Topping Process Using Different Methodologies.

J. Martínez[1], L.Ale Ruiz[2], J.M. Vega[1], M.C. Carrera[1], E. Erdmann[3] y E.E. Tarifa[4]

[1](Instituto de Investigaciones para la Industria Química - INIQUI-CONICET, Consejo de Investigaciones-CIUNSa, Facultad de Ingeniería- UNSa. Avda. Bolivia 5150. 4400 Salta – Argentina)
[2](Universidad Nacional de Salta. Facultad de Ingeniería. Avda. Bolivia 5150. 4400. Salta – Argentina)
[3] (Instituto Tecnológico de Buenos Aires- ITBA, Instituto de Investigaciones para la Industria Química - INIQUI (CONICET-UNSa). Av. Eduardo Madero 399 (C1106ACD) – Buenos Aires – Argentina)
[4](CONICET. Universidad Nacional de Jujuy. Facultad de Ingeniería. Gorriti 237. 4600 San Salvador de Jujuy - Argentina)

Abstract: - The first process which crude oil undergoes in a refinery is the atmospheric distillation o topping, whose aim is to fraction the crude oil in their products: naphtha, kerosene, gas oil, etc. The stages involved are: process analysis, carrying out laboratory analysis, identification of parameters, determination of operating conditions and analysis of operating procedures, data processing, stationary simulation and development of results. The column to simulate consists of a rectification column with three side separators and a condenser. For modeling this process a conventional procedure was performed using the Excel spreadsheet. And for simulating stationary was used HYSYS software. First the crude was characterized by laboratory analysis. Starting from these analyses, the TBP curve was built and global properties of crude were defined. Resolved the crude oil physicochemical, a model of the column was built. The stationary states found from simulation were compared with the results of the conventional procedure; the comparison had an acceptable degree of consistency. Finally, a series of alternatives were studied for the design and operating conditions. By resolving the designing process of the topping column, the advantages and disadvantages of a simulator can be compared with a conventional resolution method.

Keywords: - *Atmospheric Distillation, Crude Oil Plant, HYSYS, Model, Simulation*

I.   INTRODUCTION

The topping is the initial distillation of crude oil or petroleum, and consists in fractioning crude oil into the following cuts: gas, naphtha, JP or kerosene, light and heavy gas oil and atmospheric residue. Generally, it used to make the separation in one single column, which operates under a pressure slightly higher than the atmospheric one, possessing side extractions.

Nowadays, distillation unit of crude oil is a fractionation single unit, on contrast with a set of units that were the first fractionation units.

The crude oil feed to a fractional distillation tower is heated by flow through pipe arranged within a large furnace. The heating unit is known as a pipe-still furnace, and the heating unit and the fractional distillation tower make up the essential parts of a distillation unit or pipe-still. The crude oil feed is heated by furnace to a predetermined temperature. The vapor is held under pressure in the pipe-still furnace until it discharges as a foaming stream into the fractional distillation tower. Here the vapors pass up the tower to be fractionated into light and heavy gas oil, JP or kerosene, and naphtha. While the nonvolatile or liquid portion of the feed descends to the bottom of the tower to be pumped away as a bottom product.

The stripping section is the part of the tower below the point at which the feed is introduced, the more volatile components are stripped from the descending liquid. Above the feed point, the rectifying section, the concentration of the less volatile components in the vapor is reduced. The tower is divided into a number of horizontal sections by metal trays, and each of them is equivalent to a distillation unit. The feed to a tower may be at any point from top to bottom with trays above and below the entry point, depending on the kind of feedstock and the characteristics desired in the products.

The temperature of the trays is progressively cooler from bottom to top. The bottom tray is heated by the incoming heated crude oil feed. As the hot vapors pass upward in the tower, condensation occurs onto the trays until refluxing (simultaneous boiling of a liquid and condensing of the vapor) occurs on the trays. Vapors pass upward through the tower, whereas the liquids spill onto the tray below, and so on. Until the heat at a particular point is too intense for that fluid to remain in liquid condition; then it becomes vapor and joins the other vapors passing upward through the tower. The whole tower thus simulates a set of several distillers, with the composition of the liquid at any point or any tray remained fairly constant. This allows part of the refluxing liquid can be collected at various points as side stream product.

The fractional distillation tower is characterized by the chemical complexity charge. The existence of multiple side streams of product whose tenor of light hydrocarbon is adjusted in stripping columns. The heat needed to vaporize the products are given with feeding to through to load furnace and exchange with products of tower[1][2][3].

Two towers types are analyzed in this paper, Type I and Type II Towers. In Type I Tower all the heat is extracted in the condenser head, so that the condensate does not have appropriate heat level to achieve a good heat recovery. Because of the reflux is generated in head, liquid and vapor internal charges in the towers are increased from the bottom to the top substantially. To overcome this problem using a Type II Tower, where is extracted fluid from a plate between two side extractions of product, it makes exchange heat with another stream of the process (usually crude oil) and return to introduce some trays above; in the same tower section, this is a circulating reflux. This operation reduces the area for the condenser head, the fuel consumption of load furnace, internal charges of liquid and vapor above the return of circulating reflux in the tower, thus enabling the use of a minor section tower in that zone. By allow standardization of internal flow and that the diameter of the tower is the most suitable for the entire length of the said tower.

To model a process containing crude oil, the properties of the oil must be known or at least estimated. Determining the exact composition of crude oils is not feasible because of the enormous number of components comprising these mixtures. The crude oil is characterized then by a distillation curve done according to standard procedures (TBP, ASTM D86 or D1160 distillation, etc.). Then, existing correlations are used to estimate the physical properties of crude oil from temperatures that conform the distillation curve. In this way, a crude oil mixture may be modelled as a group of hypothetical components whose properties are estimated from different portions of the boiling distillation curve. Experimental measurements such as density, viscosity, etc. generate new curves that allow adjustment of crude oil estimates. To complete the characterization, molecular weight and density are measured for the overall crude oil sample.

