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Improving Oil Refinery Productivity through Enhanced Crude Blending using Linear Programming Modeling

[1]M.K. Hassan, [2]A. Kandeil and [3]A. Elkhayat

[1]Department of Industrial and Management Engineering, the Arab Academy for Science, Technology and Maritime Transport, P.O. Box 1029, Alexandria, Egypt
[2]Dean of Graduate Studies, the Arab Academy for Science, Technology and Maritime Transport, Egypt
[3]The Egyptian Projects Operation and Maintenance Company, Egypt

Corresponding Author: M.K. Hassan, Department of Industrial and Management Engineering, the Arab Academy for Science, Technology and Maritime Transport, P.O. Box 1029, Alexandria, Egypt Tel: +20181558140

ABSTRACT

Linear Programming (LP) models and technique among various mathematical optimization techniques have evolved through the years to optimize the crude blending and refining operations. The operations may include the crude evaluation, selection, scheduling and product logistics planning. The objective of this study was to develop a mathematical programming model for solving a blending problem in a major refinery in Alexandria, Egypt with the objective of maximizing Naphtha productivity. Refinery planning and optimization is basically addressed through special purpose linear programming software packages that remain a black box for the users and that are very costly for the organizations. The model developed in this work was proved to be highly effective at the level of solving the blending problem. This study has shown that the developed linear programming model for the blending problem has yielded better overall Naphtha productivity for the case of the oil refinery studied, as compared to results obtained by the commercial software.

Key words: Crude blending, naphtha productivity, modeling, linear programming, operations research

INTRODUCTION

Since the discovery of petroleum, the rational utilization of the fractions that compose it has strongly influenced the development of refining processes as well as their arrangement in refining flow sheets. In the late 60's, oil refining has significantly transformed being linked to the continuous increase in the need of oil light products at the expense of heavy products (Wauquier, 1995).

Whereas owing to its flexibility, the refining industry has proven its ability to adapt and being capable to meet changes in demand and quality, in the light of introducing new anti-pollution standards as well as preset limitations for the chemical composition of finished products. So it's obviously seen that current refinery flow sheets, especially beyond the year 2000 would be more restricted to the new specifications using new processes.

Refining processes are basically divided into two main categories: Separation and Conversion. Separation processes include primary distillation of crude oil, secondary distillation or Vacuum Distillation absorption processes, extraction processes, crystallization processes and adsorption processes. Conversion processes include processes for the improvement of properties by: molecular rearrangement, using co-reactants, thermal cracking, catalytic processes, finishing processes and environmental protection processes (Wauquier, 2000).

With all the complexities of the above-mentioned processes, linear programming stands out as a practical convenient and effective tool for solving refinery optimization problems, based on data of unit yields, unit capacities, utilities consumption and the like, as well as product blending operations of the refinery by means of linear objectives and constraints.

THE PETROLEUM REFINING INDUSTRY

Petroleum refining has evolved continuously in response to the always changing consumer demand of energy starting with the whale oil to the development of the internal combustion engine led to the production of gasoline and diesel fuels. Present-day refineries produce a variety of products including many required as feedstock for the petrochemical industry (Meyers, 2003; Prasad, 2002).

A refinery is basically a factory that takes crude oil as raw material and transforms it into hundreds of other useful products. A typical large refinery costs billions of dollars to build and millions more to maintain and upgrade. It runs around the clock 365 days a year, employs between 1,000 and 2,000 people and occupies as much land as several hundred football fields. Roughly, a refinery process can be divided into three parts: crude oil operations, production and product blending as shown in Fig. 1.

In crude oil operations, a variety of crude oil is fed into production devices for refining. In the production process, crude oil is decompounded into different product components. The components then are blended into numerous products.

Crude oil characteristics: Crude oil, also called petroleum, is a complex mixture of carbon and hydrogen (hydrocarbons), which exists as a liquid in the earth's crust. Crude oil has many forms; some is black, thick and tar like, while other crude oils are lighter in color and thinner. The carbon and hydrogen in crude oil are thought to have originated from the remains of microscopic marine organisms that were deposited at the bottom of seas and oceans and were transformed at high temperature and pressure into crude oil and natural gas in petroleum traps. Several different types

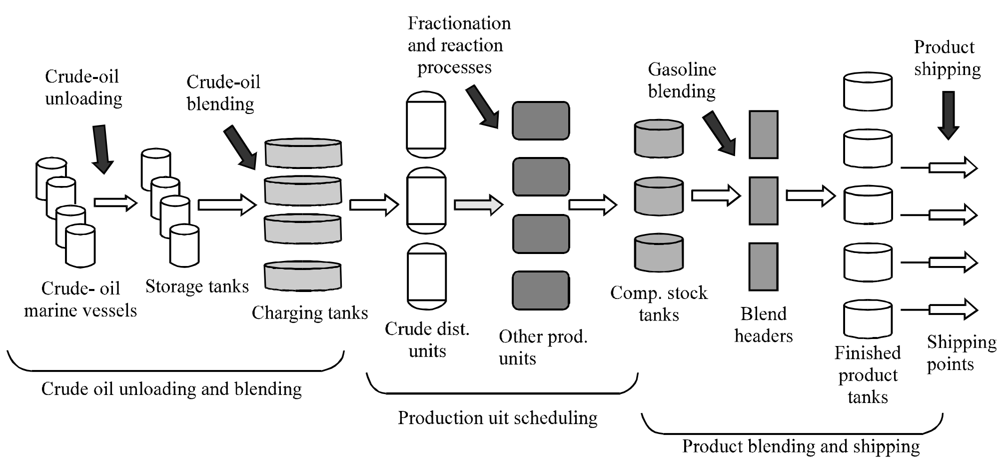

Fig. 1: Illustration of a standard refinery system

of oil and gas traps exist; a common dome is formed by folded sedimentary rocks. Crude oil is obtained by drilling a hole into the reservoir rock (sandstone, limestone etc.) and pumping it out.

What is Naphtha? Naphtha is a complex mixture of hydrocarbons with boiling range of 38 to 205°C, which corresponds to carbon numbers of C5 to C14. Naphtha is produced from petroleum crude oil by fractionation at crude distillation tower working under atmospheric pressure. Crude distillation residues are fed to conversion process units like hydrocracker and cooker units. Residues from atmospheric crude distillation are fed to vacuum distillation, which separates them to vacuum gas oil and vacuum residue. Vacuum oil is fed to hydrocracker unit, where through utilizing Ni and Mo catalysts at an atmosphere of hydrogen, vacuum residue is thermally cracked in cooker unit. Naphtha from crude atmospheric distillation, hydrocracking and cooker units has low octane number, also it contains a big amount of contaminants like sulfur, nitrogen, oxygen, olefins and metals. So Naphtha is fed to hydotreating unit, where hydrogen is reacted with contaminants on catalysts of CoMo and NiMo, giving H_2S-NH_3-H_2O at an atmosphere of 315 to 340°C and pressure of 50 bar.g. Treated Naphtha is sent for further processing by splitting it into:

- Light Naphtha that consists of C5 and C6, that passes to isomerization unit
- Heavy Naphtha with carbon number of C7 to C14, that passes to a catalytic reforming unit

After that both resulting streams are being blended together in a process called gasoline blending, to produce gasoline fuel as a final product.

