall conversations via special arrangements of the hydrotreating and hydrocracking zones is found to provide a more economical option for producing ultra-clean gas oil than conventional processing .

Integration of a separation unit with post-treatment / finishing units of the hydrotreating or hydrocracking reaction stage can enhance desulfurization and cetane upgrading.

This setup can be used to co-process heavy feeds and secondary low-quality distillates to maximize the production of high-quality gas oil.

Hydrocracking Process

Hydrocracking feed stocks were low-value light Coker gas oils and light cycle oils, straight-run vacuum gas oils (VGO's) , Coker gas oils and light waxy distillates .

Hydrocracking feed stocks are complex mixture of hydrocarbons. Reaction process can be best understood by considering the three types of hydrocarbon compounds; paraffins, naphthenes and aromatics.

Hydrocracking reaction are inhibited by the presence of nitrogen compounds in the feed and ammonia in the circulating hydrogen-rich gas.

The overall reaction rate was nearly 30 times faster when the impurities were not present. Even the paraffin conversion was significant. This is a feature of isocracking which has led to a variety of process configurations which take advantage of the enhanced reaction rates in a clean system.

Most of the sulphur and nitrogen impurities exist in the aromatic component of the feed. They are eliminated at relatively low residence times at normal hydrocracking operating conditions. There is a significant difference, however, in structure between the sulphur and the nitrogen-containing molecules sulphur-containing molecules can be removed quite easily at low pressure, but nitrogen compounds need higher pressure for complete removal.

Feedstocks and Operating Conditions [15]

The heavier the feed, the higher the design pressure level is, the lower the Liquid Hourly Space Velocity (LHSV). The LHSV represents the Reciprocal of the residence time in the reactor.

Residuum feed stocks often require very long residence times. This is partly due to the low reaction rates of these large molecules and partly due to poisoning effect of the metal contaminants present in such feeds.

Typical Hydrocracking operating conditions are represented by table (3)

Table (3) :Typical Hydrocracking operating conditions [16]

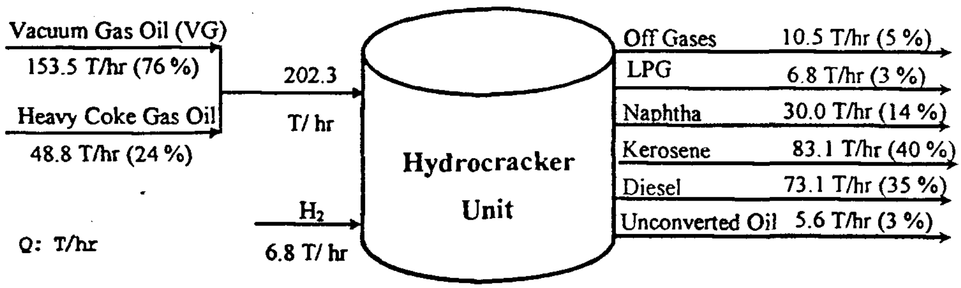
	Hydrogen Consumption (Scfb)	Temperature (^{0}c)	Pressure (Psig)	LHSV h^{-1}
Mild Hydrocracking	1000 – 2400	260 – 480	500 – 7000	0.5 – 4
Deep Hydrocracking	1200 – 1600	375 – 430	1500 – 3000	0.15 - 1

Hydrocracking for Fuels

As the conversions of VGO feed and heavy aromatics increases, the yield of the jet fractions increases and their character changes. They become less aromatic and more naphthenic and paraffinic. This trend leads to a significant increase in the burning quality. Aromatic levels drop and naphthenes increase as the VGO conversion increases. The product paraffin's are largely branched compounds, resulting in out sanding cold flow properties.

These data were generated with an amorphous catalyst. When highly active zeolitic components are added to amorphous catalysts, a higher ratio of light to heavier liquids results. The refiner can change the recycle cut point in his product fractionator. The recycling cut point controls the heaviest product that is drawn from the unit.

