IS PRE-FLASH IMPLEMENTATION A RELIABLE RETROFITTING ALTERNATIVE FOR A CRUDE DISTILLATION SYSTEM? – A CASE STUDY

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Raw crude oil distillation is an energy intensive process, taking place in a specific topology of distillation column sequences unit. The specificity of the topology depends on the characteristics (especially the composition) of the processed crude oil. A crude oil distillation system usually comprises a crude oil preheating train, including a desalter unit, followed by two main and complex distillation columns with different lateral operations (side distillation operations as side strippers or side distillation operations, pump-arounds, feeds and draws), the first operating at pressures slightly above the atmospheric one and the second one working in vacuum conditions. In the conditions in which the processed raw material presents rather frequent variations of composition, the adequate adaptation of both process installations and operating conditions in a more compatible and economically efficient way represents the main task of the process engineers. Techniques based on process integration, intensification, modeling, and simulation are the most used tools to find the right decisions to achieve the aforementioned objectives. In the case study presented in this paper, it is proved that the introduction of a supplemental pre-flash device at the end of the pre-heat train, before the atmospheric tower fired heater, improves the crude distillation process performance, in terms of utilities need.

Keywords: process modeling, heat integration, performance improvement, pre-flash drum, crude oil distillation

1. Introduction

A raw crude oil distillation unit is a critical component in petroleum refining, as it processes high flow rates and, therefore, its size and the associated operating costs are the highest in the refinery. Considering this, there is a continuous interest to find ways to improve the process efficiency of the existing plants. Retrofitting a raw crude oil distillation process unit is a more common engineering task worldwide nowadays than building a new one, due to the high investment costs involved in the latter [1]. A raw crude oil distillation plant has

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many units, such as crude oil furnaces, distillation towers, and heat exchanger networks. The production capacity of the whole refinery is basically determined by the crude oil distillation unit production capacity. An atmospheric distillation column is typically designed to operate in nominal conditions for an approximative 80% capacity of its maximum loading (flooding precaution), which indicates that the unit can be operated (theoretically in less safe flooding conditions) at 20% higher throughput than the designed nominal conditions. The capacity increase of the column is limited by the level of its internal vapor flow rate. Therefore, especially in the case of processing light raw crude oil, one of the popular process revamping solution is to expand the crude oil production capacity and improve process efficiency throughout installing a pre-flash device, mainly before the atmospheric distillation column, sometimes even before the furnace of this column [1]. Crude pre-flash is one of the promising methods to enhance the process, which can help the crude oil atmospheric distillation process unit to better adapt to different operating conditions. The basic idea of this configuration is to separate the light components by flashing the preheated crude oil into a separate pre-flash drum before atmospheric tower's fired heater. Then, the light fraction stream obtained is either mixed with the furnace outlet or introduced separately into the atmospheric column at an appropriate location. Therefore, furnace heat transfer efficiency can be increased, while its fuel use could be slightly reduced [2]. Intuitively, we expect that the biggest gain of using the pre-flash should be determined by overcoming the limitation of the separation capacity of the atmospheric tower by improving its hydraulic regime because of taking over its vapor charge by the introduced pre-flash dome.

This research considers a case study of a crude oil distillation unit performance analysis in variants without and with pre-flash dome, throughout process modeling and simulation in AspenTech Hysys®. The crude oil used in the two mentioned variants is the same. Based on operating and plant data for a typical crude distillation unit (CDU), by using process modeling and simulation, the aim of this work is to evaluate the possibility of enhancing process efficiency/performance by integrating a pre-flash device, without changing/modifying the basics related to the feedstock and physical constraints of the unit, such as column actual diameter, pump-arounds and side-columns locations, exchangers matches and areas, maximum heat loads for fired heater, etc.

