Evaluation and simulation of different crude oil in Delayed Cocker Unit (DCU) in Khartoum Refinery

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Abstract

In this study, a computer simulation and salutary analysis was conducted for two types of crude oil produced in Sudan in order to reach the goal of the ideal mixing ratio between the heavy crude (Dar / Fula) to be used as a substitute for crude oil (Fulla blend) in the delayed coke production unit at the Khartoum Refinery in Sudan which is now use only fulla blend. Based on the results of laboratory testing and computer simulations and lab analysis performed, blending 50% of DAR blend with 50% of Fula blend ore in the delayed coke production unit achieved a good improvement in the specifications and quantities of the products such as xyz which is better than that achieved by using fulla blend in DCU with the potential for a certain negative impact and limitations according to the variance in the physical and chemical properties of the two types of ore used that must be consider.

Keywords: coke, refinery, DCU, blend, crude oil.

1. INTRODUCTION

Crude oil is one of the raw materials that formed under the layers of the earth according to specific conditions and composed of hydrocarbon materials which are distilled to recover many types of products which have wide range of use in the life, according to a range of molecular weights (e.g. petrol and diesel)[1]. The heavier bottom component of crude oil is generally vended into the burner energy or cellarage energy request (for shipping) as a Heavy Energy oil as fuel (HFO). Therefore, only a cut of the crude oil after refining can be distilled to serve the automotive energy needs. It's still possible using a process called delayed coking to crack the heavy petroleum remainders to lighter products, thereby supplementing the product of automotive energies

and producing a solid carbon called coke. Delayed coking is one of the chemical engineering unit processes used in numerous petroleum refineries. The main idea of the delayed coking unit is to convert the residual products with low value to lighter products have high value and to produce a coke product. The coking process consists of thermal cracking, condensation, and polymerization process that do both in sequence and contemporaneously. The coker furnace supplies the necessary heat to initiate vaporization and cracking, while the cracking and polymerization process are completed in the coke reaction vessel, therefore the term (delayed coking). The high- molecular-weight gasoil cut and asphaltene materials are cracked into small cuts of hydrocarbons and heavy carbon products (coke). The light intermediate products formed during the cracking process are further cracked, producing low- molecular- weight components, similar as hydrogen, a wide range of other light gaseous component, and light liquids in the gasoline and distillate boiling range. Due to high coke reaction vessel (known as cocker drum) temperatures, the liquids vapors and gas products pass above to the fractionator, leaving the solid coke in the coke vessel. The structure of the green coke product is dependent upon the residue or feedstock form transferred to the Coker. In brief, the process heats the residual oil from the vacuum distillation unit in a petroleum refinery to its thermal cracking temperature in the heat transfer tubes of a furnace. This incompletely vaporizes the residual gasoil and initiates cracking of the long chain hydrocarbon components of the residual oil painting into hydrocarbon gas, coker naphtha, and coker gasoil and petroleum coke. The heater effluent discharges into veritably large perpendicular vessels(called" coke drum") where the cracking process continue to completion, forming solid petroleum coke which deposits out and accumulates in the coke drums from which the product coke is latterly removed.[2][3] This study conducted in Khartoum refinery in Sudan which use one type of blend of crude oil for DCU, this study amid to make new mix from this type of blend produce in Sudan to ameliorate the specifications of final product.

1.1 Literature review of the delayed coking process

Delayed coking evolved steadily over the early to mid-1900s to enable refiners to convert high boiling, residual petroleum fractions to light products such as gasoline. Pound for pound, coking is the most energy intensive of any operation in a modern refinery. Large quantity of energy are needed to heat the thick, poor-quality petroleum residue to the 900 - 950 degrees F required to crack the heavy hydrocarbon component into lighter, more valuable products. One common misconception of delayed coking is that the product coke is a disadvantage. Although coke is a low valued (near zero economic value) byproduct, compared to transportation fuels, there is a significant

worldwide trade and demand for coke as it is an provident fuel. Coke production has increased steadily over the last ten years, with farther increases read for the foreseeable future. Current domestic production is near 111,000 tons per day.