LINEAR PROGRAMMING IN REFINERY SCHEDULING

In the context of oil industry, a Linear Programming (LP) model is a mathematical model of the refinery, representing all refinery unit yields, unit capacities, utility consumption, as well as product blending operations of the refinery by means of an objective function subject to a number of constraints (Speight, 2002).

Refinery linear programming modeling: As in most industries operating in a competitive environment, refineries must maximize their economic results, to do that; they must maximize their margins, i.e., the difference between their revenues from the products they manufacture and their costs (Gao *et al.*, 2008). However, the refining industry has a particular characteristic, which is that the products manufactured from the various crude oils are interdependent. It is not possible to manufacture just one product from the treatment and conversion units that make up a traditional refinery. The relative proportions of the resulting products depend on the characteristics of the different feed stocks, process units used and the unit operating parameters set (Wu *et al.*, 2005). The proposed work focuses on the refinery process unit configurations to meet the market demand for products (Al-Othman *et al.*, 2008). Thus the task is to find the combination of feed stocks and the appropriate way of processing them, for the objective of giving the best overall margin (Speight, 2002).

Refinery linear program development: In the 1950s, a standard input format to describe a matrix was agreed on, opening the market to LP software from different vendors, as Modeling the oil refinery process and carrying out a Simplex procedure by hand was very tedious and time consuming (Parkash, 2004). With a Simplex algorithm available as a computer program, interest quickly developed in optimizing via a linear programming. Prior to the advent of LP techniques, all optimization studies were done by calculating several hand balances, moving toward an optimal solution by trial and error. The typical refinery LP model used for planning has approximately 300-500 equations and 800-1500 activities to optimize.

Linear programming as a performance monitoring tool in oil refineries: The use of Linear Programming (LP) tools for both long-term planning and day to day scheduling is fairly common in any refinery; still there is a gap between the LP run results and the actual operations. The related variations can be attributed to both external or market forces and internal operational constraints. There are many approaches followed for finding and attributing the money values of these variations in performances. Use of some of the LP tools is one of the best ways for identifying and quantifying these variations. The refining business is piloted on the basis of a Business Plan which is usually developed using a Linear programming package (Pan *et al.*, 2009). The LP tool is an optimizer whose objective function is the GRM (Gross Refining Margin) of the refinery and attempts to maximize this GRM subject to many constraints which are modeled (Balasubramanian, 2003).

Optimization refinery models: Studies were always searching and will continue to search for the optimum working conditions for optimizing the output of production processes (Lan, 2008; Lan *et al.*, 2008; Nja and Udofia, 2009).

Scheduling and planning of the flow of crude oil is a very important problem in a petroleum refinery due to the potential realization of large cost savings and improved feeds. Linear Programming (LP) models have been historically used in the analysis of scheduling and planning problems. Refinery planning problems have been addressed using computational tools such as AspenTech® PIMS (Process Industry Modeling System) that are largely based on Successive Linear Programming (Uri, 1985). However, it is difficult to model refinery operations since they involve units operating in both batch and continuous modes along with multiple grades of crude oil and products. Furthermore, detailed scheduling models often require a continuous time representation and a more general treatment of nonlinear equations, as well as binary variables to model discrete decisions which give rise to Mixed Integer Nonlinear Programming (MINLP) models. These models impart additional flexibility to the problem allowing the modeling of discrete decisions and constraints (Karuppiah *et al.*, 2008).

Refining industry is under immense pressure to produce cleaner products that may lower the economic margins because of stricter environmental regulations and depressed market demand. In this situation, refinery planning becomes very important as it can exploit all potential opportunities to push the economic margins to the maximum limit. The problem of optimizing refinery activities is very complex in its own right. To make the problem of overall optimization solvable, the refineries' practice adopts a decomposition approach, in which material processing is optimized first using Linear Programming (LP) techniques to maximize the overall profit (Hofferl and Steinschorn, 2009). Supporting systems, including the hydrogen network and the utility system, are optimized to reduce operating costs using LP optimization. Essentially these systems are dealt with separately, which usually leads to non-optimal solutions for refinery operations (Zhang *et al.*, 2001).

MATHEMATICAL MODEL BUILDING

The meant objective of this work focuses on improving the Naphtha productivity, through the utilization of better crude oils mixing procedure. The highlighted case study was carried out during

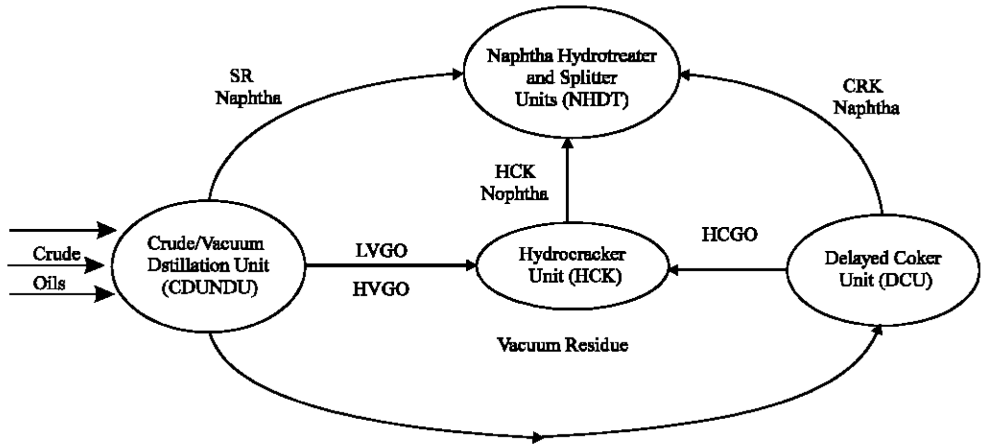

Fig. 2: Crude-naphtha route inside oil refinery

the period from 2007 to 2009 in one of the biggest oil refineries in the Middle-East with a total refining capacity of 100,000 Barrels/day, equivalent to about 665 $m^3 h^{-1}$ and is classified as a Complex Oil Refinery producing a variety of products, such as LPG, gasoline, jet fuel, gas oil, green coke and sulfur, etc. The study deals with the processing of Naphtha feeds directed from different process units, such as Crude Distillation Unit (CDU), Hydrocracker Unit (HCK) and Delayed Cooker Unit (DCU) to the Naphtha Hydrotreater Unit (NHT), then to the Splitter Unit; being referred to as the first stage of Naphtha Complex as shown in Fig. 2.