2. Process description

The crude oil atmospheric distillation system enables the first processing step that occurs within a petroleum refinery. The process flow diagram (PFD) is presented in Fig. 1. Crude oil from the tank storage farm is pumped through the first group of apparata of the heat exchanger network (HEN), where it is preheated by the hot lightest product streams from the main column to a temperature of about 150 °C, then is mixed with the wash water in a mixing valve/device and feeds the desalter, to remove the inorganic salts, impurities, etc. The desalted crude oil flows through the second group of HEN's apparata. Due to a high degree of thermal integration with hot stream distillation products, the desalted crude reaches a temperature of about 223 °C, recovering, as well, the heat from the atmospheric column inside liquid stream via pump-arounds. At this temperature level, the crude oil heat content is still low, therefore a furnace is needed to increase its temperature, to the heat level needed to achieve the degree of vaporization for the expected favorable cut fraction products separation to occur in the column. At the furnace exit, which uses fuel gas as a heating utility source, the desalted crude oil gets a temperature of about 360 °C. Atmospheric towers are not designed with reboilers, therefore all the heat needed for separation is provided in the furnace [4]. The high temperature difference between the furnace inlet and outlet together with the high flow rate throughput indicate that the atmospheric fire heater is the heat exchanger through which one of the largest required energy imports used by the refinery is added and which has a high influence on the overall operating costs of the process. The heated crude oil stream leaving the furnace is then fed to the bottom of the atmospheric distillation tower (Fig. 1), which is a complex configuration column, having the following structure:

- A main column endowed with 42 valve trays (real), having a stripping section at its bottom, below the feed tray, where steam is injected, to advanced strip out the light cuts' products from the crude oil;
- Three pump-arounds (PA): 1. to adjust the naphtha (top) product cut point interval, 2. to adjust kerosene (middle) product cut point interval and 3. to adjust diesel (bottom) product cut point interval – they are used to achieve the effective expected separation and removing residual heat from the distillation process in the column;
- Four side products withdrawals; from the top to the bottom: kerosene, light diesel (LT diesel), diesel and AGO (atmospheric gas oil). The side draw stream products are then stripped in the lateral stripping column, divided into 4 sections with 6 valve trays (real) on each section. The resulting vapors from any side stripper are returned to the main column few trays above their corresponding withdrawn. Stripping is necessary to fine tune the boiling points interval of the side products and is done with steam injected at each lateral stripper bottom;
- A partial condenser connected to a three-phase separator is located at the top of the main column and is used to separate the naphtha products from the non-condensable components and sour water. A part of the naphtha products is recycled back to the column as internal reflux.

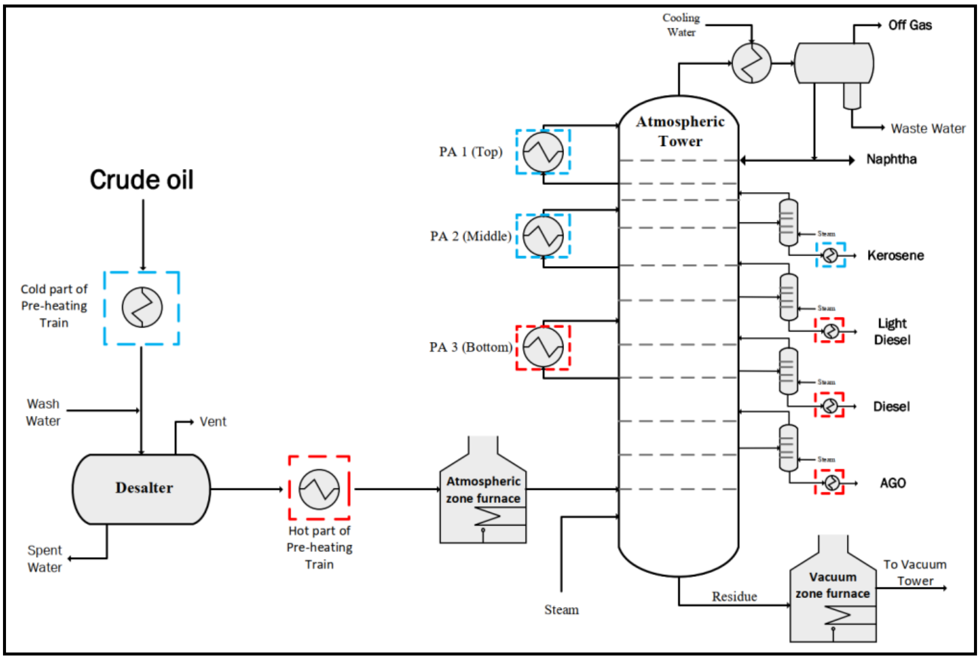

Fig. 1. Crude oil distillation process flow diagram

3. Process modeling

To analyze the implications of the process modification by integrating the Pre-flash dome before the atmospheric distillation column, the process had to be, first, rigorously modeled, simulated, and validated as representative for the analyzed process case in a crude oil refinery. The case study is based on data for a typical crude oil refinery unit that processes 15,300 t/day (745 m^3/h) raw crude oil. Process model is implemented in AspenTech HYSYS® process simulator in two variants, without (base case) and with the Pre-flash operation included in the process topology scheme. The results of the process simulations in the two variants, mentioned above, are compared to see if the Pre-flash integration represents a reliable retrofitting alternative for this analyzed crude distillation system.

