EXERGETIC ANALYSIS OF ATMOSPHERIC DISTILLATION PLANT: A CASE STUDY OF WRPC REFINERY

D.A. Fadare[a], A.O. Oni[a] and M.A. Waheed[b]
[a]Mechanical Engineering Department, University of Ibadan, P. M.B. 1, Ibadan, Oyo State, Nigeria
[b]Mechanical Engineering Department, University of Agriculture, P. M.B. 2240, Abeokuta, Ogun State, Nigeria
*E-mail of Corresponding Author: fadareda@yahoo.com

ABSTRACT

Exergetic analysis of atmospheric distillation plant of Warri Refinery and Petrochemical Company (WRPC) was conducted to evaluate exergy efficiencies and irreversibilities in each unit of the system with aim of identifying potential areas for improvement. The process simulation was carried out using commercial simulator HYSYS® 2003. The results of the simulation compared reasonably well with actual plant process parameters with a relative error of 7.22%. The highest irreversibility occurred in the main fractionator followed by TPA, IPA, kerosene stripper, BPA, LGO stripper and HGO strippers with their respective values of 80.17, 5.05, 5.01, 0.77, 0.76, 0.42 and 0.12 GJ/h. of processed crude. The main cause of thermodynamic irreversibility in the system was due to uncontrolled mixing of process streams without due consideration of their potential to produce work. Base on the assessment of the system, the stage by stage exergy profile generates ideas about how the improvement can be made in order to reduce irreversibilities in the components. After modification, the exergy efficiencies show an increase of 21.8 % for the fractionator, 19.5% in KERO stripper, 15.3% in the HGO stripper and 10.3% in the LGO stripper.

Keywords: Energy, Exergy, Atmospheric distillation unit, Irreversibility, Refinery operations

1. INTRODUCTION

Crude distillation process is an essential plant component of refinery operation. The Atmospheric Distillation Unit (ADU) is one of commonly used distillation plant in conventional crude distillation. Among others refinery operations, the ADU requires large amount of energy for separation of crude into different specification of products for direct use or for use in other part of the refining systems. Research works from different parts of the world have shown that enormous amount of energy is lost during the process of separation in the ADU. The complex configuration of ADU i.e. multiple products, reboilers, side-strippers and pump-arounds makes the task of improving energy efficiency a great challenge [1]. In addition, increasingly stringent regulations over environmental impact, complex nature of crude (light or heavy crude) and meeting demanded product specifications have created conflicting challenges for plant operators. It is very important for refiner to respond quickly and efficiently to these challenges. Therefore, there is need for to employ a thermodynamic method that deviate toward reversibility behavior of the process situation while meeting the objectives of designing and operating efficient, safer and profitable process plants. The exergy method provides these opportunities by aiming at the highest possible technical efficiency under the prevailing technical, economic and legal conditions, but also with regard to ethical, ecological and social consequences [1]. The exergy method makes this work a great deal easier. Thus, exergy method offers a unique insight where losses and possible improvements can be determined, suggesting a method to better meet environmental conditions and a sustainable development [2].

The need to improve the performance of crude oil refining systems by aiming at the highest possible efficiencies has being the focus of many researchers [1, 3 and 4]. Until recently all of these efforts are based on first law of thermodynamics analysis which state that energy is always conserved irrespective of the process and its conditions. The adequacy of this concept was challenged by the use of exergy method based on the second law of thermodynamics. Unlike energy, exergy is not always conserved but destroyed.

The goal of exergy methods is to reveal areas of imperfection by pinpointing the magnitude and locations of energy losses in an energy system. Many of the prominent thermodymicists has pointed the use of the exergy concept based on second law of thermodynamics. The use of exergy methods to analyze industrial plant results in accurate evaluation of the available energy dissipation hence more meaningful information as compared to those obtained from the first law. This has lead to increased interest in application of exergy method in process chemical industries, many of which dealt with energy saving [5, 6], plant improvement performance [2], modifications [7, 8 and 9] of various systems. One of the reasons for this is that a great number of processes have been successfully improved and their application to commercial scale is well established [1]. However, the application of exergy concept for the modification of ADU for better performance has not been receiving the attention it deserves. Although many studies have been undertaken to conduct energy analyses of various thermodynamic systems and processes in petroleum and petrochemical industries, very limited work has been done on the exergy analysis of ADU. Cornelissen [10] has conducted exergy analysis of crude distillation units and reported overall rational efficiency of 0.052. The atmospheric distillation was identified as the second main cause of irreversibility in the system. Muslim *et al* [11] has reported that total irreversibility losses in crude refining operations are 608 MW for a flow rate of 507 kg/s.