This paper describes the development of a topping stationary simulation, for Type I and Type II Tower. The method is presented by using HYSYS simulator. Starting from the True Boiling Point (TBP) distillation curve obtained on experimental basis global properties of crude oil were defined. With TBP curve and the global properties, defined the hypothetical components required for HYSYS (University License. Version 7.3-2013) were defined to model physicochemical crude oil. Once the physicochemical was determined, the tower model was built. The results were compared with those obtained by conventional method[3] using Excel spreadsheet.

II.   METHODOLOGY
2.1 PETROLEUM CHARACTERIZATION
Laboratory standard methods for distillation of crude oil and fractions are of great value in refining and formulating fuels. Three of the most important assays are [4] [5] [6] ASTM (considered the universal, normalized method), TBP (whose results are difficult to reproduce), and Flash (instantaneous vaporization) [7].

2.1.1 The American Society for Testing and Materials (ASTM) curve
The ASTM distillation employs batch fractionation apparatus using non-refluxed Engler flasks, which possesses standard dimensions. The heat flow is adjusted to produce 5 cm^3 of distillate per minute. The temperature is registered when the first drop of distillate falls, and then every time that 10% of distillate is collected.

2.1.2 The TBP distillation curve
The TBP distillation is not standardized. The distilled is obtained from the division into fractions of the load components. A getter device for crude oil and its derived products is used, which is loaded with 1000 to 5000 cm^3 of liquid. The fractioning column is surrounded by a heater that compensates the lost heat. A condenser at the top produces the vapor condensation, and the condensate descends; when it makes contact with the vapor, the separation takes place. The temperature of the first drop, as well as for each 10% of distillate, is registered.

2.1.3 The Equilibrium Flash Vaporization (EFV) distillation curve
The sample is loaded on a receptacle; from there it is sent to a serpentine that is in a hot enclosure. The vapors that are generated do not separate the liquid and remain in contact with it, until equilibrium is reached. Then, the liquid is picked up and the vapor is condensed. Different operations are made with the same sample amount and time period; but at different temperatures. The curve is obtained by plotting the corresponding the liquid-vapor equilibrium temperature vs. percentage of distilled liquid volume.

2.2 TOPPING PROCESS
The designing of a topping column involves process design and equipment dimensionally. The process design understands to calculate all operation conditions: head, bottom and feeding pressures; temperature of side

streams, feeding and products flows. Matter and energy balances should be performed. The column to be analyzed is composed by a reflux absorber, with three lateral separators and a condenser head.

The analysis is carried out using two methods: conventional procedure and simulation with HYSYS. In the first place, crude oil is characterized through laboratory analysis execution. Starting from these analyses, the corresponding TBP curved is built (to 760 mmHg). This curve is presented in the Fig. 1.
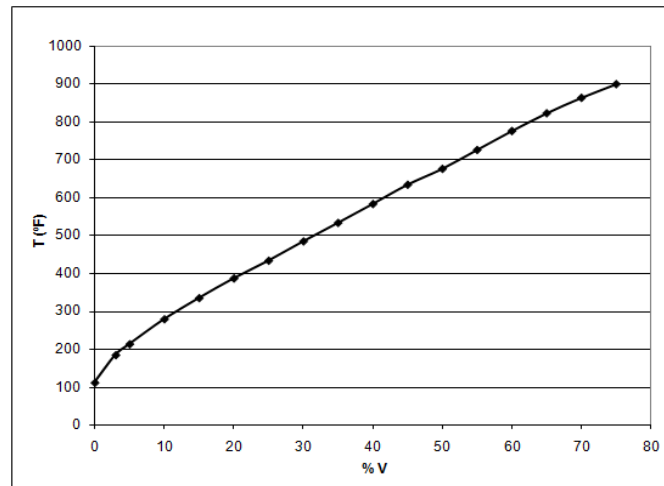

Fig. 1. TBP DISTILLATION CURVE.

2.2.1 CONVENTIONAL PROCEDURE

For topping design by conventional procedure is necessary to transform the data distillation curve from TBP to EFV, Tarifa et al. (2009) [7] presents a new method to estimate the equilibrium flash vaporization (EFV) distillation curve from standard laboratory analytical assay procedures. The cuts of this process should take place with ASTM maximum temperature specifications presented in Table 1.

Table 1. ASTM Maximum Temperatures Of Cuts.

Cuts		Max Temperature
N	NAPHTHA	170°C-338°F
JP	JP (kerossene)	245°C-473°F
GOL	GAS OIL (Light)	300°C-572°F
GOP	GAS OIL (Heavy)	345°C-653°F

2.2.1.1 Tower Design

A Type I Tower serves as basis. The number of tray is defined after having found the cutting temperatures and performance; for this problem the adopted trays quantity is 29. The head pressure is adopted in 5.5 to 6 psi keeping in mind that between the last tray and the separator there is a load loss of 5 psi and, a load loss 0.2 psi per tray.

2.2.1.2 Products Characterization

The TBP cut point is determined following the Watkins method on the crude TBP curve, obtaining the products volumetric performances starting from the specifications given. Afterwards, the following tasks are accomplished:
➢ Construction of products TBP curves.
➢ Verification of products specifications.
➢ Determination of products properties.
➢ Realization of matter balance.

