Energy Optimization and Performance Improvement for Crude Distillation Unit using Pre-flash System

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Abstract: Crude distillation units are designed to handle either heavy or light crudes, or both of them. But when processing light crudes, there are some parameters to be taken into account, such as crude heater duty, atmospheric tower overhead condensers, preheating system efficiency and so on. Flexibility of a crude unit to handle different crude types makes it easier to achieve a high product yield with desired specifications and decreases operational upsets, leading to more profits over the unit running time. In this paper, a solution has been found for a crude distillation unit in one of the refineries to efficiently handle a light crude, avoiding unit bottlenecks by establishment of a pre-flash tower in the upstream of crude heater. A simulation model has been built by using Aspen HYSYS to obtain results and to make an economic evaluation for the new modifications; comparing them to the old case.

Keywords: Aspen HYSYS, crude distillation, energy optimization, pre-flash

1. Introduction

Petroleum is the most important primary energy source in the whole world. The importance of petroleum continually rises to meet the high global demand worldwide. Furthermore, exploration, development, and lifting are mostly referred to as the upstream activities of the petroleum industry. On the contrary, refining and marketing are considered as the downstream operations, [1]. The first oil well drilled was by Edwin Drake in 1859 in Pennsylvania USA. In 2020, there were more than 70,000 oil and gas wells all over the world. Crude oil production is 80 million barrels a day. Moreover, the three top countries that produce oil and gas are: USA, Saudi Arabia and Russia. Also, India is ranked 25th on the list; as it produces 0.8 million barrel per day, [2].

Regarding crude oil refinery, it is the process of refining crude oil to produce products like gasoline, naphtha, kerosene, and light oil through refining plants such as distillation, heavy oil cracking, and desulfurization. Thus, crude oil in its natural state is not valuable to consumers; therefore, it has to go through refining process to be transformed into products used in marketplace, [3]. The first one is the distillation process. In distillation, the crude oil is divided into categories of different hydrocarbon compounds according to specific boiling-points scope. The second one is the conversion process. The third one is the treatment process. In this process, the hydrocarbon streams are treated to be prepared for further processing. Finished products are also exposed to chemical or physical separation. It also includes the usage of various processes like desalting, hydro-desulfurization, sweetening, solvent refining, solvent extraction, and dewaxing. The fourth kind is the blending and merging process. The last kind includes other refining operations like light-ends recovery, sold waste and wastewater treatment, hydrogen production, sour-water stripping, sulfur recovery, and acid and tail gas treatment, [4]. Different measures are used to characterize crude oil like density and distillation curves. Some of

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these classifications are relevant to the employment of simulation and design. These are density, K factor as well as TBP and ASTM distillation curves. Regarding the K factor classification method, it was first presented by researchers at Universal Oil Products Company (UOP). Additionally, it means that to guarantee that each component is detached at a time, a collection of distillation columns of numerous tray and big reflux ratio is used. Moreover, a test method called ASTM is used. ASTM is an acronym of American Society for Testing and Materials. It incorporates using a distillation column of 15-18 tray running at a 5:1 reflux ratio. Also, through this test, the temperature in the reboiler is noted and the distilled volume is evaluated, [5:9].

First of all, in the beginning of the 1970s, it was the beginning of simple refinery like motor fuels and heavy fuels. However, at the end of the 1980s, it was the time of the first and primary phase of introducing heavy ends' conversion. Finally, in 2010-2020, it included the refinery complex keeping up with environmental regulations and guaranteeing total conversion of the heavy ends. Therefore, the refining demands a wide variety of processes to cope with this trend. These processes include separation, conversion, finishing, and environmental protection processes, [10]. Regarding the separation processes, they consist of distillation, absorption, extraction, and crystallization. Concerning distillation process, the primary distillation is the atmospheric pressure of the crude oil. The first step on the crude is desalting, which focuses on washing the crude by water and caustic. The extraction of salts minimizes acid corrosion as well as failures in equipment, [11].

