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Optimization of Delayed Coker Unit Process Variables for Enhancement of Product Yields

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Abstract

This study has been constructed using Aspen HYSYS ver.12.1 and regression analysis had been performed by MICROSOFT EXCEL 2010 to obtain a new correlations to predict the product yields from delayed coker unit with a wide applicable range of operating variables which is more reliable with refineries data and a simulation molding of delayed coker unit has been accomplished to maximize the gas oil yield for a refinery data by optimization of process variables. The findings from optimization by linear programming performed by MICROSFT OFFICE EXCEL 2010 indicated that gas oil yield could be increased by 4 wt% instead of coke byproduct by lowering the recycle ratio to 5% wt from fresh feed and increasing the heater outlet temperature to 510°C.Also the results showed that the change in coke drum pressure has a minimal effect in product yield. The outcome from the modified process conditions studied and a profit estimated at approximately 40 million dollar yearly.

Introduction

Delayed Coking History

There is an issue with heavier crude oils with a high density that have been available over the lighter crudes in recent decades as the oil industry requires more gasoline and diesel products today. In all refineries, the necessity to process heavy oil has taken precedence. Processing heavy oils calls for more advanced physical separation, conversion, and distillation equipment. The delayed coker unit, which transforms the heaviest and least desirable components of crude bottoms, such as heavy sour vacuum residue, into marketable products that are further processed to higher economic value products like jet fuel, gasoline, and diesel fuel that are highly demanded in global markets, is one of the most significant conversion units. The delayed coking process yields coke, liquefied petroleum gas (LPG), light gas oil (LGO), heavy gas oil (HGO), naphtha (gasoline), and sour fuel gas. It has been come to consider ways to reduce coke formation while increasing the liquid yields of gasoline and gas oil from delayed coker units.

In the 1860s, the first oil refineries in Northwestern Pennsylvania produced petroleum coke. These ancient, rudimentary refineries used crude oil to make kerosene, a valuable and essential fuel, in tiny iron stills. The oil towards the bottom of the stills was heated and coked by wood or coal fires that were built underneath. The still was allowed to cool once evaporation was finished so the workers could remove the coke and tar before the subsequent run. Up to the 1880s, crude was distilled using single horizontal shell stills, albeit the process was occasionally terminated before the bottoms coked to produce heavy lubricating oil. Running the stills in series while the first still produced the coke allowed for the processing of additional fractions. The bubble cap distillation trays, which Koch had developed, were used in tube furnaces with distillation columns that were constructed in the 1920s. The bottoms of these stills, which were made of wrought iron, were in direct contact with the flue gases over their entire exterior. The most heavy gas oil was generated as a result. After World War II, several of these units continued to be used [1].

The vertical coke drum likely developed as a result of the thermal cracking of gas oil used to make gasoline and diesel fuel. The Burton process, created by Standard Oil in Whiting, Indiana, transformed gas oil to gasoline while also producing petroleum coke from 1912 to 1935. Petroleum coke was also made by thermal cracking technologies such as Dubbs [2].

Due to a shortage of crude oil supply and a heavy oil market, landlocked Middle American refineries used a delayed coker to process heavy fuel oil (atmospheric distillation bottoms and vacuum distillation bottoms) to create additional gasoline and diesel fuel [3].

Delayed coking incorporated several elements and advancements from the development of the thermal cracking process. The combination of pressure and heat for cracking and separating the heater from the coker, as well as the use of two drums, allowed the delayed coker to work continuously. The number of cokers built prior to 1955 was minimal, with a 6% annual increase in delayed coker building from 1955 to 1965 and an 11% growth rate from 1965 to 1970 [4].

The saviour delayed coker unit is a low-pressure thermal cracking technique. It gets its name from the fact that coke is formed in coke drums rather than furnace tubes, where it may be stored and removed as a saleable product. During this thermal process, the vacuum residue from vacuum crude distillation is batch-heated in a furnace, which is considered the unit's cornerstone, and then confined in a reaction zone or coke drum under proper operating conditions of temperature and pressure until the unvaporized portion of the furnace effluent is converted to vapors and coke [5].

Delayed coking is an endothermic reaction, with the furnace generating the necessary heat for the reactions to occur. The reactions in the delayed coker are complex and were initially random, with no studies or predictions for the product yields. Today, many researchers are concerned with the product yields from the delayed coker process. Using equations or simulated instances, you can target proximity for your coking yields [6].

Process description:

Fresh feed to the coker fractionator:

The delayed coker unit usually receives fresh residue feed from the hot vacuum unit residue and cold feed from storage. The two feed streams are then merged, preheated in heat exchangers, and introduced at around 290°C to the bottom of the main fractionator, which serves as a surge drum for the coker furnace figure (1) [7].

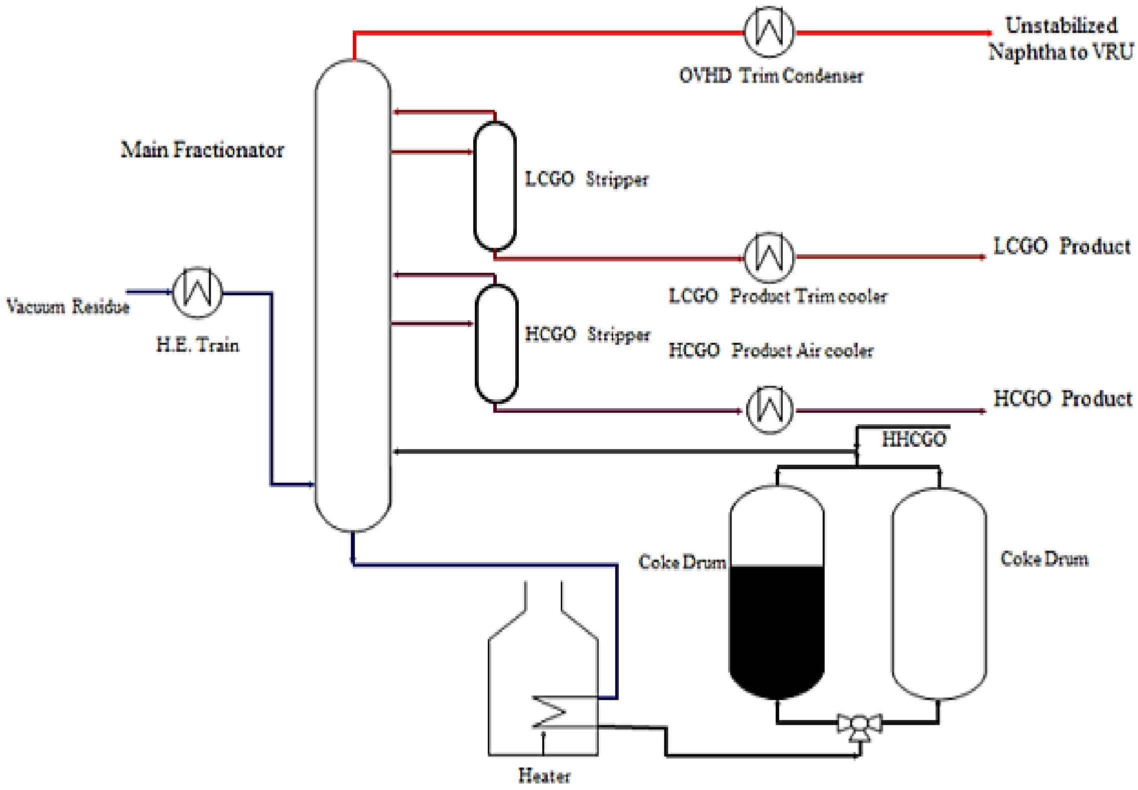

Figure 1 Delayed Coker Unit Distillation section simple process flow diagram [7-8].

Coker furnace and coke drums

The liquid that gathers at the bottom of the fractionator is pushed to the coker furnace from the bottom of the tower. Just before entering the furnace cells, a recycle naphtha or middle distillate heavy gas oil stream may join the main hydrocarbon stream. The blend mixture is then heated to the coking reaction temperature (about 500° C) [8]. Each pass has a high pressure steam pipe called the velocity steam that connects to it. The velocity steam role is critical since it is utilized to improve velocity in each pass and lower coking rate in the furnace [9]. Figure (1) shows a coke drum charge entering the bottom of one of the coke drums. The cracking and condensation reactions start to take place inside the drum forming coke and lighter components that exit the top of the drum in a vapor state [10].

Fractionator section

The coke drum vapors are introduced in the flash zone part of the fractionator. Heavy heavy gas oil (HHGO) or fresh feed hoover residue is sprayed on the vapors. Some of the flash zone gas oil condenses and falls into the flash zone draw pan as a result of the spray. The flash zone gas oil (FZGO) runs from the flash zone draw pan to the FZGO pump suction filters, which remove the entrained coke from the coke drum and transport it to the main fractionator. The FZGO is then recycled to the bottom of the fractionator with the fresh feed vacuum residue [11].

Heavy gas oil (HGO) product withdrawn from the tray is stripped with steam in the HGO stripper to modify the product flash point before being supplied hot to the Hydrocracker unit or cooled to storage, as shown in figure (1) [12].

Light gas oil (LGO) withdrawn from the tray and sent to the LGO stripper, where it is steam stripped before being transferred to storage or the distillate Hydrotreater. Part of the LGO product stream is used as lean oil in the gas recovery unit (GRU) sponge oil absorber tower [13].

At the top of the tower, unstabilized naphtha is produced, separated in the fractionator overhead receiver, and then transferred to the Gas Recovery Unit. The endpoint of the naphtha is determined by the temperature and pressure at the top of the tower [14].