Once the products properties are obtained, the calculations by zones are begum. The involved areas are those shown in the Fig. 2.
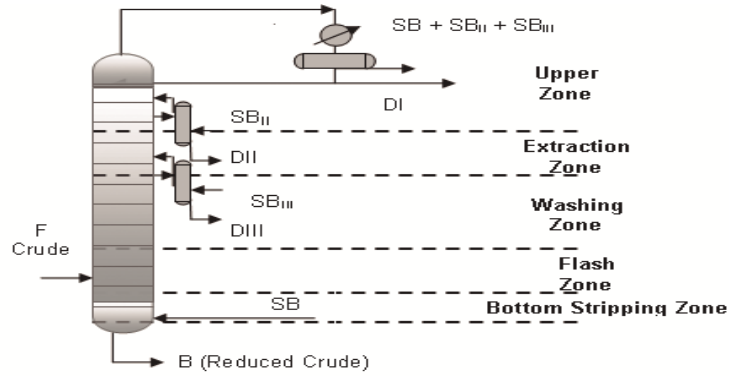
Fig. 2. TOPPING TOWER ZONES.

2.2.1.3 Flash Zone

      This zone has direct connection with the exterior. The crude oil feed (F) enters in the feeding tray that corresponds, in this case, to tray 4, to a temperature that does not exceed the 350 ºC (660 ºF), to avoid cracking. The temperature (t_{ZF}) should be such that allows that the products to obtain for head, for lateral strippers and a certain crude oil additional quantity (call over flash: OF) are vaporized. The calculation procedure is presented below:

Data: F (crude oil flow) and B (residual flow).

Unknown Variables: SB (water vapor of stripping), V_{SB} (pulled hydrocarbons by bottom vapor), OF (over flash), V_0 (vaporized crude oil), t_{ZF} (flash zone temperature), Lo (crude oil without vaporizing), L_{ZF} (liquid that enters to stripper), V_{ZF} (vapor that enters to washing area), t_4 (tray 4 temperature) and t_5 (tray 5 temperature). Equations of matter balances, (1) (2) (3).

$$V_{ZF} = V_0 + V_{SB} \tag{1}$$
$$V_{ZF} = N + JP + GOL + GOP + OF \tag{2}$$
$$V_0 = N + JP \tag{3}$$

2.2.1.4 Bottom Stripper Calculation

      Matter and energy balances are established in the bottom tower with the purpose of calculating the bottom temperature, t_B.

Data: F; B; SB; V_{SB}, OF, V_0, t_{ZF}, Lo, L_{ZF}, V_{ZF}, t_4, t_5.

Unknown Variables: t_{SB} (temperature of water vapor of the stripping), t_B (residual (B) temperature).

      Knowing the properties and the temperatures corresponding to each flow, the required enthalpies are obtained to solve the energy balance in the zone and to find t_B.

2.2.1.5 Washing Zone

      The washing zone is comprised between the feeding and the last side stream (from tray 5 to tray 7). The calculation procedure is presented below:

Data: S_B, OF, t_5, products performances (N, JP, GOL, GOP), t_{ZF}, V_{ZF}.

Unknown Variables: S_i (water vapor of lateral stripping), V_{Si} (despoiled vapor in the lateral stripping), t_7 (temperature of the side stream tray), t_6 (vapor temperature that ascends toward extraction tray), L_7 (liquid flow that leaves the extraction tray).

Equations of matter balances, (4) (5)

$$V_{ZF} = N + JP + GOL + GOP + OF \tag{4}$$
$$V_6 = N + JP + GOL + GOP + L_7 \tag{5}$$

Equation of energy balance, (6)

$$Q_{SB} = Q_{VZF} + Q_{L7} = Q_{V6} + Q_{SB} + Q_{OF} \tag{6}$$

      To obtain t_7, the range of this current (V_{SGOP} + GOP), is extended from 0 - 100%, starting from the TBP curve of the crude oil. The 0% atmospheric Flash temperature of the current is obtained; with this value (starting from graphics) we obtain the value of the lateral extraction temperature, t_7. For lineal interpolation between t_5 and t_7, t_6 is calculated. The variable L_7 is determined making an enthalpy balance in the lateral stripper. Knowing the properties and temperatures corresponding to each flow it is possible to obtain the necessary enthalpies to solve this energy balance.

2.2.1.6 GOP – GOL Extraction Zone

      The calculation procedure is presented below: equations of matter balance, (7) (8) and equations of energy balance, (9).

$$V_{13} = N + JP + GOL + L_{14} \tag{7}$$
$$V_6 = N + JP + GOL + GOP + L_7 \tag{8}$$
$$Q_{V13} + Q_{SB} + Q_{SGOP} + Q_{L7} + Q_{GOP} = Q_{L14} + Q_{V6} + Q_{SB} + Q_{SGOP} \tag{9}$$

Estimate of GOL extraction temperature, t_{14}: to obtain t_{14} (bubble temperature), the range of this current ($V_{SGOL} + GOL$) is extended from 0 - 100%, starting from the TBP curve of the crude oil. The 0% atmospheric Flash temperature of the current is obtained, with this value (starting from graphics) is obtained the value of the lateral extraction temperature, t_{14}. For lineal interpolation between t_7 and t_{14}, t_{13} is calculated. The variable L_{14} is determined making an enthalpy balance in the lateral stripper. Knowing the properties and temperatures corresponding to each flow is possible to obtain the necessary enthalpies to solve this energy balance.

2.2.1.7 GOL - JP Extraction Zone
The calculation procedure is presented below: equations of matter balance, (10) (11).
$$V_{19} = N + JP + L_{20} \tag{10}$$
$$V_{13} = N + JP + GOL + L_{14} \tag{11}$$

In order to estimate the extraction temperature of JP, t_{20}, the procedure is identical to the previous extraction zone, the same as the energy balance to find L_{20}. All the estimated temperatures should be verified, if the calculated temperature and the estimated one differ by 10 ºF, the calculation should be restarted, which is iterative until obtaining the difference of permissible temperature.