Model assumptions: The model is built on the following assumptions:

- The model is general for middle-east refineries with similar refining capacities and utilizing same process technology
- The case-study refinery is free to buy the selected crude mix supplies
- The refinery as well is free to deliver products without demand constraints
- Crude purchases cost limitation is neither taken into consideration in this model nor products sales as well
- No operating or maintenance costs, either fixed or variable are included in the study
- The process flow is not utilizing storage capacities except for received crude blends and overall produced Naphtha to NHT unit
- Crude distillation and Vacuum distillation units compromise a collective structure as the receiving area for crude oil fractionation
- Blending for used crudes is being performed online
- The decision to be taken regarding Naphtha further processing into gasoline or selling it as hydro-treated Naphtha stream is to be taken independently
- Crude yields are taken from laboratory results
- Process and units' yields are taken from technical, operations and production planning departments
- Minimum and maximum crude participation percentages are given as per production planning department recommendations
- A decrease in minimum crude participation level in recipe by 1% shall be committed and analyzed from 15% and down to 10% while keeping the maximum values

- An extra proposed case with no constraints on minimum participation values for used crudes in recipe was introduced as well

Mathematical model formulation
Decision variables:

Xi: quantity of the ith crude type in $m^3\ h^{-1}$
where: i = 1, 2, 3, ……… 6; representing the six main types of crudes used in the case study, which are:

X1: ARH Arab Heavy Crude
X2: ARL Arab Light Crude
X3: QRN Qaroun Crude
X4: VDA Val D'Agri Crude
X5: BLT Basra Light Crude
X6: OMN Oman Export Crude

Objective function: The overall Naphtha produced is the sum of the straight run Naphtha, the cooker Naphtha and the hydrocracker Naphtha and the objective function is to maximize this overall amount.

Overall Naphtha produced (Ntotal) = Straight run Naphtha (Nsr) + Cooker Naphtha (Nckr) + Hydrocracker Naphtha (Nhck).

Thus the objective function can be written as:

$$\text{Maximize, } Z = \sum_{i=1}^{6} aiXi + m\sum_{i=1}^{6} biXi + n\sum_{i=1}^{6}\sum_{j=1}^{3} cijXi \quad (1)$$

where, Z : overall Naphtha produced (Ntotal).

$$\text{Straight run Naphtha (Nsr) from Crude Distillation Unit (CDU)} = \sum_{i=1}^{6} aiXi$$

$$\text{Cooker Naphtha (Nckr) from Delayed Cooker Unit (DCU)} = m\sum_{i=1}^{6} biXi$$

$$\text{Hydrocracked Naphtha (Nhck) from Crude Hydrocracker Unit (HCK)} = n\sum_{i=1}^{6}\sum_{j=1}^{3} cijXi$$

where, ai is Naphtha (NAP) yield in crude I, bi is vacuum residue (VR) yield in crude i, cij is light vacuum gas oil (LVGO), heavy vacuum gas oil (HVGO) and heavy cooker gas oil (HCGO) yields, respectively in crude i, where i =1, 2,3,……6 and j = 1, 2, 3. Values for ai, bi and cij for case study are given in Table 1. m is naphtha yield percentage in cooker unit, n is naphtha yield percentage in hydrocracker unit, m and n values (for case study) are equal to 11% (0.11)and 20% (0.2), respectively.

Table 1: Crude yields (vol. %) for naphtha, LVGO, HVGO, VR and HCGO yields

Crude type	NAP yield	LVGO yield	HVGO yield	VR yield	HCGO yield
ARH	14.50	10.50	11.50	27.40	8.22
ARL	18.20	11.90	11.10	16.70	5.01
QRN	14.30	14.70	12.50	16.90	5.07
VDA	23.10	11.10	10.20	12.60	3.78
BLL	18.33	5.16	21.50	19.90	5.97
OMN	16.50	11.80	12.70	20.10	6.03
Cut point	150°C	370-450°C	450-550°C	550-max.°C	

Constraints: A number of constraints apply to this oil mixing problem; these are grouped by unit capacities, crude shares in blend recipe, productivity limitations and unit participations in overall production, all are being presented as follows:

Group 1: Refining capacity constraints: Maximum crude refining capacity, represented by the overall crudes blend $\sum_{i=1}^{6} Xi$.

$$\sum_{i=1}^{6} Xi \le Qmax \tag{2}$$

Minimum crude refining capacity, represented by the overall crudes blend $\sum_{i=1}^{6} Xi$

$$\sum_{i=1}^{6} Xi \ge Qmin \tag{3}$$

Where:
Qmax: Maximum refining capacity for CDU, in case study = 665 m^3 h^{-1}
Qmin: Minimum refining capacity of CDU, in case study = 450 m^3 h^{-1}

Group 2: Crude participation constraints: Minimum value of participation for individual crude type of Xi, such that i = 1, 2,3,... ... . 6:

$$Xi \ge k\sum_{i=1}^{6} Xi \qquad i = 1, 2, 3, \ldots\ldots\ldots 6 \tag{4}$$

where, k is minimum percentage of individual crude participation in blend. For case study runs, k is taking the values 15% (0.15), 14% (0.14), 13% (0.13), 12% (0.12), 11% (0.11) and 10% (0.1)

N.B: This constraint is eliminated as a special case in the last model run.

Maximum value of participation for individual crude types:

$$Xi \le \sum_{i=1}^{6} diXi \tag{5}$$

For case study: d1, d2, d5 and d6, where = 50% (0.5), while d3 and d4 where = 25% (0.25).

Group 3: Naphtha Productivity Constraints for case study: Minimum Naphtha production capacity is L = 109 $m^3\ h^{-1}$.

$$\sum_{i=1}^{6} aiXi + m\sum_{i=1}^{6} biXi + n\sum_{i=1}^{6}\sum_{j=1}^{3} cijXi \geq L \tag{6}$$

Minimum production value for straight run Naphtha (Nsr) from CDU is U1 = 109 $m^3\ h^{-1}$; so as to enable the operation of Naphtha Processing Units in case of DCU and HCK shutdown:

$$\sum_{i=1}^{6} aiXi \geq U1 \tag{7}$$

Maximum production value for straight run Naphtha (Nsr) from CDU is U2 = 125 $m^3\ h^{-1}$:

$$\sum_{i=1}^{6} aiXi \leq U2 \tag{8}$$

Maximum production limit for LVGO from VDU is T1 = 216 $m^3\ h^{-1}$:

$$\sum_{i=1}^{6} ci1Xi \leq T1 \tag{9}$$

Maximum production limit for HVGO from VDU is T2 = 230 $m^3\ h^{-1}$:

$$\sum_{i=1}^{6} ci2Xi \leq T2 \tag{10}$$

Maximum production limit for VR from VDU is T3 = 240 $m^3\ h^{-1}$:

$$\sum_{i=1}^{6} biXi \leq T3 \tag{11}$$

Maximum collective production limit for LVGO, HVGO and VR from CDU to Vacuum Distillation Unit (VDU) is V = 361 $m^3\ h^{-1}$:

$$Xi\left(\sum_{i=1}^{6} ci1 + \sum_{i=1}^{6} ci2 + \sum_{i=1}^{6} bi\right) \leq V \tag{12}$$