The highest irreversibility losses occurred in the atmospheric distillation unit. Of the losses, 6.2% was due to chemical exergy losses associated with the separation process itself. The rest of the losses were due to the physical exergy losses mainly because of the temperature difference. Rivero *et al* [12, 13] showed that most important factor affecting the transformation, operation and production costs of the products is the cost of the crude oil raw material; utilities, salaries, maintenance and even capital investment costs are less important. Anaya *et al* [14] has reported the exergy analysis of a refinery in Mexico. The study revealed that the atmospheric and vacuum distillation units are the main causes of irreversibility of the system. Most available works are focused on energetic analysis of conventional crude oil plant towards increasing product yield and energy savings using the first law concept. Ji [15] conducted detailed analysis on crude distillation plant using a rigorous targeting procedure that utilizes the heat demanded supply diagrams along with commercial software simulation. The study was based on energy audit in term of fuel consumption ratio to crude oil processed which was about 2%. Okeke *et al* [16] used in-built sequential quadratic programming (SQP) in HYSYS® for energy optimization for production of gasoline in the four refineries in Nigeria. The result yielded 8% increase in the production of gasoline. However, the approach adopted (first law) could be misleading, its gives no valuable information about the true location, magnitude and sources of inefficiencies as regards each components of the entire refinery system.

The understanding and practical application of exergy analysis has revealed that the energetic analysis alone does not give the true efficiency of the system. As such, a holistic study is essential, which incorporates energetic and exergetic analysis of the entire crude distillation system is required. The objectives of this present study are to: evaluate exergy efficiencies and irreversibilities in atmospheric distillation unit. The essence of study is to enable plant operators to maintain efficient, safer and profitable operations while meeting the stringent demand specifications and operating conditions.

2. METHODOLOGY

Exergetic analysis was conducted on the Atmospheric Distillation Unit (ADU) of Warri Refinery and Petrochemical Company (WRPC), Nigeria. A schematic diagram of the plant with its various components considered in this study is shown in Figure 1.

2.1 Process description

The main column (Atmospheric tower) is equipped with 46 trays, thermal profile is kept by three pump-around and three side cuts are withdrawn. The crude feed is heated in a series of heat exchanger and gas-fired furnace to approximately 350°C and subsequently fed into the fractionating or distillation tower in a packed tray called flash zone from the sixth tray where it comes in contact with the stripping vapors from the bottom stripping section and the liquid reflux (overflash) from the tray above. The crude oil vaporizes more as it enters and rises up. As the vapour rises, it cools as it passes through the tray and comes in contact with liquid coming down the column. Crude fractions settle in trays in the rectifying section and are drawn off at five side cut depending on their average boiling point. In the tower, the different products are separated based on their boiling points.

The boiling point is a good measure for the molecule weight (or length of the carbon chain) of the different products. The light products, which have low boiling points, tend toward the top and the heavier products, with relatively higher boiling points tend toward the bottom. The side cuts are, from heavy to light: heavy diesel, medium diesel, light diesel and kerosene. The very light products, e.g., butane and lighter in addition to light naphtha, exit as vapor at the top of the column. The three side products are withdrawn from the main column, sent to the stripping columns and pumped to storage as follow: Heavy gas oil (HGO) is withdrawn from tray 11 and flows to stripper 10-C-02. In 10-C-02 stripping steam is injected to remove light ends. Light ends and steam from top of 10-C-02 return to main column above tray 12. HGO from 10-C-02 bottom is cooled and sent to the storage. Light gas oil (LGO) is withdrawn from tray 26 and flows to stripper 10-C-03. In 10-C-03 stripping steam is injected to remove light ends. Light ends and steam from top of 10-C-03 return to main column above tray 27. HGO from 10-C-03 bottom is cooled then to storage. Kerosene is withdrawn from tray 35 and flows to stripper 10-C-04. In 10-C-04 stripping steam is injected to remove light ends. Light ends and steam from top of 10-C-04 return to main column above tray 12. Kerosene from 10-C-04 bottom is cooled and sent to the storage. The product (Naphtha) from the overhead column of 10-C-01 is preheated in heat exchanger and fed splitter distillation column where light naphtha is separated from heavy naphtha. The three side strippers are characterized with five (5) trays along it column. The ADU has three pump-around circuits (Bottom BPA, Intermediate IPA, and Top TPA).The pump-around circuit draws liquid from a certain tray, cools the liquid in heat exchangers and returns the liquid to a specified tray

