University of Sadat city,
Environmental Studies and Research Institute
The4Th International Conference
Environmental for Studies and Research Smart Sustainable Environment
27-29 March 2017

Journal of Environmental Studies and Researches (2017), 7(2-D): 883-905

ENVIRONMENTAL AND ECONOMICAL OPTIMUM INTEGRATED DESIGN OF CRUDE OIL DISTILLATION UNIT.

Ahmed M. M.Samy[1], Mamdouh A. Gadallah[2], Asmaa Abdelsattar[1], Mahmoud Abdelhalim[1]

[1] Dept. of Chemical Engineering, Minia University, Minia 61511, Egypt
[2] Dept. of Chemical Engineering, the British University in Egypt, Cairo11837, Egypt

ABSTRACT:

Distillation is an energy intensive process in oil refining. The huge heat supplies needed for heating the crude require burning a large amount of fuel, that results in a considerable amount of greenhouse emissions which affect the environment badly. Also hot products require large amounts of cooling water to be cooled down to the storing temperatures, which is another environmental load. Carbon dioxide presents the majority of emissions, so that we are concerned about Co2 emissions reduction and decreasing the cooling water demands. As these two environmental objectives require huge financial investments, this work tries to find a compromise between the environmental and economical sides.

This approach suggests a compromise which is lowering the overall environmental load by burning less amount of fuel per unit mass of processed crude oil. It includes Optimization of the operating conditions and overall heat integration. Besides the environmental benefits this solution reduces the cost needed for fuel, the carbon tax as the emissions are reduced and the cost of cooling water as cooler products are obtained. The approach is valid for both the retrofit and grassrout designs.

A basic design of distillation system is created to process 150,000 bpd of crude. The basic design is simulated by Aspen HYSYS software and then optimized by factorial design of trials using DOE v10.1.6 software for statistical analysis and regression. The heat integration is achieved by designing a heat exchanger network using Aspen Energy Analyzer. This leads to an optimum integrated design with the lowest possible cost.

KEYWORDS: *Distillation, Environmental and Economical Design, Crude Oil, Co2, Cooling Water, Heat Integration, Heat Exchanger Network.*

1.INTRODUCTION

An equivalent of 1% to 2% of the crude oil processed, is consumed to provide the sufficient amount of heat required for the distillation process , so that it's considered to be a highly energy intensive process . Considerable effort has been made to reduce the energy requirements of the crude oil distillation process as the price of energy goes up. These efforts in parallel with the

increasing concerns about the environment, resulted in stricter regulations on the emissions of green house gases. Consequently, both economic and environmental issues are essential factors in our design of crude oil distillation system.

The distillation column included in the crude oil distillation system is strongly interacted with the associated heat recovery system compared to the conventional design approach of crude oil distillation systems. The heat-integrated design approach using pinch analysis techniques finds way better solutions compared to the conventional design approach of crude oil distillation systems. Less energy consumption leads to less gas emissions subsequently, which is an advantage for the environment. The heat-integrated design approach is facilitated by shortcut column models and the pinch analysis method. In order to apply shortcut column models, product specifications in the refineries need to be translated into the form of specifications that the shortcut column models require. However, there are a number of limitations present in the existing translation methods. This thesis targets to extend the existing methods and overcome their limitations.

In the case of grassroots design, the general objective is to design a new eco-system , which fulfils the specified separation requirement. The design problem comprises two sub-problems: design of the separation devices and design of the preheat train. Many of design issues are interlinked with each other; for example, the feed temperature affects the vapor and liquid flow rate at the feed location, which in turn changes the separation taking place and the number of theoretical stages and the diameter of each column section. The capital cost of the distillation column depends primarily on the diameter and height of the column. The distillation columns interacts strongly with the preheat train, where pump-around streams and hot products of distillation columns heat up cold streams, especially the cold crude feed. The remaining heating and cooling requirements are fulfilled by hot and cold utilities. The grassroots design of the preheat train involves determining the configuration of the network and the required heat transfer areas. The HEN design thus determines its capital cost and utility cost. As the heating and cooling requirements are governed by the design of the distillation system, the design of the preheat train and distillation columns are inter-dependent. Designing them both in single framework, rather than sequentially can allow these links to be exploited, which will lead to a better performance of the overall system, for example in terms of energy efficiency or total annualized cost.

2. DESIGN AND SIMULATION OF CRUDE OIL DISTILLATION SYSTEM

The establishment of a crude distillation column design require some basic steps:

1- The product specifications should must be set
2- The operating conditions is to be selected, i.e operating pressure.
3- Determining the number of equilibrium step and the reflux requirements.
4- Selection of the contacting method i.e plates or packing.
5- Column sizing, the appropriate dimensions (diameter and number of stages).
6- Choosing the column internals mechanism.
7- Mechanical requirements of design (vessel and internal fittings).

According to how heat is removed the column and attachments are arranged. These arrangements are mainly top tray reflux, pump back reflux and pump around reflux. In Top tray

reflux the reflux takes place only at the top tray which means the heat is add through the bottom and removed at the top which requires a large column diameter and results in a fraction of poor quality due to the improper reflux, also this arrangement is not suitable for the economic utilization of the heat which makes it totally inappropriate for those who seek an eco-design. Pump back reflux is provided at equal intervals what makes every plate acts like a separate fractionator considering the good amount of liquid provided, as the column is uniformly loaded that results in a smaller column diameter and allow the external reflux to be utilized economically, however a design of this arrangement requires a large investments, it provides a very high performance. For pump around the reflux is drawn from a lower plate and fed into a higher plate (1-3 stages higher), that creates a local mixing problem as uneven compositions of reflux and liquids mix in a single tray, designers solve that problem by treating all the plates at the zone as a single plate.

The column consists of two sections, rectifying or enrichment section which's above the feed point and stripping section which is below the feed point. Reflux ratio is a major concept crude distillation column design. It is the ratio between the amount of liquid returned as reflux to the flow of top product (boil up rate/take-off rate). To choose an optimum reflux ratio for a design another concept should be considered which is the minimum reflux ratio and it is defined as the least value of reflux that can achieve the separation even with an infinite number of stages. The optimum reflux ratio mostly ranges between 1.2 – 1.6 times the min reflux ratio. The reflux ratio is highly dependent on the relative volatility which is given by

aij=Pi/Pj=Ki/Kj (2.1)

Where K value is the vapor liquid distribution ratio of light key and heavy key components.

The basis of any rigorous model is formed by solving a set of equations called the MESH-equation.

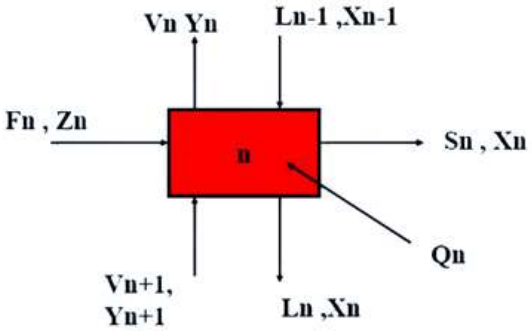


Figure (2.1) Stage system boundaries of mass and energy balance.

1- Mass balance equation

Vn+1Yn+1 + Ln-1Xn-1 +FnZn =VnYn +LnXn +snXn (2.2)

2- Energy Balance equation

Vn+1Hn+1 +Vn-1Hn-1 +Fhf +Qn =VnHn +Lnhn +Snhn (2.3)

3- Equilibrium Relation

Yi =KiXi (2.4)

4- Liquid and vapor composition summation equation

$$\sum_{i-n}^{n} X_i = \sum_{1\equiv n}^{n} Y_i = 1 \quad (2.5)$$

The process design starts with computing the hydrocarbon mass balance for inputs and outputs then determine the steam rates required in the stripping sections and the distribution of steam between overhead distillates vapors and liquids. A mass balance must also be applied on the product strippers should. Then the temperature enthalpy charts should be plotted according to the available data and the temperatures of cuts should be obtained considering the estimated strip out for each product. A schematic draw is then established for the sections of the tower and the number of trays in each section and the total number of trays in the tower.

For the product determination the desired product cuts are obtained after drawing the TBP curve. Not only the cuts temperatures also the volumetric yields. For the adjacent cuts the ASTM (5-95) GAP/OVERLAP should be calculated. The separation criteria is influenced by two terms .The first term is the degree of separation which is defined in terms of product purity or component recovery and is given by equation (2.6) , the higher the degree of separation the greater will be the recovery of the light component in the distillate and heavy component in the bottom. Secondly the degree of difficulty of separation which is defined as the relative difficulty encountered in separating the two compounds, regardless the purity requirements set by process specifications, inversely proportional to the relative volatility between the two components.