A major driving force behind this increase is the steady decline in crude quality available to pollutants. Crude slates are anticipated to grow heavier with advanced sulfur contents while Crude slates are anticipated to grow heavier with advanced sulfur contents while environmental restrictions are anticipated to significantly reduce the demand for high- sulfur residual energy oil products. Light sweet crudes will continue to be available and in indeed lesser demand than they're moment. Refineries will be faced with the choice of get light sweet crudes at a preferential price, or adding bottom of the barrel elevation capability, through   new investments, to reduce the product of high- sulfur residual energy oil   and increase the product of distillate fuel with low- Sulphur[4] . A alternate disadvantage is that liquid products from cokers constantly are unstable, i.e., they fleetly form gum and sediments. Because of intermediate investment and operating costs, delayed coking has increased in fissionability among refiners worldwide. Grounded on the 2000 Worldwide Refining Survey published in the oil and Gas Journal December 2000 issue, the coking capacity for 101 refineries around the world is 2937439 barrels/ timetable day.. These cokers produce,154607 tons of coke per day and delayed coking accounts for 88% of the world capacity.

The delayed coking charge capacity in the United States is1, 787,860 b/cd. In general, coking belongs to a class of thermal corruption, free radical chemical reactions that have been considerably studied, except for the last stages of coke conformation itself. The engineering of thermal processes has been well developed. Recent mind efforts to minimize the environmental impacts of furnace processes have lead to significant advancements. As the coke product has declined in value, alternate coking processes have been developed to consume the coke produced in- house (e.g., burn it).   The values of these processes are privately associated with the mileage requirements of a particular refinery. Delayed coking is still the favored process in new construction. [5]

1.2 Delayed coking process

The first delayed coker was erected by Standard Oil of Indiana at Whiting, Indiana in 1929[6], [7]. The development of hydraulic decoking came in the late 1930's. Shell Oil at Wood River, Illinois presented a paper on hydraulic decoking4.0 m(13 ft) periphery Dobbs units and stated that they had patents along with Worthington Pump Company on hydraulic decoking bits and snoots[8]. Standard Oil of Indiana had patents on the original cutting nozzles used by Pacific Pump [9].

A veritably analogous nozzle is presently used in the new compact combination coke cutting unit. A pilot hole is drilled down through the coke in the vessel using high pressure water, and then the coke is cut out with a drilling bit with horizontal water nozzles. Roy Diwoky while at Standard Oil Whiting was one of the crucial people in developing the hydraulic decoking in the 1930's. Diwoky in May 1952, while Administrative Vice President of Pan Am SouthernCorp. (Possessed by Standard Oil of Indiana), worked with Great Lakes Carbon Corporation to produce the first needle coke in a delayed coker. Bernard Gamson, the Director of Research and Development for Great Lakes Carbon at the time, stated in a report that Diwoky was "the father of delayed coking [**7**].

"Delayed coking combined a number of the features and advancements from the development of the thermal cracking process. The use of pressure as well as heat for cracking and separating the heater from the coker and the use of two cans enabled the delayed coker to operate on a nonstop base. The number of cokers erected before 1955 was small, with a swell in delayed coker construction between 1955 to 1975 at 6% and an 11% growth rate during the 1965 to 1970 period [**10**]. The growth of delayed cokers was in step with the growth of fluid catalytic cracking and rapid-fire decline in thermal cracking. A fluid coker, suchlike to a fluid catalytic cracker except that fluid coke is circulated rather of catalyst, was first erected in 1954 at Billings, Montana. Five further fluid cokers were erected in the late fifties, and one in 1970.
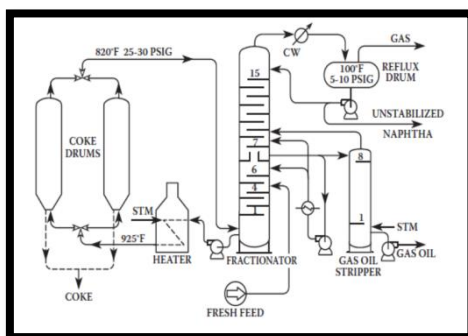
Fig.1. Typical schematic diagram of a delayed coking unit [3]