Deep Hydrocracker



Material Balance of Hydrocracking Unit

Item	Design		Actual	
	Quantity (T)	Percentage (%)	Quantity (T)	Percentage (%)
Liquid feed	202.3	97.7	187.0	96.6
H_2	6.8	2.3	6.5	3.4
Off gases	10.5	5	10.2	5.3
LPG	6.8	3	4.8	2.5
Naphtha	30	14	29.2	15.1
Kerosene	83.1	40	73.1	37.8
Gas Oil	73.1	35	71.2	36.7
Un Converted Oil	5.6	3	5.0	2.6

Fig. (5): Deep Hydrocracker Material Balance

Fig. (5) illustrates typical deep hydrocracker process unit material balance. The feed for this process unit is a mixture of heavy vacuum gas oil and heavy coke gas oil.

The operating conditions was:

- Reactor temperature (inlet) °C = 375
- Reactor pressure (kg/cm^2) = 170 kg/cm^2
- H_2/Hc ratio (Nm^3/m^3) = 2200
- L H S V (h^{-1}) = 0.47

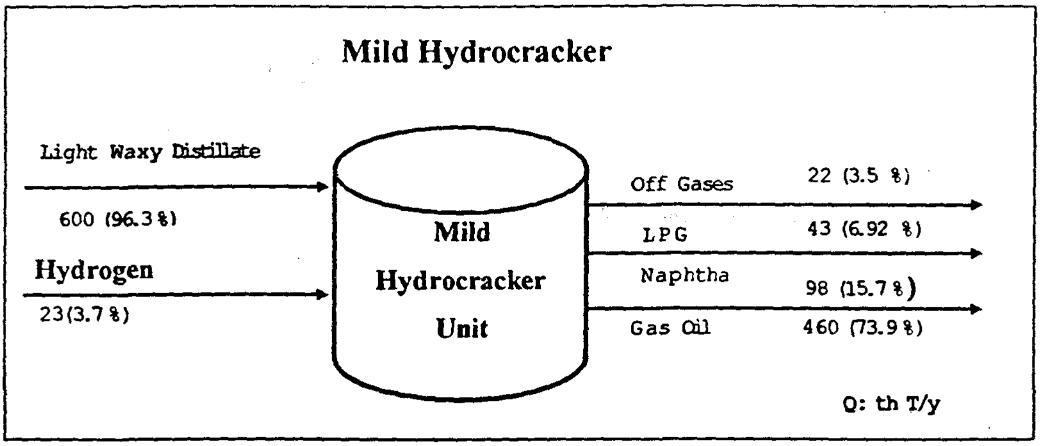
The produced gas oil has high quality properties, where sulfur content reached to 15 ppm max. While the yield was definite at 35 % max. of the feed as design case.

Reviewing the above mentioned data for Deep Hydrocracker concerning the designed actual material balance can be concluded in the following points:

- Actual feed rate reduced to 92 % of the design ratio due to the decrease in vacuum gas oil yield. Adjusting vacuum distillation unit will maximize distillate yield by 12 T / hr.

- Feed properties specially basic nitrogen content and Conrad son carbon content must be controlled to improve the performance of the catalyst , while basic nitrogen reduces the catalyst lifetime . These feed specifications can be controlled through adjust the heavy coke gas quality (reduce the recycle ratio at the delayed Coker unit).

- Decreasing or maintaining the reactor inlet temperature at the value 370 – 375 ^0c will increase the yield of middle distillate by rate reached to 5 %, while the naphtha yield will be decreased.



Item	Design		Actual	
	Quantity (T)	Percentage (%)	Quantity (T)	Percentage (%)
Liquid feed	600	96.3	423	97.7
H_2	23	3.7	10	2.3
Off gases	22	3.5	25.2	5.8
LPG	43	6.9	30.6	7.1
Naphtha	98	15.7	110.3	25.5
Gas Oil	460	73.9	266.9	61.6

Fig.(6) : Mild Hydrocracker Material Balance

Fig.(6) : illustrate typical mild Hydrocracker process unit material balance . The feed for this process unit is light waxy distillate (LWD).

