Process simulation and optimization of crude distillation unit to improve energy efficiency

ABDUL-AMEER BANDAR[1], and MUSTAFA OUN[2]
Oil and Gas Projects Department[1], Operations Department[2]
Pro-Knowledge Engineering Ltd[1], Azzawiya Oil Refining Comp., Azzawiya[2]
AUSTRALIA[1], LIBYA[2]
askadhum@hotmail.com[1], musta61@yahoo.com[2]

Abstract: This paper describes the current performance of the crude topping unit at Azzawiya refinery, Libya, and identify the operations that can improve the energy conservation without investment. Three areas of crude unit operation were studied, heater operation, overflash, and stripping steam rates. Significant opportunities were identified to reduce energy consumption in the topping unit. Thermal efficiency of topping heater increased from 71% to 78% by optimizing the operating conditions; excess air and stack temperature. Finally, steam rates to the stripping sections should be kept to the minimum required for a product to meet specifications or used only to the extend where the increase in value of product recovery justifies the cost.

Key-Words:- Crude Topping Unit, Thermal Efficiency, Fired Heater, Overflash, Stripping Steam

1 Introduction

Oil refineries are not only energy suppliers but also large energy users who spend typically 50% of cash operating costs (i.e.,, excluding capital costs and depreciation) on energy, making energy a major cost factor and also an important opportunity for cost reduction. Energy use is also a major source of emissions in the refinery industry making energy efficiency improvement an attractive opportunity to reduce emissions and operating costs [1]. Consequently refiners have been greatly concerned with more efficient energy utilization as well as other industries consuming energy.

In conventional oil refineries Crude Topping Units (CTU) are key process plants as they produce intermediate streams (gases, naphtha, kerosene, diesel, and heavy products) that are used in downstream process units. Changes in these units have a great impact on product yield and quality and, therefore it is recommended to operate these units at optimal conditions from technical and economical point of view; that means operating conditions such as temperatures, pressures, and flows of the units that maximize their economic performance (increase product yield), subject to their real physical restrictions and their design capabilities. Process simulation has become very common to develop these optimization studies. Optimization of the crude oil separation process becomes increasingly important because of the high energy costs and ecological requirements for quality oil products [2].

In the refining industry, typical energy consumption is approximately 0.32 MMBtu/bbl of crude oil processed. This translated into 2,667 MMBtu/hr for a 200,000 barrel/day refinery. Even 1% improvement in thermal efficiency translates into energy savings of $600,000 per year [3]. The authors carried out similar investigations on catalytic reforming unit at Azzawiya Oil Refining Company, Libya to optimize the energy and operational variables [4].

The leading refinery in Libya is the Azzawiya Refinery with a capacity 5 million tons per year. The refinery usually runs Sharara crude (Its API is 43.95 and very low sulphur content <0.07 wt% with very low salt content < 2 ppm). The refinery was started-up in 1974. It is of the hydroskimming type refinery consisting of atmospheric distillation and associated stabilizers, LPG recovery unit, naphtha hydrotreater and catalytic reformer and kerosene hydrotreater. The refinery burns a mixture of fuel oil and fuel gas.

The Azzawiya Refinery has been designed and built when the energy to equipment cost was significantly lower than the level standards. The rise in fuel costs makes it profitable to reconsider all factors capable of reducing energy consumption refinery processes.

The objective of the present paper is to review the current performance of the crude topping unit and identify the operations that can improve the energy conservation without investment.

2 Process Description

In all refineries, including small less complex refineries, the crude oil is first distilled, which
is followed by conversion in more complex refineries. The most important distillation processes are crude or atmospheric distillation, and vacuum distillation. Different conversion processes are available using thermal or catalytic processes, e.g., using a catalytic reformer, where the heavy naphtha, produced in the crude distillation unit, is converted to gasoline, and the fluid catalytic cracker where the distillate of the vacuum distillation unit is converted. Newer processes, such as hydrocrackers, are used to produce more light products from the heavy bottom products. Finally, all products may be treated to upgrade the product quality (e.g., sulfur removal using a hydrotreater). Side processes that are used to condition inputs or produce hydrogen or by-products include crude conditioning (e.g., desalting), hydrogen production, power and steam production, and asphalt production. Lubricants and other specialized products may be produced at special locations.

Figure 1 shows an outline schematic diagram of the crude topping unit, CTU. The unit is typical of the industry. Crude oil is heated against other streams within the unit. During heatin heating, it is pumped and desalted. Final heating is done in a furnace. The unit itself comprises an atmospheric distillation tower and three side strippers. These separate the crude into a light naphtha overhead, kerosene, light gas oil, heavy gas oil, and an atmospheric residue stream.. As these streams and the atmospheric tower pump-arounds are cooled, they provide heating for the crude. Furthermore, there is a stabilizer section which consists of an overhead gas compressor and reboiled stabilizer tower producing off-gas and SR Naphtha, suitable for the catalytic reforming unit.