(2.6)

The degree of separation *= ASTM-5% of heavier distillate – 95% of lighter distillate.*

The degree of difficulty of separation*= Difference between ASTM 50% point of distillate fractions* (2.7)

As mentioned before, the separation of adjacent cuts is controlled by the ASTM(5-95). It is the difference between 5% ASTM point of Heavy cut and the 95 % point on the ASTM curve of a lighter cut of two adjacent products. If the obtained difference is positive then it's called gap if negative it's called overlap.

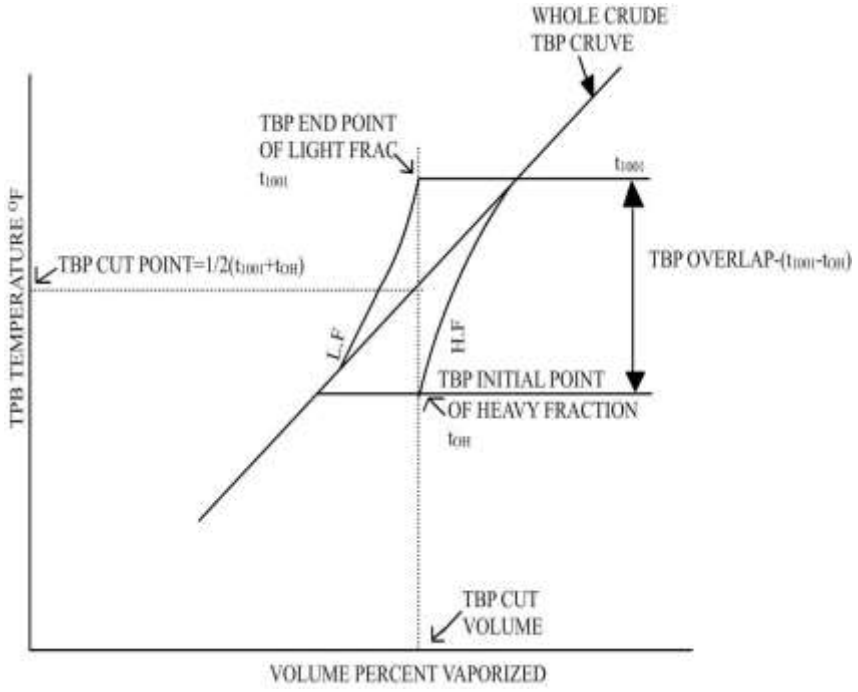


Figure (2.2) TBP Cut gap and overlap

Now all the input requirements we need for the design are:

1- Crude True boiling (major)
2- Density/API gravity (major)
3- Molecular weight (minor)
4- Viscosity (minor).

And for the column specifications requirements we need :

1- Column pressure.
2- Product specifications in terms of cut points .
3- Pump around duty
4- Column top tray temperature (optional).

Now to calculate the minimum number of stages required for the separation Fenske equation (2.8) can be used

$$N = \frac{\log\left[\left(\frac{X_d}{1-X_d}\right)\left(\frac{1-X_b}{X_b}\right)\right]}{\log \alpha_{avg}} \qquad (2.8)$$

Where,

N, is the theoretical number of stages (including the reboiler).

Xd, mole fraction of light component in the overhead distillate.

Xb, mole fraction of light component in bottoms.

aavg, the average relative volatility of the light component to the heavy component.

And to calculate the minimum reflux ratio Underwood equation can be used :

$$\sum \alpha_i X_i d / \alpha_i - \theta = Rm + 1 \qquad (2.9)$$

$$\sum \alpha_i X_i f / \alpha_i - \theta = 1 - q \qquad (2.10)$$

Where,

q , energy required to vaporize 1 mole of feed / molar latent heat of vaporization.

θ , Root of underwood equation.

And for feed tray location Kirkbirde equation is used

$$\log[Nr/Ns] = 0.206\log[(B/D)(Xf.LK/Xd.HK)2] \qquad (2.11)$$

Nr, Number of stages in enrichment section (partial condenser included).

Ns= Number of stages in stripping section

B= Molar flow bottom product.

D= Molar flow top product.

Xf.HK = concentration of the heavy key in the feed.

Xf.LK = concentration of the light key in the feed.

Xd.LK = concentration of the heavy key in the top product.

Xb.LK = concentration of the light key in the bottom product.

Avoiding the long inaccurate calculations, simulation method (shortcut method) can be applied using Aspen HYSYS simulation. All we need for a shortcut method are the specification of the crude oil (crude assay and/or bulk properties), the operating pressure (atmospheric or vacuum) and the attachments type.

3. HEN DESIGN

ASPEN HYSYS energy tool eases the task in a good way. It avoids the long manual calculations, the human errors and time consumption, besides giving an accurate set of results that helps us obtaining the optimum design. To start the simulation process Using ASPEN HYSYS energy tool some data are required so that the application can go on the simulation steps , such as the temperature of each hot and cold stream involved , the minimum temperature change required and the flow rates of the streams . Firstly we start with calculating the specific heat value of each stream involved in the heat exchanger network, and using Cp we can calculate the minimum and maximum enthalpy.

$$C_p = \frac{\triangle H}{\triangle T} \qquad (3.1)$$

$\triangle H$, enthalpy change .
$\triangle T$, temperature difference .

The data obtained from this step is used to plot the Temperature-Enthalpy curves, these curves are the guide for the rest of the design procedure. Once we have the T-H curves, ASPEN HYSYS energy analyzer can calculate the heat flow through the heat exchanger network. With known energy targets ASPEN HYSYS energy tool starts to calculate the area targets of thermal exchange and the utilities providing this area.

Cost index targets come at the end of the list of calculations to support the economical side of the study. But it requires the entry of the unit prices list. The last step of simulation is connecting the heat exchanger network grid diagram, which is a schematic diagram showing the connections of the hot and cold utility streams. After the data entry the following steps lead us to our grid diagram:

1- After entering the grid diagram environment, we choose the preset to be according to Temperature. Figure(3.1)
2- We should make sure that all streams are put in a descending order according to temperature.

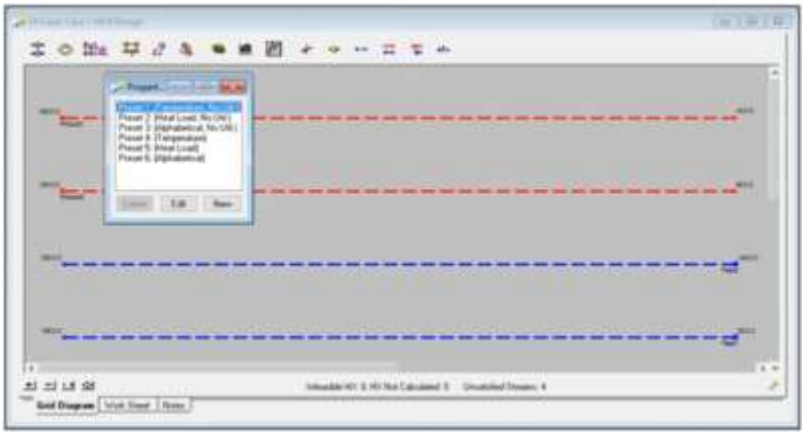


Figure (3.1) Illustrative screenshot from Aspen Energy Analyzer shows the establishment of HEN grid diagram steps.

3- Note that red color stands for hot streams and blue for cold streams.
4- Clicking on the icon (show/hide pinch line) shows the pinch line on the diagram, the above pinch and below pinch regions. Figure (3.2).

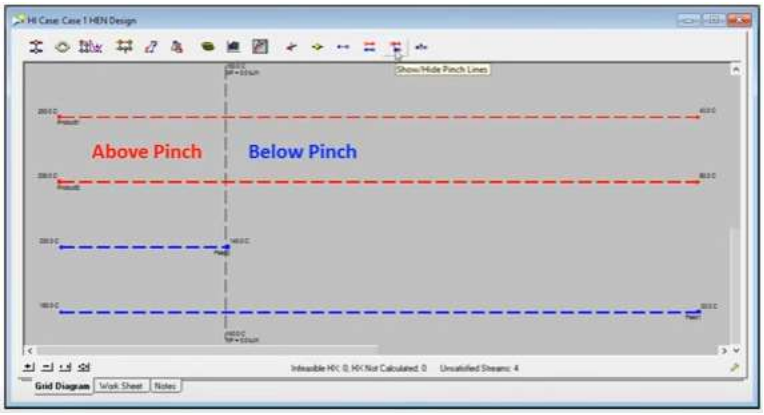
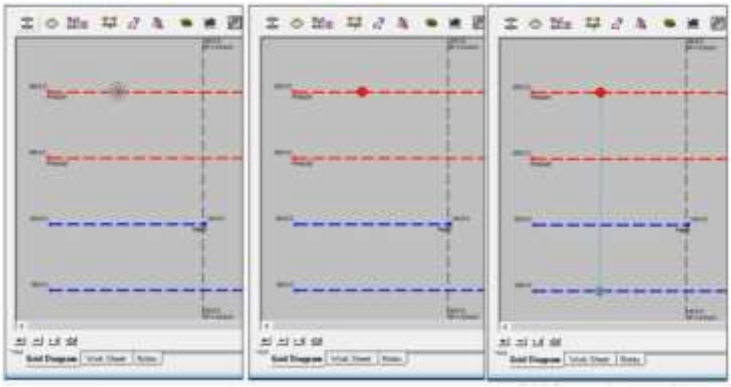


Figure (3.2) Illustrative screenshot from Aspen Energy Analyzer shows the pinch line.