3.1. Base case model implementation

The first step in the development of the simulation model is the selection of lighter components and the appropriate thermodynamic property package. The selected thermodynamic fluid package is Peng Robinson, this equation of state being the most suitable for complex hydrocarbon mixtures [5]. Moreover, raw crude oil exact composition is difficult to specify; therefore, it is defined as a combination of a mixture of a relevant number of chemical species which are taken from the simulator's pure component library (lighter components) and pseudo components, both being specified in accordance with the results of the standardized laboratory analysis of this crude oil.

The second step is the crude oil feed characterization, which is done using distillation curves obtained experimentally, through the standardized procedure by distilling a crude oil sample. In this study, the true boiling point (TBP) curve reported in Fig. 2 is used to define the feed stream composition. Furthermore, to have accurate results, it is better not to use the TBP curve as implemented in the simulation software package, but always the one obtained from the real plant data.

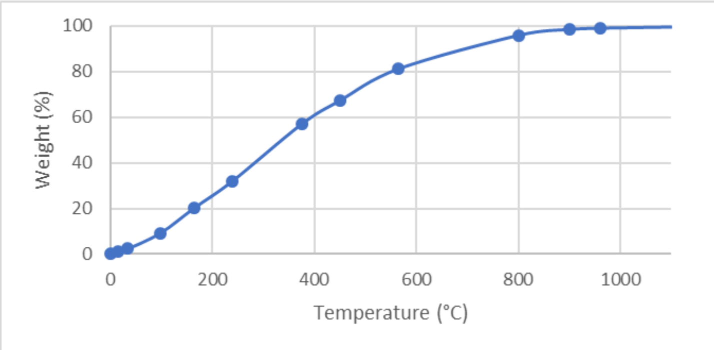

Fig. 2. Crude oil feed TBP curve

The third step in refinery process modeling is the definition of the process model flow sheet, where the unit operations contained in the simulator palette as simulator operations models are adequately placed and the logical connections between them are made. Here, the key process parameter inputs should be defined. The operating parameters used for the simulation model, such as pressures, temperatures, and flow rates were based on the values from the collection data of a typical real refinery.

3.2. Results presentation

The process topology shown in the PFD of Fig. 1, implemented in HYSYS process simulator, is exposed in Fig. 3. The characteristics of the main process parameters of the atmospheric distillation column and process streams in the simulated HYSYS model are shown in Tables 1 and 2, against these from the typical process, which are taken from a typical refinery.

Table 1

Operating variables comparison

Variable	Unit	Real Plant	Simulated
Main steam	t/h	2.9	2.9
Column feed temperature	°C	359	359
Condenser temperature	°C	84	83
Column top pressure	bar	0.84	0.84
Column bottom pressure	bar	1.12	1.12