Process variables

The key process variables are [15]:

- Type of Feedstock
- Coke Drum Temperature
- Coke Drum Pressure
- Recycle ratio
- Fractionation Section.

Type of feedstock

The crude source and kind of charge stock, as stated in the process description, have a significant impact on coke yield and quality. The primary property controlling coke yield is the feedstock's Conradson carbon content: the higher the feed's Conradson carbon concentration, the higher the coke yield. The composition of the feedstock, specifically the relative proportions of asphaltenes, resins, aromatics, and contaminants, influences coke quality [16].

Coke drum temperature

After the type of feedstock, the coke drum temperature is regarded the second most effective variable in the coker process. It is controlled by altering the coker heater temperature, which has a substantial effect on both the yield and quality of coke and liquid yields. Heater temperatures should be kept between 480°C and 520°C.

At lower temperatures, tarry coke with a high volatile component matter VCM% and a large increase in coke yield and gas plus gasoline are generated. An increase in temperature within the temperature range discussed for a particular feedstock will enhance the gas oil yield rather than coke and gas plus gasoline yield [17].

Coke drum pressure

The thermal cracking reactions in coking process are a function of time and temperature. The effect of the two variables is related. The drum pressure which determines the degree of vaporization inside the drum and the velocity through the heater can be used to vary the residence time inside the heater passes. By increasing the coke drum pressure, the residence time increased through the heater and also lowering the velocity of vapors inside the coke drums which allow more condensation reactions to occurs that lead to increase in coke yield with high VCM% and decrease in the liquid gas oil yields produced. The fractionator overhead receiver pressure controls the pressure of the coke drum. Changing in coke drum pressure has a low effect on delayed coking process product yields [18-19].

Recycle ratio

Recycle ratio wt% considered a critical factor influencing delayed coker process yields wt%. By increasing the recycle ratio wt%, the heavy gas oil yield draw will be reduced, as would the total fresh feed pulled to the unit reduced by taking up unit capacity. The flash zone temperature where the coke drum vapors effluent enters the fractionation section is the most effective factor in recycling ratio flow rate. A lower flash zone temperature allows more condensation to form initially in the bottom of the fractionator, increasing the recycle flow rate and lowering its temperature. It’s noted that increasing in recycle wt% will produce coke with a high hardgrove grindability index HGI and lower coker liquid yields [20].

Fractionation section

The end point of each product cut (naphtha, light gas oil and heavy gas oil), flash zone point, tower refluxes, pump around and side strippers steam flow rate are the primary elements that determine the fractioning section operation. Any product end point can be changed by modifying the tray temperature and varying the pump and reflux flow rates. For example, increasing the temperature of the LGO draw tray will raise the end point of the LGO and increase its yield while decreasing the HGO yield. As a result, changes in any of the variables in the fractionation section will have an influence on the other variables [21].

Correlations

Coking reactions assumptions

The delayed coker receives its feed from the crude oil delivered to the refinery, and it is integrated with the other refinery operations. One of the main benefits of the coking process is that a refinery with a coker unit is commonly referred to as a "zero-resid refinery". The inherent adaptability of this process for converting a range of feedstocks is another benefit. This allows the refinery to address the issue of a declining residual fuel demand and benefit from the attractive economics of upgrading it to more value lighter products.

Due to the complexity of coking processes, it is challenging to derive an accurate kinetic model. The capacity to accurately characterize the massive, multifunctional molecules involved is the fundamental challenge in modelling a delayed coker. A brilliant model for product distribution was created by Xiao et al. It is assumed that every reaction is a first-order reaction that the cracked products do not participate in secondary reactions and the condensation reaction does not involve a subsequent process [22].

Zhou et al. created a 12 lumped reaction model for product distribution in the thermal conversion of heavy stock. In order to modeling delayed coking, they created a predictive kinetic model and looked at group composition, including residue. The six-component strategy was found to be suitable to employ as a lumped species for residual stock [23].

Bozzano and Dente discuss the particular characteristics of this process as well as the adaptation of a mechanistic approach to liquid-phase pyrolysis of hydrocarbon mixtures to delayed coking modelling. A kinetic diagram with around 1600 equivalent reactions and 450 equivalent components was first created [24].

Tian et al. characterized the delayed coking process' response behaviors by employing the structure-oriented lumping (SOL) idea. To describe the residue, they offered 46 varieties of multicore seed molecules and 92 types of single-core seed molecules. To describe the molecular makeup of residues, 7004 different types of molecular lumps were created. These illustrations demonstrate the difficulty of the assignment, as was previously stated [25-26].

As a result, empirical modelling methods seem to be the most effective way to determine product yields and are preferred in refining practice. Different correlations for calculating delayed coker yields have been developed by petroleum industry companies and consultants; however, these correlations have recently been applied to take into account yields and product characteristics that are useful in preliminary studies for deciding when a delayed coker is desired to be incorporated in an existing or new refining scheme [27].

A delayed coker may be modelled using an empirical technique, which is based on the fact that the coke yield and the feed's CCR have a strong correlation, as shown in Table (1).

Table 1 Typical Coke Yields from Delayed Coking [28].

Carbon residue (wt%)	API gravity (deg)	Coke yield (wt%)
1	NR[a]	0
5	26	8.5
10	16	18
15	10	27.5
20	6	35.5
25	3.5	42

[a]NR = not reported

It is found that the coke yield and the other product yields correlate more favorably to the CCR because it has been shown that CCR is a superior predictor than feed American Petroleum Institute (API) gravity, it is exclusively employed for the mass balance [28].

Coking Correlations

Hankwert and Gary correlaions

Hankwert and Gary, A number of correlations are provided in the book by Gary and Hankwert to determine the yields of coke, gas (C4--), gasoline (C5-400 °F), and gas oil (400-925 °F) in weight percent and gasoline and gas oil in volume percent, respectively. Along with the API gravity, they also reported a typical split of naphtha and gas oil. The yield data from commercial and pilot plants with a coke drum pressure of 35–45 psig were utilized to create the correlations. The feed was a straight-run residual with an API of less than 18°. There are correlations created as following [29]:

$$\text{Gas (wt \%)} = 7.8 + 0.144(\text{CCR, wt \%})$$

$$\text{Naphtha (wt \%)} = 11.29 + 0.343(\text{CCR, wt \%})$$

$$\text{Coke (wt \%)} = 1.6(\text{CCR, wt \%})$$

$$\text{Gas oil (wt \%)} = 100 - \text{gas} - \text{naphtha} - \text{coke}$$

The weight and volume percent are based on the net fresh feed to the coking unit. To transform naphtha and gas oil yields from weight to volumetric basis, the following equations were used:

$$\text{Naphtha (vol \%)} = 186.5/\ (131.5 + \text{API})\ (\text{naphtha, wt \%})$$

$$\text{Gas oil (vol \%)} = 155.5/\ (131.5 + \text{API})\ (\text{gas oil, wt \%})$$

Where API is the gravity of the feed. To split the coker naphtha into light and heavy, the authors proposed:

Light naphtha = 35.1 vol %, 65° API

Heavy naphtha = 64.9 vol %, 50° API

Similarly, to split the coker gas oil, they proposed

Light gas oil (LCGO) = 67.3 vol %, 30° API

Heavy gas oil (HCGO) = 32.7 vol %, 13° API

Gary and Handwerk's correlations do not include terms to account for the operating conditions, and the only independent variable is the CCR of the feedstock. The application of this method, in general, leads to very unpractical and inaccurate results [30].

Maples correlaions

This approach also uses the residual carbon content of the feed as a single independent variable. Correlations were obtained from an extensive database collected in delayed coking plants at typical operating conditions for a wide range of feeds Figure (2). Feed properties range between 1.4 and 21.5° API gravity and CCR content between 2.84 and 25.5 wt %. The correlations are [31]:

$$\text{Gas yield (wt \%)} = 0.2745(\text{CCR, wt \%}) + 4.1264$$

$$\text{Naphtha yield (wt \%)} = -0.0082\ (\text{CCR, wt \%}) + 17.025$$

$$\text{Gas oil yield (wt \%)} = -1.9418\ (\text{CCR, wt \%}) + 79.225$$

$$\text{Coke yield (wt \%)} = 1.6755(\text{CCR, wt \%}) - 0.3765$$

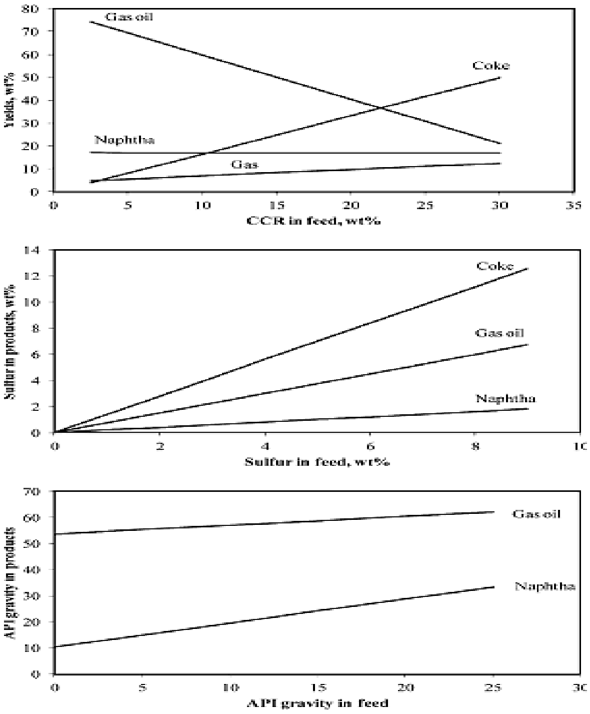

Figure 2 Maples yields of the delayed coking plant [31].