5- The criteria for adding heat exchangers on the both sides of pinch line (above pinch and below pinch regions) is as follows :

- Above pinch: Ccold > Chot.
- Below pinch: Ccold < Chot.

6- Place the heat exchangers at the points of interest as shown in Figure (3.3).



Figure(3.3) placing a heat exchanger on the grid diagram from Aspen Energy Analyzer .

7- After placing the heat exchangers in the points of interest , the suitable temperatures should be entered based on the stream data

8- Aspen Energy Analyzer calculates the missing temperatures, the area of heat exchanger, the duty of heat exchanger and the calculated DTmin (hot end, cold end).

9- Note that the calculated DTmin may not be applicable, so that it might need to be changed manually (20 °C in the case study).

10- At the end the HEN grid diagram is achieved and the work sheet of the HEN shows all the data regarding the HEN design and criteria.

4. CASE STUDY

4.1 basic design

Designing a crude oil distillation tower depends mainly on the type of crude oil to be processed, degree of separation and the type and number of required products (cuts) .In our case study an amount of 150,000 bpd of a blend consisting of Arab light and Arab heavy crudes is to be processed, the crude oil assay data was provided by Cairo Oil Refining Co. For crude oil assay data (APPENDIX A). After entering the TBP assay data to ASPEN HYSYS oil manager and creating the hypothetical components. ASPEN HYSYS creates a new TBP curve for the

obtained blend depending on the crude data we entered. Figure (4.1) shows the TBP curve of the new blend plotted by Aspen HYSYS.

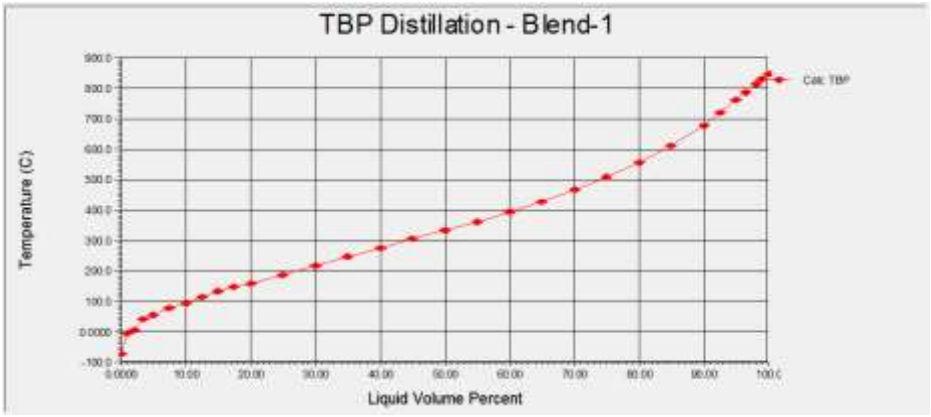


Figure (4.1) TBP curve of the new blend plotted by Aspen HYSYS.

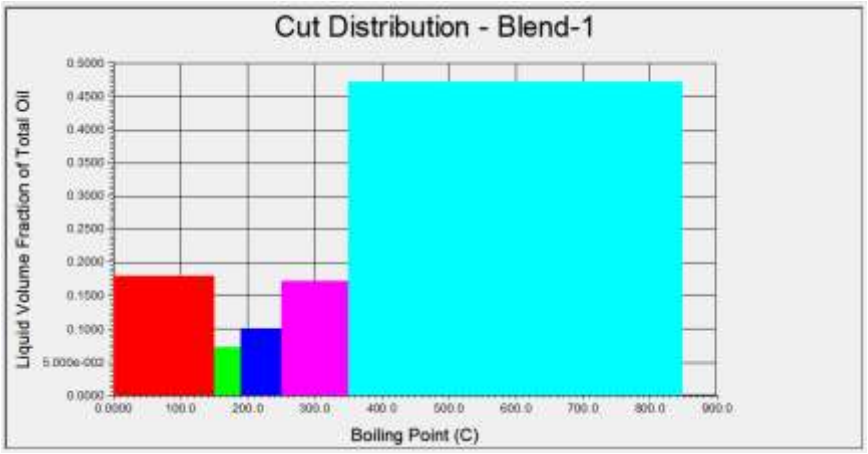


Figure (4.2) the cut distribution of the products plotted by Aspen HYSYS.

Five products are to be produced by an atmospheric distillation column. The five products are (off gases, heavy naphtha, kerosene, LGO and bottom residues) respectively. Figure (4.2) show the cut distribution of the products of the new blend plotted by Aspen HYSYS. The cut ends and the liquid-volume percents of cuts are shown in table (4.1).

Table (4.1) cut ends and the liquid-volume percents of cuts.

PRODUCT	CUT END °C	VOLUME %
Off Gases	150	21.23
Heavy Naphtha	190	4.83
Kerosene	250	6.79
LGO	350	18.51
Bottom Residue	1200	48.62

A basic design of an atmospheric distillation system is suggested for this mission. The main elements of a distillation system are as follows: Distillation column, to achieve the desired separation, Furnace to raise the temperature of the crude feed to the processing temperature (362°C in this case study) And heat Exchanger Network (HEN), to recover heat from product streams.

Assumptions of the number of trays are obtained from Fenske equation and it has been found that 27 theoretical trays are appropriate for our design. Three pump-arounds and three side strippers are to be connected to the column in order to remove the required amount of heat to maintain uniform temperature profile along the tower and to separate (Heavy Naphtha, Kerosene and LGO). The distillation column is operated by injecting high pressure steam and partial condenser, connected with three pump arounds and three side strippers as shown in Figure (4.3).

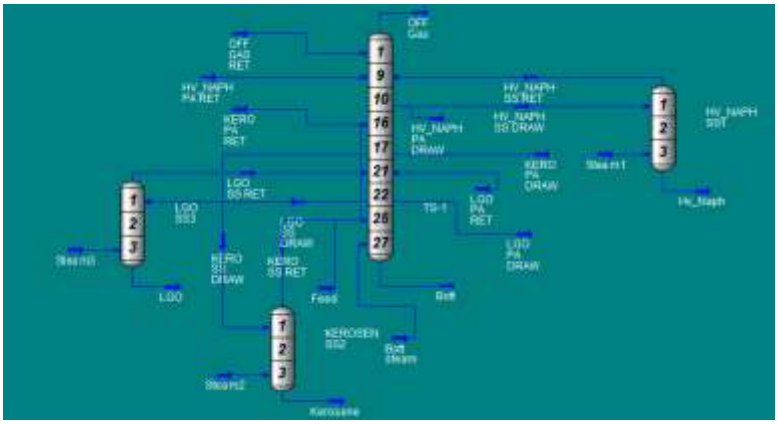


Figure (4.3) Distillation column, attachments, side draws design by Aspen HYSYS.

After achieving the basic design of the distillation system simulation, the hot and cold streams data are entered in Aspen Energy Analyzer to obtain the energy targets. In our basic design Aspen Energy Analyzer calculated the heating target to be (2.682e8 kj/hr = 74.5 MW)

Figure (4.4) show the overall design of the distillation system, including the furnace used to achieve the heating target and all the design elements and streams simulated by Aspen HYSYS.