2.2.1.8 Upper Zone.
For appreciation of T_D, it is proposed $t_{28} = t_D$, corresponding to the dew point temperature of the head product: Naphtha (N), that is to say the 100% EFV (atm) curve temperature of the cut N.
The calculation procedure is presented below:
A head temperature, somehow lower than the dew point temperature of the head product ($t_{100\%}$) is adopted.
For interpolation with t_{20}, t_{27} can be calculated.
The heat and mass balances is carried out to calculate internal reflux (Ri), that flows from the plate 28. It is considered that the properties of Ri are similar to N.
For the external reflux (Re) calculation an enthalpy balance in the last tray is carried out, fixing a temperature in the condenser. It is considered that Re has the same composition that N.
The thermal loads of condenser (Qc) is the heat that should be extracted of the condenser. By means an enthalpy balance in the condenser, the Qc is obtained.

2.2.1.9 Modification Type II Tower
Circulating refluxes between two lateral extractions are incorporated to the basic Tower. Choosing the introduction place of circulating reflux where the vapor flows upward and the descending liquid manifest a change of considerable slope. The action of the circulating reflux produces a decrease of liquid load and internal vapor on top of the reflux return to the tower.
The temperature from the immediate superior extraction tray at the recycle return is estimated; the trays temperatures involved in the recycle are calculated by lineal profile, the mass of circulating reflux and the extracted heat are fixed. The liquid current flow that leaves of superior immediate lateral extraction tray (will appear much smaller as greater is the extracted heat) is calculated by thermal balance. The calculations are repeated in similar form to those carried out for the Type I Tower to find the rest of the flows and temperatures. The conventional calculation procedure is very long and tedious, having to read permanently petroleum industry characteristic graphic and abacus. A simple change implies a great work for to turn to find flows values, temperatures and pressures again for up of this change.

2.2.2 PROCEDURE WITH HYSYS SIMULATOR
The hypothetical components that HYSYS requires to model the physical-chemistry of the petroleum were defined with TBP data and global properties. The main flowsheet of the simulator of processes HYSYS is shown in Fig. 3. For this model, the properties package choosing to be used in the fluid package is Peng Robinson [8]. The crude oil example is fed to the pre-flash drum, modeled as a *separator*, where vapor are separated from the crude oil liquid. The liquids are then heated to 350 ºC (660 ºF) in the crude oil furnace, modeled as a *heater*. The pre-flash vapors bypass the furnace and re-combined, using a *mixer*, with the hot crude stream. The combined stream is then fed to the atmospheric column for separation. The crude oil column is modeled as a *refluxed absorber*, equipped with three side-stripper operations.
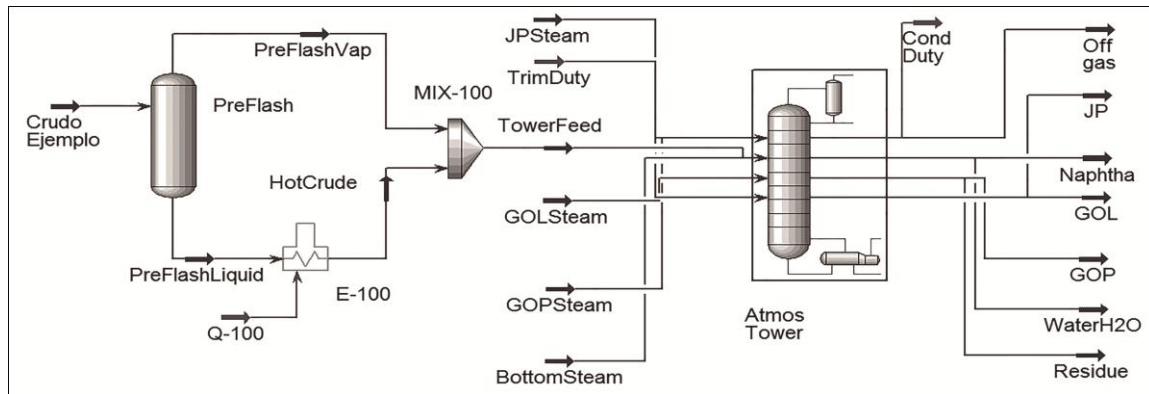
Fig. 3. PROCESS FLOWSHEET.

The data of the crude oil entered in simulator are: the molecular weight is 243, the density is 855 Kg/m^3 and the relative densities curve and TBP temperatures curve. Both curves are presented in Table 2. For this case, the physical-chemical package recommended is Peng Robinson. The feeding conditions are: volumetric flow of 500 m^3/h, temperature of 319.4 ℃ (606.9 ℉) and pressure of 14.7 psia.

Table 2. Densities (D) and TBP Temperatures (T)

% Acumulate Volume	D	T	
		℃	℉
3	0.710	83.89	183
5	0.728	100.00	212
10	0.746	136.70	278
15	0.766	167.80	334
20	0.781	196.70	386
25	0.795	222.80	433
30	0.810	251.10	484
35	0.823	278.30	533
40	0.837	306.10	583
45	0.849	334.40	634
50	0.860	357.80	676
55	0.870	385.60	726
60	0.880	413.30	776
65	0.887	439.40	823
70	0.894	462.20	864
75	0.899	482.20	900

.

III. RESULTS

The yield of each product on crude oil was obtained with the crude oil characterization data and the cut temperatures entered in the HYSYS simulator, the same ones were compared with those calculated in Excel spreadsheets, Table 3.