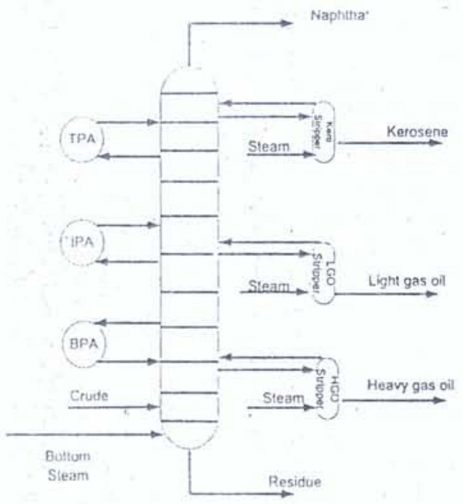

Fig.1 Atmospheric crude distillation unit

above the original tray at a lower temperature. The pump-around circuit is characterized by the withdraw tray, the return tray, the mass flow rate and the return temperature.

2.2 Process Simulation

The process simulation was developed using Aspen HYSYS®. The simulation process of the crude distillation systems was performed in order to reproduce the process used at WRPC to find the exact side cuts for distillation columns, temperature, pressure, enthalpy and entropy for all streams considered. The program has the flexibility to model refinery processes in detail. Its modeling capabilities address a wide range of applications from crude oil characterization and preheating to complex reaction and separation units in both steady state and dynamic mode. For this study, the steady state mode is recommended [4, 10]. It is also important to emphasize that the validation was done by comparing the operating conditions obtained by simulator with industrial data and their relative error R_e evaluated using equation 1.

$$R_e = \left(\frac{Industrial - Simulated}{Industrial}\right) \qquad (1)$$

The production process of crude oil from raw crude was simulated using Aspen HYSYS 3.2. Typically the modeling and simulation process takes the following steps: defining input units, defining the simulation basis, characterizing the crude oil (assay), installing unit operation and running the simulation program. For this study, each of the above steps is briefly described below.

2.2.1 Define input Unit

For ease and simplicity, SI unit was defined for all variables to be calculated in the flow sheet. The flexibility of the software also includes customized units set at any stage within the HYSYS flow sheet environment.

2.2.2 Definition of simulation basis

The first step is the selection of lighter components and the appropriate thermodynamic method. The thermodynamic fluid package selected is Peng Robinson, equation of state which is recommended for the petroleum components. Since the exact composition of the crude is unknown and is defined in terms of distillation temperatures the feed developed is a combination of pure library components (lighter components) and pseudo components. The lighter components, methane, propane, i-butane, n- butane, i-pentane, n-pentane and hexane are added to the pure component library.

2.2.3 Characterization of the crude oil

The third step in the simulation is to characterizing the crude oil from the experimental laboratory data. A complete definitive analysis of a crude oil is called crude assay. The assay (characterized crude oil) contains all the crude oil laboratory data, boiling curves, light ends, property curves and bulk property. The petroleum characterization method in the simulation software is capable of converting laboratory analysis of crude oils, condensates and petroleum cuts into series of discrete hypothetical components.

The data from the crude assay is used to define the petroleum pseudo-components. The pseudo components are the theoretical components that are not readily available in the component library and have to be defined. The data from the pure component library are used to represent the defined light components in the crude oil. It is required to input the laboratory distillation curve (TBP or ASTM data) and any bulk property such as Molecular Weight, Density, or Watson K Factor. It should be noted that the more the information is provided to the simulation, the accuracy of the property prediction is improved. In this study, the light end composition, TBP distillation curve, density and viscosity are used in characterizing the oil. Each crude type is characterized separately and finally the required crude oil blend is defined and installed into the flow sheet. The calculated TBP data by HYSYS for the given crude is compared to the input data to identify any inaccuracies.