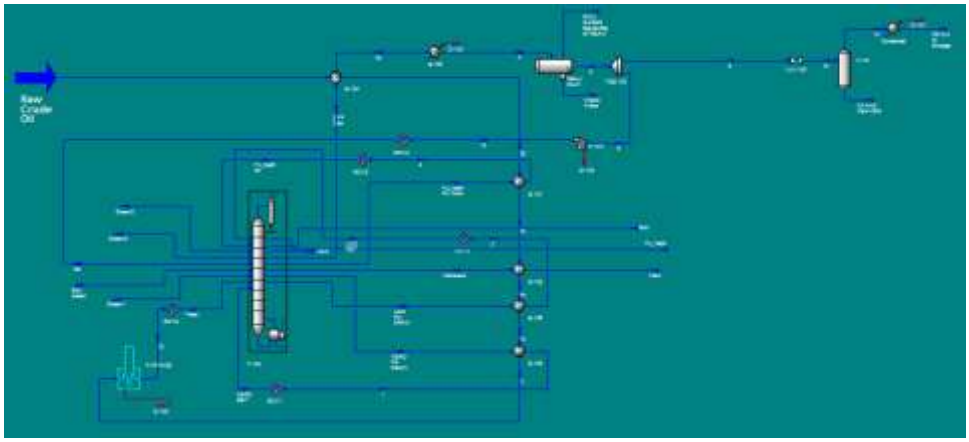


Figure (4.4) crude oil distillation system simulation by Aspen HYSYS.

4.2 optimization approach

Factorial design is the shortest accurate statistical way to relate the operating variables of the process in order to obtain the optimum targets (responses). In this case we are mainly concerned about energy saving, so that our targets will be the minimum heating enthalpies, which will be represented by the response of the factorial experiments.

A set of 20 runs of three-five factorial experiments has been applied relating the five independent variables (reflux ratio , Heavy Naphtha pump around duty , Kerosene pump around rate , LGO pump around duty and LGO pump around rate) to the heating and cooling targets

obtained by ASPEN HYSYS Energy analyzer (Qcooling , Qheating). The design matrix of the five independent variables and their 3 levels are shown in table (4.2)

Table (4.2) shows the design matrix of the factorial design

INDEPENDENT VARIABLE	CODE	RANGE / LEVEL		
		1	2	3
PA HN DUTY	A	80	100	120
PA K RATE	B	2800	3000	3250
PA LGO DUTY	C	35	45	55
PA LGO RATE	D	3100	3200	3300
R	E	0.6	0.7	0.8

Relating all these variables, gives us the value of each variable that gives the optimum response. The results of 5 runs out of 20 runs of experiments are presented in table (4.3). The table includes the values of the operating conditions of distillation column that we chose (coded from A to E) and the response heating and cooling enthalpies (Qh , Qc) .

Table (4.3) Factorial design trials results .

RUN	A: PA HN DUTY	B: PA K RATE	C: PA LGO DUTY	D: PA LGO RATE	E: REFLUX RATIO	QH X10^8	QCX10^8
1	120	3250	35	2500	0.8	2.557	1.685
2	80	3250	55	3100	0.6	2.420	1.260
3	80	3250	35	3100	0.6	2.454	1.365
4	100	3000	45	2800	0.7	2.698	1.456
5	80	2800	35	2500	0.8	2.498	1.344

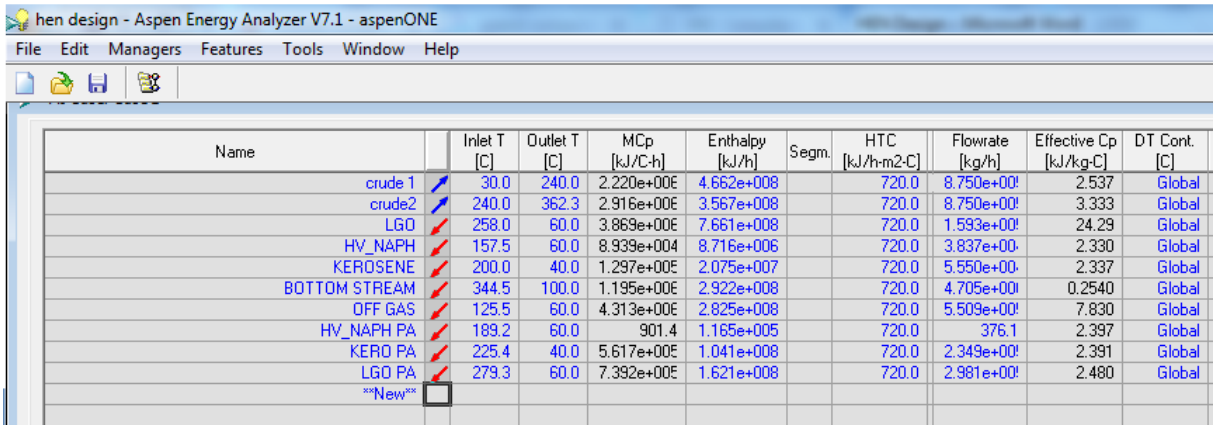
4.3 heat exchanger network

Using ASPEN HYSYS energy analyzer we can establish all the required calculations for the heat exchanger network design as well as plotting all the essential T-H curves and finally simulating the heat exchanger network. ASPEN HYSYS energy analyzer is a computer application that allows us to connect all the hot and cold streams involved in a heat exchanging process using a set of thermodynamical correlations and mathematical equations.

First step is providing the required data for ASPEN HYSYS energy tool to establish the **design and simulation. These data include the following:**

1- The flow rates of all hot and cold streams included in the heat exchanging operation (from Aspen HYSYS simulation).
2- The input temperature and the target temperature of each stream, the target temperature should be within the constrains and limitations according to the physical and thermal properties of each stream.
3- The heating and cooling mediums (cooling water, fired heat).
4- The value of HTC will be assumed (720 kj/h.m2.°C) by ASPEN HYSYS energy analyzer as long as we don't have HTC process stream values.

5-



Name	Inlet T [C]	Outlet T [C]	MCp [kJ/C-h]	Enthalpy [kJ/h]	Segm.	HTC [kJ/h-m2-C]	Flowrate [kg/h]	Effective Cp [kJ/kg-C]	DT Cont. [C]
crude 1	30.0	240.0	2.220e+00E	4.662e+008		720.0	8.750e+00!	2.537	Global
crude2	240.0	362.3	2.916e+00E	3.567e+008		720.0	8.750e+00!	3.333	Global
LGO	258.0	60.0	3.869e+00E	7.661e+008		720.0	1.593e+00!	24.29	Global
HV_NAPH	157.5	60.0	8.939e+004	8.716e+006		720.0	3.837e+00·	2.330	Global
KEROSENE	200.0	40.0	1.297e+00E	2.075e+007		720.0	5.550e+00·	2.337	Global
BOTTOM STREAM	344.5	100.0	1.195e+00E	2.922e+008		720.0	4.705e+00!	0.2540	Global
OFF GAS	125.5	60.0	4.313e+00E	2.825e+008		720.0	5.509e+00!	7.830	Global
HV_NAPH PA	189.2	60.0	901.4	1.165e+005		720.0	376.1	2.397	Global
KERO PA	225.4	40.0	5.617e+00E	1.041e+008		720.0	2.349e+00!	2.391	Global
LGO PA	279.3	60.0	7.392e+00E	1.621e+008		720.0	2.981e+00!	2.480	Global
New									

Figure (4.5) ASPEN HYSYS energy analyzer data input window with all the required data entered.

ASPEN HYSYS energy tool starts the calculations by calculating the Cp value, which is given by equation (4.1)

$$C_p = \frac{\triangle H}{\triangle T} \qquad (4.1)$$

For example, for the crude oil

Cp = (4.662e8/8.750e5)/(240-30) = 2.537 kj/kg.°C

As we calculated Cp of the raw crude oil manually, ASPEN HYSYS energy analyzer goes on calculating Cp for the rest of streams. Figure (4.5) show the results of Cp calculated by ASPEN HYSYS energy analyzer as well as some other calculated properties.

Note that the minimum temperature difference DTmin must be specified and given to ASPEN HYSYS energy analyzer, as the calculated DTmin may not be applicable in the real application. The minimum temperature approach for this design is chosen to be DTmin=20 °C.

As we previously discussed the role of the T-H curves in the design procedure, now we will show the results and the plots of the composite curves and grand composite curves of our design. The following figure (4.6) shows the composite curve of the optimum design approach plotted by ASPEN HYSYS energy analyzer.