The operating conditions were:

- Reactor temperature (inlet) °C = 435
- Reactor pressure (kg/cm^2) = 60 kg/cm^2
- H_2/Hc ratio (Nm^3/m^3) = 350
- L H S V (h^{-1}) = 2

The produced gas oil has high quality properties, where sulfur content reached to 100 ppm max., while the yield was 73.9 % (max.) of the feed as the design case .

With reference to the design and actual material balance, its clear that the applied type of catalyst not achieving the performance of mild hydro raking process .

The yield of gas oil in design case represent 73.9 % of the initial feed, while this value reached to 61.6 % at actual case. This gab represents about 50,000 ton/y of gas oil yield which can be added to the final product .

Selecting another type f catalyst with high activity and selectivity will improve the yield of gas oil production from mild hydro cracking unit

The naphtha yield can be decreased by reducing the inlet operating temperature by 10 ^{0}c, while this yield will add to gas oil yield.

Increase Gas Oil Yield Through Optimizing the Delayed Coker Operating Technique

Delayed Cockers are the most effective processes to decarbonize and demetallize heavy petroleum residues. This refining process can provide 20% to 40% of the downstream hydroprocessing feed stocks. Effective use of coker recycling can improve unit yields and the quality of coker-liquid products.

The preference to have the delayed coker produce clean hydroprocessing feedstocks may compete with the need to improve Coker operations to process heavier feed stocks. Cokers, producing by product coke, are operated to maximize liquids and minimize coke. Coker pressure, coking temperature and recycling ratio are the swing operating parameters.

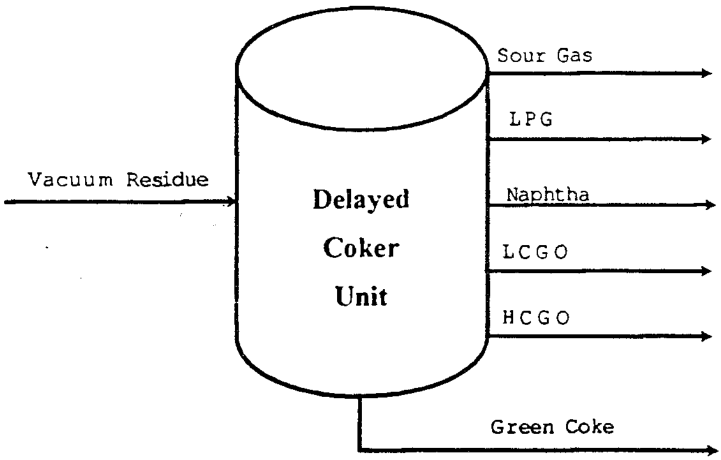
Typical material balance and products obtained from delayed Coker for feed capacity 20,000 bpsd of vacuum residue (VR) are shown in Fig (7) . These data concern the design and actual percentages of each product at 20 % recycle ratio of fresh feed and at 530 °C coking temperature.

Methods to reduce the production of low-valued coke and increase the volume of hydrocracker plant feeds, are of high interest to refiners(17,18) .

Coke Drum Temperature(19) : The hotter the coke drum, the harder the coke . Hard coke means low weight percent volatile combustible matter (VCM).
By keeping a coke drum hot, the 540° C boiling range gas oil is vaporized out of the coke. Heavy gas oil left in the coke thermally degrades to gas, coke and distillates or even worse stays in the coke in the form of VCM.

For every increase of 5°C, the production of heavy gas oil and distillate increase by 1.1vol. % on fresh feed. The proper place to monitor the coke-drum temperatures is in the coke drum overhead vapor line, up stream of the vapor-line quench injection point.

The beneficial effect of raising the coke drum temperature is greater at the end of coking cycle. This is the reason for increasing the coking heater outlet temperature by 3-5 °C the last 2 hr. before switching.