Additionally, the chemical conversion processes include: catalytic reforming, isomerization, hydrotreating, catalytic hydrocracking, catalytic cracking, and alkylation. Catalytic hydrocracking is important to produce light products in the presence of hydrogen regarding higher molecular weight fractions such as atmospheric residues and vacuum gas oils. In such a case, a dual function catalyst is employed. This catalyst is consisted of zeolite catalyst for the cracking function and rare earth metals supported on alumina for the hydrogenation function. The products of such a process are kerosene, diesel, jet fuel, and fuel oil, [12,13]. The final process of the thermal chemical conversion is visbreaking. It is considered as a mild thermal cracking process employed to lower the high viscosity and pour points of a vacuum residue to an extent that it can be used in more downstream processes. It was developed in the late 1930s in order to yield more desirable and valuable products. Moreover, the residue is either broken in the furnace or soaked in a reactor for several minutes. The yielded products of this process are gases, gas oil, gasoline and the unconverted residue, [14].

A crude oil distillation unit is consisted of three main stages: a pre-flash tower, followed by an atmospheric distillation unit then followed by a vacuum distillation unit. However, the pre-flash tower is not constantly used. It is optional and only used if needed. It is used when the feed used consists of a larger proportion of lighter components. Therefore, the usage of pre-flash tower is necessary to remove these lighter components and reduce the load on the following towers. The products of a crude distillation unit are basically naphtha, diesel, kerosene, atmospheric gas oil, light and heavy vacuum gas oil and vacuum residue, [15]. The crude desalter is very crucial in oil refinery process. Crude oil contains dissolved salts that lead to problems, failures and corrosion in the process equipment. Therefore, the desalting unit detach salt from the crude oil. However, the main problem that hinders the desalter job is the temperature of the operation. The best temperature for efficient salt removal is between 100-300 °F. Thus, the crude oil is then heated to about 250 °F as a preparation to enter the desalter unit. Usually, when a high level of salt removal is needed; therefore, three stages of desalting are usually performed, [16].

The pre-flash unit main aim is to remove the light components of the crude oil prior to entering the furnace. So, the extracted vapor stream may be gathered at the furnace outlet or in a suitable location of the main column. Consequently, heat duty of the distillation can be minimized and the development of the hydraulic performance of the heat exchanger network can be achieved. The best place of a pre-flash unit is downstream the desalting process. In this location, the pre-flash unit can also remove water carryover which may cause later corrosion in the next equipment, [17]. It was difficult to use light oils as a new feedstock as it needs great efforts of revamping traditional refinery layout that was originally made to deal with traditional crude oil feedstock. The main issues were the imbalance of capacity in downstream conversion units as well as the deficiency of overhead train processing capacity. Researchers attempted to solve these problems by figuring out the excellent suitable ratios of blending light oil and traditional used heavy crude oil. However, the result was unfortunately disappointing because this blend led to the instability of asphaltene. Therefore, the researchers focused their effort in

employing various pre-flash options in order to achieve optimal performance and minimum energy consumption in the crude distillation unit, [18].

Being the most economical option, the pre-flash drum is considered the simplest and preferred option to use. In addition, it needs no complicated modifications of equipment. It detaches vapor from the flashed crude liquid while having to keep any foam created. The only deficiency of the pre-flash drum is its lack of control on pre-flashed vapor's quality in a service subjected to foaming. Thus, it can lead to impurity problems. Inaccurate placing of pre-flashed vapor feeding exposes final product's quality to impurity. However, the main problems of the pre-flash column are the high reflux ratio flow rate as well as the existence of just a few plates between the inlet and the withdrawn, [19,20]. In this study, a pre-flash column configuration has been established to provide naphtha product with accepted specs to the downstream debutanizer system.

2. Process Description

The crude oil feed firstly enters a preheat network of five heat exchangers to be heated from 20°C to about 120-130°C, which is a suitable temperature for desalter entry. After that it is pumped by a feed pump to another preheat network of six heat exchangers to be heated from 120-130°C to about 250°C before entering the crude furnace. In the crude furnace, the crude oil is heated to about 360°C before entering the flash zone of the atmospheric distillation tower. In this stage, the crude oil is separated into products such as straight run naphtha, kerosene, diesel in addition to atmospheric residue, which is introduced as a feed to the vacuum distillation unit (VDU). Figure 1 shows the process flow before furnace.