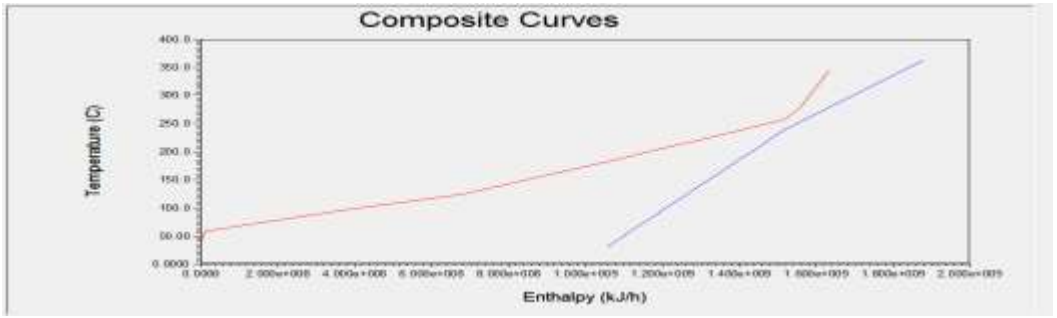


Figure (4.6) the composite curve of HEN design by ASPEN HYSYS energy analyzer

The figure shows the hot composite curve in red as well as the cold composite curve in blue on the same plot so we can point the pinch point where the heat flow is zero. The composite

curves are moved horizontally such that the minimum approach temperature on the plot equals the minimum approach temperature we specified (DTmin=20°C).

The following figure (4.7) shows the grand composite curve created by ASPEN HYSYS energy analyzer regarding our optimum design.

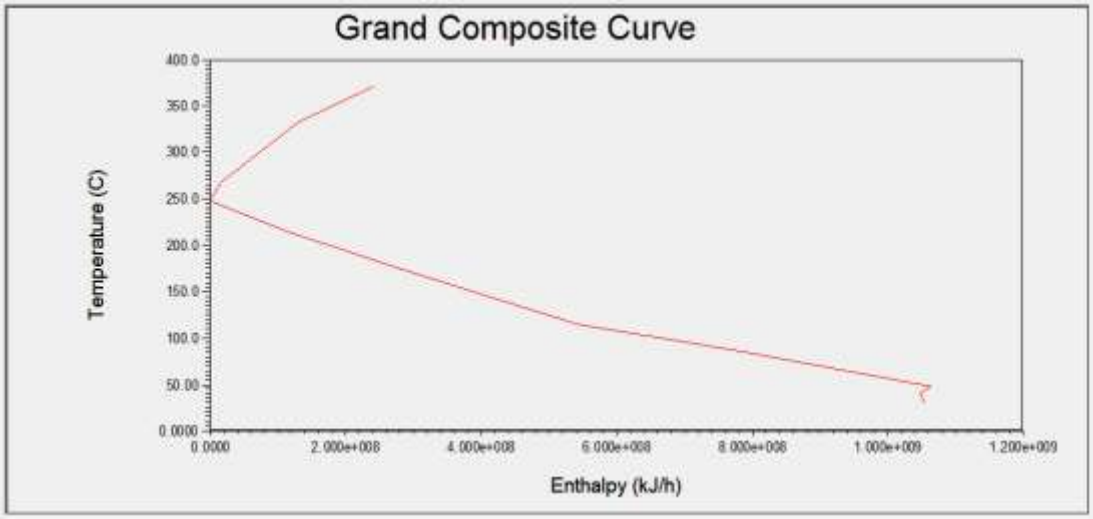


Figure (4.7) The grand composite curve of the design by ASPEN HYSYS energy analyzer.

This curve shows the shifted temperatures of streams versus the cascaded heat flow between each temperature interval. We can clearly see the heat available in various temperature interval and the net heat flow. Note that the heat flow at the pinch temperature equals zero.

4.4 ENVIRONMENTAL SIDE OF THE CASE STUDY

4.4.1 water conservation

Our work's vision regarding water is saving water by reducing the amount of cooling water entering the system as it presents the majority of water used in the whole refining process. This concept can be achieved by reducing the amount of heat that the products contain. Our optimum design of crude oil distillation system reduces the temperatures of the output streams by recovering heat energy for crude oil preheating operation. This can reduce the amount of cooling water used remarkably.

Petroleum products must be stored in a safe temperature or a temperature that is safely below the flashpoint of the product.

Table (4.4) The output and storing temperatures of petroleum fractions

PRODUCT STREAM	TOUT °C	TSTORE °C
Off Gases	125.5	25
Heavy Naphtha	157.5	30
Kerosene	200	30
LGO	258	30
Bottom Stream	344.5	100

By this optimum integrated design as we recover the largest possible amount of heat, the products temperature is lowered remarkably. For example Kerosene product stream in the basic

design exits the tower at a temperature of 140 °C by passing the kerosene through the heat recovery system of the optimized approach its temperature is lowered by 100 °C degrees down to 40 °C. Further cooling achieved by cooling water down to 30 °C as a safe storing temperature below the flash point .This procedure is done for all products as well.

4.4.2 carbon dioxide emissions reduction

The recovered heat lowers the furnace duty and the amount of burned fuel in turn. Also, optimizing the operating conditions lowers the heating target and the boilers duty which in turn reduces the amount of unburned fuel.

Too many elements influence the amount of Co2 emissions including the amount of heat supplied by fuel, the molar mass ratio of Co2, the net heating value of fuel and the carbon mass percent in fuel. That's why we chose natural gas for this case study, it provides a relatively high net heating value in a considerable price. Also the equipments efficiency plays a role in the amount of Co2 emissions. The amount of Co2 emissions is given by equation (6.3):

$$Co_{2,emiss} = (Co_{2,furnace} + Co_{2,boiler}) \quad (4.2)$$

Where ,

$Co_{2,emiss}$, is the total Co_2 emission (kg/hr).
$Co_{2,furnace}$, is the Co_2 emission from furnace (kg/hr).
$Co_{2,boiler}$, is the Co_2 emission from boilers (kg/hr).

As we previously mentioned, the emissions from each equipment depend on the amount of heat supplied by the fuel and the fuel's emission factor. Equation (4.3) gives the emissions of Co2 from furnace while equation (4.4) gives the emissions of Co2 from boiler:

$$Co_{2,furnace} = Q_{ff} . E \quad (4.3)$$

$$Co_{2,boiler} = Q_{fb} . E \quad (4.4)$$

where ,

Q_{ff}, is the heat duty supplied by the fuel in furnace (kj/h).
Q_{fb}, is the heat duty supplied by the fuel in boiler (kj/h).
E , is the fuel's emission factor (kg_{Co2}/kj).

In this case study we will deal with the terms (Qff , Qfb) as the amount of heat duty that the unburned fuel can supply on furnace and boiler respectively so we can know the reduction in Co2 emissions. The following equations (4.5), (4.6) represent the two terms. Note that the two terms are influenced by the efficiency of the equipment (furnace or boiler).

$$Q_{ff} = (Q_O - Q_{HEN}) / \eta_F \quad (4.5)$$

$$Q_{fb} = (Q_B - Q_O) / \eta_B \quad (4.6)$$

Where,

η_F, is the efficiency of the furnace.
η_B, is the efficiency of the boiler.

Equation (4.7) gives the fuel's emission factor (E):

$$E = \frac{C}{NHV} \times \frac{C\%}{100} \quad (4.7)$$

Where ,

NHV , is the net heating value of the fuel.
C% , is the carbon mass percent in the fuel.
C = 3.67.

To fulfill equations (4.5),(4.6) the value of the equipment efficiency is rather assumed based on the field data or calculated by the correlations (4.8) , (4.9).

For the furnace :

$$\eta_F = \frac{(T_{FTF} - T_{STACK})}{(T_{FTF} - T_O)} \quad (4.8)$$

Where ,

T_{FTF} , is the theoritical flame temperature of the furnace.
T_{STACK} , is the stack temperature.
T_O , is the ambient temperature.

For the boiler :

$$\eta_B = \frac{\lambda_B}{(h_B - 419)} \times \frac{(T_{FTB} - T_{STACK})}{(T_{FTB} - T_O)} \quad (4.9)$$

Where,

λ_B , is the latent heat of vaporization.
h_B , is the Enthalpy.
T_{FTB} , is the theoritical flame temeperature of the boiler.

5. RESULTS AND DISCUSSION

5.1 Optimization approach results

The most appropriate operating conditions are those which result in the minimum heating target, from table (4.3) and the rest of results of the 20 runs of factorial experiments, we can obtain the most appropriate optimum operating conditions and their response. Table (5.1) shows these conditions.

Table (5.1) The optimum operating conditions and optimum response.

CODE	VARIABLE	VALUE
A	PH HV NAPH DUTY	80 °C
B	PA K RATE	3250 lbmol/hr
C	PA LGO DUTY	55 °C
D	PA LGO RATE	3100 lbmol/hr
E	Reflux Ratio	0.6
Optimum Response		
Qh		2.420e8 kj/hr
Qc		1.260e8 kj/hr

From our basic design, and the calculations done by Aspen Energy Analyzer, we previously found that the heating target required for the basic design was (2.682e8 kj/hr). Now after preceding the factorial design we apply the new values obtained to Aspen HYSYS simulation. Figure (5.1) and (5.2) shows the new data simulated on Aspen HYSYS simulation.