Table 3. Yields on Crude Oil

Products	Excel	HYSYS	% difference
N	15.5	15.6	0.6
JP	13.5	13.3	1.5
GOL	11.0	10.0	9.1
GOP	9.5	8.4	11.6
Residue	50.5	52.7	4.2

The topping process corresponding to Type I Tower is simulated, maintaining the conditions of the conventional method. See the flowsheet of the tower in Fig. 4. The main column consists of 29 trays, plus a partial condenser. The feeding (TowerFeed current in the flowsheet) enters on tray 4, while superheated steam

(BottomSteam) is fed to the bottom stage. In addition, the trim duty is represented by an energy stream feeding onto tray 4. The naphtha product is produced from the condenser. Each of the three-stage side strippers yields a straight run product. JP, Gas Oil Light and Heavy (JP, GOL y GOP) are produced from the steam-stripped JPSS, GOLSS and GOPSS side strippers, respectively.
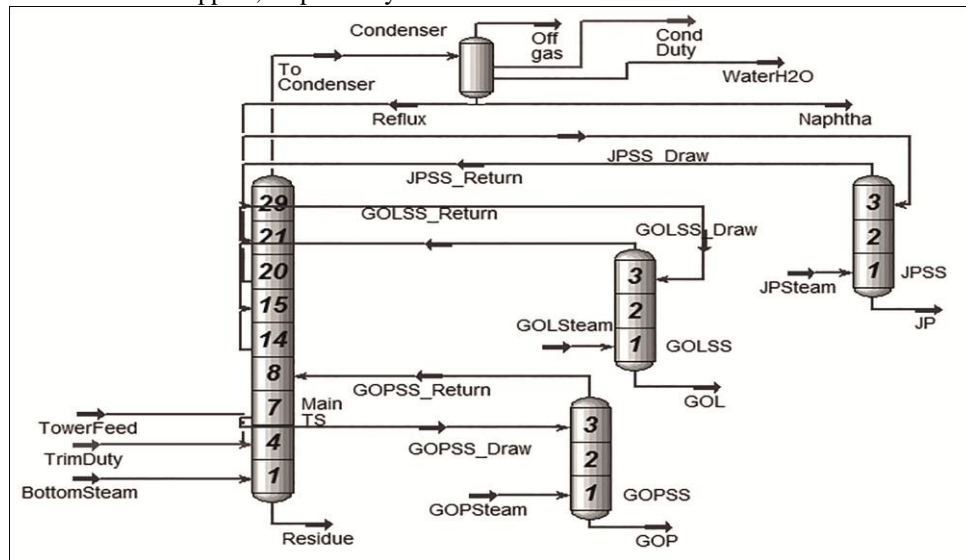

Fig. 4. FLOWSHEET OF TYPE I TOWER.

The variation of liquid and vapor molars flows (kgmol/h) along the tower is observed in Fig. 5, obtained by means HYSYS simulation and conventional procedure. The curves with rhombus correspond to the liquid molar flows and the curves with square to the vapor molar flows. The conventional procedure provides fewer points in the curves, since they are carried out balances of matter and energy by zones. While the simulator carries out equilibrium and balances calculations tray to tray.
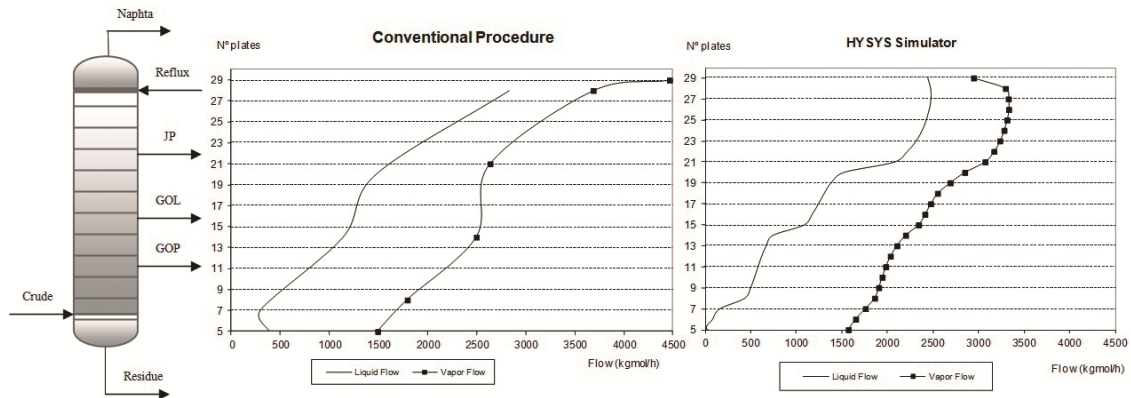

Fig. 5. VARIATION OF MOLAR FLOWS WITH RESPECT TO THE NUMBER OF TRAY. TYPE I TOWER.

Because the whole reflux is generated in the head, the liquid and vapor internal loads in the tower are considerably increased from bottom toward top, therefore it is necessary to recover the heat of head vapor. One of the ways is by exchanging with cold crude oil, through circulating refluxes, leading us to the resolution of a Type II Tower. The bombs are placed to get the circulating reflux; this reflux enters to a heat interchanger where the current cools down and returns to the tower some trays up above. In order to analyze Figure 5 it is important to highlight a change in the slope of molars flows curve in plate 14.