Figure 1. Process flow before furnace

A two stage desalting system is used in series to guarantee a good process efficiency. The atmospheric distillation tower uses stripping steam in the tower bottom and has two side strippers; kerosene side stripper, which uses a reboiler, and light diesel side stripper, which uses stripping steam. The atmospheric overhead system uses two overhead drums with a two stage condensation; the first drum is the tower reflux drum, which is free of water to guarantee a dry column reflux, and the second drum separates water from the straight run naphtha produced. Figure 2 shows atmospheric distillation tower system.

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Figure 2. Atmospheric distillation tower system

3. Case Study

A simulation model for the existing CDU has been built by using Aspen HYSYS, in which Arabian Heavy crude is processed (°API=28.2). That model has a capacity of 100 000 bbl/day (662.4 m³/hr). The obtained products material balance according to TBP curve is shown in table 1.

Product	Flow rate (m ³ /hr)
Off-gases	4
Naphtha	155
Kerosene	75
Diesel	150
Atmospheric Residue	295

Table 1. Product flow rates with Arabian Heavy crude

A second model has been built in which Arabian Heavy has been replaced by Arabian Light crude (°API=37.9), in which some bottlenecks have appeared with the full capacity, such as the duty of crude heater and atmospheric tower overhead condensers. So, the unit capacity was decreased to about 86 000 bbl/day (575 m³/hr) so as not to exceed these limits. The obtained products material balance according to TBP curve in that case is shown in table 2.

Product	Flow rate (m ³ /hr)	
Off-gases	1.1	
Naphtha	155	
Kerosene	90	
Diesel	170	
Atmospheric Residue	165	

Table 2. Product flow rates with Arabian Light crude

According to the results shown, there has been a need for a new modification to restore or to increase the unit efficiency to deal with Arabian Light crude. A pre-flash stage has been the solution to accomplish that job by extracting the light naphtha cut from Arabian Light crude before entering the furnace then to the atmospheric tower; decreasing the duty of both crude heater and atmospheric tower overhead condensers. In the new simulation case of pre-flash modification, capacity increase has been achieved within the unit design limits.

The best place for pre-flash tower has been in the second preheat train at a feed temperature of 160°C, which meets the lowest temperature to make enough flash to obtain the light naphtha cut. In addition, this place meets the lowest reflux ratio for the tower overhead; which means an overhead condenser with lower duty. Of course this provides a good choice on the technical and economic levels.

The pre-flash tower has been selected with 3m diameter, 0.6m tray spacing, 12 (sieve) trays and a feed entry at the 9th tray; which guarantees a good stripping for crude oil and a good rectifying for light naphtha. It is also selected with a two stage overhead condensation. The column reflux is withdrawn from the first stage accumulator in which there is no water condensation to guarantee a dry reflux to the tower. The product light naphtha is withdrawn from the second stage accumulator to be mixed with the heavy naphtha produced from the atmospheric tower. This resulting full range naphtha is then stabilized in the debutanizer section.

This case study is divided into three cases to be compared to each other. The first case (Case A) is 100% capacity (100 000 bbl/day) using Arabian Heavy Crude. The second case (Case B) is 86% capacity using Arabian Light Crude, which is the maximum capacity of the traditional unit that can handle the light crude avoiding the unit bottlenecks. This capacity has been selected according to the comparison made in table 3, figure 3 and figure 4 to be matching with the design duty of the crude heater which is 53 MMKcal/hr and that of atmospheric tower overhead condensers which are 21 MMKcal/hr for the first stage condenser and 15 MMKcal/hr for the second stage condenser. This reduced capacity has resulted in a heater duty of 52.3 MMKcal/hr and a condenser duty of 14.3 MMKcal/hr for the first stage and 18.7 MMKcal/hr for the second stage.