Castiglioni correlaions:

A graphical technique was given by Castiglioni to calculate delayed coker yields as a function of two feed characteristics (API gravity and CCR) and three operating factors (combined feed rate, drum pressure, and drum temperature). Dry gas, gasoline, gas oil and coke are the products of the delayed coker. The lighter fraction and propane are the two components of dry gas, whereas butanes and the C5-400 °F fraction are the two components of gasoline. The process consists of three steps. Using the feed CCR and the operation temperature, the coke yield is first estimated using a reference pressure of 0 psig and the actual drum pressure. A number of correction factors are calculated in the second stage based on the calculated combined feed rate (CFR). In the third stage, a second series of correction factors are obtained as a function of the operation CFR. Finally, gasoline and coke factor corrections are obtained as a function of the yield of gasoline and coke, respectively. Castiglioni's charts do not allow for prediction at pressures above 30 psig or feedstocks with CCR higher than 25%; therefore, for such conditions, extrapolation is necessary, which is an important limitation of this approach [32-33]:

Smith et al correlaions:

The basis of the Smith et al. correlation comes from Gary and Handwerk. They developed equations based on the feed CCR to estimate the yields of coke, gas, gas oil, and naphtha. The effect of pressure (P) was considered in the correlations as seen as follows [34]:

$$\text{Gas (wt \%)} = 7.4 + 0.1CCR + 0.8((P - 15)/20)$$

$$\text{Naphtha (wt \%)} = 10.29 + 0.2CCR + 2.5((P - 15)/20)$$

$$\text{Coke (wt \%)} = 1.5CCR + 3((P - 15)/20)$$

$$\text{Gas oil (wt \%)} = 100 - \text{gas} - \text{naphtha} - \text{coke}$$

Where P is the coke drum pressure in psig.

Volk et al correlaions

Volk et al. proposed a set of linear correlations to predict the product yields as function of the micro carbon residue (MCR, in wt %), temperature (T, in °F), pressure (P, in psia), and liquid space velocity (LSV, in min^{-1}). The range of operating conditions used to develop the correlations is 900−950 °F, 6−40 psig, and MCR from 16 to 29 wt %. The correlations are [35]:

$$\text{Liquid (wt \%)} = - 1.1139MCR + 0.0419T - 0.2897P + 1103.08LSV + 41.59$$

$$\text{Coke (wt \%)} = 0.9407MCR - 0.0609T + 0.1529P - 319.759LSV + 65.075$$

$$\text{Gas (wt \%)} = 0.1729MCR + 0.0191T + 0.13646P - 786.319LSV - 6.762$$

$$\text{Naphtha (wt \%)} = - 0.3086MCR + 0.0137T + 0.1571P - 819.63LSV + 16.461$$

$$\text{Diesel (wt \%)} = - 0.3339MCR - 0.02635T - 0.0392P + 70.957LSV + 50.452$$

$$\text{Gas oil (wt \%)} = - 0.4714MCR + 0.0546T - 0.4076P + 1851.76LSV - 25.315$$

The authors stated that the correlations could not be used to predict yields from industrial cokers, because of the lower liquid yields obtained in the micro reactor, as compared to those observed in refineries, which becomes worse at the lowest feed rate. Also, the correlations include the effects of LSV, which has a different meaning than that for commercial units. For these reasons, the following correction was proposed to derive product yields [35]:

$$\text{Coke* (wt \%)} = 0.91\text{coke}$$
$$\text{Gas* (wt \%)} = 0.82\text{gas}$$
$$\text{Liquid* (wt \%)} = 100 - (\text{coke*} + \text{gas*})$$
$$\text{Naphtha* (wt \%)} = 0.75\text{Naphtha (liquid*/liquid)}$$
$$\text{Diesel* (wt \%)} = 0.90\text{diesel (liquid*/liquid)}$$
$$\text{Gas oil* (wt \%)} = \text{liquid*} - (\text{gasoline*} + \text{diesel*})$$

Previous researches show many correlations that examined the applicable functions to estimate the products yield of the delayed coking process; they are mainly connected to conradson carbon residue CCR wt% value as a feed properties and operational variables (Recycle Ratio wt% , Heater outlet Temperature °C and Coke drum pressure KPa).

Delayed coker unit simulation:

This study has been constructed using Aspen HYSYS ver.12.1 and regression analysis by MICROSOFT EXCEL 2010 to obtain a new correlations to predict the product yields from delayed coker unit with a wide applicable range of operating variables which is more reliable with refineries data and a simulation molding of delayed coker unit has been accomplished to maximize the gas oil yield for the Middle East Oil Refinery by optimization of process variables at the expense of decreasing the coke product yield that is lead to a net economic profits.

To optimize the delayed coking process variables, it has been studied the effects of each process variable on product yields while keeping the essential concepts of process safety terms in mind. The Aspen HYSYS modelling simulation technology tool Ver 12.1 and MICROSOFT OFFICE 2010 Linear programming were used for this study's evaluation of process factors and their influence on delayed coker unit product yields to find the optimal conditions for maximization of gas oil yield wt% and minimization of coke yields wt%.

Feed properties and compositions:

The simulated feed to delayed coker unit based on the Middle East Oil Refinery at Alexandria in Egypt which receives its crude oil from a blend of 50%: 50 % of Arabian Light Crude Oil and Arabian Heavy Crude Oil. The vacuum residue produced with the following composition and properties shown in Table (2) the feed cut point is 538 C° and fed with a temperature average 200 C° to the delayed coker unit.

Table 2 Design feed properties and composition.

Item	API °	Rate Kg/hr	Viscosity @99C°,Cst	Sulfur, wt%	Nitrogen, wt%	Nickel, wppm	Vanadium, wppm	Conradson, wt%
Value	5.22	156169	2500	5	0.43	44	143	22.87

Product specifications:

The running delayed coker unit in Middle East Oil refinery product specifications was constructed in the simulation tool to fit the refinery policies as following in Table (3-4):

Table 3 Design products distillation.

Distillation	Naphtha, C°	Light Gas Oil, C°	Heavy Gas Oil, C°
IBP	17	167	255
5%	55	194	358
10%	64	203	385
30%	83	228	433
50%	101	261	463
70%	120	293	489
90%	148	331	507
95%	162	344	516
EP	179	358	549

Table 4 C3 / C4 product compositions.

Component	LPG, wt%
H_2O	0.002
H_2	0.0
H_2S	5.5
Methane	0.0
Ethylene	0.004
Ethane	1.2
Propylene	14.7
Propane	33.8
1-butene	18.6
i-butane	4.8
n-butane	19.9
C_5+	1.5
TOTAL	100.0

Completed modeling of delayed coker unit:

The process of building a delayed coker model covered in the starting guide, which will include setting up a heavy crude feed with a petroleum assay, configuring a delayed coker unit operation, calibrating the coker unit, and putting together a recycle network [36].

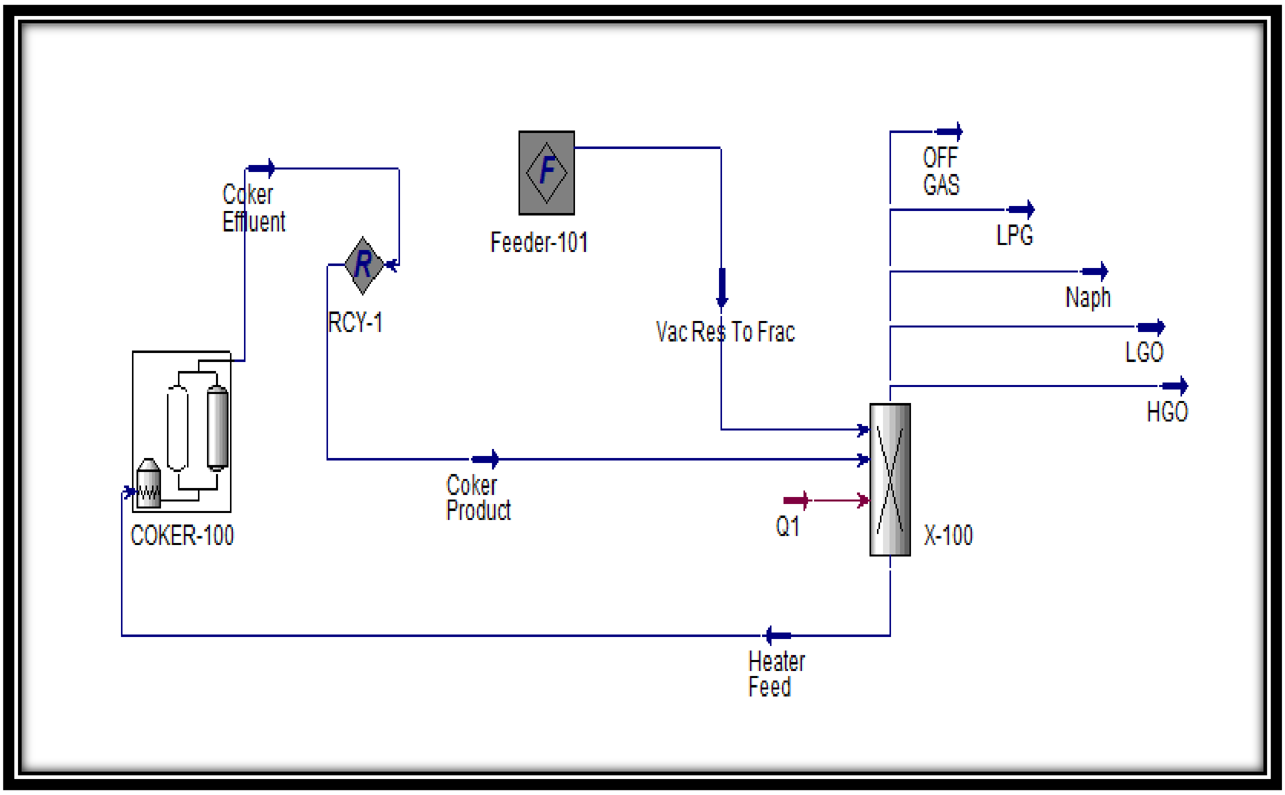

Figure 3 Completed flowsheet for a delayed coker process Combined Feed Ratio.