3 Results and Discussion

Base case simulation of the process CTU is set up using HYSYS Process program to model the CTU operation. The model has been set up using the operating conditions and mass balance recorded on the unit log sheets, and calibrated to match the product quality analyses reported by the laboratory. The operating conditions used are close to those in the original refinery design case.

Three areas of crude unit operation were studied, and these are dealt with in turn below. The topping unit is running at design and actual conditions as shown in the Table 1.

3.1 Optimize Heater Operation

The crude unit heater is currently operated with coil outlet temperature of 328 C ; this compare with a design coil outlet temperature of 330 C. Design duty of heater was exceeded
because of:

- Lower heater inlet temperature
- Higher feed
- Comparatively lighter fractions needing more latent heat.

The efficiency results 71% and the excess air is about 50%. For energy saving it is considered a practical goal for heater to reduce the oxygen in the flue gas down to 4-4.5%; this will allow to increase the efficiency of the topping heater up to 78% and consequently the fuel saved comparing the actual conditions will be 18.2 Ton/day of fuel oil.

The main causes of high excess air are normally due to:

- air leakage through peep holes, shutdown burners, inspection doors
- improper draft control
- faulty burners operation
- operation without permanent oxygen analyzer.

3.2 Overflash Optimization

Simulation of the cude unit shows that it is operating at overflash of 4.3 vol% on feed. Higher levels of overflash can increase distillate yields by improving the fractionation between products. The higher internal reflux rates in the tower will result in an increased gas oil yield at the expense of fuel oil.

The calculated overflash figure is slightly below the normally recommended optimum of 5 LV% on feed. The effect of increasing the heater outlet temperature (and hence flash zone temperature) to achieve this level of overflash.

The main operating impact of such a change is the higher heater duty and additional heat entering the tower. The impact on the refinery yields is to marginally increase the yield of distillate products. The refinery net fuel saving is about 4.1 ton/day of fuel oil.

3.3 Stripping Steam Rates

Stripping steam is used to improve the initial boiling points and flash point of the sidestream and for stripping the reduced crude in the lower part of the tower. The stripping effect depends on the amount of steam injected and the efficiency of the stripping operation also depends on the correct degree of

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superheat of stripping steam. Lowering the stripping steam rates will result in loss of lighter oil in the atmospheric residue; excess steam may cause entrainment of residue into the higher cut, due to excessive velocity of the vapours.

The steam rates to the stripping section of the main tower and the distillate side-strippers are low in comparison with both standard industry practice and unit design rates. The new conditions of operation , decrease the heater outlet temperature down to 326 C and increase of stripping steam up to practical refinery industry have been investigated with the capacity of 410 m^3/hr. However, actual stripping steam rates investigated up to (24 m^3/hr) and (12 m^3/hr) for residue and distillates respectively.

The main impact of this change will be the improved separation between products through removal of light material. This is most effective for kerosene and residue stripping. The improved fractionation is reflected in the refinery yield pattern, where a reduction in fuel oil yield is seen. This reflected in the refinery profitability comparing the actual conditions where an increase of 13.3 Ton/day of fuel oil is seen after discounting the cost of additional steam.

It can be seen from the Table 1 that the temperature drop across the residue stripping section of the tower (from flash zone to bottom) is very small (typically 2 or 3 C). A temperature difference of around 10-15 C would be expected for correctly functioning stripping section. The higher steam rate only increased the temperature drop and improved the stripping in the residue section.

4 Conclusion

Significant opportunities were identified to reduce energy consumption in the topping unit. Thermal efficiency of topping heater increased from 71% to 78% by optimizing the operating conditions; excess air and stack temperature. Finally, steam rates to the stripping sections should be kept to the minimum required for a product to meet specifications or used only to the extend where the increase in value of product recovery justifies the cost.