The recirculation is coincidently introduced where the changes of slopes of the Type I Tower curve are observed. The bomb feeds of tray 14 and the cooled reflux returns to the column in the tray 16, this situation can be observed in Fig. 6.
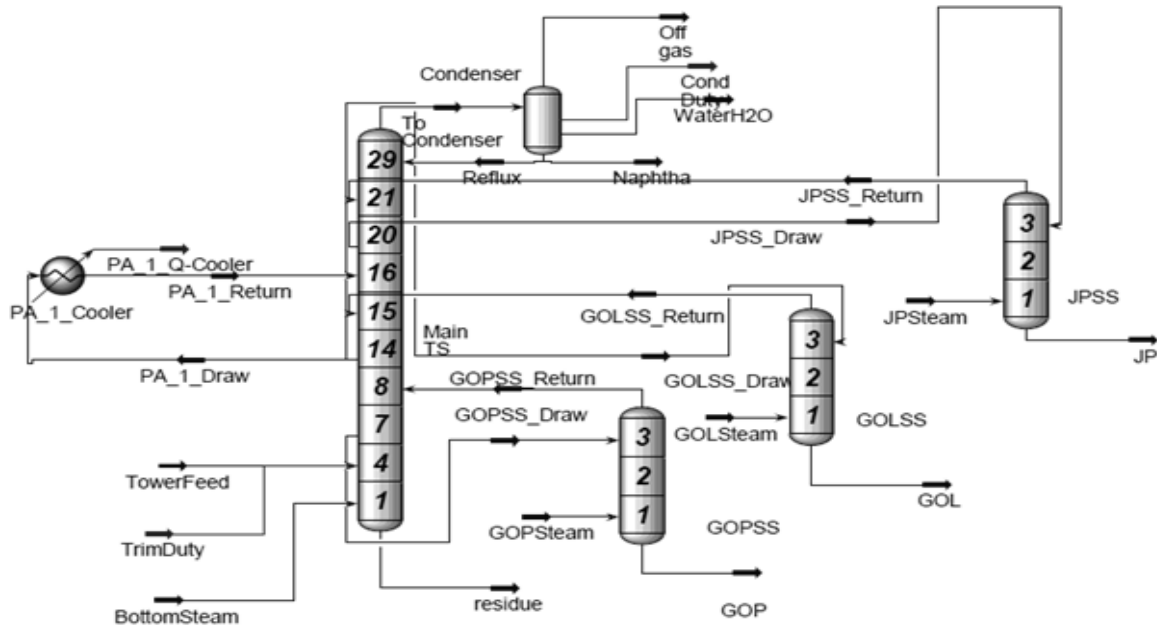
Fig. 6. FLOWSHEET OF TYPE II TOWER WITH A RECIRCULATING FLOW.

The data required to install a bomb in the tower are the recirculating flow and temperature drop (ΔT) among the entrance current to the bomb and the current that returns to the tower, is considered as recirculating flow to 40% of liquid volumetric flow that these tray and a temperature drop of 100 ℃.

A marked slope change in the curves of the flows of Type II Tower in tray 20 is also observed. Thus, a second bomb is added, Fig. 7. This bomb feeds from tray 20 and the cooled reflux returns to the column in tray 22.
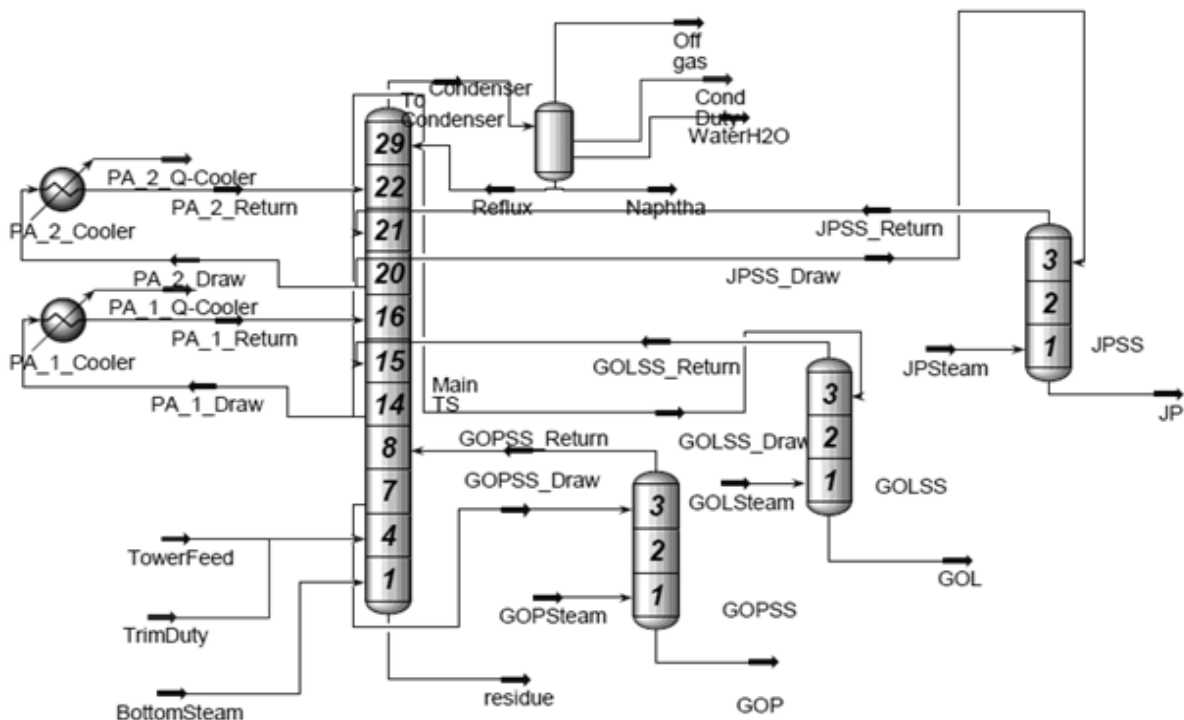

Fig. 7. FLOWSHEET OF TYPE II TOWER WITH TWO CIRCULATING REFLUXES.

The same considerations as in the first one are carried out to install this second bomb. A Type II Tower with two bombs by HYSYS simulator is solved, the flow variation with respect to number of trays for a Type II Tower with 2 pumps can be seen in Fig. 8.
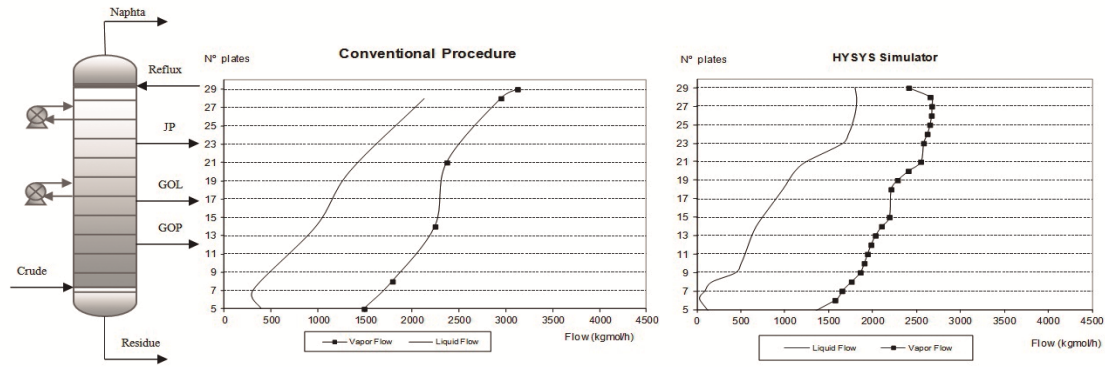
Fig. 8. VARIATION OF MOLAR FLOW WITH RESPECT TO THE NUMBER OF TRAY. TYPE II TOWER.