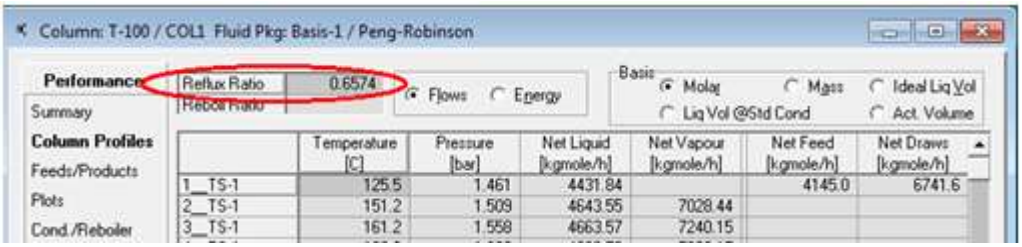
Name	HV_NAPH PA [	HV_NAPH PA [	DUTY	LGO PA DRAW	LGO PA RET	DUTY
Vapour	0.0000	0.0000	≈ 80 C	0.0000	0.0000	≈ 55 C
Temperature [C]	189.2	108.6		279.3	227.2	
Pressure [bar]	1.896	1.896		2.475	2.475	
Molar Flow [kgmole/h]	2.704	2.704		1394	1394	
Mass Flow [kg/h]	376.1	376.1		2.981e+005	2.981e+005	
Std Ideal Liq Vol Flow [m3/h]	0.4779	0.4778		350.6	350.5	
Molar Enthalpy [kcal/kgmole]	-6.017e+004	-6.691e+004		-7.965e+004	-8.719e+004	
Molar Entropy [kJ/kgmole-C]	252.9	186.2		537.1	477.3	
Heat Flow [kcal/h]	-1.627e+005	-1.809e+005		-1.110e+008	-1.216e+008	

Name	KERO SS DRA'	LGO SS DRAW	HV_NAPH PA [	KERO PA DRA'		
Vapour	0.0000	0.0000	0.0000	0.0000		
Temperature [C]	225.4	279.3	189.2	225.4		
Pressure [bar]	2.234	2.475	1.896	2.234		
Molar Flow [kgmole/h]	440.8	938.1	2.704	1459		
Mass Flow [kg/h]	7.097e+004	2.006e+005	376.1	2.349e+005		
Std Ideal Liq Vol Flow [m3/h]	87.64	235.9	0.4779	290.1	1459 kgmole/h	
Molar Enthalpy [kcal/kgmole]	-6.588e+004	-7.965e+004	-6.017e+004	-6.588e+004	3217 lbmole/hr	
Molar Entropy [kJ/kgmole-C]	334.8	537.1	252.9	334.8	Calculated by: COL1	
Heat Flow [kcal/h]	-2.904e+007	-7.471e+007	-1.627e+005	-9.613e+007		

Name	LGO PA DRAW
Vapour	0.0000
Temperature [C]	279.3
Pressure [bar]	2.475
Molar Flow [kgmole/h]	1394
Mass Flow [kg/h]	2.981e+005
Std Ideal Liq Vol Flow [m3/h]	350.6
Molar Enthalpy [kcal/kgmole]	-7.965e+004
Molar Entropy [kJ/kgmole-C]	537.1
Heat Flow [kcal/h]	-1.110e+008

1394 kgmole/h
3074 lbmole/hr
Calculated by: COL1

Figure(5.1) The optimized values of operating conditions simulated by Aspen HYSYS .



Figure(5.2) Shows the optimized value of reflux ratio simulated in Aspen HYSYS .

Comparing the heating target of the basic design to that of the optimized design we find:

Qs-o%= (2.682e8 – 2.420e8) / 2.682e8

= 9.7%

Where,

Qs-o, is the energy saved by the optimization.

This percent saving of 9.7% shows an achievement gained by the optimization of operating conditions of the column, the saving of energy achieved in this part of design has huge environmental and economical effects.

Calculating the probability range (p-value) allows us to form a full understanding about how significant is each variable. All these statistical values are computed by the DOE v10.1.6 software to avoid long mathematical calculations

.

Table (5.2) Significance and statistical probability

MODEL PARAMETER	P-VALUE	SIGNIFICANCE
E	0.00012	Highly Significant
D	0.00210	Significant
C	0.00282	Significant
B	0.00351	Significant
ED	0.02250	Possibly Significant
A	0.08934	Not significant

Note that the P-value for A model is 0.08934 which make this parameter totally not significant, this physically means that the effect of heavy naphtha pump around duty on the process is not significant on its own. Referring to the ASPEN HYSYS simulation, we note that the flow rate of the heavy naphtha pump around stream is 376.1 kg/hr which is relatively low enough to cause this insignificance. On the other hand we note that the P-value of the E model is highly significant 0.00012 which shows the great effect of the reflux ratio on the design.

5.2 Heat recovery by HEN results

Now that we have all the data required as well as the calculated values, ASPEN HYSYS energy analyzer creates the main features of the design. Figure (5.3) shows the features of the new optimum design approach of the heat exchanger network based on pinch analysis.

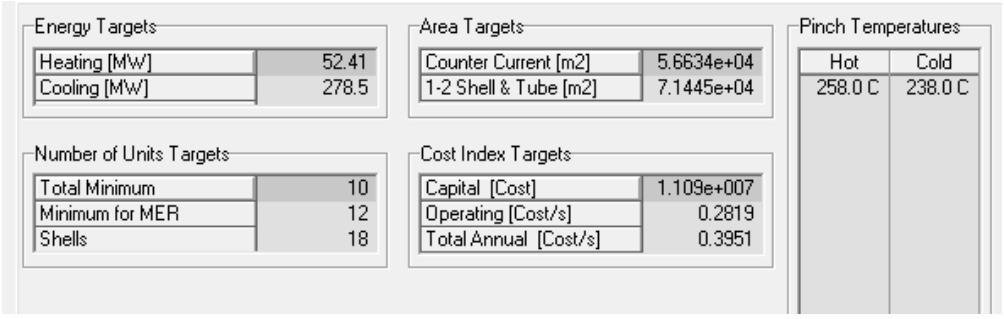


Figure (5.3) The main features of the design of the HEN by ASPEN HYSYS energy analyzer.

Figure (5.3) shows the hot and cold pinch temperatures 258°C, 238°C respectively. Note the 20°C difference between the hot and cold pinch temperatures that we specified. It also shows that we need 5.6634e4 m2 of area for a counter current heat exchanger network and 7.1445e4 m2 of area for the 1-2 shell and tube heat exchanger network to achieve the heating target of 52.41 MW(1.88676e8 kj/h) and cooling target of 278.5 MW (1.0026e9 kj/h).

For the number of units targets, ASPEN HYSYS energy analyzer calculated the total minimum units which is given by :

TMU = Ns + Nd – 1 (5.1)

Where,

TMU, is the total minimum units.

Ns, is the number of streams.

Nd, the total number of distinct hot and cold utility sources.

TMU = 9 + 2 – 1 = 10

Finally for the minimum energy recovery units target MER, ASPEN HYSYS energy analyzer calculated it to be 12.

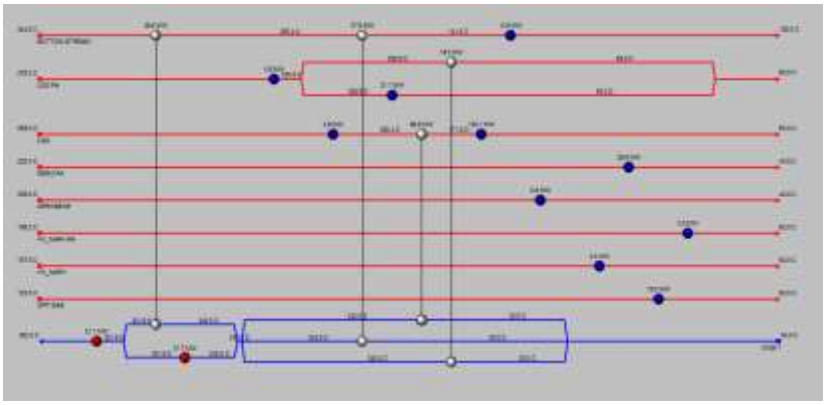


Figure (5.4) Heat exchanger network grid diagram by ASPEN HYSYS energy analyzer.

Now the final step of the Heat exchanger network design is the heat exchanger network grid diagram. It shows the matching of all hot and cold streams that finally lead to the integrated heat exchange that takes the advantage of every possible recoverable heat. Figure (5.4) shows the grid diagram of the heat exchanger network we are working on.