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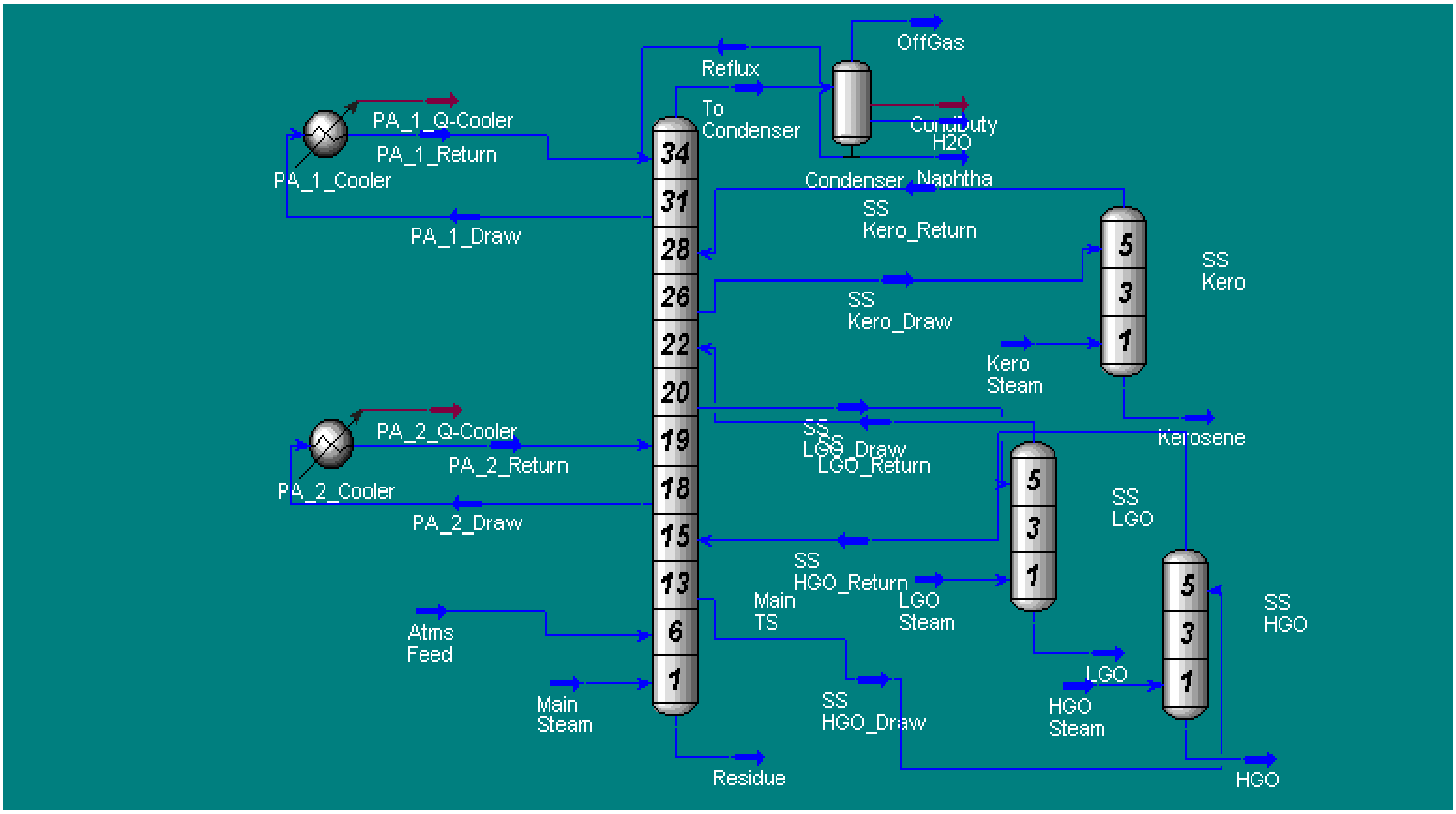

Figure 1, Process Schematic of Crude Topping Unit, CTU.

Table 1- Design and actual operation conditions of crude topping unit.

Parameter	Design	Actual
Crude oil	Es Sider	Sharara
Crude flow rate (m^3/hr)	380	410
Heater inlet temperature (C)	214	212
Heater outlet temperature (C)	330	328
Duty absorbed (Mkcal/hr)	35.87	41.97
Overhead temperature (C)	107	115
Overhead accumulator temperature (C)	43	42
Overhead accumulator pressure (kg/cm^2)	0.2	0.2
Flash zone temperature (C)	326	324
Flash zone pressure (kg/cm^2)	0.95	1.05
TPA flow rate (m^3/hr)	704.89	540.0
B.P.A flow rate (m^3/hr)	287.8	280.0
Overflash (LV% on feed)	3.0	4.4
Kerosene flow rate (m^3/hr)	60.5	90.0
Steam flow for kerosene stripper (kg/hr)	798	1000
Light gas oil flow rate (m^3/hr)	56.4	55
Steam flow rate for L.G.Oil (kg/hr)	742	700
Heavy gas oil flow rate (m^3/hr)	26.6	40.0
Steam flow rate for H.G.Oil (kg/hr)	350	500.0
Residue flow rate (m^3/hr)	150.1	95.0
Bottom temperature of main tower	315	322
Steam flow rate for main tower (kg/hr)	3595	1750

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