In 1958, the head of petroleum refining engineering at Colorado School of Mines, J.O. Ball, stated that there would not be any further belated cokers erected. Ball allowed all new cokers would be fluid cokers, and that a delayed coker was just a scrap can in the refinery, the delayed coking process was developed to minimize refinery yields of residual energy oil painting by severe thermal cracking of stocks similar as vacuum residuals, aromatic gas oils, and

thermal tars [**10**]. In early refineries, severe thermal cracking of similar stocks redounded in unwanted deposit of coke in the heaters. By gradational elaboration of the art it was set up that heaters could be designed to raise residual stock temperatures above the coking point without significant coke conformation in the heaters. This needed high rapidity (minimal retention time) in the heaters. furnishing an insulated swell barrel on the heater effluent allowed sufficient time for the coking to take place before posterior processing, hence the term "delayed coking".

2. STATEMENT OF THE RESEARCH PROBLEM

In the Khartoum refinery using only the crude from Fulla field in Sudan as feed stock for DCU unit and all design set depending on the fulla blend, now the Fulla crude Decreased from 40000 barrel/day to 20000barrel/day for some reasons . this work planned to study possibility of Blending fulla crude with other crude available in Sudan (Dar blend)with ideal ratio should be considered to guarantee amount of feed stock without any change or major change in the unit design.

2.1 Location
Khartoum Refinery Co., Ltd, which is located 75 Km North of Khartoum - Sudan country

3. OBJECTIVES

3.1 General Objectives:
The main objective of this thesis is to reach the ideal blending percentage between Dar Blend /Fula heavy crude mix in order to be used as the alternative to Fulla crude oil in Khartoum refinery DCU.

3.2 Specific Objectives:
1. Evaluation and simulation studies to estimate the possibility processing different ratios of Dar   Blend /Fula heavy crude mix.
2. To disquisition of the effects on some physic-mechanical properties of heavy and light crude oil.

4. MATERIAL AND METHODS

 Khartoum Refinery Company is the biggest Refinery in Sudan which is located about seventy-five kilometers north of Khartoum; the engineering construction was completed in January 2000. In May 2000 the refinery started production.

The study method has two step included lab scale run done for different mixing ratio between Dar blend /Fulla blend the result used as input data for simulation program (hyses) which descripted in the below:

4.1 Sampling

The sampling taken from the inlet of DCU unit and the sample was free of any disposal. The sample was observed to be semi-liquid at room temperature and it was shaking to homogenization, and then a representative portion of the sample was taken and tested.

4.2 Distillation:

The Distillation Apparatus used is:

i-FISCER C FR467 V1.72 (10/22/08)

The distillation procedure is carried out according to the following:

[A] ASTM D2892 [11]:

 (1) The distillation cut from (IBP – 165) ℃ are separated at Vacuum 1, pressure (100 Torr).

(2) The distillation cut from (165 – 220) ℃ are separated at Vacuum 2, pressure (10 Torr).

(3) The distillation cut from (220 – 280) ℃ are separated at Vacuum 2, pressure (10 Torr).

(4) The distillation cut from (280 – 350) ℃ are separated at Vacuum 3, pressure (2 Torr).

[B] ASTM D5236 [11] (POTSTILL Method)

(1) The distillation cut from (350 – 410) ℃ are separated at Vacuum pressure (1 Torr).

 (2) The distillation cut from (410 – 500) ℃ are separated at Vacuum Pressure (0.1 Torr).

4.3 Fractions and Lab Analysis:

The products as Naphtha, kerosene, diesel fraction, vacuum distillation fraction, atmospheric distillation residue fraction and vacuum distillation residue fraction by true boiling point distillation instrument and then tests the properties of crude and these fractions under ASTM and GB standard.