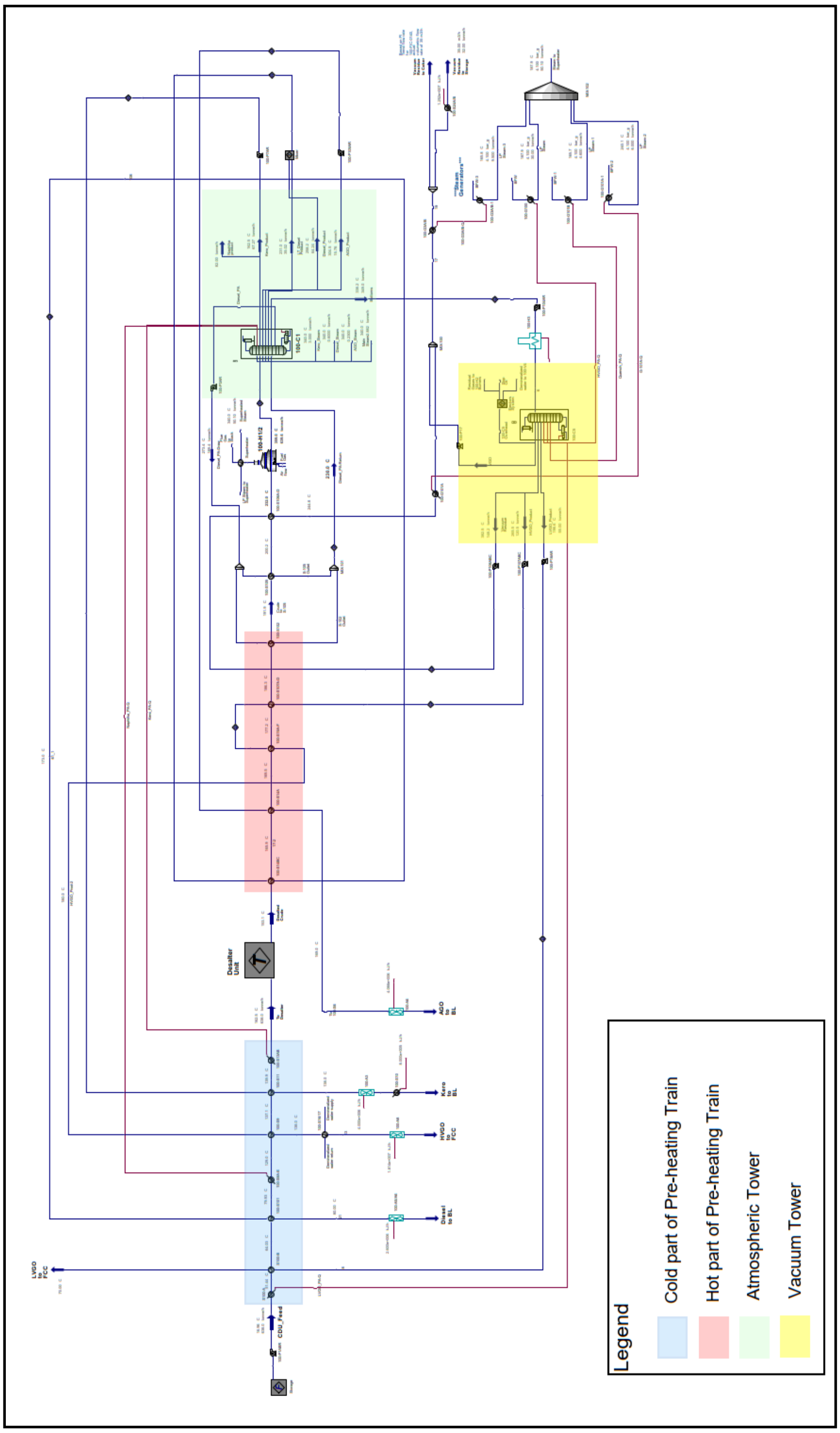

Fig. 3. Crude distillation process model flowsheet in HYSYS

Table 2

Products temperatures and quality comparison

Product	Real Plant		Simulated	
	℃	ASTM T95 %	℃	ASTM T95 %
Naphtha	84	153	83	153
Kerosene	170	224	168	224
LT Diesel	232	268	231	268
Diesel	260	323	258	323
AGO	307	378	305	378

Fig. 4 (a) presents the comparison between the real plant and simulated model temperature profiles inside the main tower, while Fig. 4 (b) illustrates column internal vapor and liquid flow rates. Furthermore, in Table 3, the key tray capacity percentages until flooding are shown.

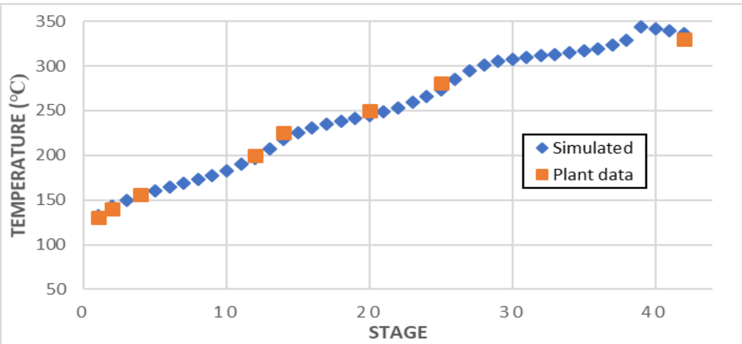

Fig. 4 (a). Main tower temperature profile (comparison of measured and simulated data)