The completed modelling of the delayed coker unit in Figure (3), the study targeted the optimization of process variables in order to maximize the gas oil yield produced by the delayed coker unit by evaluating the effect of changing of each variable on the coker product yields. The design conditions were used based on the Middle East Oil refinery data, then finding and logging the results product yields.

Compatibility of HYSYS results

The design conditions that were relevant to the study, such as Feed rate, Feed specifications, Product specifications, Drum outlet temperature, Drum pressure, Recycle ratio, and Cycle time provided into the delayed coker complex simulation and solving it. As shown in Table (5) the provided design process variables and the design product yields wt% shown in Table (6). The outcomes are depicted in Figure (4).

Table 5 Design condition of process variables.

Variable	Feed Rate, m^3/hr	Coke Drum Pressure, barg	Coke Drum Overhead Temperature, C°	Recycle Ratio, wt%	Cycle Time, hr
Value	156	1.034	446	20	16

Table 6 Design product yields.

Product	Gases + Naphtha, wt%	Gas Oil, wt%	Coke, wt%
Value	19.99	54.88	25.13

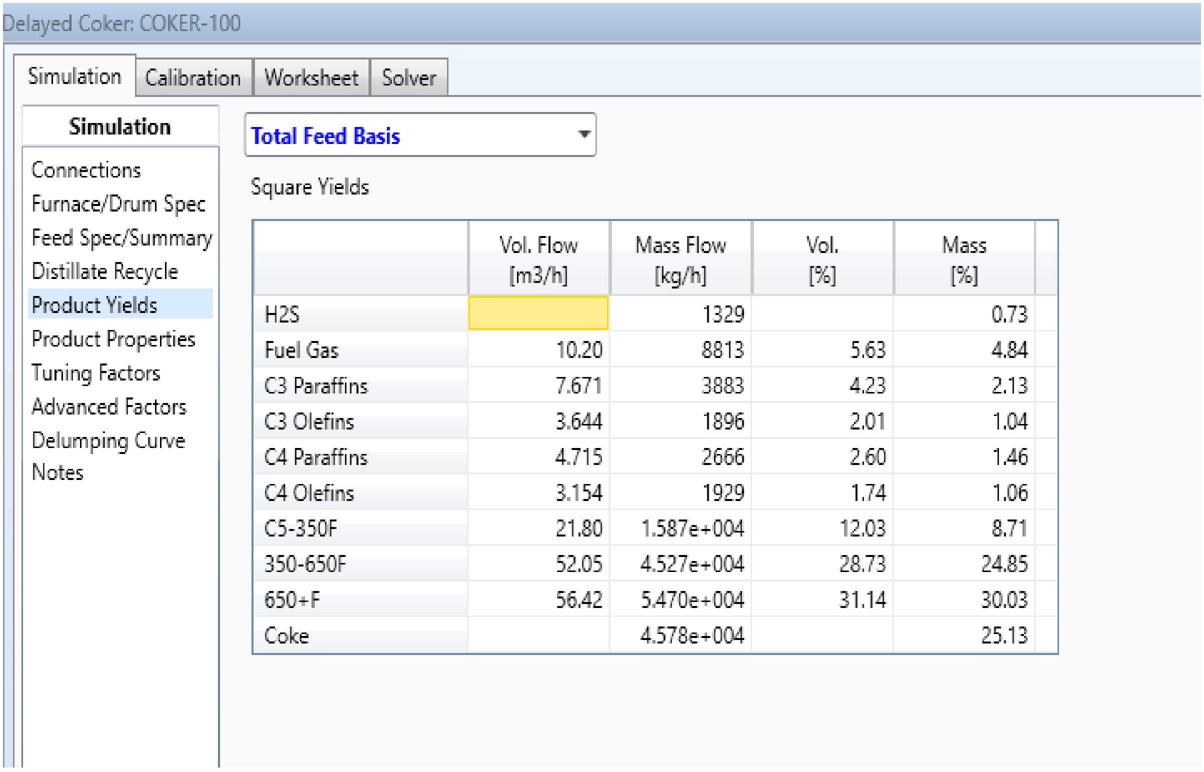

Figure 4 Product yields resulted from simulation using the design parameters by HYSYS.

The indicated product yields mass % from HYSYS simulation of delayed coker unit with design basis of Middle East Oil refinery concluded as follow Table (7) which is similar to the actual design product yields.

Table 7 Product yields from delayed coker unit simulated by HYSYS at design conditions.

Product	(C_5--) + (C_5 - 350 F) ,Gases + Naphtha, wt%	(350-650 F) + (650+), Gas Oil, wt%	Coke, wt%
Value	19.99	54.88	25.13

Research Methodology

The effect of change in process variables:

The study involved a change in each process variables (Recycle ratio wt%, Heater outlet temperature C° and Coke drum pressure KPa) and log out the resulted product yields from simulation.

Recycle ratio

The Aspen HYSYS simulation results from changing the recycle ratio wt% on the delayed coker product yields while maintaining all other variables as constant shown in Figures (5-6).

The effect of the recycle ratio wt% showed in Table (8) and Figures (5-6). It's indicated that lowering the recycle ratio wt% increases the gas oil yield wt% produced by the delayed coker unit. It had been noted that further more lowering in the recycle ratio wt%, the gas oil yield wt% increases more rabid.

It is noted that keeping the recycle ratio as close to 3 to 5% would add a benefits from the delayed coker cracking. There are also concerns about running delayed coker heaters without a distillate recycle stream because the velocity of flows inside the heater passes will begin to drop, but this can be solved by increasing the steam flow rate or decreasing the heater passes diameter.

Table 8 Recycle ratio wt% change versus delayed coker product yields wt%.

Recycle Ratio, wt %	20%	15%	10%	5%	3%
Gases + Naph Yield, wt%	19.99	19.75	19.47	19.23	19.14
Gas Oil Yield, wt%	54.88	55.12	55.58	56.10	56.34
Coke, wt%	25.13	25.13	24.95	24.67	24.52

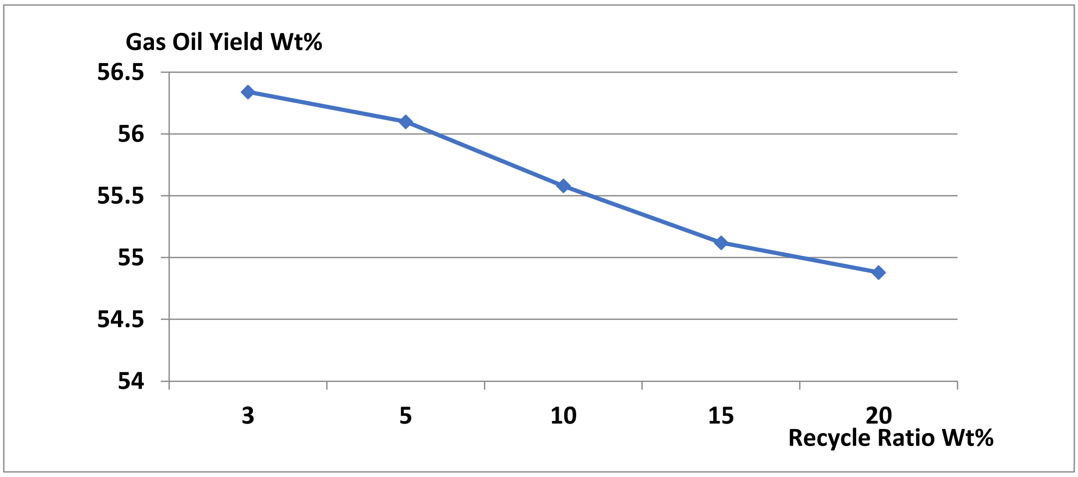

Figure 5 The effect of change in recycle ratio wt% to delayed coker gas oil yield produced wt%.

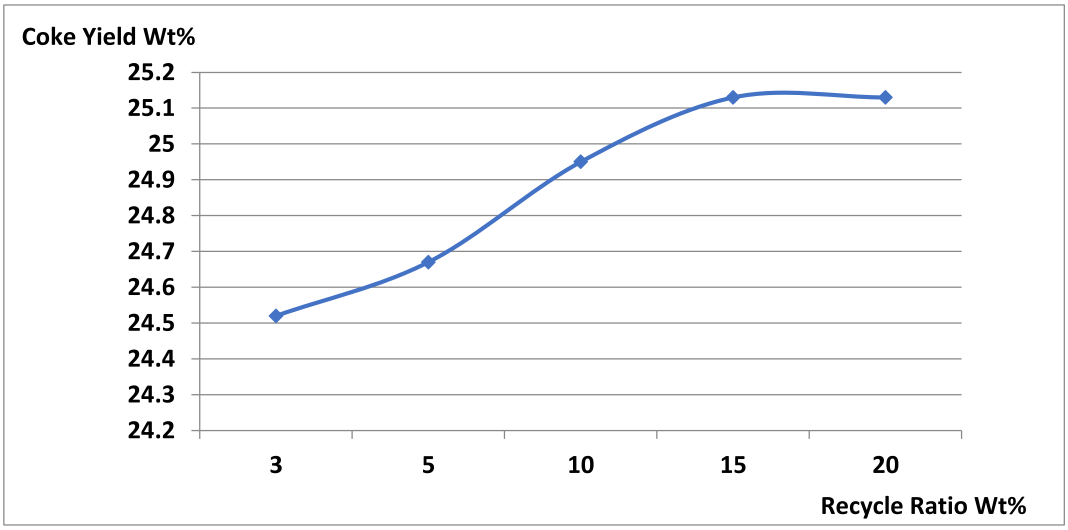

Figure 6 The effect of change in recycle ratio wt% to delayed coker coke yield produced wt%.