2.2.4 Installing Unit Operations

The commonly used unit operations which are installed in this case are column, heat exchangers, coolers, heaters and side strippers. For each unit operation it is required to specify certain parameters to satisfy the number of degrees of freedom. Each parameter specification will reduce the degrees of freedom by one. The number of active specifications must equal the number of unknown variables to solve. The detailed modeling procedure of each section in the unit is described in Aspen HYSYS® operations guide.

2.2.5 Running the simulation program

Once the required operating parameters and thermodynamic-related properties have been set, the simulation can proceed when the initial conditions of each process stream given. In running the simulation it is of great importance to ensure that proper initial values be used for each stream as failure in doing so may lead to convergence to different values, which is not desirable due to the non-linearity and unstable characteristics of the process. Once the initial conditions have been specified, iterative calculations are automatically performed until all the values in the calculated streams match those in the assumed stream within some specified tolerances. After creating the process flow diagram (PFD) completely, the simulation is run for results to be analyzed.

2.3 Exergy analysis

Exergy is more precisely based on the application of the first and second laws of thermodynamics and is a measure of energy quality. Szargut [17] stated that the exergy of a material is the work or electrical energy necessary to produce that material in its specified state from materials common in the environment in a reversible way, heat being exchanged only with the environment. At full equilibrium between a system and an environment, the exergy of that system is equal to zero. Such a system state is called zero or dead state.

The exergy of a stream material of consists of the physical exergy and chemical exergy terms.

$$\varepsilon_t = \varepsilon_{ph} + \varepsilon_{ch} \qquad (2)$$

The physical exergy is the work obtainable by taking the substance through reversible processes from its initial state temperature T and pressure P, to the state determined by the temperature T_0 and the pressure P_0 of the environment. It can be calculated (Cornelissen, 1997) using equation (2) with:

$$\varepsilon_{ph} = m[h - h_0 - T_0(s - s_0)] \qquad (3)$$

where h is the enthalpy and s the entropy.

Chemical exergy is equal to the maximum amount of work obtainable when the substance under consideration is brought from the environmental state [10], defined by the parameters T_0 and P_0, to the reference state by processes involving heat transfer and exchange of substances only with the environment.

The chemical exergy for mixtures can be calculated as follows:

$$e_{ch,mix} = \sum_i x_i e_{ch,i} + RT_0 \sum_i x_i \ln \gamma_i x_i \qquad (4)$$

where x_i is the mol fraction of the i-th component, R is the molar gas constant and γ_i is the activity coefficient. For ideal solutions the activity coefficient is equal to one.

The irreversibility rate, also called exergy destruction rate or exergy loss rate, is calculated by setting up the exergy balance and taken the difference between all incoming and outgoing exergy flows or in formula form

$$I = \varepsilon_{loss} = \sum_{in} \varepsilon_i - \sum_{out} \varepsilon_j \sum \left(1 - \frac{T_o}{T_i}\right) Q_i - \dot{W}_{cv} = T_o \dot{S}_{gen} \qquad (5)$$

Another way of calculating the irreversibility can be done by the Gouy-Stodola formula,
in which the entropy increase is multiplied by the environmental temperature, in formula form

$$\dot{I} = T_o.(\sum_{out} \dot{S}_j - \sum_{in} \dot{S}_i - \sum \frac{Q_r}{T_r}) = T_o \dot{S}_{gen} \qquad (6)$$

The exergy efficiency is express by equation

$$\psi = \frac{\sum \varepsilon_{out}}{\sum \varepsilon_{in}} \qquad (7)$$

3. APPLICATION OF THE EXERGY METHOD

The mathematical models are applied to each component and the overall system to find heat added irreversibility rate and exergy efficiency. The following assumptions were made: all systems are operating at steady state steady flow conditions; variations of potential and kinetic energies in all equipments are neglected; all equipments operate adiabatically and reference state conditions are at $T_0 = 25^o C (298.15^o K)$ and $P_0 = 101 kPa$. With these assumptions potential and kinetic energy term are null, remaining physical and chemical exergy. The contribution of chemical exergy in the exergy analysis of crude oil distillation systems was found to be 6.2% [4]. Since the major object of discussion is the irreversibility of the system studied, a simplifying assumption was made that the chemical exergy term in the distillation process studied in this work would not be appreciable and the chemical exergy term was, therefore, not considered.

3.1 Exergy balance for the system

Based on mentioned procedures, exergetic efficiency and irreversibility rate are obtained for each component of the atmospheric distillation unit according to stream wise approach.