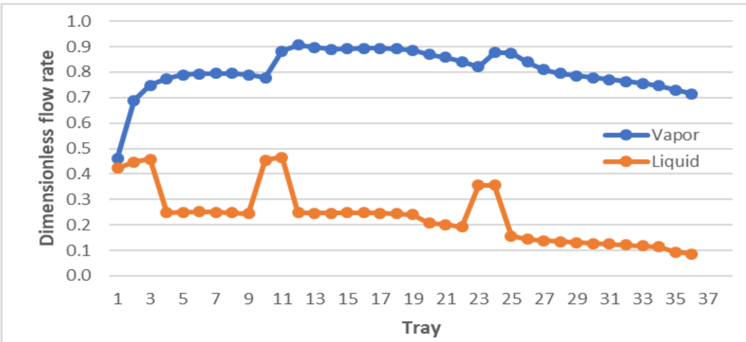

Fig. 4 (b). Liquid and vapor mass flow profiles

Table 3

The degree of flooding of the key trays of the atmospheric distillation column

Tray	% Capacity left until flooding
1 (Naphtha PA Return)	**24.0**
4 (Naphtha Product)	**14.6**
10 (Kerosene PA and SS Return)	**5.3**
12 (Kerosene Product)	**9.0**

19 (LT Diesel SS Return)	**20.1**
20 (LT Diesel product)	**22.1**
23 (Diesel PA and SS Return)	**16.3**
25 (Diesel Product)	**21.4**
34 (AGO SS Return)	**33.8**
35 (AGO Product)	**35.2**

Since the main column runs below the flooding threshold, separation efficiency and product quality requirements are not affected, whilst working near the operational limits. Still, tray 10 seems rather close to flooding and this could be a supplemental reason to use a Pre-flash unit, to enhance the performance of the main tower.

Crude distillation process requires a great deal of high-quality heat input, therefore, to increase the efficiency of the process, residual heat should be recovered and used inside of the process (heat integration). As is shown in Fig. 1, in this crude distillation process configuration, the residual heat is recovered in two distinct parts of the pre-heating HEN system: *Cold part of pre-heating train*, preceding the Desalter unit, and *Hot part of pre-heating train*, located after the Desalter unit. The rigorous process model is built to provide high fidelity simulation results for these two parts of the HEN, which recover the residual heat from main column via pump-arounds (PA) streams and from its hot distillation products, as shown in Table 4.

Table 4

Relevant temperatures and temperature changes in HEN key locations

Temperature (°C)	**Simulation**	**Plant**
Cold part of pre-heating train outlet	149	145
Hot part of pre-heating train outlet	223	221
Furnace coil outlet temperature	359	359
Naphtha PA ΔT	63	63
Kerosene PA ΔT	64	65
Diesel PA ΔT	43	45

The crude oil distillation process main data, *i.e.*, products and pump-arounds flowrates, heat exchangers outlet temperature and atmospheric zone furnace outlet temperature, are shown in Fig. 5 and Table 4. A good simulation result is obtained, with a maximum error below 3% against the real plant data, confirming the possibility to use the implemented model for analysis and retrofit purposes.

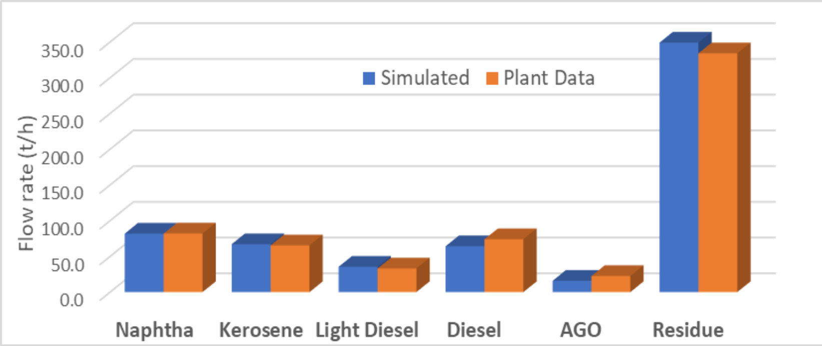

Fig. 5. Products flow rate comparison between the real plant data and simulation results

3.3. Pre-flash implementation

Based on partial results given in Table 3 and in order to improve the efficiency of the fired heater, as well as to have a more judicious distribution of the vapor phase inside the atmospheric tower, the integration of a Pre-flash dome before atmospheric tower's fired heater is decided, as a retrofit procedure, to enhance the behavior of the atmospheric crude oil distillation process unit. In the HYSYS process model, the above-mentioned integration is based upon the introduction of a Pre-flash operation unit at the end of the hot part of the pre-heating train, before the atmospheric distillation column furnace, as shown in Fig. 6.