Group 4: Units Participation Constraints: We have:

$$\frac{N_{ckr}}{N_{total}} \leq E$$

where, E is the maximum Cooker Naphtha participation percentage in overall Naphtha and for case study = 11% (0.11)

$$\frac{m\sum_{i=1}^{6} b_i X_i}{\sum_{i=1}^{6} a_i X_i + m\sum_{i=1}^{6} b_i X_i + n\sum_{i=1}^{6}\sum_{j=1}^{3} c_{ij} X_i} \le E \tag{13}$$

Also, we have:

$$\frac{N_{ckr} + N_{hck}}{N_{total}} \le F$$

where, F is the maximum Cooker and Hydrocracker collective Naphtha participations percentage in overall Naphtha and for case study = 40% (0.4)

$$\frac{m\sum_{i=1}^{6} b_i X_i + n\sum_{i=1}^{6}\sum_{j=1}^{3} c_{ij} X_i}{\sum_{i=1}^{6} a_i X_i + m\sum_{i=1}^{6} b_i X_i + n\sum_{i=1}^{6}\sum_{j=1}^{3} c_{ij} X_i} \le F \tag{14}$$

Finally, we apply the: Non-negativity constraint:

$$Xi \ge 0 \quad i = 1, 2, 3, \ldots, 6 \tag{15}$$

The script of the model is shown in appendix.

Table 2 lists the average prices for crudes in refinery case during the year of 2007, which was the year taken as the base for the study.

Table 3 lists the average petroleum products' prices during the base year of 2007.

Model scenarios: Seven scenarios have been proposed for the model runs based on the proposed blend cases. Each scenario puts a constraint on the minimum individual crude participation that

Table 2: Average prices for crudes used in refinery during year 2007

Crude type	BBL price in USD ($)
Arab heavy (ARH)	67.040
Arab light (ARL)	70.240
Qaroun (QRN	71.740
Val D'Agri Blend (VAL)	72.740
BLTLAB (BLL)	68.740
Oman export (OMA)	69.240

Table 3: Average products prices (Platts) for year 2007

	LPG	MOGAS	Jet Fuel	DSL 0.2S	EN 590	Naphtha	FO
Average prices ($/Ton)	629.87	684.98	689.68	641.24	660.81	646.46	342.21

is allowed. The minimum participations have been constrained between 10 and 15%, as no crude percentage is allowed to be less that 10% to be able to judge its effect of the overall Naphtha productivity; and when all the seven crudes are almost equal in participation ratios, this will give a percentage of about 15%. For the seventh run, the minimum crude participation constrain was eliminated to give the freedom for the model to calculate the ratios according to the other set of constraints.

Scenario (1): Model run with min. individual crude participation value of 15%
Scenario (2): Model run with min. individual crude participation value of 14%
Scenario (3): Model run with min. individual crude participation value of 13%
Scenario (4): Model run with min. individual crude participation value of 12%
Scenario (5): Model run with min. individual crude participation value of 11%
Scenario (6): Model run with min. individual crude participation value of 10%
Scenario (7): Model run by elimination of min. crude participation% constraint

For all scenarios, the model was run with max. individual crude participation% constraint for crudes X1, X2, X5 and X6 being 50% of total blend recipe, whereas for crudes X3 and X4 respectively the max. was 25%. These ratios are governing ratios depending on the design of the refinery and the availability of the crudes.

RESULTS AND DISCUSSION

The developed linear programming model has been solved using the data corresponding to the seven proposed cases detailed above. The proposed cases have shown that, output improvements have been achieved through better crude mixing procedures; the traced results have shown the improvement of Naphtha productivity through varying the blend recipes for the seven model runs. The seventh run was observed as the highest record for Naphtha productivity in $m^3\ h^{-1}$ unit for the year 2007 case study. Consequently, all records have shown better results than the study case refinery crude mixing schemes, but still the first two of them had shown figures that were a bit less than Naphtha productivity rate regarding the refinery's actual run scenario for the meant year of study. The Naphtha Productivity for the Refinery Design Scheme (1): 50%ARH-50%ARL was 172 $m^3\ h^{-1}$, while for Design Scheme (2): 100%ARL was 171 $m^3\ h^{-1}$, whereas for the Actual Run Scheme the output was 173.50 $m^3\ h^{-1}$.

The Refinery LP model is such a complex and sophisticated model that relates crude purchases to operation parameters and used technologies together with cost and availability of purchased crudes in conjunction with product sales, taking into account the main aim of achieving minimum cost to maximum profitability with lots of fixed and variable costs to take into consideration regarding assisting units and operation costs including cooling water, steam, hydrogen, natural gas, chemicals , catalysts, laboratory examinations, inspection work, maintenance activities, spare parts, labor, asset depreciation, downtimes and etc.

It takes also into account the hierarchy of petroleum derived products in great attention, coping as well with supply chain functions and relating all of that to a strategic scheduling plan. As being in great concern for handling a continuous process such as oil refining and taking into account that it deals with all refining activities as a whole unit; thus being a global refinery optimization model. On the other hand, the developed LP model represents just a highhght of the Crude-Naphtha route inside the case study refinery, taking into attention parameters relating to the studied system, including crude characteristics and yields, process and units' yields, constraints on crude participation, sources of Naphtha inside refinery, influenced units and methods used for productivity calculation.

Table 4, represents the different crude oils' participations in the seven proposed model runs for mixing recipes. As noticed on the table, any decrease in the ARH, ARL, QRN and OMN participations will be an increase in the participation of BLL and that the participation of VDA is constant at 25%.

VDA has the maximum Naphtha yield and its percentage is already maximized at 25% and BLL has the second maximum yield, therefore any decrease in the other crudes participations will go the BLL ratio as till it reaches the maximum allowed ratio of 50%.

Table 5 shows that further improvement using the proposed mathematical model and the accompanying linear program are possible through better crude mixing scheme.

Results have shown the improvement of Naphtha productivity through the consecutive records of the model runs. Productivity of the first and second model runs have fallen below the actual rate of Naphtha productivity (173.50 m^3 h^{-1}), but still showing better figures for Naphtha productivity compared to the design cases for the refinery crude operations schemes. Table 5 also shows enhancement in the collective crude purchases from run 1 to 7.