Pump-around (PA)

The PA's are essentially heat exchangers designed to remove heat from the distillation column. The exergy balance for PA's is expressed by the equation given below.

The energy and exergy balances of the PA are given by [1]

$$\dot{Q}_{PA} = \dot{m}(h_{in} - h_{out}) \qquad (8)$$

where Q_{PA} is the heat rejected to the cooling fluid.

The irreversibility in the PA's can be written as:

$$\dot{I}_{PA} = \dot{m}(\varepsilon_{in} - \varepsilon_{out}) + \dot{\varepsilon}_{QPA} \quad (9)$$

The exergy efficiency is given by

$$\psi_{PA} = \frac{\dot{\varepsilon}_{out}}{\dot{\varepsilon}_{in} + \dot{\varepsilon}_{QPA}} \quad (10)$$

where $\varepsilon_{QPA} = \left(1 - \frac{T_o}{T}\right) Q_{PA}$ (11)

Columns

The three side strippers are smaller distillation column attached to the main distillation column (ADU) to ensure sharpness of the product. For this purpose they are also treated as columns. The exergy balance is given as [18]

$$\sum_{into\,system} \left(\dot{n}\varepsilon + \dot{Q}\left(1 - \frac{T_o}{T_s}\right) + W_s \right) - \sum_{out\,of\,system} \left(\dot{n}\varepsilon + \dot{Q}\left(1 - \frac{T_o}{T_s}\right) + W_s \right) = I_{irr} \quad (12)$$

where W_s is the shaft work. The total rate of exergy loss represents the overall thermodynamic imperfections, and directly proportional to the rate of entropy production due to irreversibilities in a column operation. As the exergy loss increases, the net heat duty has to increase to enable the column to achieve a required separation. Consequently, smaller exergy loss means less waste heat or thermodynamic imperfections, which include pressure drop, heat and mass transfer due to finite driving forces, and mixing of flows with different compositions temperatures and pressures.

The minimum exergy required for separation is defined as the difference between the exergies of products and feed streams and is given as [18]

$$E_{min} = \sum n\dot{\varepsilon} - \sum n\dot{\varepsilon} \quad (13)$$

The thermodynamic efficiency is given as [18]:

$$\eta = \frac{E_{min}}{Irr + E_{min}} \quad (14)$$

The denominator of Eq. 15 represents the total exergy input into the system.

The potential for improvement is therefore given as the difference between exergy loss or irreversibility and the minimum exergy required for separation [1].

$$\dot{IP}c = I_{irr}(1 - \eta) \quad (15)$$

3.2 Stage exergy loss

The stage by stage exergy losses represent inefficient use of available energy due to irreversibility. The exergy loss on tray can be evaluated with an exergy balance over the tray.

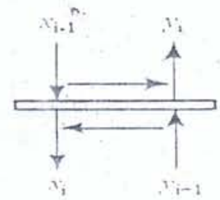

Fig. 2 Exergy balance at a column stage.

The exergy balance applied to the tray represented in fig 2 takes into account of the various streams interacting with it. The irreversibility is thus given as [20]

$$\dot{I}_{irr} = \dot{\varepsilon}_{in}^{V} + \dot{\varepsilon}_{in}^{L} + \sum \dot{\varepsilon}_F - \dot{\varepsilon}_{out}^{V} - \dot{\varepsilon}_{out}^{L} - \sum \dot{\varepsilon}_S + \dot{Q}\left(1 - \frac{T}{T_{heat}}\right) \quad (16)$$

Where ε^V and ε^L are exergies of vapour and liquid respectively.

4. RESULTS AND DISCUSSION

All operational conditions were provided by WRPC. The validation of the simulation was done by comparing the simulated results with the actual operating parameters (Table 1). The relative error was found to be 7.22%. It is important to point out that the ADU is a complex system containing three pump-arounds, three side strippers, and the main fractionators that are inter-connected with streams flowing in and out to ensure that quality products. These are therefore potential areas for high exergy destruction due to momentum loss, thermal loss (temperature driving force/mixing) and chemical potential loss (mass transfer driving force/mixing) that occur within the system [21]. The exergy analysis conducted gave insight to the inefficiencies and the opportunities for exergy loss minimization of each of the unit operation found in ADU. The results obtained for the each of the operation and the overall unit was evaluated and discussed below.