After this optimum heat exchanger network design is done now it`s time to calculate how much energy saved by installing this heat exchanger network to the distillation system. The net energy saved by this heat exchanger network is calculated as follows

(2.420e8 - 1.88676e8) / (2.420e8) = 20.6%

The saving of 20.6% of energy required for heating the crude oil feed stream is very considerable and no doubt the effect of this saving on the economical and environmental sides will be very remarkable as we will see later on this chapter.

5.3 ENVIRONMENTAL RESULTS

5.3.1 Cooling water

Cooling water consumption is lowered as a result of lowering the amount of heat that the products contain before storing. The amount of heat saved by the optimum integrated design was supposed to be a cooling load that requires more cooling water. By saving this amount of heat we save an equivalent amount of required cooling water. This amount of water is given by

Qs = Mwater Cpwater (Tow – Tiw) (5.2)

Where,

Qs = Qb – Qo (5.3)

Qs, is the amount of heat recovered by the optimum integrated design**.**

Qb, is the heating target of the basic design.

Qo, is the heating target of the optimum integrated design.

Mwater, is the amount of cooling water saved.

Tiw, Tow, Are the input temperature and output temperature of cooling water.

From Aspen HYSYS energy analyzer, we can obtain all the data required to calculate the amount of saved water Mwater.

ame		Inlet T [C]	Outlet T [C]	Effective Cp [kJ/kg-C]
Cooling Water		20.0	40.0	4.183
Fired Heat (1000)		1000.0	400.0	1.000
MP Steam Generation		174.0	175.0	1981
HP Steam Generation		249.0	250.0	1703
LP Steam Generation		124.0	125.0	2196
<empty>				

Figure(5.5) Tow, Tiw and Cpwater , by Aspen HYSYS energy analyzer .

Now all the required data are collected then we can start the calculations
Qs = 2.682e8 - 1.88676e8 = 7.953e7 kj/hr

7.953e7 = Mwater (4.183)(40-20)

Mwater = 9.506e5 kg/hr

From the previous equation we got the amount of cooling water saved by our approach and it's a considerable amount that has its reflections on the environmental and economical sides. Cooling water in a country like Egypt doesn't cost too much, as the cost of it varies between 0.6 - 1 US$ per 1000 gallon, so that the economical effect may not be remarkable enough. But on the other hand, the environmental effect is highly considerable, as a huge amount of water is saved even if it doesn't cost that much.

5.3.2 Carbon dioxide emissions reduction

Using the equations and correlations reviewed in section (**4.4.2**), we can calculate the Co2 emission reduction. Equation (5.4) gives the total amount of carbon emissions from the unburned fuel.

$$Co_{2,emiss} = (Co_{2,furnace} + Co_{2,boiler}) \quad (5.4)$$

So the calculations will be divided into two sections reduction in furnace and reduction in boilers. All the data are collected from Aspen HYSYS Energy Analyzer, Aspen HYSYS simulation, fuel specifications and the field observations. Table (5.3) shows a set of data collected.

. Table (5.3) a set of data used in Co2 calculations.

TERM	VALUE
TFTF	1960 c
Tstack	297.88 c
To	30 c
C%	75.3%
NHV	46117 kj/kg
Constant C	3.67
Estimated ηF	0.806

5.3.2.1 Co2 emissions reduction from furnace unburned fuel

We will start the calculations by calculating the emission factor of the natural.

$$E = \frac{C}{NHV} X \frac{C\%}{100} \quad (5.5)$$

E = (3.67/46117) (75.3/100)= **5.99e-5 kg/kj**

Now the emissions of Co2 from furnace is given by equation (5.6) :

$$Co_{2,furnace} = Q_{ff} . E \quad (5.6)$$

To calculate the value of the heat that the unburned fuel can supply if it burned in the furnace we need first to calculate the furnace efficiency.

$$\eta_F = \frac{(T_{FTF} - T_{STACK})}{(T_{FTF} - T_O)} \quad (5.7)$$

From table(5.3) we get

. ηF = (1960 – 297.88)/(1960 – 30) = 0.86

Now we can calculate the value of the heat that the unburned fuel can supply if it burned in the furnace which is given by:

$$Q_{ff} = (Q_O - Q_{HEN}) / \eta_F \quad (5.8)$$

= (2.420e8 - 1.88676e8) / 0.86 = 6.200e7 kj/hr

Now we can fulfill equation (5.6)

$Co_{2,furnace} = Q_{ff} . E$ = 6.200e7 * **5.99e-5** = 3713.8 kg/hr

5.3.2.2 Co2 emissions reduction from boiler unburned fuel

Factorial design reduced the duty of the boilers what in turn reduced the amount of unburned fuel, the value of the heat that the unburned fuel can supply if it is fired in the boiler is given by equation (5.9) knowing that the value of boilers efficiency is assumed to be (0.806) based on the field observations:

$$Q_{fb} = (Q_B - Q_O) / \eta_B \quad (5.9)$$

= (2.682e8 - 2.420e8) / (0.806) = 3.251e7 kj/hr

From the previously calculated value of E and equation(5.10) ,we get

$$Co_{2,boiler} = Q_{fb} . E \quad (5.10)$$

= 3.251e7 * 5.99e-5 = 1948 kg/hr

The sum of the emissions reduced in furnace and boilers gives the total quantity of reduction

$$Co_{2,emiss} = (Co_{2,furnace} + Co_{2,boiler}) \quad (5.4)$$

= (3713.8 – 1948) = 5661 kg/hr

The total amount of carbon dioxide emitted from the basic design is given by:

$$Co_{2,emiss,b} = Q_{F,Basic} * E \quad (5.11)$$

= (3.118e8)(5.99e-5) = 18681 kg/hr

The percent of reduction:

(5661)/(18681) % = 30.3%

5.4 Cost calculations

Shell cost is given by the correlation (5.12)

$$\text{Section cost} = 14970 \left(D_{sec}^{1.066} \right) \left(H_{sec}^{0.802} \right) \quad (5.12)$$

Where ,

D_{sec} , is the section's diameter.

H_{sec} , is the section's hight.

While trays cost is given by the correlation (5.13**)**

$$\text{Section cost} = 366\ (D_{sec}^{1.55})\ (H_{sec}^{1.4}) \qquad (5.13)$$

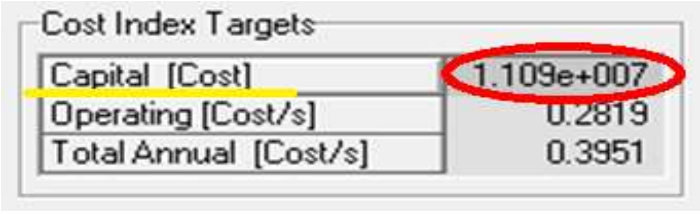
These correlations are also valid for cost calculations of side strippers the following table (5.4) summarize the cost calculation results of the tower

Table (5.4) cost calculation results of the tower

COLUMN COST	
Shell cost	1266990 US$
Trays cost	363060 US$
HVNAPH S-stripper	58142 US$
KEROSENE S-stripper	70855 US$
LGO S-stripper	61965 US$
Total cost of column = 1821012 US$	

The capital cost of the heat exchanger network was calculated by Aspen Energy Analyzer to be 11,090,000 US $.

Table (5.5) The capital cost of the HEN by Aspen Energy Analyzer.



The following correlation (5.14) gives the cost of furnace:

$$\text{Cost Of Furnace} = a + b\ (Q_{FURNACE})^{C} \qquad (5.14)$$

Where,

$Q_{FURNACE}$, is the furnace duty.

a, b, c , Empirical modifing factors.

Table (5.6) Empirical modifying factors of Equ (5.14).

EMPIRICAL FACTOR	VALUE
A	1*10^5
B	1000
C	0.8

Cost of furnace = (1*10^5) + (1000) (1.88676e8)^0.8 = 4,175,094,524 US$

The total capital cost of the distillation system now can be calculated as follows:

Total capital cost = column cost + HEN cost + furnace cost (5.15)

= 1821012 + 11,090,000 +4,175,094,524= 4,188,005,536 US$

Also from table (5.5) Aspen Energy analyzer calculated the operating cost to be 0.2819 US $/s (8,889,998 US $/yr).

5.4.1 Cost reduction

Cost reduction of fuel = Qs x unit price of natural gas .