Temperature effect on thermal cracking

One of the essential variables in the delayed coker process is the cracking temperature, more specifically the furnace outlet temperature. Some processes are dependent on the temperature of the coke drum inlet, but the cracking reaction is endothermic, thus there will be a temperature difference between the coil outlet or furnace output temperature and the temperature of the coke drum inlet.

As previously stated, the furnace exit temperature is maintained between 480°C and 520°C. Changing in the furnace outlet temperature C°, as it rises, the delayed coker yields begin to fluctuate. The gas oil yield will increase significantly, while the gases and coke outputs will decrease.

Following this increase in the furnace outlet temperature C° beyond the discussed limits, more coke layer will build inside the heater coils as cracking will be faster and may begin to occur inside the heater coils. To avoid going via the decoker systems like steam / air decoking or online spalling in short intervals, it's recommended to raise the velocity of the flow inside the heater passes through the velocity steam facilities.

Using Aspen HYSYS simulation and changing in furnace outlet temperature C° and log out the delayed coker product yields wt% while holding all other variables as constant.

The Results from changing of heater outlet temperature C° of the delayed coker heaters through 480°C to 520°C and its effects on the delayed coker product yields wt%. From Table (9) and Figures (7-8) it's indicated that by raising the heater outlet temperature by 5°C will increase the gas oil yield wt% by an average of 0.64 wt% with a highest increase rate in the initial of the raising in heater outlet temperature C° starting from 480°C. Furthermore increase in heater outlet temperature C° will also increase the gas oil yield wt% but with lower rate near 520°C. It's also indicated that while increasing in the heater outlet temperature by 5°C, the gases plus naphtha yield increased by an average of 0.26 wt% and coke yield wt% decreased by an average of 0.9 wt%.

The rate of decreasing in coke yield wt% produced from delayed coker unit while increasing the cracking temperature will be higher in the first raising in the temperature than in further increasing. From the study, it's noted that keep the heater outlet temperature as high as possible within the discussed range will allow higher degree of cracking reactions to take place for vacuum residue while taking care from the coke formation rate inside the heater passes. It should be noticed and lowered by increasing the velocity a little more by the means of steam.

Table 9 Heater outlet temperature °C change versus delayed coker product yields wt%.

Heater Outlet Temperature, °C	**480**	**485**	**490**	**495**	**500**	**505**	**510**	**515**	**520**
Gases + Naph Yield, wt%	19.42	19.71	19.99	20.27	20.53	20.78	21.04	21.28	21.52
Gas Oil Yield, wt%	53.53	54.21	54.88	55.53	56.18	56.83	57.46	58.09	58.71
Rate of increase in gas oil yield, wt%	0.68	0.67	0.65	0.65	0.65	0.63	0.63	0.62	0.62
Coke, wt%	27.05	26.08	25.13	24.20	23.29	22.39	21.50	20.63	19.77
Rate of decrease in coke yield, wt%	0.97	0.95	0.93	0.91	0.90	0.89	0.87	0.86	0.85

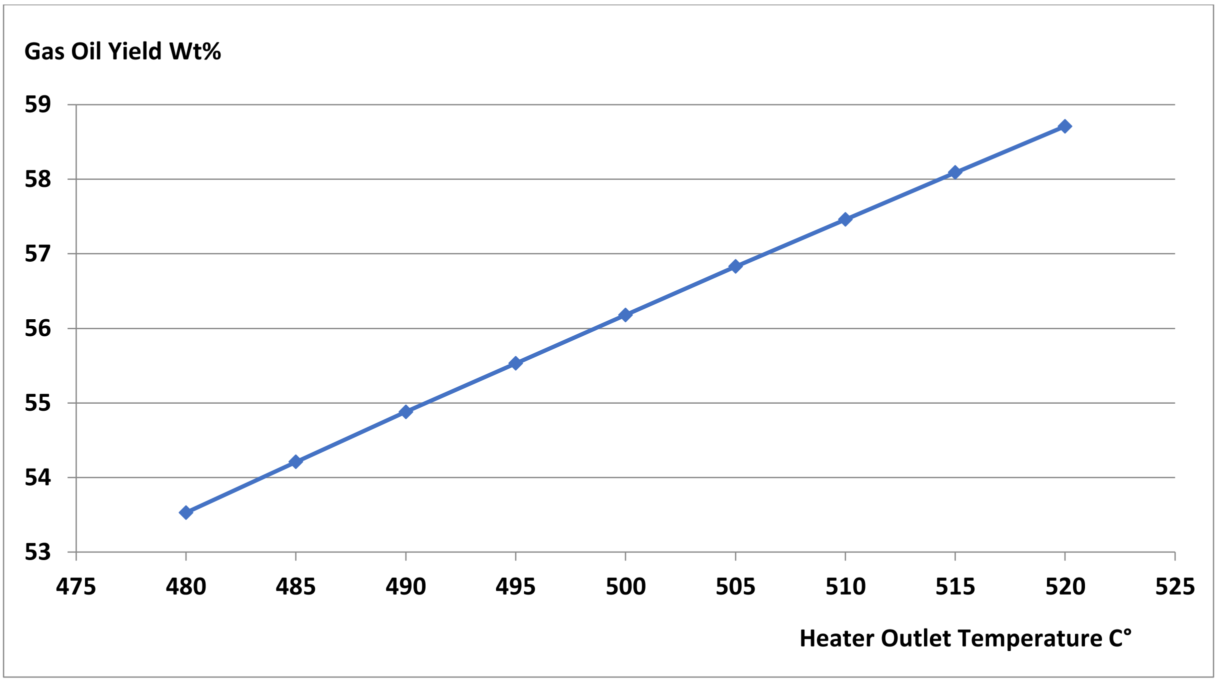

Figure 7 The effect of change in Heater outlet temperature °C to delayed coker gas oil yield produced wt%.

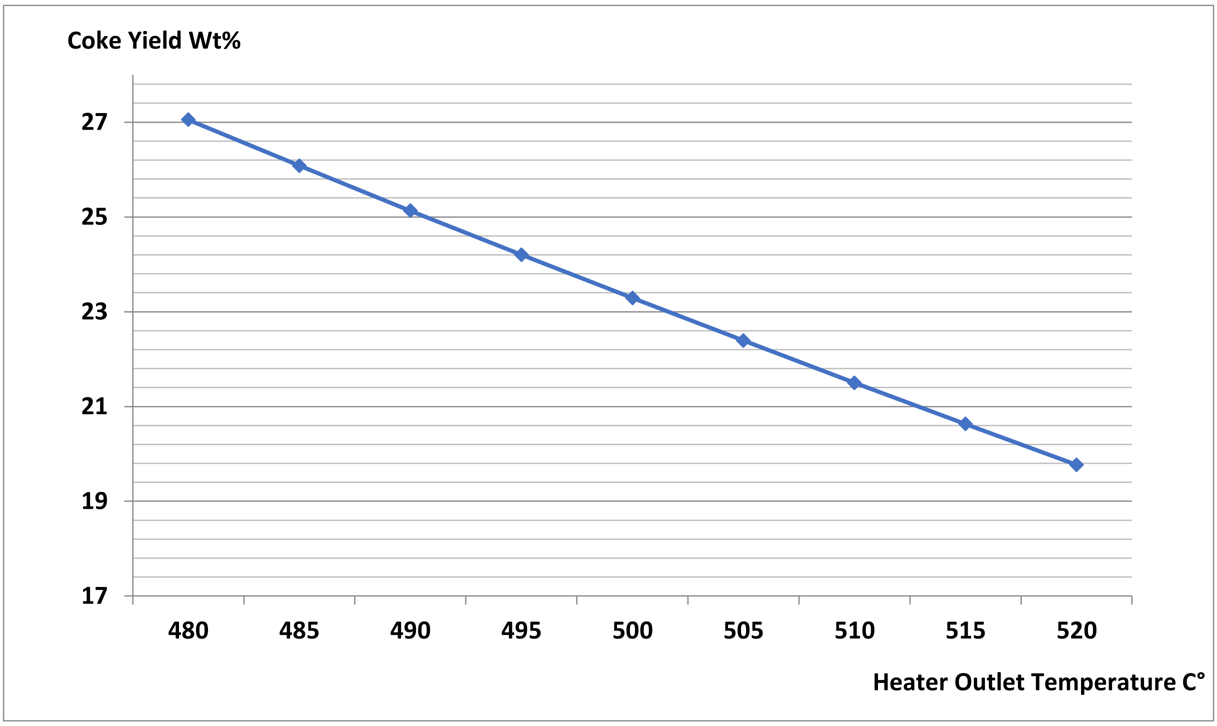

Figure 8 The effect of change in Heater outlet temperature °C to delayed coker coke yield produced wt%.