Calculating the productivity-cost index, which is the ratio of the overall produced Naphtha (m^3 h^{-1}) to the collective crude purchases ($1000) also shows improvement in the index from run 1 to 7. Figure 3 shows the decrease of crude purchases from run 1 to run 7. While Fig. 4 shows an increase in Naphtha productivity and in the mean time decrease in crude purchases from model run1 to 7.

Table 4: Crude participation percentages for model runs' mixing recipes

Crude type	Model run (1)	Model run (2)	Model run (3)	Model run (4)	Model run (5)	Model run (6)	Model run (7)
ARH	15%	14%	13%	12%	11%	10%	13.5%
ARL	15%	14%	13%	12%	11%	10%	0.0%
QRN	15%	14%	13%	12%	11%	10%	0.0%
VDA	25%	25%	25%	25%	25%	25%	25.0%
BLL	15%	19%	23%	27%	31%	35%	50.0%
OMN	15%	14%	13%	12%	11%	10%	11.5%

Table 5: Model Scenarios output and calculated data

Model scenarios	Run 1	Run 2	Run 3	Run 4	Run 5	Run 6	Run 7
Overall produced naphtha (m^3 h^{-1})	172.41	173.18	173.95	174.72	175.49	176.26	179.11
Naphtha yield (%)	25.93%	26.04%	26.16%	26.27%	26.39%	26.51%	26.93%
Collective crude purchases (1000$)	293.75	293.62	293.48	293.34	293.21	293.07	290.98
Productivity-cost index (m^3/1000$.h)	0.587	0.590	0.593	0.596	0.599	0.601	0.616

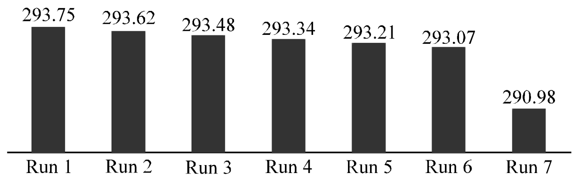

Fig. 3: Crude purchases (1000$) for lingo model runs

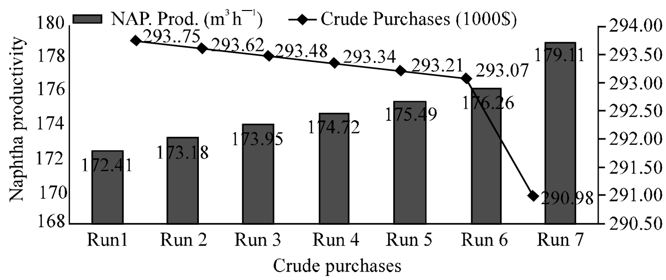

Fig. 4: Naphtha productivity versus crude purchases for lingo model runs

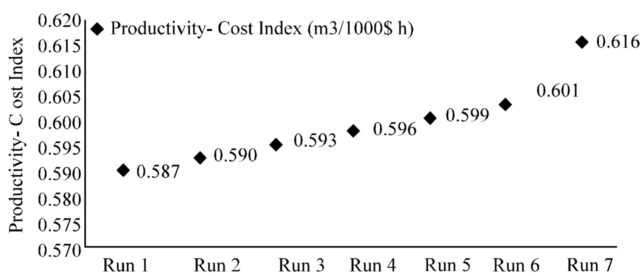

Fig. 5: Productivity-cost index for lingo model case runs

Figure 5 shows the productivity-cost index for the different model runs. It is noticed that there is an increase (improvement) in the cost index going up from model run 1 to 7.

MODEL VERIFICATION AND VALIDATION

After the formulation of the mathematical model had been completed and the generated LP model was run for seven consecutive times relating to the seven proposed blend cases that were highlighted earlier. Taking into consideration all model assumptions and system requirements that were given. The results were then compared to the actual case for an operating day of the study year 2007. The actual processed blend was originally composed of nine different crude types, from which three were excluded in our study case due to their minor participations in the crude recipe. The left six dominating types were analyzed through LP techniques to see the extent of how much Naphtha productivity could be improved with the different scenarios. The sensitivity analysis for the LP model was performed by the LINGO package itself. A full review for the model verification, validation and runs is following, that shall be used in further interpretation and analysis.

Model verification: The model was reviewed mathematically regarding its linear programming format for verification and consistency, the suggested route and proposed assumptions were also given concern. The model context was then verified by the refinery LP team leader and confirmed for adequacy by the refinery production planning manager. The model was tested for validity and conformity, the highlighted case study criteria were also considered. Finally the model was validated by the LP team leader and confirmed by the refinery production planning manager.

Model validation: In the validation step of the model the crude participations for the different runs that were input to the model were also input to the refinery LP software to compare the results of both. Table 6 shows the comparison of the Naphtha productivity and yield calculated from the developed Lingo model and calculated also using the refinery LP software.

As shown on the table, there was a margin of error between the developed model and the refinery LP in productivity values that decreases from about 5 $m^3 h^{-1}$ for run 1 to less than 1 $m^3 h^{-1}$ for run 7. Comparing the yield calculations, there was a margin of error between the developed model and the refinery LP that decreases from 0.92% for run 1 down to 0.24% for run 7.

The error margins are due to the factors that were not taken into consideration in the construction of the model. These factors are directly related to the model assumptions mentioned earlier. Model run 7 has showed the maximum productivity, the best yield and highest productivity-cost index. This should be correct as the constraint of minimum crude participation percentage have been removed, as a results, the model will utilize only the crudes with the maximum Naphtha yields. This is why we see on Table 4 that, both the ARL and the QRN crudes have been eliminated, BLL and VDA have reached their maximum allowable participation percentage at 50 and 25% respectively and the rest of the 100% went to the ARH and OMN with a higher percentage for ARH as it has a higher yield than the OMN crude. But unfortunately, this case is not always achievable as the refinery has to utilize the available crudes coming out of the oil fields and also has to satisfy the cost and productivity constraints as the refinery should be operating full time around the clock.

Figure 6 shows the comparison between the developed model and the running refinery LP model regarding the Naphtha productivity. Figure 6 shows the decreased difference between the two methods from run 1 to 7.