Material Balance of Delayed Cooler Unit

Item	Design		Actual	
	Quantity (T)	Percentage (%)	Quantity (T)	Percentage (%)
Feed Rate	156	100	158	100
Sour gas	8.6	6	9	6
LPG	6.6	4	4	2.5
Naphtha	12.2	8	15	9.5
LCAO	35.7	23	49	31
HCGO	48.7	31	31	19.4
Coke	44	28	50	31.6

Fig. (7) Typical Delayed Coker Material Balance for Design and Actual Cases

Steam-out to Fractionator: Using 4000-5000 lb/hr of coke-drum quench steam for an hour on a 20,000 b/sd Coker, while the drum is still lined-up to the fractionator, may recover an extra 100 b/sd. of distillates.

Operators who are abbreviating this "small-steam" step due to fractionator flooding problems or reduced cycles, often count on recovering the residual gas oil and distillates in the coke drum steam-out quench blow down scrubber .

In practice, this is somewhat of an illusion. The recovered hydrocarbons are partially lost to atmospheric evaporation, or settle to the bottom of water-recovery ponds during the subsequent stop processing.

Light Distillate Recycle [20]**:** Recycling kerosene and gas oil boiling range Coker distillate reduces both coke yield and the coke's VCM. A unit processing 20.000 b/sd of a vacuum residue, initiated a recycle of 2,500 b/sd of a light distillate and the coke declines by 2-3 wt. % on feed.

Steam Injection: Most refiners begin injection of steam to the full coke drum prior to completing the switch. Some start the steam flow before breaking down the Wilson-Synder valve.

Continue steam injection to coke drum during most of the coking cycle reduce coke drum hydrocarbon partial pressure that reduce coke yield and maximize gas oil production.

Drum Pressure Reduction: The best known method to minimize coke yield from a barrel of vacuum residue (VR) is to reduce the coke-drum pressure.

Each 8 psi reduction in drum pressure increases liquid yields by 1.3 vol. % and cuts coke yield by 1 wt. % of fresh (VR) feed.

Recycle Ratio [21] : Recycle is formed by condensing the high boiling components that vaporizes and entrains within the coke drum vapors . This back-end "tail" contains high metals, Conrad son carbon and C7 insoluble contents, which have deleterious impacts on downstream hydroprocessing and other catalytic units. Condensed recycle stream is re-injected into the feed stream and recycled to extinction. This conversion of heavy, high-boiling components enables producing lighter products and some coke.

Refiners will operate Cokers at a recycle ratio of 10% to further reduce the Conradson carbon residue (CCR), heptanes insoluble (HIS) and metals in the HCGO to protect down stream units.

Charging heavy Coker gas oil to a hydrocracker typically reflux gas oil below the trap-out pan to control ASTM end-point and Conrad son carbon (not more than 0.5 %). Both of these parameters are functions of the contacting efficiency, as well as the liquid rate across the wash trays.

There is a reasonable probability that installation of a structured-type packing wash section below the heavy gas oil trap-out pan would cut the internal reflux rate required to control gas oil production. Co carbon and endpoint of such a reduction in reflux would, in turn, minimize the coking heater through-put ratio (TPR) where:

$$TPR = \frac{\text{Coking heater feed}}{\text{Fresh feed}}$$

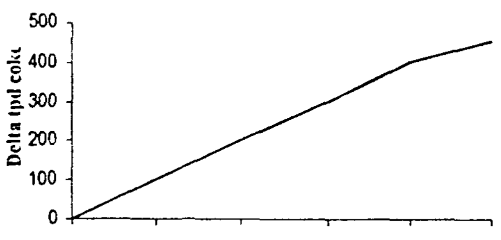


Fig. 8 : Recycle and Delta coke production for a 20.000-bpsd Coker using recycle concepts