4.1.3. Pressure effect on thermal cracking

The fractionator overhead pressure control valve could be used to regulate the drum overhead pressure. The cracking reactions get longer residence time inside the coke drum as the drum pressure increases. Condensation and polymerization reactions are accelerated, allowing for more coke formation and a decrease in gas oil production due to higher condensation inside the coke drum. Lowering the coke drum overhead pressure allows vapors of product liquid yields to build more inside the main fractionator, resulting in more gas oil wt% and lowering the volatile component matter (VCM) in coke product, preventing hydrocarbon liquid loss. Changes in drum pressure should be closely monitored because a decrease in drum pressure may result in the initiation of foam level appearance, particularly at the end of a coke drum cycle, because foam is formed based on the amount of VCM and liquid layer inside the coke drum, and these amounts of VCM and liquid layer inside the coke drum are much higher at the end of each cycle. Starting antifoam injection and consumption should be raised if the pressure is dropping towards the end of the drum cycle, thus it is preferable to make the drum pressure as steady as possible during the cycle period.

The effect of change in coke drum pressure on the delayed coker product yields wt% modeled by Aspen HYSYS simulation while keeping all the other variables as constant figures.

It has been observed that adjusting the coke drum pressure has little effect on product yields wt% when compared to changing other process variables. From Table (10) and Figures (9-10),its noted When the coke drum pressure was increased by 50 KPa, The coke yield wt% increased by 0.07% on an average, the gas oil yield wt% declined by 0.025% on an average and the gases wt% increased by 0.015% on an average. It is noted that the pressure be kept as low as feasible in order to maximize the gas oil production while lowering the coke yield produced by the delayed coker unit. Due to the minor effect of change in coke drum pressure on the delayed coker unit product yields it is preferred to keep the pressure as constant to prevent any disturbance in coke drum pressure that might causing foam carry over to main fractionating section especially at the end of coke drum cycle.

Table 10 Coke drum pressure in KPa change versus delayed coker product yields wt%.

Coke Drum Pressure, KPa	100	150	204.7	250	300	350	400	450
Gases + Naph Yield, wt%	19.95	19.98	19.99	20	20.02	20.03	20.05	20.05
Gas Oil Yield, wt%	54.93	54.9	54.88	54.86	54.83	54.82	54.79	54.78
Coke, wt%	25.12	25.12	25.13	25.14	25.15	25.15	25.16	25.17

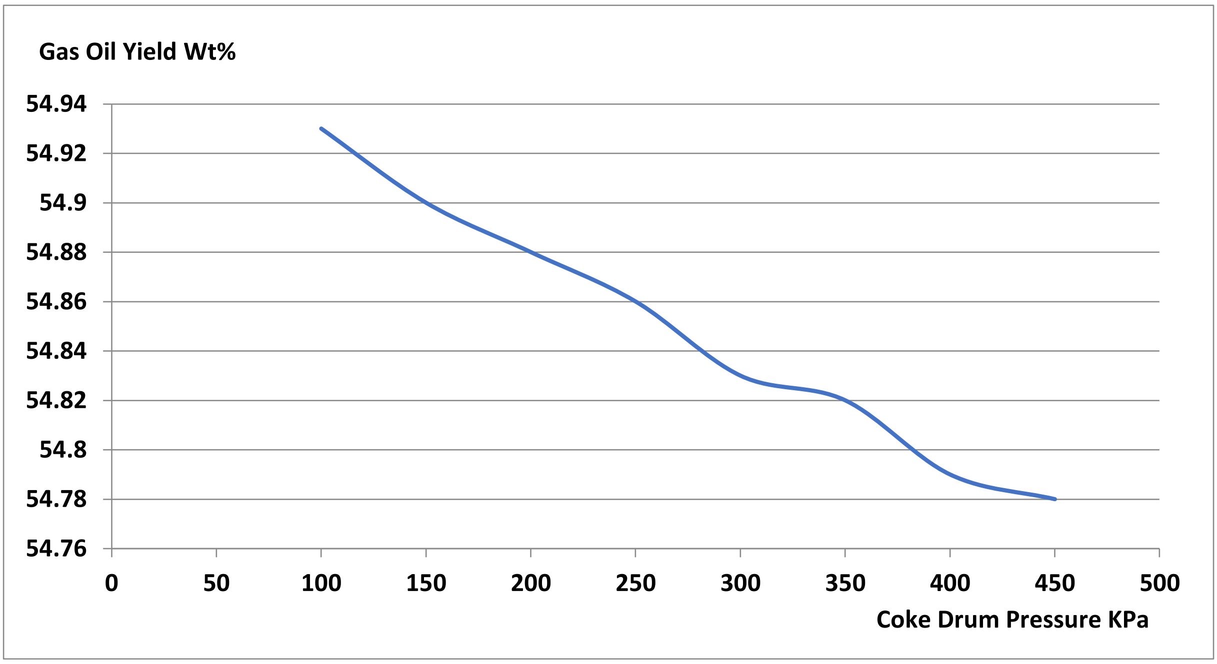

Figure 9 The effect of change in coke drum pressure KPa to delayed coker unit gas oil yield produced wt%.

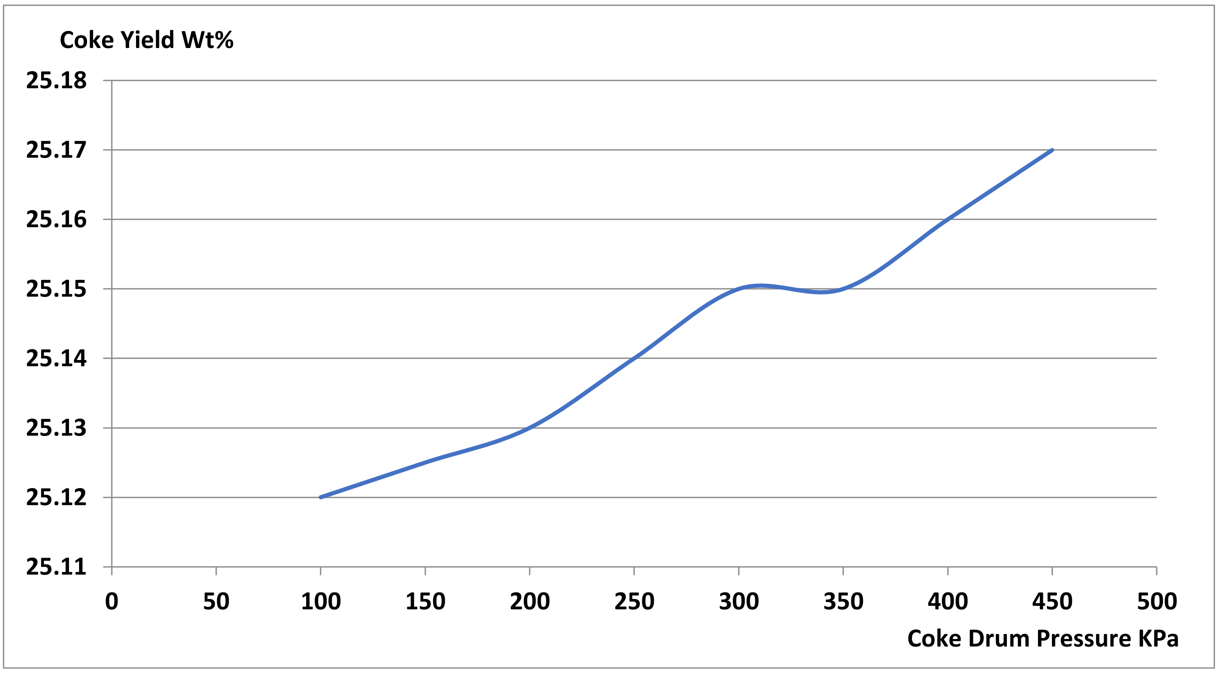

Figure 10 The effect of change in coke drum pressure KPa to delayed coker unit coke yield produced wt%.

Regression Analysis

Regression analysis

The resulted data from the previous simulation collected in Table (11) and allow for regression analysis to be performed finding the best fit correlations to predict the delayed coker unit product yields wt%. The table shows the effect of change in all process variables together on the delayed coker unit product yieldswt% 50 trials for random change between operating conditions had been logged out with the resulted yields.

Table 11 The HYSYS results coker product yields wt% while changing in process variables

No. of Trial	Recylce Ratio, wt%	Temperature, C°	Pressure, KPa	Gas Oil, wt%	Coke, wt%	Gases+Naph, wt%
1	3	480	100	55.05	26.4	18.55
2	3	480	200	54.97	26.43	18.6
3	3	480	300	54.91	26.46	18.63
4	3	480	400	54.85	26.48	18.67
5	3	480	500	54.8	26.51	18.69
6	3	490	100	56.41	24.49	19.1
7	3	500	100	57.76	22.61	19.63
8	3	510	100	59.1	20.77	20.13
9	3	520	100	60.39	18.97	20.64
10	5	480	100	54.82	26.54	18.64
11	10	480	100	54.29	26.82	18.89
12	15	480	100	53.83	27.02	19.15
13	20	480	100	53.57	27.04	19.39
14	5	490	200	56.11	24.66	19.23
15	5	500	200	57.45	22.79	19.76
16	5	510	200	58.78	20.95	20.27
17	5	520	200	60.08	19.16	20.76
18	10	480	200	54.23	26.84	18.93
19	10	490	200	55.58	24.95	19.47
20	10	500	200	56.91	23.09	20
21	10	510	200	58.21	21.27	20.52
22	10	520	200	59.49	19.49	21.02
23	15	500	200	56.43	23.28	20.29
24	20	500	200	56.19	23.29	20.52
25	15	510	200	57.72	21.48	20.8
26	20	510	300	57.41	21.52	21.07
27	20	520	400	58.62	19.79	21.59
28	3	490	204.7	56.34	24.52	19.14
29	4	490	204.7	56.22	24.6	19.18
30	10	490	204.7	55.58	24.95	19.47
31	15	490	204.7	55.12	25.13	19.75
32	20	490	204.7	54.88	25.13	19.99
33	20	480	204.7	53.53	27.05	19.42
34	20	485	204.7	54.21	26.08	19.71
35	20	490	204.7	54.88	25.13	19.99
36	20	495	204.7	55.53	24.2	20.27
37	20	500	204.7	56.18	23.29	20.53
38	20	505	204.7	56.83	22.39	20.78
39	20	510	204.7	57.46	21.5	21.04
40	20	515	204.7	58.09	20.63	21.28
41	20	520	204.7	58.71	19.77	21.52
42	20	490	100	54.93	25.12	19.95
43	20	490	150	54.9	25.12	19.98
44	20	490	204.7	54.88	25.13	19.99
45	20	490	250	54.86	25.14	20
46	20	490	300	54.83	25.15	20.02
47	20	490	350	54.82	25.15	20.03
48	20	490	400	54.79	25.16	20.05
49	20	490	450	54.78	25.17	20.05
50	3	480	100	55.05	26.4	18.55

