JOURNAL OF BIOPROCESSING AND CHEMICAL ENGINEERING

Journal homepage: http://scienceq.org/Journals/JBCE.php

Research Article Open Access

Development of Optimal Strategy for the Design and Operation of a Crude Petroleum Distilation (Topping) Unit

Tsunatu D. Yavini[1*], James M. Namo[2], and Inuwa L. Sani[3]
[1]Chemistry Department, Taraba State University, P.M.B 1167, Jalingo – Taraba State. Nigeria.
[2]Chemical Engineering Department, Ahmadu Bello University, Zaria – Kaduna State, Nigeria.
[3]Chemical Engineering Department, Kaduna Polytechnic, Kaduna – Kaduna State, Nigeria.
*Corresponding author: Tsunatu D. Yavini E-mail: tsunatu.danlami@yahoo.co.uk
Received: May 12, 2014, Accepted: June 26, 2015, Published: June 26, 2015

ABSTRACT
An optimal strategy of Crude Petroleum Distillation Unit (CDU) was developed to carry out process optimization for maximizing oil production and net profit for the plant. The optimization technique suggested the employment of Sequential Quadratic Programming (SQP) Method using ASPEN PLUS. The input operating variables of the CDU were properties of the crude oil obtained from Kaduna Refinery and Petrochemical Company (KRPC) and manipulated variables such as the flow rates, reflux ratio, temperature and pressure. The objective function of the optimization was to maximize the net profit of the desired products using the SQP method. Variation of Bottom Oil flow rate and Reflux Ratio on Net Profit calculated in Naira per year on a molar flow rate and reflux ratio reflects the expected results which should increase the profit. The outcome should indicate and suggest whether the plant is operating efficiently and the model can provide an effective planning and operation tool.
Keyword: *Optimization, ASPEN PLUS, Strategy, CDU, Net Profit, Variables, Model*

INTRODUCTION
Crude Distillation Units are key process plants in a petroleum refinery as they produce intermediate streams that are used in downstream process units. Changes in these units have a great impact on product yield and quality and, therefore, it is recommended to operate these units at optimal conditions from technical and commercial points of view; that means operating conditions such as temperatures, pressures and flow rates of the units that maximize their economic performance (increasing product yield), subject to their real physical restrictions and their design capabilities. Mathematical modeling has become very common to develop these optimization studies.

STRATEGIES FOR OPTIMIZATION
There are several strategies that can be used to perform mathematical optimization processes.

Linear Programming (LP) is generally used in production planning and programming due to their simplicity. However, its precision has been the subject of discussion for decades due to the simplified linear formulation of non-linear processes that can lead to non real solutions. One of the most important strengths of linear models is that they always converge in reasonable calculation times.

Another optimization strategy consists in the utilization of rigorous models, based on thermodynamic principles linked to an optimizer [1]. Even though this type of optimization begins with a precise CDU representation, it exhibits serious problems for its practical implementation since it does not ensure convergence and the calculation time increases significantly for very complex models.

Another strategy in order to take advantage of synergy among processes. The idea applies again the LP model structure, although this time proposes the inclusion of non-linear process models within the global model through linearization techniques.

The last strategy mentioned here was proposed by the Brazilian researchers [2] who have shown their interest in the improvement of planning models by enhancing process representation using non-linear optimization. They propose the inclusion of process complexity within the model, in order to improve the accuracy of the results. In other words, they consider the non-linear representation of units such as CDU within the macromodel in order to take advantage of the interaction among processes and provide operational guidelines to the units through a simple but robust model.

DESCRIPTION OF KADUNA REFINING AND PETROCHEMICAL COMPANY LIMITED (KRPC)
Kaduna Refining and Petrochemical Company Limited (KRPC) is a subsidiary of Nigeria National Petroleum Corporation (NNPC) that was established in 1980 to efficiently and profitably process crude oil into refined petroleum products and manufacture of linear alkyl benzene (LAB) for domestic consumption and export [3].

The CDU process poses a challenging control and optimization problem since it is multivariable, nonlinear and encounters a variety of process disturbances. In an attempt to increase unit profitability, many refiners have replaced conventional process control schemes with advanced multivariable systems [4]. Having gained the economic benefits of advanced control schemes, many companies around the world are considering process optimization to further increase CDU profitability. Increased unit profitability provides the impetus to develop a process optimization strategy for the CDU.