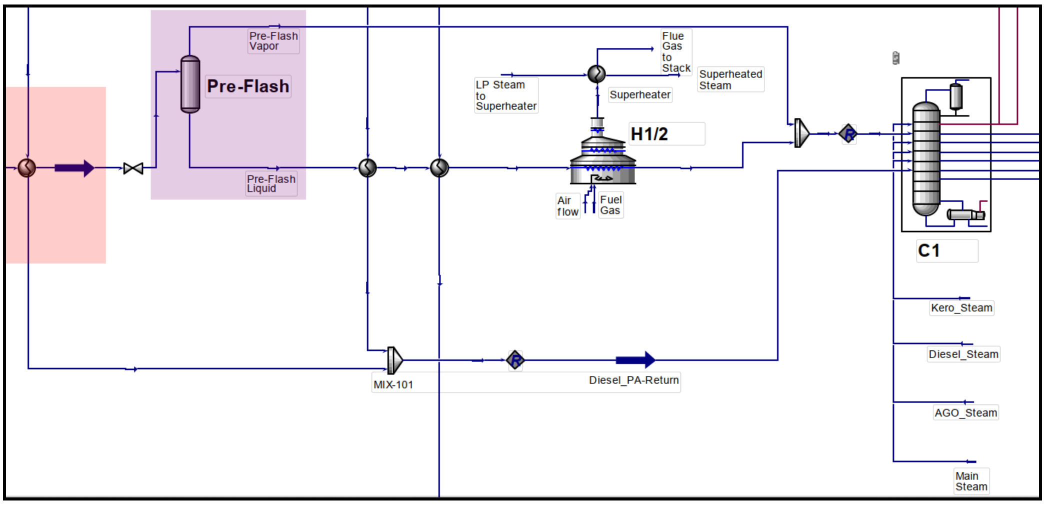

Fig. 6. Pre-flash dome location in the CDU flowsheet

The crude fired heater is one of the bottlenecks of the process and, by using a Pre-flash drum, it could be possible to enhance heat transfer efficiency of the atmospheric tower fired heater (the vapor phase is removed), thus creating the possibility for the latter to work at a slightly higher production capacity. Also, the mentioned improvement can determine a slight reduction of the fuel flow used to finalize the preheating of the crude oil that will be fractionated by distillation. A case study analysis demonstrates the capabilities of this approach and illustrates that the introduction of the Pre-flash unit can improve the crude oil distillation process performance.

To have accurate results, products quality constraints need to be defined. The D86% cut point of naphtha, kerosene, light diesel (LT Diesel), diesel, and AGO are used as active specifications in the simulation model. They are held constant at 153 ℃, 224 ℃, 268 ℃, 323 ℃ and 378 ℃, respectively, to ensure that each product meets the composition specification.

Furnaces are the major users of imported hot utilities also in the atmospheric and vacuum distillation units [3]. The impact of higher residue yields on the combined atmospheric and vacuum distillation system is considered by including the vacuum distillation furnace in the total energy demand.

The implementation of Pre-flash helps to separate vapor phase of the initial crude oil light fraction content from the liquid phase before the latter enters the atmospheric tower fired heater. The location of the destination in the process of the new vapor phase stream withdrawn from the Pre-flash should be carefully analyzed because it constitutes an important parameter in the selection of the Pre-flash design. Therefore, two possibilities could be envisaged:

- Pre-flash configuration I, in which flashed vapor stream is mixed with the furnace outlet stream and fed to the main column as in base case,
- Pre-flash configuration II, where the flashed vapors are sent to the adequate tray where the initial and end cut-points of the flashed vapor nearly match the internal liquid composition (as correspondent cut-points), improving column performance by creating additional capacity on tray 10, which operates near to the flooding point in the base case (see Table 3), while after Pre-flash retrofit, it operates 16.7 % away of flooding.

The bottom product, from which the light fractions are separated as vapors, is sent to the fired heater, in both cases, where the same outlet temperature (359 ℃), is achieved, using a slightly lower heat flow. However, for the Pre-flash Configuration II, the furnace should operate at a higher outlet temperature in order to achieve the required over flash specification. Therefore, the furnace outlet temperature was increased to 365 ℃ to compensate for the lower carrier effect, due to the vapors removed from the atmospheric column feed.