A complete exergy analysis of the overall unit operation has been conducted, using the general methodology presented in section 3.1. The exergy efficiencies of each unit are shown in Figure 2. It was observed that the BPA has the highest exergy efficiency of 99%. The high exergy efficiency was due to relatively high flow rate and low temperature difference between the hot and cold stream passing through the BPA respectively. Although the cold streams are not modeled here, it is represented with the cooling duty. The relatively low temperature difference is justified by the bottom temperature that is needed to be maintained constantly at a relatively high temperature. It is important to also note that aside removing heat from the distillation column the cooling effect of the BPA improves the vapour-liquid interaction that enhance the of good quality of the heavy gas oil products.

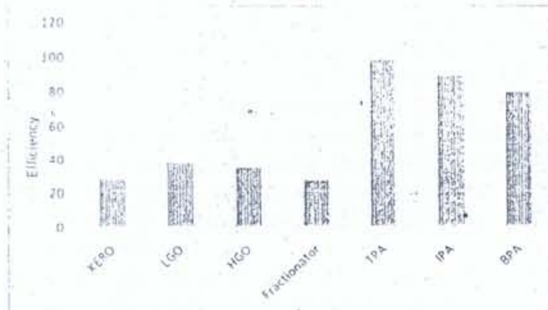

Figure 3 Exergy efficiencies of the components in the crude oil plant

The relatively high temperature difference in the IPA accounts for the relatively low exergy efficiency of the system compared with the BPA. Of the three pump-arounds, the TPA has much load, it maintains the column top temperature aside improving the

Table 1: Comparison between the actual and simulated process parameters

Process parameters	Base Case	Simulated Case
Number of Stages	46	46
Column Top Pressure KPa	58.84	58.84
Column Bottom Pressure KPa	147.1	147.1
Column Top temperature ^{0}C	135	135
Column Bottom Temperature ^{0}C	340	341.7
Crude Feed flow rate Kg/h	612178	612178
Crude temperature ^{0}C	350	350
Bottom Steam flow rate Kg/h	6120	6120
Bottom Steam Temperature ^{0}C	350	350
HGO Steam flow rate m^3/h	180	180
LGO Steam flow rate m^3/h	720	720
Kero Steam flow rate m^3/h	1152	1152
BPA flow rate m^3/h	400	400
BPA Draw Temperature ^{0}C	304	309.8
BPA Return Temperature ^{0}C	274	274
IPA flow rate m^3/h	400	400
IPA Draw temperature ^{0}C	208	210
IPA Return Temperature ^{0}C	163	163
TPA flow rate m^3/h	600	600
TPA draw temperature ^{0}C	160	159
TPA Return Temperature ^{0}C	92	92
HGO Product Temperature	310	320
HGO flow rate m^3/h	25	25
LGO Product Temperature ^{0}C	280	280
LGO Flow rate m^3/h	90	90
Kero Product Temperature ^{0}C	190	190
kero flow rate m^3/h	143	143

vapor-liquid interactions and removing heat from the column. The lowest exergy efficiency of 28.2% occurred in the fractionating unit. The low exergy efficiency in this unit is due the fact that an uncontrolled mixing of the streams is done, without considering their potential to produce work, since they have different temperatures and pressures, resulting in a great exergy destroyed. Of the three side strippers the lowest exergy efficiency was generated in the kerosene side-stripper followed by LGO side-stripper and HGO side-stripper. The corresponding exergy efficiency for each unit is 29.5, 38.7 and 35.7%, respectively. The losses in the side strippers are due to irreversibilities within the system as a result of mixing and high temperature difference between the inlet and the outlet streams. The result also revealed that for all the four columns present in the system the heating processes were inefficient. This is always the case for exergy calculations and is due to the fact that the exergy value of heat is often much lower than its energy value, particularly at temperatures close to reference temperature.

Figure 4 Irreversibilities of the component in the crude oil plant

Figure 4 shows the level of exergy destroyed in each unit operations of the ADU. On a clear note the main fractionator have the largest irreversibilities followed by TPA, IPA, kerosene stripper, BPA, LGO stripper and HGO strippers with their respective values of 80.17, 5.05, 5.01, 0.77, 0.76, 0.42 and 0.12 GJ/h. Figure 3 and 4 showed that the pump around units has high exergy efficiency and relatively high thermodynamic irreversibility compared to the three side strippers. Also, the three side strippers has low exergy efficiency and relatively low thermodynamic irreversibility According to Anozie at al [19] this situation arise because exergy efficiency values are quantitative measurement derived as the ratio of two numbers with the constraint that the ratio is not greater than 1, whereas irriversibilities are quantitative measurement derived as the difference between two numbers. Thus it is possible for the ratio of two large numbers to be high and the difference to be as well, and the ratio of two small numbers to be high and the difference to be small.