METHODOLOGY

Process Simulation Procedure

Collection of Data: Operating Data and Piping and Instrumentation Diagram of Crude Distillation Unit (CDU) of Kaduna Refinery and Petrochemical Company (KRPC) were collected from KRPC.

Construction of CDU Model in a Process Simulator: Building the CDU model of KRPC in ASPEN PLUS using the data collected from 1 above.

Computer Simulation: Computer Simulation of the model constructed in 2 was carried out using ASPEN PLUS simulator.

Process Description

The modeling and optimization was performed using ASPEN PLUS version 11.1 as shown in Figure 1. Simulations was performed using KRPC data to validate the whole simulation procedure and subsequently optimize using the ASPEN PLUS optimizer. The procedures for process simulation mainly involve defining chemical components, selecting a thermodynamic model, determining plant capacity, choosing proper operating units and setting up input conditions (flow rate, temperature, pressure, and other conditions). Data on most components, such as water, hydrocarbons, oxygen, CO, CO_2, NO_2, SO_2, is available in the ASPEN PLUS component library. Due to the presence of the highly polar and non-polar components thermodynamic and activity models, Peng Robinson was recommended to predict the activity, densities coefficients of the components in a liquid and vapor phase [5]. Most of the heat utilities information will be assumed in order to develop the model. The main processing units include separator, distillation column, valves, cooler and heaters. Single stage distillation will be used as main fractionators for vapor effluents and separation into end products recovery such as the flue gases, gasoline (C5+) and bottom oil. After the input information and operating unit models were set up, the process steady- state simulation will be executed by ASPEN PLUS. Mass and energy balances of each unit, as well as operating conditions and model of CDU will be obtained.

The Crude Distillation Unit of the Kaduna Refinery and Petrochemical Company is a crude oil processing facility consisting of a pre-fractionation train used to heat the crude liquids, and an atmospheric crude column to fractionate the crude into its straight run products. Preheated crude (from a preheat train) is fed to the pre-flash drum, where vapours are separated from the crude liquids. The liquids are then heated to 650°F in the crude furnace. The pre-flash vapours bypass the furnace and are re-combined, using a mixer, with the hot crude stream. The combined stream is then fed to the atmospheric crude column for separation. The crude column is a refluxed absorber, equipped with three pump-around and three side stripper operations [6].

The main column consists of 29 trays plus a partial condenser. The tower feed enters on stage 28, while superheated steam is fed to the bottom stage. In addition, the trim duty is represented by an energy stream feeding onto stage 28. The Naphtha product, as well as the water stream Waste water, is produced from the three-phase condenser. Crude atmospheric Residue is yielded from the bottom of the tower. Each of the three-stage side strippers yields a straight run product. Kerosene is produced from the reboiled Kerosene side stripper, while Diesel and AGO (Atmospheric Gas Oil) are produced from the steam-stripped Diesel and AGO side strippers, respectively.

Mathematical Model Optimization

Optimization of Modeled Base Case in the Optimizer

For the purpose of this task, the following will be the approaches to the optimization problem. The optimization technique employed was the **Sequential Quadratic Programming** (SQP) method. This method handles inequalities and equality constraint. SQP is considered by many to be the most efficient method for maximizing general linear and nonlinear constraint, provided a reasonable initial point is used and the number of primary variables are small [7].

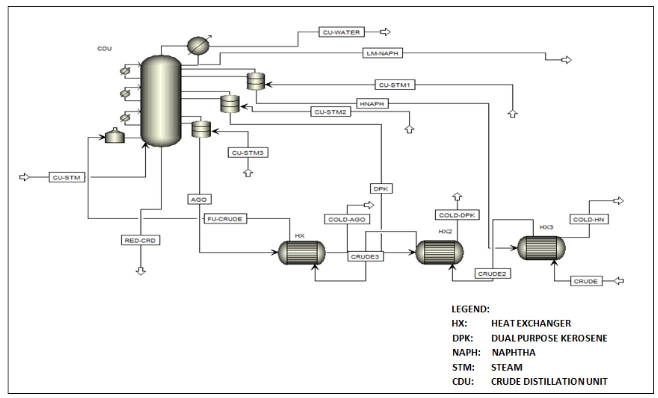

Figure 1: Simplified Diagram of the Crude Distillation Unit.