4.4 Theoretical Evaluation and Simulation

4.4.1 Simulation software:

In this project we are using aspen Hysys (v8.8) which is one of best simulation software in the field of downstream processing.
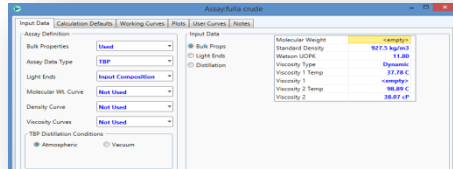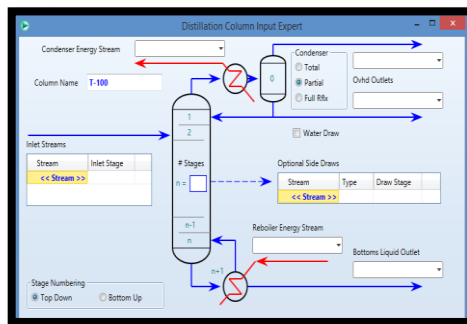

Fig.2. Aspen Hysys software input data

Fig.3. Aspen hysys software fractionation distillation
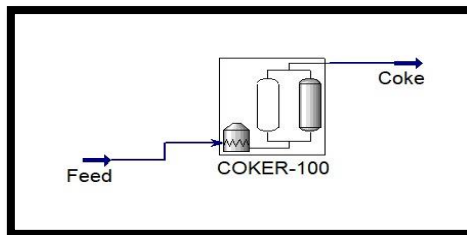

Fig.4. Aspen Hysys software DCU simulation

4.4.2   Material balance:

An overall material balance will be calculated for the whole process using the simulation software. And a correlation method will be used to predict the products yields manually.

4.4.2.1   Material balance by Prediction:

Estimation of product yields can be carried out using correlations based on the weight percent of Conradson carbon residue determine the trend of coke production [**12**] (wt% CCR) in the vacuum residue.

$$\text{Gas}(C_4^-)\text{wt\%} = 7.8 + 0.144 \times (\text{wt\% CCR})$$
$$\text{Naphtha wt\%} = 11.29 + 0.343 \times (\text{wt\% CCR})$$
$$\text{Coke wt\%} = 1.6 \times (\text{wt\% CCR})$$
$$\text{Gas oil wt\%} = 100 - \text{Gas wt\%} - \text{Naphtha wt\%} - \text{Coke wt\%}$$
$$\text{Diesel wt\%} = 64.5 \text{ \%* Gas oil wt}$$
$$\text{HCGO wt\%} = 35.5 \text{ \% * Gas oil wt}$$

5. RESULTS AND DISCUSSION

1. DCU in Khartoum Refinery is designed for Fulla crude, which is characterized by the followings:

   a) High acid number

   b) Heavy crude oil (0.93 specific gravity)

   c) High Calcium content (about 1200ppm)

   d) High Viscosity (117mm²/s at 100 °C)

   e) Sweet crude (sulfur content 0.15% wt)

2. Dar blend crude is also a heavy crude (0.91specific gravity), sweet (0.11% wt sulfur content), and low calcium content, more paraffinic than Fulla (its pour point 38°C while Fulla about 5 ° C), both (Fulla and Dar) blend are high TAN No., (> 4 mgKOH/g). Several test runs were carried in Khartoum refinery DCU with different Dar blend ratio 23% , 26% and 36% respectively in the past during period from February 2012 up to May 2016 and the results were as in Table .1.

Table 1. Result of blend with different ratios comparing with pure Fula blend.

Item	FULA 100%	16%	23%	26%	36%
Diesel +Gasoline	64.98	67.8	68.3	68	67.4
Coke	14	13	15	16	14.6
Wax oil	16	12.2	9.8	10.48	12.6
LPG	1.6	2.9	2.9	1.84	2
Dry gas	3	3.6	3.6	3.18	3
Loss	0.42	0.5	0.4	0.4	0.4
Total	100	100	100	100	100
Middle distillate yield (Diesel+Gasoline+HCGO)	80.98	80	78.1	78.58	80.98

5.2 Simulation results:

In the simulation program using input data as in fig 5.1 and fig 5.2 with mixing ratio 50% for each blend Fula and Dar after applying input the simulation result below as in Fig. 5, Fig. 6. Fig.7.and Fig. 8. the results for mixing ratios of 50/50 for Fula and Petrodar shown in Table .1.