Capacity%	Heater duty (MMKcal/hr)	Atm tower ovd condenser duty (1 st /2 nd) Stage (MMKcal/hr)
100	56.3	16.7/22.5
95	55	15.9
90	53.5	14.1/22.9
85	52	13.6/20.1
80	50	12/19

Table 3. Maximum capacity for old CDU with Arabian Light crude



Figure 3. Traditional unit capacity vs crude heater duty

The third case (Case C) is 110% capacity using Arabian Light Crude after pre-flash modification. Some operational data for pre-flash and atmospheric towers such as temperature, pressure, stripping steam and pump-arounds are shown in Table 4.



Figure 4. Traditional unit capacity vs atmospheric condensers duty

Parameter	Pre-flash tower	Atm tower		er
	С	А	В	С
Top Temperature,°C	130	157	159	160
Top Pressure, Barg	4	1	1	1
Bottom Pressure, Barg	4.1	1.4	1.4	1.4
Top Reflux, m ³ /hr	19	146	263	235
Stripping Steam, t/hr	4.5	8	7.5	7.5
Kero PA, m³/hr	-	620	620	590
Diesel PA, m³/hr	-	550	550	530

Table 4. Operational data for both pre-flash and atmospheric towers

There are many possible locations for the pre-flash stage. From economical point of view, that stage should be at the lowest temperature where light naphtha cut can be obtained. This can be through the second preheat train in the downstream of desalter section. A comparison has been made for these possible positions as shown in the tables 5, 6 and 7.

Table 5. Pre-flash tower different locations conditions

Location	Feed T (°C)	Atm tower ovd cond duty (MMKcal/hr)	Heater inlet T (°C)	Heater duty (MMKcal/hr)
HE6 Out	139	0.9/6	264	55
HE7 Out	160	1.6/7	255	52.3
HE8 Out	190	4.7/8	238	56
HE9 Out	212	11.6/11	225	57

Table 6. Pre-flash tower location against heater feed

Location	Heater feed (t/hr)	Heater feed (m ³ /hr)	
HE6 Out 577.5		686	
HE7 Out	574.5	670	
HE8 Out	563	655	
HE9 Out	533.4	615	

Location	Pre-flash naphtha (m³/hr)	Pre-flash top product% aginst feed	Pre-flash 95%T (TBP) (°C)	Reflux rate (m³/hr)
HE6 Out	42	6.1	119	5
HE7 Out	58	8.2	146	19
HE8 Out	73	10	173	75
HE9 Out	113	15.4	229	157

Table 7. Pre-flash tower product naphtha and reflux rate

From the tables shown, the most suitable position is obviously in the downstream of the heat exchanger no.7 in the second preheat train. The feed pump location has been also changed to be in the pre-flash tower bottom instead of second stage desalter. For the new capacity selection, it should be the maximum capacity that the unit can handle the Arabian Light Crude without any bottlenecks. These bottlenecks are the duty of crude heater and the atmospheric tower overhead condensers. This comparison is shown in table 8, figure 5 and figure 6.

Table 8. New capacity selection

Capacity%	Heater Duty (MMKcal/hr)	Atm tower ovd cond duty (1 st /2 nd) Stage (MMKcal/hr)
86	41	13/15.7
100	47.4	13.8/17.4
105	49.8	14.1/18
110	52.3	14.3/18.7
115	54.8	14.7/19.3



Figure 5. Unit capacity vs crude heater duty



Figure 6. Unit capacity vs atmospheric condensers duty

From the previous comparison, a capacity of 110% (110 000 bbl/day) is selected to be the new one in which the duty of both crude heater and atmospheric tower overhead condensers are kept within design limits. Some parameters for the unit heat exchangers, such as inlet temperature, pressure and duty, are shown in the coming tables.

From the study, it is clear that all heat exchangers conditions meet the design ones except for the second case (case B) which breaks the duty design limits for the duty of the atmospheric tower overhead condensers. The calculated duty is 39.2 MMKcal/hr and the design duty is 37 MMKcal/hr. The crude heater duty, coil pressure drop and outlet temperature in each case are shown in table 9.