The differences between the molar flow values can be compared in Figures 9a and 9b with respect to the number of trays for Type I and II Towers. An improvement in the curve slope of Type II Tower with regard to Type I in trays 14 and 20 is seen, coincident with the introduction of the respective circulating reflux.
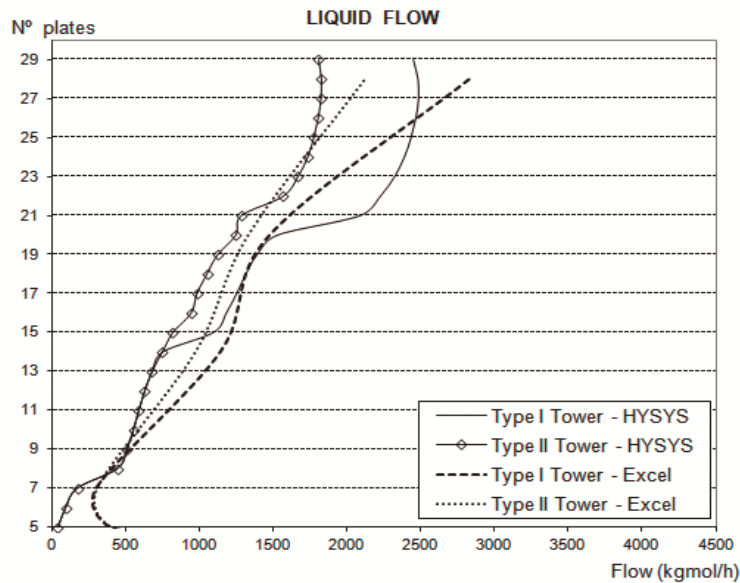

Fig. 9a. Variation Of Liquid Flow With Respect To The Number Of Trays For Type I And Type Ii Towers.
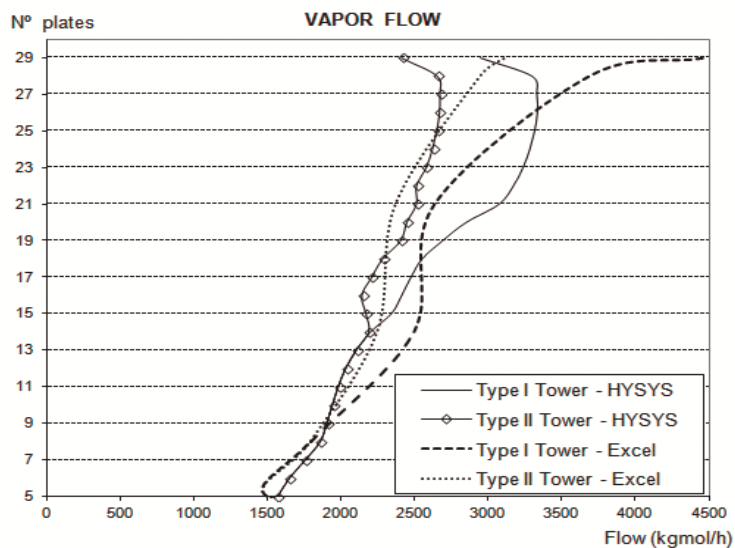

Fig. 9b. Variation Of Vapor Flow With Respect To The Number Of Trays For Type I And Type Ii Towers.

The purpose of carrying out a modification of Type I to Type II is reflected in Fig. 9, where a decrease in the flow slope in function of the tower height can be seen. This means that the liquid and vapor internal decrease beyond the return from circulating reflux to the tower, thus allowing the possibility of using a smaller section in this area. The temperatures of vapor - liquid equilibrium (Equilibrium Flash Vaporization EFV) are shown in Table 4, both the Type I and Type II are very similar. The molar flow of products obtained by simulation and conventional procedure (HYSYS and Excel) for Type I and II Towers are presented in Table 5.

Table 4. Products Temperatures (ºF).

Products	Type I Tower. Without Recirculation.	
	HYSYS	Excel
Naphtha	273	283
JP	399	371
GOL	539	498
GOP	663	598

Table 5. Products molar flow (kgmol/h).

Products	Type I Tower. Without Recirculation.		Type II Tower. With Recirculation.	
	HYSYS	Excel	HYSYS	Excel
Naphtha	553.9	543.7	553.6	543.7
JP	335.8	326.9	335.9	326.9
GOL	211.1	213.6	211.3	213.6
GOP	142.6	154.5	144.0	154.5

In the conventional calculus the specified temperatures are the starting point for the calculation of both towers. These temperatures are taken as the final temperatures of the product and they give the yields of the crude for the later calculation of molar flow values. The difference between Type I and II Tower can be seen mainly in the variation of internal flows and in the energy requirement of each condenser. It can be observed that the Type II tower requires less energy.

The HYSYS simulator can calculate a Type I and Type II Tower and the temperatures of the final products obtained by simulation fulfill the specifications given in Table 1. The values obtained in the simulation are not very different from those calculated by conventional procedure.