Apart from the difference in the heat flow exchanged in the fired heater for proposed configurations, another aspect that should be considered in analyzing this design performance is how the Pre-flash implementation influences the distribution of the main atmospheric tower products and pump-arounds. Therefore, the comparison between the base case and the two Pre-flash configurations is made, according to the following aspects: the distillate flow rates, pump-arounds duties and temperature drops, and minimum hot utility demand. The results presented in Tables 5 and 6 show that, by introducing a Pre-flash drum, the naphtha production is substantially increased, while that of light ends has a considerable decrease doubled by a slight decrease in kerosene. At the same time, middle distillates have a slight increase while residue flow rate decreases for the configuration II, when compared to the base case design. Furthermore, as shown in Fig. 7, the Pre-flash Configuration II model obtained better results in term of products recovery on mass basis when comparing the distillation theoretical and actual yields.

Table 5

Main column products flow rates

Product	Unit	Base case	Pre-flash Configuration I	Pre-flash Configuration II
Light ends		27.3	10.3	0.6
Naphtha		81.8	97.8	110.3
Kerosene	t/h	67.0	63.9	61.3
Diesel		99.6	99.8	102.0
AGO		15.7	16.1	15.4
Residue		349.1	351.0	348.7
Sour water		5.6	7.37	7.88

Table 6

Summary of configurations results

Variable	Units	Base Case	Pre-flash Configuration I	Pre-flash Configuration II
Main steam flow rate	t/h	2.9	2.9	2.9
Naphtha PA duty	MMkJ/h	47.0	46.4	46.2
Kero PA duty		34.1	34.1	34.1
Diesel PA duty		22.6	20.6	25.0
Furnace duty		278.0	248.0	259.0
Q_{Hmin} (minimum heating duty)		294	282	277
Column feed temperature	°C	359	359	365

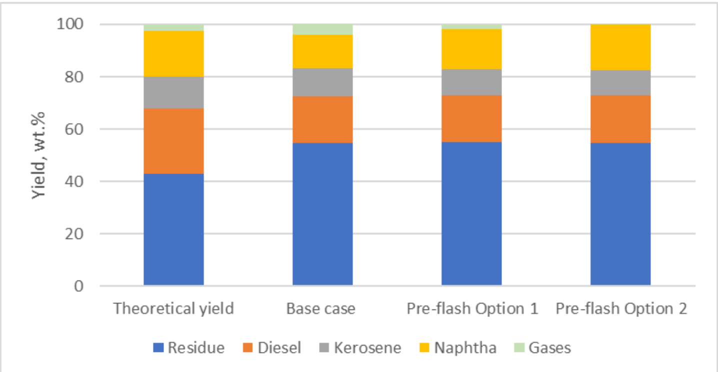

Fig. 7. Distillation product yield comparison

Therefore, the Naphtha product recovery has increased from 72.2% (base case) to 97.5% on the Pre-flash Configuration II, Diesel was increased by 1.2% and residue yield had a slight decrease (1.2%). As a drawback, there is a decrease in the kerosene product recovery by 7.5%.

Another aspect that should be considered is the influence of the product yield change on the downstream processing units and, on the market requirements as well. Improving the Naphtha production will have a limited effect on the downstream Naphtha Hydrotreater, as the D86% cut point of Naphtha is kept constant to meet the composition specification. Diesel product yield increase creates opportunity to improve refinery profitability, since many FCC or hydrocracker feeds contain 25-35% or more diesel boiling-range material, there is a significant opportunity to improve recovery. A decrease in Kerosene yield will lower the refinery jet fuel production, as shown in Fig.7. Kerosene has the lowest market value of the four products. Another aspect is the jet fuel market demand, as a reference in 2020, during the pandemic, demand has collapsed with more than 70% from the year before and is expected to still be down with some 22% until 2021 [7]. Therefore, the kerosene reduction can have a very limited effect on the refinery profitability.

HYSYS tray sizing utility is used to determine the influence of removed vapors from the feed to the atmospheric column hydraulics. Column trays are defined based on technical data available, therefore an accurate model of the column internals is obtained. Fig. 8 shows that for the Pre-flash Configuration II vapor load in the column is significantly decreased, however furnace temperature is increased to counter the lower carrier effect of vapors being removed and prevent hydrodynamic problems in the column, such as tray weeping.