OBJECTIVE FUNCTION

The cost of operation per annum is assumed to be directly proportional to Q, because the maintenance and cooling costs are relatively small and the capital costs per annum are already fixed. Consequently, the objective function is relatively simple, hence;

The proposed optimization problem was the maximization of the Net Profit:

$$Net\ Profit\ \left(\frac{N}{year}\right) = \sum_{i=1}^{n}\left[\left(\text{Flow Rate (i)} \times \text{Price of Products(i)}\right)\right] - \left(\text{Flow Rate of Gas Oil} \times \text{Price of Gas Oil}\right) - \sum_{j=1}^{n}(\text{Utility Costs})$$

(1)

Where; n = number of products

Flow rate of fuel gas (m^3/day -gas)

Gasoline (m^3/hr),

Heat flow (kJ/hr)

$$\sum(\text{Utility Costs}) = \text{Sum of the costs related to heaters, coolers.}$$

The above equation (1) takes into account the product recovery values (product flow rate multiplied by the price of the products) and the operating costs. All these will be put into account in the optimizer spreadsheet.

PRIMARY VARIABLES

These are variable in the process flow diagram (PFD) that is chosen to be adjusted to maximize net profit in the CDU (objective function). These variables are imported from the PFD. The lower and upper bounds for all of the primary variables set, and used to set the search range, as well as normalization.

These primary variables which are adjusted include:

Table 1: The adjusted (primary) Variables

Object	Variable Description	Current Value
Bottom Oil Main Fractionator	Molar Flow (m^3/yr) Reflux Ratio	2190000 6.76

CONSTRAINT FUNCTION

These are inequality functions that were defined in the spreadsheet. The objective function is set to a constraint. The constraint set on the primary variable is Reid vapor pressure (RVP) of 96.4kPa. This is a Cold property of product, which determines the volatility of the gasoline (C5+) Current value for RVP (kPa) < RVP (kPa) specification. 195.5 kPa < 96.4 kPa

The model will be subject to semi-rigorous process (SQP), restrictions such product specification, operational variable boundaries, limits of other equipments and the energy of the pre-heating crude trains (Heat Exchangers).

The inequality constraints: Various kinds of inequality constraints exist, such as requiring that all of the $x_{i.k}$, $y_{i,k}$, Q_k, F_k, W_k and so on be positive, that upper and lower bounds be imposed on some of the product stream concentrations, and specification of the minimum recovery factors. These are represented in equation 2 to 9.

The Objective function of equation 1 will be subjected to mass balance on the tower (equation 2) and its upper and lower limits of feed flow to the tower (equation 3), to boundaries of operation and quality variables (equation 4), to restrictions of products and pump around flow according to pump capacity (equation 5 and 6), to the maximum operation limits of duties of the atmospheric furnace (equation 7), to the maximum limits of the fluid temperature (h) entering and leaving the exchanger (equations 8 and 9).

$F_t = \sum \mathrm{F_{s,t}}$ (Mass Balances) (2)

$\mathrm{F_t^L} \leq \mathrm{F_t} \leq \mathrm{F_t^U}$ (Limits of Feed flows) (3)

$\mathrm{V_{Op_t}^L} \leq \mathrm{V_{Opt}} \leq \mathrm{V_{Op_t}^U}$ (Boundaries of Operations) (4)

$\mathrm{F_{st}^L} \leq \mathrm{F_{st}} \leq \mathrm{F_{st}^U}$ (Product flows) (5)

$\mathrm{F_{pa}^L} \leq \mathrm{F_{pa}} \leq \mathrm{F_{pa}^U}$ (Pump Capacities) (6)

$\mathrm{Q_{f.t}^L} \leq \mathrm{Q_{f.t}} \leq \mathrm{Q_{f.t}^U}$ (Duties of Atmospheric Furnace) (7)

$T_{hx.in} \leq T_{hx.in}^U$ (Fluid temperature entering heat Exchangers) (8)

$T_{hx.out} \leq T_{hx.out}^U$ (Fluid temperature leaving heat Exchangers) (9)

The L and U superscripts indicates the lower and upper level that can take the flow or involved variables. The Operation Variables (V_{Opt}) of the Atmospheric tower t, are independent variables and their amount depends on tower design.