Table 6: Developed model versus refinery model statistics

Model run	1	2	3	4	5	6	7
NAP productivity $m^3 h^{-1}$							
Lingo LP model	172.41	173.18	173.95	174.72	175.490	176.25	179.11
Refinery LP	177.528	177.829	178.266	178.445	178.847	178.607	179.897
NAP yield %							
Lingo LP model	25.93%	26.04%	26.16%	26.27%	26.39%	26.51%	26.93%
Refinery LP	26.85%	26.89%	26.95%	26.98%	27.04%	27.08%	27.17%

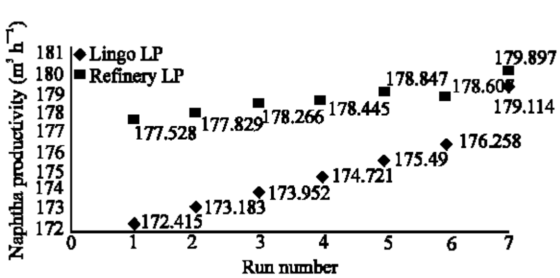

Fig. 6: Naphtha productivity for lingo LP versus refinery LP

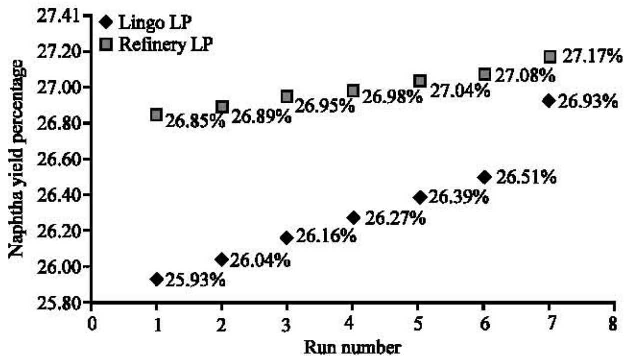

Fig. 7: Naphtha% yield for lingo LP versus refinery LP

Figure 7 shows the comparison between the developed model and the running refinery LP model regarding the Naphtha yield. Again as shown in Fig. 6 and 7 the decreased difference between the two methods from run 1 to 7.

CONCLUSIONS

The developed linear program has been solved using data corresponding to the seven proposed cases detailed. Generally, the proposed cases have shown that further improvements can be achieved through better crude mixing procedures; the traced results have shown the improvement of Naphtha productivity in most of the runs being compared to the actual case. Starting from the third run and up to the seventh, the resulting values have exceeded the actual value for the year 2007 case. Productivity of the first and second runs have fallen below the actual rate of Naphtha productivity for actual case, but still showing better figures for Naphtha productivity rates regarding design cases for the proposed refinery crude operations schemes. To analyze the results for the performed model runs comparisons were made between their Naphtha's Productivities and Yields and their crude purchases were calculated for each model case run, also a Productivity-Coat Index was generated to show which scenario would have been more preferable to operate. The run 7 that of maximum Naphtha productivity and yield, as well as showing the biggest Productivity-Cost Index value among other model case runs was compared with the actual case run for the study year 2007 on Refinery LP results basis due to the removal of the minimum participation level constraints. This comparison showed that this case would have been more beneficial to the Refinery regarding Naphtha Productivity and Yield issue. But this case is not always feasible due to the constraints on the available crudes, prices and refinery capacity utilization issues. The proposed model would be very beneficial when trying to optimize single components of the refinery outputs. As a simple and easy to construct and run, the developed model showed consistency, robustness and low error margin compared to the very sophisticated and very expensive refinery model. As a recommendation, the model can further be enhanced to compensate for the other variables that were not taken into consideration.

APPENDIX

Model script:

[OBJECTIVE_max_Produced_Naphtha in $m^3\ h^{-1}$] MAX= N_TOTAL;
[total_crude _quantity in $m^3\ h^{-1}$]Q - X_1 - X_2 - X_3 - X_4 - X_5 - X_6=0;

[non-zero_constraint_for_crude1] X_1 >= 0;
[non-zero_constraint_for_crude2] X_2 >= 0;
[non-zero_constraint_for_crude3] X_3 >= 0;
[non-zero_constraint_for_crude4] X_4 >= 0;
[non-zero_constraint_for_crude5] X_5 >= 0;
[non-zero_constraint_for_crude6] X_6 >= 0;

[low_participation_for_crude1] X_1 / Q >= 0;
[low_participation_for_crude2] X_2 / Q >= 0;
[low_participation_for_crude3] X_3 / Q >= 0;
[low_participation_for_crude4] X_4 / Q >= 0;
[low_participation_for_crude5] X_5 / Q >= 0;
[low_participation_for_crude6] X_6 / Q >= 0;

[high_participation_for_crude1] X_1 / Q <= 0.50;
[high_participation_for_crude2] X_2 / Q <= 0.50;
[high_participation_for_crude3] X_3 / Q <= 0.25;
[high_participation_for_crude4] X_4 / Q <= 0.25;
[high_participation_for_crude5] X_5 / Q <= 0.50;
[high_participation_for_crude6] X_6 / Q <= 0.50;

[Straight run_Naphtha_from_Crude1 in m^3 h^{-1}] - 0.145 * X_1 + N_SR_1 = 0;
[Straight run_Naphtha_from_Crude2 in m^3 h^{-1}] - 0.182 * X_2 + N_SR_2 = 0;
[Straight run_Naphtha_from_Crude3 in m^3 h^{-1}] - 0.143 * X_3 + N_SR_3 = 0;
[Straight run_Naphtha_from_Crude4 in m^3 h^{-1}] - 0.231 * X_4 + N_SR_4 = 0;
[Straight run_Naphtha_from_Crude5 in m^3 h^{-1}] - 0.183 * X_5 + N_SR_5 = 0;
[Straight run_Naphtha_from_Crude6 in m^3 h^{-1}] - 0.165 * X_6 + N_SR_6 = 0;

[Vacuum_Residue_from_Crude1 in m^3 h^{-1}] - 0.274 * X_1 + VR_1 = 0;
[Vacuum_Residue_from_Crude2 in m^3 h^{-1}] - 0.167 * X_2 + VR_2 = 0;
[Vacuum_Residue_from_Crude3 in m^3 h^{-1}] - 0.169 * X_3 + VR_3 = 0;
[Vacuum_Residue_from_Crude4 in m^3 h^{-1}] - 0.126 * X_4 + VR_4 = 0;
[Vacuum_Residue_from_Crude5 in m^3 h^{-1}] - 0.199 * X_5 + VR_5 = 0;
[Vacuum_Residue_from_Crude6 in m^3 h^{-1}] - 0.201 * X_6 + VR_6 = 0;

[Cooker_Naphtha__from_Crude1 in m^3 h^{-1}] N_CKR_1 - 0.11 * VR_1 = 0;
[Cooker_Naphtha__from_Crude2 in m^3 h^{-1}] N_CKR_2 - 0.11 * VR_2 = 0;
[Cooker_Naphtha__from_Crude3 in m^3 h^{-1}] N_CKR_3 - 0.11 * VR_3 = 0;
[Cooker_Naphtha__from_Crude4 in m^3 h^{-1}] N_CKR_4 - 0.11 * VR_4 = 0;
[Cooker_Naphtha__from_Crude5 in m^3 h^{-1}] N_CKR_5 - 0.11 * VR_5 = 0;
[Cooker_Naphtha__from_Crude6 in m^3 h^{-1}] N_CKR_6 - 0.11 * VR_6 = 0;