Furnace Specifications

Furnace Outlet Temperature [C]	490.0
Furnace Outlet Pressure [kPa]	510.0
Feed Steam Mass [tonne/h]	0.2200
Feed Steam Temperature [C]	200.0
Feed Steam Pressure [kPa]	1150
Drum Volume [m3]	<empty>
Steam to Feed Ratio	9.932e-004

Drum Conditions

Drum Outlet T [C]	404.4
Theoretical Drum Outlet T [C]	414.8
Fill Time (hrs)	<empty>
Drum Outlet P [kPa]	510.0
Heat of Reaction at 900F [kJ/kg]	252.6

Combined Feed Volume Ratio (CFR)
◉ Specify CFR   ○ Reference Flowsheet Streams

1.14	Coker Feed from Fractionator
	Fresh Feed to Fractionator

Fig .5. Hysys operating condition

Feed properties input from stream

Specific Gravity	0.9856
Sulfur [wt %]	6.026
ConCarbon [wt %]	8.206
Nitrogen [ppmwt]	7803
D2887 IBP [C]	389.9
D2887 5% [C]	432.2
D2887 10% [C]	474.6
D2887 30% [C]	533.6
D2887 50% [C]	588.7
D2887 70% [C]	641.6
D2887 90% [C]	715.5
D2887 95% [C]	818.0
D2887 FBP [C]	861.2

Fig.6. Hysys feed properties input

	C5-230F	230-350F	350-650F	650F+Liquid
Specific Gravity	0.6542	0.7354	0.8089	0.8373
Sulfur [wt %]	0.6542	1.830	2.903	4.983
Nitrogen [ppmwt]	9.754	19.51	94.85	247.4
Paraffins [vol %]	28.98	36.88	24.25	16.25
Olefins [vol %]	66.14	48.99	12.50	8.500
Naphthenes [vol %]	3.173	5.490	24.25	16.25
Aromatics [vol %]	1.710	8.638	39.00	59.00

Coke Properties

Carbon [wt %]	79.62
Hydrogen [wt %]	2.444
Sulfur [wt %]	13.87
Nitrogen [wt %]	4.068
Vanadium [ppmwt]	0.0000
Nickel [ppmwt]	0.0000
Volatile Matter [wt %]	31.84
HGI	72.23
Specific Gravity	1.252
Coke Type	Sponge

Fig.7. Hysys simulation properties results

Product	Vol. Flow [m3/h]	Mass Flow [tonne/h]	Vol. [%]	Mass [%]
H2S		1.361		0.61
Fuel Gas	15.08	12.53	6.71	5.66
C3 Paraffins	6.522	3.301	2.90	1.49
C3 Olefins	2.288	1.191	1.02	0.54
C4 Paraffins	3.315	1.911	1.47	0.86
C4 Olefins	1.996	1.221	0.89	0.55
C5-350F	35.88	25.35	15.95	11.45
350-650F	41.93	33.88	18.64	15.30
650+F	118.3	98.99	52.61	44.69
Coke		41.68		18.82

Fig.8. Hysys simulation cuts results

Table 2. Mixing ratio 50% Fula blend and 50% Dar blend

Name		Yield m%	Flow Rate kg/hr
input	Fula blend	50%	125,000
	Dar blend	50%	125,000
	Subtotal	100%	250,000
Output	Diesel + Gasoline	66.2	165,500
	Coke	12.82	32,050
	Wax oil	13.3	33,250
	LPG	3.44	8,600
	Dry Gas	4.27	10SSS675
	Loss	0	

5.3 Delayed Coker Yield Prediction:

5.3.1 Using correlation [11]

Gas wt% = 7.8 + 0.144 * (wt% CCR) Gas wt% = 7.8 + 0.144 * 4.4 =8.4336 %.