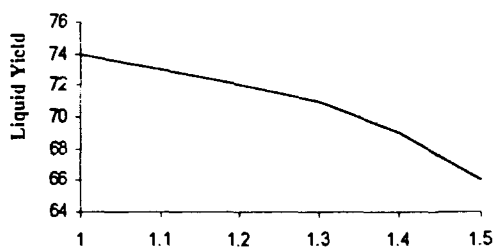


fig 9 : Recycle effects on liquid products of a delayed Coker

Fig.'ˢ. (8) and (9) illustrate operation performance for a typical 20,000-bpsd Coker operation. Additionally, as the yield of liquid products increases, yield distribution changes. The liquid products become heavier with less naphtha and light-Coker-gas oil (LCGO) and more heavy-Coker-gas oil (HCGO). Furthermore, the HCGO has a higher end point and mid-boiling point as shown in **Fig. (10)**. Reducing recycle lowers capital and operating costs. The single largest variable factor in operating costs is fuel consumption. The incremental reduction for fuel cost with lower recycle for a typical 20,000-bpsd Coker is shown in **Fig. (11)**.

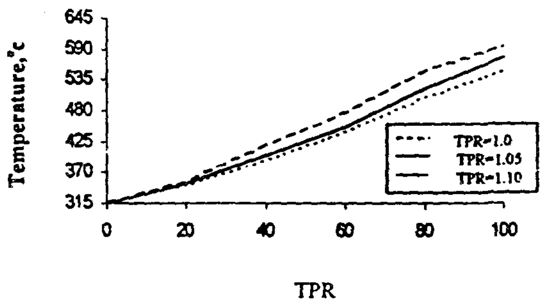


Fig. 10: As recycling rations increase, the heavy Coker gas oil becomes a heavier product with a higher boiling temperature

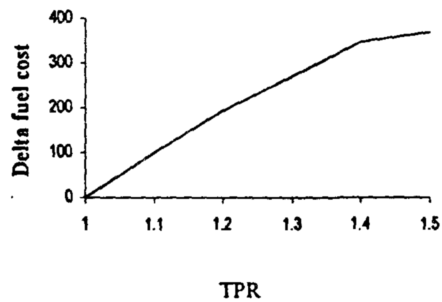


Fig11: With lower recycle efforts fuel consumption

Reducing TPR from 1.20 to 1.0 reduces coke production by approximately 1.2 wt % based on fresh (VR) feed.

Vacuum Reduction: Reducing the volume of Coker fresh feed by 10 % through enhanced removal of heavy virgin gas oil in the upstream vacuum tower, will reduce coke yield by 1.5 wt % on fresh (VR) Coker feed. Although, this factor alone represents big profit savings.

Drum Insulation: These are a host of superficially extraneous factors which tend to be overlooked in minimizing coke yields:

- Deteriorated coke drum insulation, water from the coke-cutting operation gets underneath the drum weatherproofing and degrades the insulation integrity. Not only is energy lost but the cooler drums retain 540 °C, gas oil in the green coke product.
- Tray flooding due to ammonia sulfide sublimation, high nitrogen content crude oils promote plugging of the tray decks at the top of the Coker fractionator with ammonia salts. The trays then flood prematurely with a resulting increase of several psi in back pressure against the coke drum. Periodic, on-stream water washing, corrects this problem. Residues, produced from the heavy oil extracted from tar sands seem, especially susceptible to this problem, possibly due to the nitrogen content of these residues.
- Overhead condenser fin fan external fouling, coke fines and road dust accumulating around the outside of the cooling fins reduce air flow and heat transfer in the tube bundle . This increased the vapor load to the wet gas compressor to such an extent that the compressor suction pressure rose to 8 psig from 3 psig.

The coke drum pressure also increased to 31 psig from 28 psig, with a concurrent and unnecessary escalation in coke yield. This situation was easily corrected by an external, on stream water wash of the fin-fan tube bundle by the operating crew.

Shot Coke[21] **:** The residues from crude's containing a high percentage of asphaltenes will tend to from shot coke when processed in a delayed Coker . It appears as if successful methods to reduce coke yields (with the probable exception of reduced pressure) will enhance the tendency to produce shot coke.

Minimizing the salt content and the exposure of hot Coker feed to oxidation has been suggested as a method to partially inhibit the formation of shot coke.

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