From the resulted previous Table (11) with regression analysis using MICORFOST OFFICE EXCEL shown in Figures (11:13) it's allowed to determine the correlations between delayed coker unit process variables and product yields wt%.

A1 fx SUMMARY OUTPUT

	A	B	C	D	E	F	G	H	I
1	SUMMARY OUTPUT								
2									
3	*Regression Statistics*								
4	Multiple F	0.999019							
5	R Square	0.99804							
6	Adjusted I	0.997909							
7	Standard E	0.080898							
8	Observati	49							
9									
10	ANOVA								
11		*df*	*SS*	*MS*	*F*	*gnificance F*			
12	Regressio	3	149.9452	49.98174	7637.152	6.49E-61			
13	Residual	45	0.294505	0.006545					
14	Total	48	150.2397						
15									
16		*Coefficient*	*andard Err*	*t Stat*	*P-value*	*Lower 95%*	*Upper 95%*	*ower 95.0%*	*pper 95.0%*
17	Intercept	-7.79018	0.447955	-17.3905	1.35E-21	-8.69241	-6.88795	-8.69241	-6.88795
18	Recycle Ra	-0.08547	0.001635	-52.2645	6.23E-42	-0.08876	-0.08218	-0.08876	-0.08218
19	Temp	0.131472	0.000903	145.5347	8.25E-62	0.129653	0.133292	0.129653	0.133292
20	Press	-0.00048	0.000127	-3.79228	0.000442	-0.00074	-0.00023	-0.00074	-0.00023
21									
22									
23									

Figure 11 Regression analysis results gas oil product yields wt% while changing in process variables.

	A	B	C	D	E	F	G	H	I
1	SUMMARY OUTPUT								
2									
3	*Regression Statistics*								
4	Multiple F	0.99939							
5	R Square	0.998781							
6	Adjusted I	0.9987							
7	Standard E	0.085882							
8	Observati	49							
9									
10	ANOVA								
11		*df*	*SS*	*MS*	*F*	*gnificance F*			
12	Regressio	3	272.0054	90.66845	12292.73	1.47E-65			
13	Residual	45	0.33191	0.007376					
14	Total	48	272.3373						
15									
16		*Coefficient*	*andard Err*	*t Stat*	*P-value*	*Lower 95%*	*Upper 95%*	*ower 95.0%*	*pper 95.0%*
17	Intercept	114.5429	0.475553	240.8627	1.21E-71	113.5851	115.5007	113.5851	115.5007
18	Recycle Ra	0.034775	0.001736	20.03074	4.76E-24	0.031278	0.038272	0.031278	0.038272
19	Temp	-0.18382	0.000959	-191.675	3.49E-67	-0.18575	-0.18189	-0.18575	-0.18189
20	Press	0.000142	0.000135	1.055221	0.296957	-0.00013	0.000413	-0.00013	0.000413
21									
22									
23									

Figure 12 Regression analysis results coke product yields wt% while changing in process variables.

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A1 | fx | SUMMARY OUTPUT

	A	B	C	D	E	F	G	H	I
1	SUMMARY OUTPUT								
2									
3	*Regression Statistics*								
4	Multiple F	0.999751							
5	R Square	0.999502							
6	Adjusted I	0.999469							
7	Standard I	0.018622							
8	Observati	49							
9									
10	ANOVA								
11		*df*	*SS*	*MS*	*F*	*gnificance F*			
12	Regressio	3	31.34759	10.4492	30131.74	2.6E-74			
13	Residual	45	0.015605	0.000347					
14	Total	48	31.36319						
15									
16		*Coefficients*	*tandard Err*	*t Stat*	*P-value*	*Lower 95%*	*Upper 95%*	*ower 95.0%*	*pper 95.0%*
17	Intercept	-6.75269	0.103116	-65.4867	2.78E-46	-6.96038	-6.54501	-6.96038	-6.54501
18	Recycle Ra	0.050695	0.000376	134.6694	2.69E-60	0.049937	0.051453	0.049937	0.051453
19	Temp	0.05235	0.000208	251.7426	1.66E-72	0.051931	0.052768	0.051931	0.052768
20	Press	0.000339	2.92E-05	11.60792	3.98E-15	0.00028	0.000398	0.00028	0.000398
21									
22									
23									

Figure 13 Regression analysis results gases plus naphtha product yields wt% while changing in process variables.

Regression analysis resulted correlations

For the same feed properties and in the following range of operating conditions Furnace outlet temperature (480 – 520 C°), Recycle ratio wt% (3-20) and Coke drum pressure (100-500) KPa, The resulted correlations were obtained to predict the coker product yields wt% as follow:

Gas Oil wt% = 0.131472 T – 0.08547 R – 0.00048 P – 7.79018

Coke wt% = 0.034775 R – 0.18382 T + 0.000142 P + 114.5429

Gases + Naphtha wt% = 0.050695 R + 0.05235 T + 0.000339 P – 6.75269

Where R = Recycle Ratio wt%, T = Furnace Outlet Temperature C°, P = Coke Drum Pressure KPa

Optimization of the process variables

Linear Programming were solved by MICROSFOT OFFICE EXCEL 2010 to find the optimum operating conditions to achieve the maximum gas oil yield wt% and minimize the coke yield wt% Figure (14).

F19 | fx |

	A	B	C	D	E
1	Vairables				
2	Temperat	520			
3	Recylce Ra	3			
4	Pressure	100			
5					
6	Objective				
7					
8	Maximize	60.27085			
9	Minimize	19.07503			
10					
11	Constraints				
12			Inequality	RHS	
13	1	520	>=	480	
14	2	520	<=	520	
15	3	3	>=	3	
16	4	3	<=	20	
17	5	100	>=	100	
18	6	100	<=	500	
19					

Figure 14 The optimization using simplex method and linear programming results by MICROSFT EXCEL.

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The results by optimization using MICROSOFT EXCEL 2010 for the adjustment in process variables to achieve the maximization of gas oil yield wt% and minimizing the coke yield wt% from the delayed coker unit in the Middle East Oil Refinery. It's found that the Furnace outlet temperature 520 C°, Recycle ratio 3% wt% and Coke drum pressure 100 KPa achieving the maximum gas oil yield 60.27 wt% and minimum Coke yield 19.07 wt%.

Results and discussion

Optimum operating variables

As previously stated the operating conditions that achieving the maximum gas oil yield wt% and minimum Coke yield wt%, It had been selected the most effective points to work in moderate conditions not the severe one as a process safety wise to prevent any disturbances in the delayed coking process operation. Regarding the recycle ratio effect, as shown in Figure (5-6), it's recommended that the optimum recycle ratio is around 3-5%. According to Figures (7-8), the ideal heater outlet temperature is the maximum temperature ranging from 510-520°C that can be attained while maintaining safe operation for delayed coker heater. The change in coke drum pressure should be as low as possible to avoid any disturbance that might result in foam carry over of coke drum to the main fractionator. As a result, it could be concluded that the optimum coke drum pressure is 150 KPa, which achieves higher liquid yields and lowers coke yield wt% while maintaining safe coke drum operation. Table (12) and Figure (15) depicts the most likely outcomes of adjusting process variables to achieve maximization of gas oil output wt% and reduction of coke wt%.