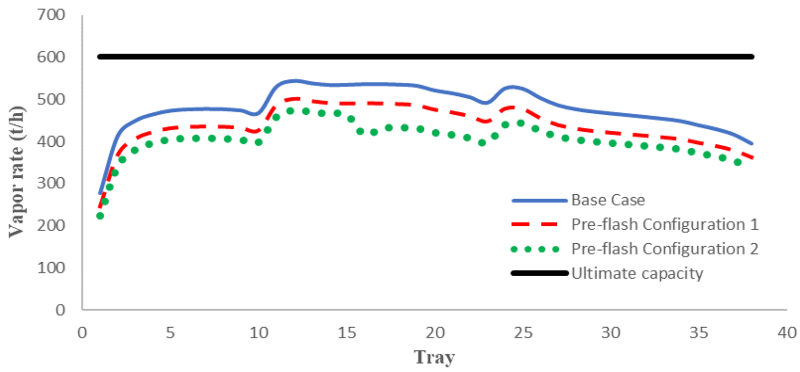

Fig. 8. Atmospheric column vapor loading

4. Economic considerations

The retrofit strategy proposed, which consists in adding a Pre-flash drum before the furnace, can improve the crude distillation process performance. As shown in Table 7, this method improves the crude distillation process operations, in terms of energy demand and carbon emissions, resulting in a certain net profit. As mentioned before, energy demand of a crude distillation process is mainly from furnace. In Pre-flash Configuration I, a 3.6% reduction in the overall energy demand and 19.6% increase in Naphtha yield is achieved. However, the Pre-flash Configuration II obtained better results, an energy demand reduction by 4.5% with a naphtha yield increase by 34.8%, as shown in Tables 5 and 7. In both cases, the energy savings attained are related to a reduction in kerosene distillate yield.

Table 7

Utility demands comparison between crude distillation system configurations

	Unit	Base Case	Pre-flash I	Pre-flash II
Cold utility	MMkJ/h	179.6	178.3	177.0
Hot utility		382.2	363.2	359.4
Carbon emission	t/h	31.4	30.3	30.0

The additional cost corresponding to the Pre-flash implementation will include the costs of the Pre-flash drum and a supplemental pump. The Pre-flash drum is assumed to be vertical, and its estimated cost is obtained from the HYSYS model as US$95,300 and Pre-flash drum bottoms pump US$10,000. The total retrofitting cost estimated as required is US$105,300. The economic impact of the proposed configurations shows an operating cost reduction of 13 % for Pre-flash Configuration I and 17% for Pre-flash Configuration II. The addition of a Pre-flash to the existing design showed substantial utility cost savings of US$2.9 MM per year for first configuration, respectively US$3.6 MM per year for second configuration. Furthermore, more Naphtha is produced in the Pre-flash presence; since the price of Naphtha is the highest of the four products (see Fig. 9), this can further increase the refinery profitability, by generating an additional estimated income of US$1.3 MM per year [6].

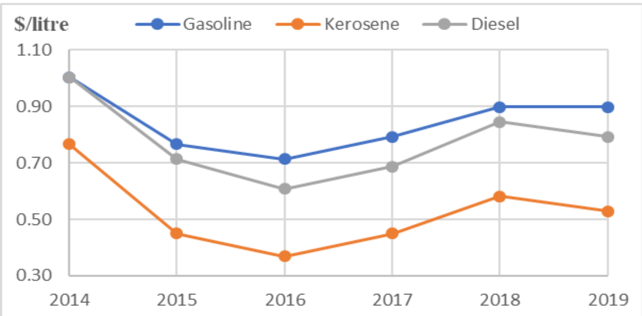

Fig. 9. Refined products average prices in Europe

The energy demand, operating costs, capital investments, and equipment cost for the two configurations were estimated using Aspen HYSYS built-in features, Aspen Energy Analyzer, and Aspen Process Economic Analyzer [5].

5. Conclusions

Two Pre-flash configurations were studied in this paper, as retrofit solution. The simulation model results were compared, and the most efficient configuration was selected. The Pre-flash configuration II showed a sustainable way to configure and operate the crude oil distillation, specifically to produce more high value products with less energy demand. The selected model showed an average of 4.5% of energy saving, associated with a 35% Naphtha yield increase. The operating cost can be reduced with 13%, doubled by a utility cost saving of US$3.6 MM per year. The required capital investment for the Pre-flash system is US$105,300 with a payback time of less than a month. In future work, the analysis can be extended to consider different types of feedstocks, light crude oil, and heavy crude oil, and determine the impact on the performance of proposed models.

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