[Light_Vacuum_Gas_Oil_from_Crude1 in m^3 h^{-1}] - 0.105 * X_1 + LVGO_1 = 0;
[Light_Vacuum_Gas_Oil_from_Crude2 in m^3 h^{-1}] - 0.119 * X_2 + LVGO_2 = 0;
[Light_Vacuum_Gas_Oil_from_Crude3 in m^3 h^{-1}] - 0.147 * X_3 + LVGO_3 = 0;
[Light_Vacuum_Gas_Oil_from_Crude4 in m^3 h^{-1}] - 0.111 * X_4 + LVGO_4 = 0;

[Light_Vacuum_Gas_Oil_from_Crude5 in m^3 h^{-1}] - 0.052 * X_5 + LVGO_5 = 0;
[Light_Vacuum_Gas_Oil_from_Crude6 in m^3 h^{-1}] - 0.118 * X_6 + LVGO_6 = 0;

[Heavy_Vacuum_Gas_Oil_from_Crude1 in m^3 h^{-1}] - 0.115 * X_1 + HVGO_1 = 0;
[Heavy_Vacuum_Gas_Oil_from_Crude2 in m^3 h^{-1}] - 0.111 * X_2 + HVGO_2 = 0;
[Heavy_Vacuum_Gas_Oil_from_Crude3 in m^3 h^{-1}] - 0.125 * X_3 + HVGO_3 = 0;
[Heavy_Vacuum_Gas_Oil_from_Crude4 in m^3 h^{-1}] - 0.102 * X_4 + HVGO_4 = 0;
[Heavy_Vacuum_Gas_Oil_from_Crude5 in m^3 h^{-1}] - 0.215 * X_5 + HVGO_5 = 0;
[Heavy_Vacuum_Gas_Oil_from_Crude6 in m^3 h^{-1}] - 0.127 * X_6 + HVGO_6 = 0;
[Heavy_Cooker_Gas_Oil_from_Crude1 in m^3 h^{-1}] - 0.0822 * X_1 + HCGO_1 = 0;
[Heavy_Cooker_Gas_Oil_from_Crude2 in m^3 h^{-1}] - 0.0501 * X_2 + HCGO_2 = 0;
[Heavy_Cooker_Gas_Oil_from_Crude3 in m^3 h^{-1}] - 0.0506 * X_3 + HCGO_3 = 0;
[Heavy_Cooker_Gas_Oil_from_Crude4 in m^3 h^{-1}] - 0.0378 * X_4 + HCGO_4 = 0;
[Heavy_Cooker_Gas_Oil_from_Crude5 in m^3 h^{-1}] - 0.0597 * X_5 + HCGO_5 = 0;
[Heavy_Cooker_Gas_Oil_from_Crude6 in m^3 h^{-1}] - 0.0603 * X_6 + HCGO_6 = 0;

[Hydrocracker_Naphtha__from_Crude1 in m^3 h^{-1}] N_HYDRO_1 - 0.2 * LVGO_1 - 0.2 * HVGO_1 - 0.2 * HCGO_1 = 0;
[Hydrocracker_Naphtha__from_Crude2 in m^3 h^{-1}] N_HYDRO_2 - 0.2 * LVGO_2 - 0.2 * HVGO_2 - 0.2 * HCGO_2 = 0;
[Hydrocracker_Naphtha__from_Crude3 in m^3 h^{-1}] N_HYDRO_3 - 0.2 * LVGO_3 - 0.2 * HVGO_3 - 0.2 * HCGO_3 = 0;
[Hydrocracker_Naphtha__from_Crude4 in m^3 h^{-1}] N_HYDRO_4 - 0.2 * LVGO_4 - 0.2 * HVGO_4 - 0.2 * HCGO_4 = 0;
[Hydrocracker_Naphtha__from_Crude5 in m^3 h^{-1}] N_HYDRO_5 - 0.2 * LVGO_5 - 0.2 * HVGO_5 - 0.2 * HCGO_5 = 0;
[Hydrocracker_Naphtha__from_Crude6 in m^3 h^{-1}] N_HYDRO_6 - 0.2 * LVGO_6 - 0.2 * HVGO_6 - 0.2 * HCGO_6 = 0;

[Purchases_for_Crude1 in USD] - 421.630 * X_1 + C_CRD_1 = 0;
[Purchases_for_Crude2 in USD] - 441.761 * X_2 + C_CRD_2 = 0;
[Purchases_for_Crude3 in USD] - 451.195 * X_3 + C_CRD_3 = 0;
[Purchases_for_Crude4 in USD] - 457.484 * X_4 + C_CRD_4 = 0;
[Purchases_for_Crude5 in USD] - 432.372 * X_5 + C_CRD_5 = 0;
[Purchases_for_Crude5 in USD] - 435.471 * X_6 + C_CRD_6 = 0;

[Collective_Cost_for_Crude_Blend in USD] SUMCOST - C_CRD_1 - C_CRD_2 - C_CRD_3 - C_CRD_4 - C_CRD_5 - C_CRD_6 = 0;

[Total_Straightrun_Naphthas in m^3 h^{-1}] SUMN_SR - N_SR_1 - N_SR_2 - N_SR_3 - N_SR_4 - N_SR_5 - N_SR_6 = 0;

[Total_Cooker_Naphthas in m^3 h^{-1}] SUMN_CKR - N_CKR_1 - N_CKR_2 - N_CKR_3 - N_CKR_4 - N_CKR_5 - N_CKR_6 = 0;

[Total_Hydrocracker_Naphthas in m^3 h^{-1}] SUMN_HYDRO - N_HYDRO_1 - N_HYDRO_2 - N_HYDRO_3 - N_HYDRO_4 - N_HYDRO_5 - N_HYDRO_6 = 0;

[Collective_Naphtha_Feeds in m^3 h^{-1}] N_TOTAL - N_SR_1 - N_HYDRO_1 - N_CKR_1 - N_SR_2 - N_HYDRO_2 - N_CKR_2 - N_SR_3 - N_HYDRO_3 - N_CKR_3 - N_SR_4 - N_HYDRO_4 - N_CKR_4 - N_SR_5 - N_HYDRO_5 - N_CKR_5 - N_SR_6 - N_HYDRO_6 - N_CKR_6 = 0;

[Naphtha_Yield_Productivity] NAP_PROD_YIELD - N_TOTAL / Q = 0;

[Total_Vacuum_Residues in m^3 h^{-1}] SUMVR - VR_1 - VR_2 - VR_3 - VR_4 - VR_5 - VR_6 = 0;