= 7.537e7 Btu/hr x 3.308 US$/million BTU

= 250 US$/ hr = 6000 US$/day = 2.19e6 US$/yr

Cost reduction of cooling water = water saved (kg/hr) x minimum price of the unit
= 9.506e5 kg/hr X 0.15US$/1000 kg
= 143 US$/hr = 3412 US$/day = 1.252e6 US$/yr

Carbon Tax Saving = Co2 reduction x Tax
= 49275 ton/yr * 5 US$/ton = 246375 US$/yr

CONCLUSION

This work presented a new model of atmospheric distillation column. In the new simplified model, given conventional refining product specifications, the key components and associated recoveries are identified systematically without carrying out rigorous simulation. The simulation is then made by Aspen HYSYS software. Compared with the existing approach developed by Gadallah et al. (2002a), the new method is simple and straightforward; no complex steps are carried out and no extra user judgments are required. These features of the new approach allow simplified models to be applied to the design and analysis of crude oil distillation columns.

Accurate values of the chosen operating conditions are obtained by optimization procedure based on factorial design which is one of the most appropriate statistical methods by which we can obtain optimum values of the operating conditions of the crude oil distillation column to give the optimum response. This was clearly proved in this work as it was able to achieve a reduction of 9.7% of the heating target.

A heat exchanger network design is also achieved using Aspen Energy Analyzer software. A considerable amount of heat is recovered in order to benefit the heat integration of the system. The HEN was able decrease the heating target by 20.6% which is a very remarkable ratio.

Environment and economics are not contrary , it's all about finding an appropriate way to benefit both sides .That's what we obtained in this wok , an amount of 8.327e6 t/yr of cooling water is saved and 49275 t/yr of Co2 emissions are reduced , at the same time an amount of 2.19e6 US$/yr is saved due to the unburned fuel, and 1.252e6 US$/yr is saved due to the saved cooling water , and another amount of 246375 US$/yr is saved due to carbon tax reduction.

REFERENCE

- Gadalla, M. A., Retrofit of heat-integrated crude oil distillation systems, PhD Thesis,
- UMIST, Manchester , UK (2002)
- Gadalla, M., M. Jobson, and R. Smith, Shortcut models for retrofit design of
- distillation columns, *Trans IChemE,* September 81(A), 971-986 (2003a)
- Gadalla, M., M. Jobson, and R. Smith, Optimization of existing heat-integrated
- refinery distillation systems, *Trans IChemE,* January, 81(A), 147-152 (2003b)
- Gadalla, M., M. Jobson, and R. Smith, Increase capacity and decrease energy for
- existing refinery distillation columns, *Chemical Engineering Progress,* April, 44-50
- (2003c)
- Gadalla, M., M. Jobson, R. Smith, and P. Boucot, Design of refinery distillation
- processes using key component recoveries rather than conventional specification

- methods, submitted once to *Chem. Eng. Res. Des.* in November (2002a), under
- revision
- Nevine Kamal , The Feasibility of Imposing a Carbon Tax in Egypt ,Institute Of National Planning, in July (2015)
- http://kkft.bme.hu/htms/kornykozp/Cost.of.Equipments.pdf
- James L. Bergmann, Combustion Analysis , (2005)
- Al-Muslim, H. and I. Di ncer, Thermodynamic analysis of crude oil distillation systems, *Int. J. Ener gy Res.*, 29, 637-655 (2005)
- http://ocw.unican.es/ensenanzas-tecnicas/procesos-quimicos-de-fabricacion/materiales/Subject%207.%20Equipment%20Sizing%20and%20Costing%20OCW.pdf
- Ahmad, S., B. Linnhoff and R. Smith, Cost optimum heat exchanger networks. 2: Targets and design for detailed capital cost models, *Comp. Chem. Eng.*, 14 (7), 751-
- 767 (1990)
- Arora, J. S., *Introduction to Optimum Design*, Academic Press, 2nd edition (2004)
- Asante, N. D. K. and X. X. Zhu, An automated approach for heat exchanger network
- retrofit featuring minimal topology modifications, *Comp. Chem. Eng.*, 20, S7-S12
- (1996)
- http://www.met.kth.se/asialink/Curriculum/Royal%20Institute%20of%20Tech-KTH/lecture4.pdf

Table (A.1) The true boiling data of Arab light and Arab heavy crudes.

Arab Light TBP Distillation Curve		Arab Light Density Distribution		Arab Heavy TBP Distillation Curve		Arab Heavy Density Distribution	
Distilled %	Temp. (C)	Distilled %	S. G.	Distilled %	Temp. (C)	Distilled %	S. G.
3.79	40	3.79	637.08	4.97	50.00	4.97	656.81
4.51	50	4.51	650.23	6.32	60.00	6.32	672.65
5.14	60	5.14	659.99	7.83	70.00	7.83	687.73
7.06	70	7.06	682.86	8.06	80.00	8.06	689.85
7.97	80	7.97	691.59	9.45	90.00	9.45	701.81
8.78	90	8.78	698.65	11.00	100.00	11.00	713.76
10.89	100	10.89	714.54	11.81	110.00	11.81	719.54
11.82	110	11.82	720.70	13.21	120.00	13.21	728.95
12.79	120	12.79	726.73	14.14	130.00	14.14	734.80
15.33	130	15.33	740.95	15.76	140.00	15.76	744.47
17.11	140	17.11	749.81	17.38	150.00	17.38	753.56
18.88	150	18.88	758.03	18.98	160.00	18.98	761.99
21.10	160	21.10	767.56	20.55	170.00	20.55	769.83
23.11	170	23.11	775.61	22.08	180.00	22.08	777.18
25.13	180	25.13	783.29	23.59	190.00	23.59	784.11
26.99	190	26.99	790.02	25.08	200.00	25.08	790.72
28.86	200	28.86	796.50	26.57	210.00	26.57	797.09
30.54	210	30.54	802.15	28.05	220.00	28.05	803.29
32.41	220	32.41	808.22	29.55	230.00	29.55	809.38
34.26	230	34.26	814.09	31.08	240.00	31.08	815.40
36.12	240	36.12	819.81	32.62	250.00	32.62	821.37
37.97	250	37.97	825.38	34.19	260.00	34.19	827.29
39.81	260	39.81	830.83	35.77	270.00	35.77	833.14
41.64	270	41.64	836.18	37.37	280.00	37.37	838.93
43.47	280	43.47	841.43	38.97	290.00	38.97	844.66
45.37	290	45.37	846.80	40.57	300.00	40.57	850.31
47.18	300	47.18	851.90	42.18	310.00	42.18	855.89
48.99	310	48.99	856.94	43.78	320.00	43.78	861.40
50.78	320	50.78	861.93	45.38	330.00	45.38	866.82
52.57	330	52.57	866.87	46.97	340.00	46.97	872.16
54.35	340	54.35	871.79	48.54	350.00	48.54	877.42
56.11	350	56.11	876.68	50.09	360.00	50.09	882.57
57.90	360	57.90	881.63	51.61	370.00	51.61	887.62
59.61	370	59.61	886.42	53.10	380.00	53.10	892.54
61.28	380	61.28	891.11	54.56	390.00	54.56	897.34
62.90	390	62.90	895.71	55.99	400.00	55.99	902.05
64.48	400	64.48	900.22	57.39	410.00	57.39	906.65
66.01	410	66.01	904.67	58.76	420.00	58.76	911.17
67.50	420	67.50	909.03	60.10	430.00	60.10	915.60
68.94	430	68.94	913.33	61.41	440.00	61.41	919.96
69.96	440	69.96	916.41	62.70	450.00	62.70	924.25
71.32	450	71.32	920.59	63.96	460.00	63.96	928.48
72.65	460	72.65	924.74	66.42	480.00	66.42	936.79
75.23	480	75.23	933.02	68.79	500.00	68.79	944.92
77.68	500	77.68	941.28	71.07	520.00	71.07	952.93
80.02	520	80.02	949.56	73.27	540.00	73.27	960.80
82.24	540	82.24	957.91	75.36	560.00	75.36	968.48
84.19	560	84.19	965.69	77.37	580.00	77.37	976.04
85.88	580	85.88	972.89	79.28	600.00	79.28	983.49
87.45	600	87.45	980.03	83.67	650.00	83.67	1001.72
90.90	650	90.90	997.78	87.53	700.00	87.53	1019.47
93.72	700	93.72	1015.61	100.00	850.00	100.00	1113.07
100.00	850	100.00	1096.99				

Table (A.2) The Arab light and Arab heavy light ends and standard density.

Arab Light

Light Ends	
Methane	0.00
Ethane	0.05
Propane	0.69
i-Butane	0.31
n-Butane	1.30
Standard Density	884.3 Kg/m3

Arab Heavy

Light Ends	
Methane	0.00
Ethane	0.01
Propane	0.17
i-Butane	0.18
n-Butane	0.99
Standard Density	859.9 Kg/m3