So mass flow of the gas = 250000 * 0.084336 = 21084 kg/h Naphtha wt% = 11.29 + 0.343 * (wt% CCR)

Naphtha wt% = 11.29 + 0.343 * 4.4 = 12.8 %

So mass flow of the naphtha = 0.128 * 250000 = 32000 kg/h Coke wt% = 1.6 * (wt%CCR)

Coke wt% = 1.6 * 4.4 = 7.04%

So mass flow of the coke = 0.0704 * 250000 = 17600 kg/h Total gas oil = 100 - Gas wt% - Naphtha wt% - coke wt% Total gas oil= 100 - 7.04 - 12.8 - 8.43 = 71.73 %

Total gas oil = 0.7173 * 250000 = 179325 kg/h HCGO wt% = 0.355 * total gas oil

= 0.355 * 179325 =63660 kg/h

Diesel % = 0.645 * 179325 = 115664 kg/h

5.4 Discussion:

1. All GDHT stream to DCU calculated and segregated from DCU material balance, about 26% of DCU LPG and 21% of DCU dry gas comes from GDHT, considerable amount should be taken in account during material balance calculation.

2. As seen from table (5.1) material balance, total liquid yield (naphtha+ diesel+ heavy coker gas oil) is slightly increased with the increasing Dar

blend crude ratio and coke yield also slightly decreased, the reason is that the liquid component of Dar blend crude is little more than Fulla crude based on both crudes TBP assay.

3. Distillates (naphtha+diesel) are increased from 64.98% to 67.4%～68% compared with before processed Dar in 2014 test run , but slightly decreased compared with different blending Dar ratio because the IBP of wax oil for 36% Dar blend crude is 247℃ while the IBP of wax oil for 26% is 254℃.

4. Coke yields decreased as Dar CCR less than Fulla., also Coke quality changed (see table-4), ash and calcium content decreased due to lower calcium content in Dar blend.

5. Naphtha paraffinic content increased, and diesel cetane number improved number due to the paraffinic nature of Dar blend (see Table.2.).

6. For rich gas quality, only H2S concentration increased (about 4200 ppm during the test, before around 2000 ppm) , so during processing Dar Blend more chemical(MDEA) is needed to treat H2S to the required value (0.5 ppm).

7. Refer to table 5.2 the study proved that the mixing ratio 50% from each blend fulla and Dar blend resulted in product with in the range of DCU products and expand the chance of filling the gap of fulla blend decrease to keep refinery work as normal conditions and the became clear in hysys result for simulation of products specifications such as (sulfur, nitrogen, vanadium) content and other properties approved by refinery as standard. Fig 9. Below shows the comparison between different mixing ratio and the mixing ratio under study (50%/50% fulla & Dar blend).
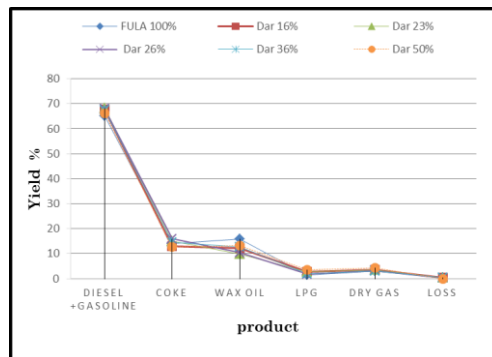
Fig.9. Comparing of output between different mixing ratios of fulla and Dar blend

6. CONCLUSION & RECOMMENDATION

6.1 Conclusion:
According to the lab test and simulation results, blending of 50% Dar crude with Fula crude in DCU achieved a good improvement in term of product qualities and quantities with the probability of certain negative impact and limitations according to the differences in physical &chemical properties of the two crude oils used.

6.2 Recommendation:
1. KRC could increase the percentage of Dar /Fula blending processing in DCU up to 50% with close monitoring for operation parameters and product specification
2. Increase Dar ratio more than 50% Dar /Fula is affecting to some extend positively on the product but also may affect negatively on the equipment load.
3. Dar is classified as medium crude oil , the light content is obviously much higher than Fulla crude ,the negative expected impact that should be considered during operation of 50% Dar ratio as follows:
   a. Change many operating conditions to minimize the negative impact (to some extend effective).
   b. Starvation at fractionator bottom level and surge drum
   c. High load at the top of the fractionator.

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