[Total_Light_Vacuum_Gas_Oils in m^3 h^{-1}] SUMLVGO - LVGO_1 - LVGO_2 - LVGO_3 - LVGO_4 - LVGO_5 - LVGO_6 = 0;

[Total_Heavy_Vacuum_Gas_Oils in m^3 h^{-1}] SUMHVGO - HVGO_1 - HVGO_2 - HVGO_3 - HVGO_4 - HVGO_5 - HVGO_6 = 0;

[Total_heavy_Cooker_Gas_Oils in m^3 h^{-1}] SUMHCGO - HCGO_1 - HCGO_2 - HCGO_3 - HCGO_4 - HCGO_5 - HCGO_6 = 0;

[Minimum_Crude_Flow in m^3 h^{-1}] Q >= 450;
[Maximum_Crude_Flow in m^3 h^{-1}] Q <= 665;
[Minimum_Produced_Straigtrun_Naphtha in m^3 h^{-1}] SUMN_SR >= 109;
[Maximum_Produced_Straigtrun_Naphtha in m^3 h^{-1}] SUMN_SR <= 125;
[Maximum_Produced_Light_Vacuum_Gas_Oil in m^3 h^{-1}] SUMLVGO <= 216;
[Maximum_Produced_Heavy_Vacuum_Gas_Oil in m^3 h^{-1}] SUMHVGO <= 230;
[Maximum_Produced_Vacuum_Residue in m^3 h^{-1}] SUMVR <= 240;
[Maximum_Produced_Vacuum_Gas_Oils_and_Residue in m^3 h^{-}] SUMVR + SUMLVGO + SUMHVGO <= 361;

[Maximum_Participation_for_Collective_Cooker_Naphthas_versus_Overall_Naphthas in m^3 h^{-1}] 0.11 * N_TOTAL - SUMN_CKR >= 0;

[Maximum_Participation_for_Collective_Cooker_and_Hydrocracker_Naphthas_versus_Overall_Naphthas in m^3 h^{-1}] 0.4 * N_TOTAL - SUMN_CKR - SUMN_HYDRO >= 0;
END

!LINGO Model to maximize Naphtha Productivity for the case study Refinery with a maximum refining capacity of 100,000 barrels, equivalent to 665 m^3 h^{-1}

Crude Types: X1 Arab Heavy (ARH), X2 Arab Light (ARL), X3 Qaroun (QRN), X4 Val D'Agri (VDA), X5 Basra Light Lab (BLL), X6 Oman Export (OMN)

N_TOTAL: Overall Produced Naphtha in m^3 h^{-1} (OBJECTIVE function is to maximize)

N_SR: Straight run Naphtha in m^3 h^{-1}
N_ckr: Cooker Naphtha in m^3 h^{-1}
N_hck: Hydrocracker Naphtha in m^3 h^{-1}
C_crd: Cost of Individual Crude in $

Cost: cost of crude in $/m3
VR: Vacuum Residue in $m^3\ h^{-1}$
LVGO: Light Vacuum Gas Oil in $m^3\ h^{-1}$
HVGO: Heavy Vacuum Gas Oil in m/hr
HCGO: Heavy Cooker Gas Oil in $m^3\ h^{-1}$
NAP_yield: Naphtha yield in Crude (Vol.%)
LV_yield: LVGO yield in Crude (Vol.%)
HV_yield: HVGO yield in Crude (Vol.%)
VR_yield: VR yield in Crude (Vol.%)
HC_yield: HCGO yield in Crude (Vol.%)
High_part:Maximum Participation of Crude in Blend(%)
low_part:Minimum Participation of Crude in Blend(%)
CKR_Yield:Cooker Unit Naphtha Yield(%)
HCK_Yield:Hydrocracker Unit Naphtha Yield(%)
NAP_Prod_Yield: Calculated Overall Naphtha to Total Crude Feeds

REFERENCES

Al-Othman, W.B., H.M. Lababidi, I.M. Alatiqi and K. Al-Shayji, 2008. Supply chain optimization of petroleum organization under uncertainty in market demands and prices. Eur. J. Operat. Res., 189: 822-840.

Balasubramanian, K., 2003. Performance monitoring in a refinery using linear programming tools. Infosys. http://www.infosys.com/offerings/industries/energy/white-papers/Documents/performance-monitoring-refinery.pdf.

Gao, Z., L. Tang, H. Jin and N. Xu, 2008. An optimization model for the production planning of overall refinery. Chinese J. Chem. Eng., 16: 67-70.

Hofferl, F. and D. Steinschorn, 2009. A dynamic programming extension to the steady state refinery LP. Eur. J. Operat. Res., 197: 465-474.

Karuppiah, R., K. Furman and I. Grossmann, 2008. Global optimization for scheduling refinery crude oil operations. Comput. Chem. Eng., 32: 2745-2766.

Lan, T.S., 2008. Decision criteria for optimum machining control. J. Applied Sci., 8: 1898-1904.

Lan, T.S., C.Y. Lo and C.I. Hou, 2008. Optimum production control and workforce scheduling of machining project. J. Applied Sci., 8: 836-841.

Meyers, R., 2003. Handbook of Petroleum Refining Processes. 3rd Edn., McGraw-Hill, New York, ISBN: 0-07-139109-6.

Nja, M.E. and G.A. Udofia, 2009. Formulation of the mixed-integer goal programming model for flour producing companies. Asian J. Math. Statistics, 2: 55-64.

Pan, M., X. Li and Y. Qian, 2009. New approach for scheduling crude oil operations. Chem. Eng. Sci., 64: 965-983.

Parkash, S., 2004. Refining Processes Handbook. Elsevier Publisher, London, ISBN-13: 978-0-7506-7721-9.

Prasad, R., 2002. Petroleum Refining Technology. 2nd Edn., Khanna Publishers, New Delhi, ISBN: 81-7409-064-9.

Speight, J., 2002. Handbook of Petroleum Product Analysis. Wiley-Inter Science, New York, ISBN: 0-471-20346-7.

Uri, N., 1985. Modeling oil refinery operations. Socio-Econ. Plann. Sci., 19: 35-40.

Wauquier, J.P., 1995. Petroleum Refining: Crude Oil, Petroleum Products, Process Flowsheets. 1st Edn., Editions Technip, Paris, ISBN-10: 2710806851.

Wauquier, J.P., 2000. Separation Processes. 1st Edn., Editions Technip, Paris, ISBN-10: 2710807793.

Wu, N., M. Zhou and F. Chu, 2005. Short-term scheduling for refinery process: bridging the gap between theory and applications. Int. J. Int. Control Syst., 10: 162-174.

Zhang, J., X. Zhu and G. Towler, 2001. A simultaneous optimization strategy for overall integration in refinery planning. Ind. Eng. Chem. Res., 40: 2640-2653.