Case	Heater duty (MMKcal/hr)
Α	50.9
В	52.3
С	52.3

Table 9. Crude heater data

From this table, all cases meet the design duty of 53 MMKcal/hr, the design pressure drop of 10 Bar and the design outlet temperature of 367 °C, except for a clear heater overload appears in the second case (56.3 MMKcal/hr) compared to the design duty, which has been reduced to be 52.3 MMKcal/hr after capacity reduction. Also, the power consumption of pumps is shown in table 10.

Pump	Power (kW)		
	А	В	С
Feed pump (old)	327	284	-
Kero product	58	70	92
Diesel product	75	85	97
Atm residue	128	69	100
Naphtha product	68	77	64
Feed pump (new)	-	-	514
Pre-flash naphtha	-	-	28

Table 10. Pumps data

From the pumps data shown, it is clear that a new feed pump with a higher power is required (514 kW) compared to the existing feed pump (400 kW). In addition, the product conditions such as flow and draw temperature are shown in table 11.

Product	Flow (m³/hr)		Draw	temper (°C)	rature	
	А	В	С	А	В	С
Naphtha	136	156	188	157	158	160
Kerosene	75	90	119	200	205	207
Diesel	150	170	195	288	292	291
Atm residue	296	158	227	351	346	341

Table 11. Products conditions

Finally, the products distillation results according to TBP are shown in table 12.

Table 12. Products TBP distillation results

Spec	Case	Naphtha	Kerosene	Diesel	Atm residue
	A	-9	163	219	339
5%T	В	2.2	162	222	344
	С	68	151	213	335
	A	183	247	348	-
95%T	В	180	248	355	-
	С	176	243	346	-

The previous tables (11 and 12) reflect a good tray dynamics of the atmospheric distillation tower after pre-flash modification and processing a lighter crude than the heavy one. These dynamic characteristics are shown in table 13; comparing them before and after modification.

4. Results

The mentioned case study has shown a good performance for both the pre-flash and crude distillation towers, which is obvious in table 13. An overall increased capacity by 24% has been approached with minimum possible modifications to be modified. The only item to be changed is the feed pump, in addition to the pre-flash section.

Parameter	Atm tower		Pre-flash tower
	Before pre-flash	After pre-flash	
DC residence time (sec)	6.2	6.5	5.7
DC backup (%)	52	64	50
Tray pressure drop (Bar)	0.010	0.011	0.009
Tray jet flood (%)	57	53	50
Minimum vapor velocity	9	9.5	9.8
(weeping point) (m/sec)			
Actual vapor velocity (m/sec)	12.8	12	13
Liquid height over weir (How)	19	20	17.5
(mm Liq)			
Entrainment factor	0.075	0.070	0.850

 Table 13. Tray dynamic characteristics of atmospheric distillation tower

The table shows a satisfying performance for the atmospheric tower after the new modification for all tray parameters. The residence time has increased from 6.2 to 6.5 Sec through the downcomer as a result of decreased vapor load. As a result, the downcomer backup has increased from 52 to 64%. The tray pressure drop has slightly increased from 0.01 to 0.011, but it is still in the accepted limits. The actual vapor velocity is still above the weeping point, which provides a good tray stability and separation efficiency. The tray jet flood, entrainment factor and liquid height over weir are also within accepted limits.

The total equipment cost of these modifications is about USD \$18.05MM, and the operating cost is about USD \$4.16MM per year according to Aspen HYSYS Economic Evaluation. The profit for refining 1 bbl oil is about 2 \$ and the running time is considered to be 11 months/yr. So profits are expected to be increased by USD \$15.84MM per year after increasing the capacity by 24000 bbl/day. The Return On Investment is 64.75%, and the payback time is 1.55 year(s).

5. Conclusion

A pre-flash model has been established for a crude distillation unit in one of the refineries to make it possible to handle light crudes, increasing unit flexibility. The established model resulted in keeping the duty of both crude heater and atmospheric tower overhead condensers within design limits, leading to overall capacity increase by 24%. The Return On Investment is 64.75%, which means a payback time of about 1.55 year(s).

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