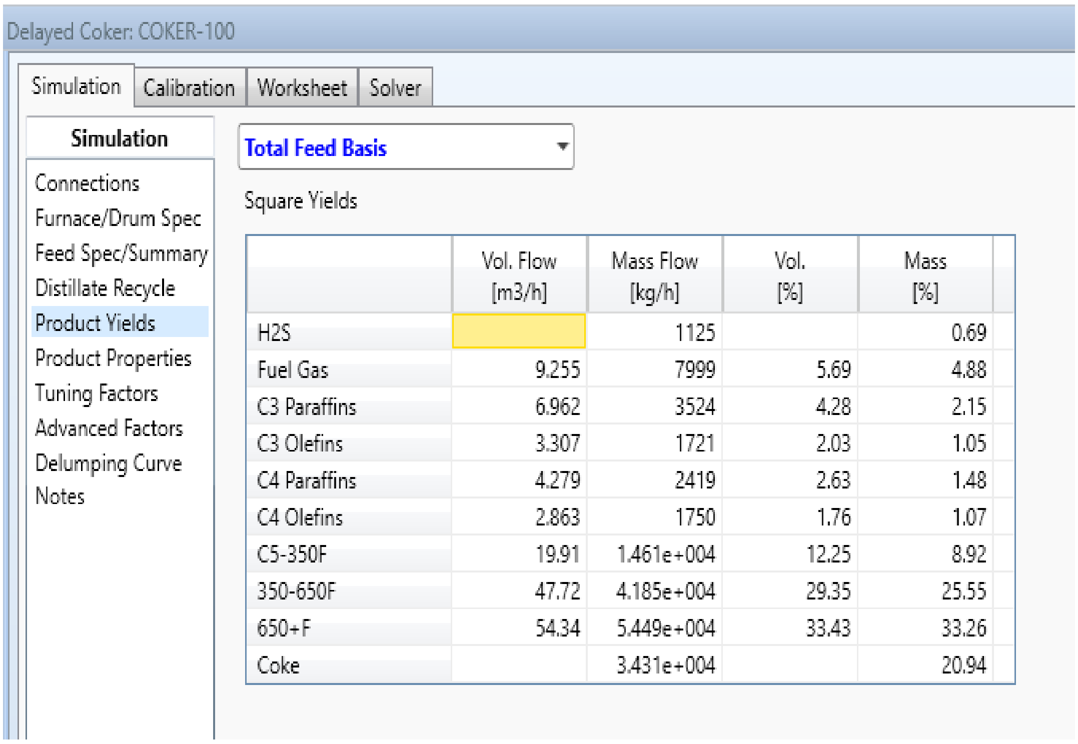

Figure 15 Product yields by HYSYS at the optimum suggested operating conditions.

The indicated product yields mass weight % from HYSYS simulation with suggested operating conditions for Middle East Oil Refinery delayed coker unit concluded as follow Table (12) which is similar to the actual design product yields.

Table 12 Product yields from delayed coker unit simulated by HYSYS at suggested conditions.

Product	**(C_5--) + (C_5 - 350 F) ,Gases + Naphtha, wt%**	**(350-650 F) + (650+), Gas Oil, wt%**	**Coke, wt%**
Value	20.25	58.81	20.94

Comparison between changing in process variables:

Table (13) and Figure (16) showed a comparison of the current operating conditions and the suggested operating variables it's noted that gas oil yield wt% increased by 4 wt% at the expense of coke yield wt%.

Table 13 Comparison between design conditions and suggested conditions.

Operating parameters	**Design condition**	**suggested condition**
Furnace Outlet Temp C°	490	510
Drum Press KPa	204.7	150
Recycle wt%	20%	5%
Feed m3/hr	156	156
Gases + Naph Yield wt %	19.99	20.25
Gas Oil Yield wt %	54.88	58.81
Coke wt%	25.13	20.94

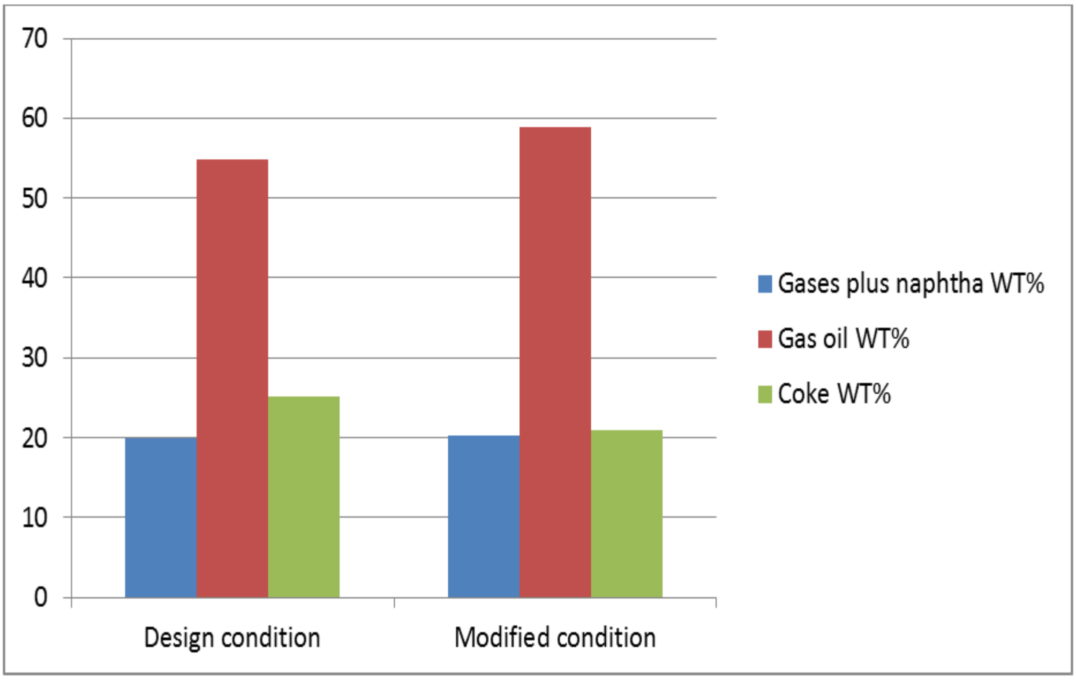

Figure 16 Product yields for design and modified conditions.

Economic study:

Economic case:

It had been studied the reflection of the suggested conditions on the economic case and how the maximizing of the gas oil yield wt% produced from delayed coker unit would effect on the economic ways. Table (14) showed the amounts of product yields in kg/hr for both design case and suggested case.

Table 14 Product yields in kg/hr in design and suggested cases.

Products in Kg/hr	Design Case	Modified Case
Gases and Naphtha	32726.83	33151.82
Gas Oil	89907.37	96349.03
Coke	41176.54	34310.08

The average of the prices of gas oil and coke products in 2022 is shown in Table (15).

Table 15 The average of the prices of gas oil and coke products in 2022 [37].

Price in $/Ton	Coke	Diesel	HGO
May	387.52	1,087.30	837.22
June	401.05	1,285.00	989.45
July	393.50	1,098.05	845.50
August	380.10	997.05	767.73
Average	390.54	1,116.85	868.975

The average of the prices for products is 992.91 $ per Ton for diesel oil and 390.54 $ per Ton for petroleum coke based of the world oil prices and the ministry of petroleum and mineral resources in Egypt announcements till August 2022. [37-38]

It's calculated the net benefits from increasing the gas oil yield and decreasing the coke yield from the delayed coker unit while neglecting the minor change in gases plus naphtha wt%.

Calculations:

Gas oil calculations:

- The average density of Gas Oil is 926.54 kg/m3.
- The amount of Gas Oil produced per year in design case is 89,907.37 * 24 * 360 = 776,799.68 Ton per year.
- The amount of Gas Oil produced per year in modified case is 96,349.03 * 24 * 360 = 832,455.62 Ton per year.
- The net difference in Gas Oil Produced in ton is 832,455.62 – 776,799.68 = 65,655.94 Ton per year.
- The net profit from increasing in Gas Oil yield is 65,655.94 * 992.91 = 65,190,439.39 $ per Year.

Petroleum coke calculations:

- The amount of Coke produced per year in design case is 41,176.54 * 24 * 360 = 355,765.3 Ton per year.
- The amount of Coke produced per year in modified case is 34,310.08 * 24 * 360 = 296,439.09 Ton per year.
- The decreasing in coke yield per year is 355,765.3 - 296,439.09 = 59,326.21 Ton per year.
- The net loss from decreasing in coke yield is 59,326.21 * 390.54 = 23,169,258.05 $ per year.

Utilities consumption calculations:

- The estimated amount of **Fuel Gas** required to increase the heater outlet temperature = 300 m3/hr * 24* 360 = 2,592,000.0 m3
- The amount of Fuel Gas = 2,592,000.0 / 28.32 = 91,525.42 MMBtu
- The total cost of Fuel Gas consumption = 91,525.42 * 7.27 = 665,389.83 $ per year.
- The estimated amount of **Steam** consumption required to increase the pass velocity = 320 Kg/hr * 24 * 360 = 2,764,000.0 Kg
- The amount of steam in lbs = 864,000.0 / 0.45 = 6,144,000.0 lbs
- The total cost of steam consumption = 6,144.0 * 12.29 $ = 75,509.76 $ per year.

The total net approximate profit from increasing the gas oil and decreasing the coke yield is 65,190,439.39 - 23,169,258.05 - 665,389.83 - 75,509.76 = **41,280,281.75 $ per year.**

Conclusion

- The importance of the conversion units in oil processing usually calls for a challenge in improvement in production, economic, environmental and safety studies. This case study is established to enhance the delayed coker unit production by increasing the gas oil yield wt% instead of coke yield wt%.
- The study has been done by modeling the delayed coker unit using Aspen HYSYS Ver 12.1 software with a design condition based on refinery data of Middle East Oil Refinery that provided to study the optimized operating variables to achieve the maximization of gas oil yield from delayed coker unit. Simulation and analysis of each operating parameter has been performed by Aspen HYSYS Ver 12.1.
- This work included the study of changing each operating variable and its effect in the delayed coker unit product yields and regression analysis had been performed By MICROSOFT OFFICE EXCEL 2010 which resulted with new correlations to predict delayed coker unit product yields.
- The study included the optimization selection of the most appropriate operating variables for delayed coker unit to achieve maximization of gas oil yield wt%. The optimization has been performed using linear programming and Simplex method by MICROSOFT OFFICE EXCEL 2010.

The suggested modified operating variables show that it could be achieved to increase the gas oil yield by around 4 wt% instead of coke by-product. Economic study had been performed and the results showed that it could save up to 40 million $ per year by optimized the operating variables.

Funding sources

None

Conflicts of interest

There are no conflicts to declare.

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