The TBP curves of each product of topping must be transformed into EFV curves in order to verify if the simulation produces the commercial cuts previously defined. By definition, the EFV curve is generated by a serie of experiments carried out at constant pressure (usually at 1 atm). The result linking equilibrium temperature with percentage of distilled liquid volume where the total vapor is in equilibrium with the liquid that is not evaporated. This curve is widely used in the design of equipment for the petrochemical industry [6]. The experimental procedure for obtain the EFV curve is complicated; this curve is usually estimated from other curves that involve simpler experimental procedures. The Edmister and Maxwell methods are commonly used to estimate the EFV curve. These methods use a graphic procedure which require data obtained from diagrams and tables.

The comparison of EFV curves of each products obtained from the conventional method and simulation (Tarifa et al. (2009) [7]) was performed. The EFV curve of naphtha is presented in Fig.10, where a small difference between the two methodologies is observed, matching in the 100% value of the mentioned curve.
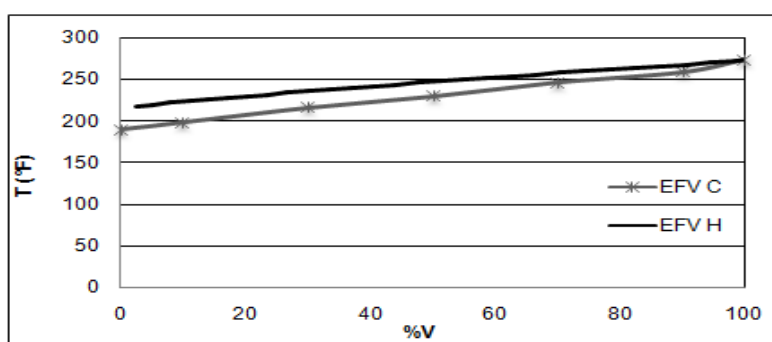

Fig. 10. Efv Curves Of Naphtha. (H) Obtained From Hysys (C) Obtained From Conventional Procedure.

IV.   CONCLUSION

The distillation is a fundamental process in the crude oil refining industry. It allows fractioning hydrocarbons of a crude cut taking advantage of their different boiling points. The first process which crude oil undergoes in a refinery is the atmospheric distillation. In that stage, crude oil is separated in fractions which, after some additional processing, will give origin to the main products that are sold in the market: LP gas (used in domestic furnaces), gasoline for automobiles, turbosina for airplanes jet, gas oil for automobiles. Another process related to topping is vacuum distillation of the bottom product of atmospheric distillation.

The designing of a Type I tower allows, furthermore, the analysis of the internal flows and uniform them through a Type II tower, with the purpose of obtaining an adequate diameter of the tower.

Working with the HYSYS simulator allows, without mayor efforts, to carry out changes, for example, something very usual in the industry, the exchanging the properties of the crude oil to be processed and, at the end, the distillation curve.

The process design of the Type I and Type II topping columns is performed using conventional methods and a process simulator. In this work, the advantages and disadvantages of each method have been analyzed.

The simulation can be applied to carry out a parametric sensibility study in order to fractionate a crude oil, from a petroleum characterization curve.

REFERENCES

[1]   R.N. Watkins, Petroleum Refinery distillation. (Gulf Publishing Company. Houston Texas. 1981).
[2]   R.N. Watkins, How to design crude distillation. (Special Report. Hydrocarbon Processing. December 1969).
[3]   Ojeda, A.L. et al. Destilación Aplicada a Sistemas de Hidrocarburos (YPF-PGM. 1978. Tomos I y II).
[4]   P. Wuithier, Refino y Tratamiento Químico: el petróleo. Cepsa S.A. and Marcel Dekker Inc. Madrid, Spain. 1971.
[5]   A Cerutti,. La Refinación del petróleo (Ed. IAPG. Buenos Aires, Argentina.2002. Tomo I).
[6]   L. Cohen, Diseño y Simulación de Procesos Químicos (Editor León Cohen Mesonero. Segunda Ed. Algeciras, Spain. 2003).
[7]   E.E Tarifa, E. Erdmann, D. Humana y J. Martínez. A New Method for Estimating the EFV Distillation Curve. Petroleum Science ND Technology, 27:3,331-344. 2009.
[8]   Aspen Hysys v. 7.3.2.- Aspentech. Aspen Technology, Inc. Cambridge, USA. 2012.
[9]   Speight, J. The Chemestry and Technology of Petroleum. Marcel Dekker Inc. New York. 1999.
[10]   J.P.Wauquier, El Refino del petróleo. Petróleo Crudo. Productos Petrolíferos. Esquemas de Fabricación. 1era edición. Instituto Superior de la Energía, ISE. Ed. Diaz de Santos. 2004.
[11]   M. S. Lopes, M Saviolo Lopes, R. Maciel Filho, M.R. Maciel, L. C. Medina. Extension of the TBP curve of petroleum using the correlation DESTMOL. Procedia Engineering 42 (2012) 726- 732.
[12]   V. Kumar, A. Sharma, I. R. Chowdhury, S.Ganguly, D. N. Saraf. A crude distillation unit model suitable for online applications. Fuel Processing Technology 73 (2001) 1-21.
[13]   T. Cheatterjee, D. N. Saraf. On-line estimation of product properties for crude distillation units. Journal of Process Control 14 (2004) 61-77.
[14]   M. Errico, G. Tola, M. Mascia. Energy saving in a crude distillation unit by a preflash implementation. Applied Thermal Engineering 29 (2009) 1642- 1647.
[15]   G. Argirov, S. Ivanov, G. Chalokov. Estimation of crude oil TBP from crude viscosity. Fuel 97 (2012) 358- 365.
[16]   M. A. Fahim, T. A. Alsahhaf and A. Elkilani, Fundamentals of Petroleum Refining. (Imprint Elsevier Science and Technoloy. 2009).