The equality constraints: The process model comprises the equality constraints. For a Crude Distillation Unit, we have the following typical relations representing the expansion of equations 2 to 9 expressed in terms of the component x and y for the liquid and vapour phase respectively:

F_k Flow of feed into stage k, moles
H_k liquid enthalpy (a function of P_k, T_k and x_k) on stage k
H_k vapour enthalpy (a function of P_k, T_k and y_k) on stage k
k stage index number, k = 1,…,n
L_k flow of liquid from stage k, moles
m number of components, i = 1,…,m
P_k pressure on stage k
Q_k heat transfer flow to stage k (positive when into stage)
T_k temperature on stage k
V_k flow of vapour from stage k, moles
W_k withdrawal stream from stage k, moles
$x_{i,k}$ mole fraction of component i on stage k in the liquid phase
$y_{i,k}$ mole fraction of component i on stage k in the vapour phase

The Total material balances (one for each stage, k)

$$F_k^L + F_k^V + V_{k-1} + L_{k+1} = V_k + L_k + W_k^V + W_k^L \qquad (10)$$

The Component material balances (one for each component i for each stage k)

$$x_{i,k}^F F_k^L + y_{i,k}^F F_k^V + y_{i,k-1} V_{k-1} + x_{i,k+1} L_{k+1} = y_{i,k} V_k + x_{i,k} L_k + y_{i,k} W_k^V + x_{i,k} W_k^L \qquad (11)$$

The Energy Balance (one for each stage)

$$Q_k + h_k^F F_k + H_{k-1} V_{k-1} + h_{k+1} L_{k+1} = H_k V_k + h_k L_k + H_k W_k^V + h_k W_k^L \qquad (12)$$

The Equilibrium relations for liquid and vapour at each stage (one for each stage)

$$y_{i,k} = K_{i,k} x_{i,k} \qquad (13)$$

The relation between equilibrium constant and p, T, x, y (one for each stage)

$$K_{i,k} = K_i(p_k, T_k, x_k, y_k) \qquad (14)$$

Relation between enthalpies and p, T, x, y (one for each stage)

$$h_k = h(p_k, T_k, x_k) \qquad (15)$$

$$H_k = H(p_k, T_k, y_k) \qquad (16)$$

EXPECTED RESULTS

The results to be obtained should include:-

Variation of Bottom Oil Flow Rate on Net Profit (Calculated in N/year on a molar flow rate (m^3/yr)). When Net Profit is plotted against molar flow rate, an optimum point should be obtained.

Variation of Reflux Ratio on Net profit (Calculated in N/year on a Reflux Ratio). When Net Profit is plotted against Reflux

Ratio, an increase in Reflux ratio will be observed, this is attributed to increase in distillate flow rate (Gasoline and Reflux flow Rate).

CONCLUSION

This task carried out considered the modeling and simulating of CDU based on designed flow sheet, and then playing around with operational conditions to bring about optimization through the application of Modular approach, such as Sequential Quadratic Programming. Optimizing CDU using ASPEN PLUS will provide the detailed flow information that is related to the performance of the system. The outcome should indicate and suggest whether the company is operating efficiently and model can provide an effective planning and operations tool. Hence the base case and optimized case should be evaluated.

REFERENCES

1. Basak, K. and Abhilash, K. S. (2002). "On-Line Optimization of a Crude Distillation Unit with Constraints on Product Properties". Ind. Eng. Chem., 41: 1557-1568
2. Pinto, J.M., Joly, M. and Moro, LFL. (2000). "Planning and Scheduling Models for Refinery Operations". Comput. Chem. Eng., 24: 2259-2276
3. Abubakar I. J (2005) Student Industrial Work Scheme (SIWES). "A Technical Report at NNPC"
4. Elamurugan P; Dinesh Kumar D; (2010). "Modeling and Control of Fluid Catalytic Cracking Unit in Petroleum Refinery"; Proceeding of the International Joint Journal Conference on Engineering and Technology (IJJCET)
5. www1: "Aspen HYSYS operation Guide", 2008, User's Manual, Aspen Technology, Inc. www./aspentech.com/Accessed on [December, 2014]
6. Jibril M. (2008). "Kaduna Refinery and Petrochemicals Company's (KRPC) Crude Distillation Unit I (CDU I) Optimization Using HYSYS software".
7. Himmeblau, D.M. and Thomas, E. (2005) Process Optimization and Simulation. Prentice Hall, pp. 63-90, New Jersey. U.S.A.

Citation: Tsunatu D. Yavini etal. (2015). Development of Optimal Strategy for the Design and Operation of a Crude Petroleum Distilation (Topping) Unit. J. of Bioprocessing and Chemical Engineering. V3I1. DOI: 10.15297/JBCE.V3I1.01