After the identification and quantification of irreversibilities and the minimum exergy required for separation, the value of improvement potentials establish the possibilities for the system improvement. The stage exergy loss profile further generates ideas about how the improvement can be made in order to reduce irreversibilities in the components. For the main fractionator the profile, fig 5 shows two sections with very large exergy losses. The section around BPA (Stages

24 and 25) and IPA (stage 32 and 34) respectively. For the three side strippers fig 6 the exergy losses occurred at the top and bottom section of the columns. The large exergy destruction in these sections are caused by a mismatch of temperature and compositional differences occurring due to inefficient mixing. Base on this analysis the following proposals for optimization were made.

1. Decrease the flow rate of BPA and IPA pumparounds.
2. Change return location of BPA from stage 24 to 23
3. Reduce steam flow rate of all side strippers

The simulations were repeated for the proposals ensuring that in each case the stringent operating condition were observed in order to keep the product quality within allowable range i.e.

- pump around return temperatures were kept constant. Only changes in flow were considered.
- Bottom steam and crude feed properties was not changed.
- Product specifications were kept within allowable range.

The exergy analyses were conducted for the proposals. Figure 7 shows the comparison of the base case and the modified case. The decrease in BPA and IPA flow rate shows reduction of irreversibilities in those sections. Irreversibility reduction was also noticed around section 10 to 11 which is as a result of corresponding decrease in steam flow rate of the HGO stripper. Changing the BPA return location further reduces the exergy destruction due heavy mixing in stages 24 and 25. Figures 8-10 reveal the effect of reducing steam flow rate which in turn reduces the exergy loss at the inlet stage of the side strippers (KERO, LGO and HGO). The corresponding effect of these reductions is slightly felt at the top section of the strippers. For a significant reduction of exergy destroyed at top section, irreversibility due to mixing could be reduced by changing stream withdraw location, this was not possible in the present study due to limitation of the software used. The reduction of irreversibility rate in the modified cases generated corresponding increase in exergy efficiency (Figure 11).

The exergy efficiencies show an increase of 21.8 % for the fractionator, 19.5% in KERO stripper, 15.3% in the HGO stripper and 10.3% in the LGO stripper. The exergy loss of the PA need no further modification in that the exergy losses are used for preheating the cold crude oil streams in the refinery.

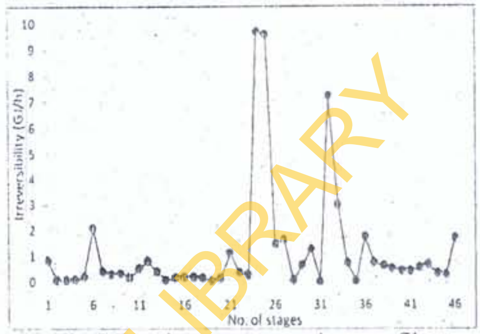

Fig 5 Fractionator exergy loss profile

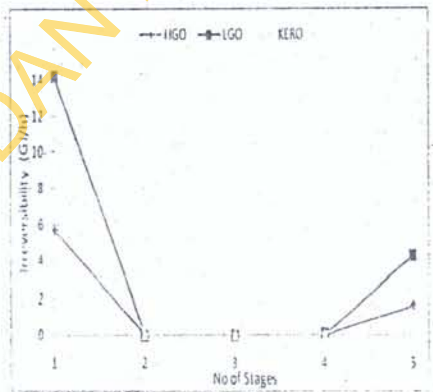

Fig 6 HGO, LGO and KERO exergy loss profile

Table 2. Exergy performance for Individual Columns in the ADU

Components	Minimum Work (KJ/h)	Irreversibility (KJ/h)	Improvement potential
Kerosene Stripper	324598.8	775975.5437	547112.5599
LGO Stripper	264505.7	418054.763	256050.2558
HGO Stripper	66319.8	119432.8901	76791.41452
Main Fractionator	31518782.8	80176506.59	57551864.94

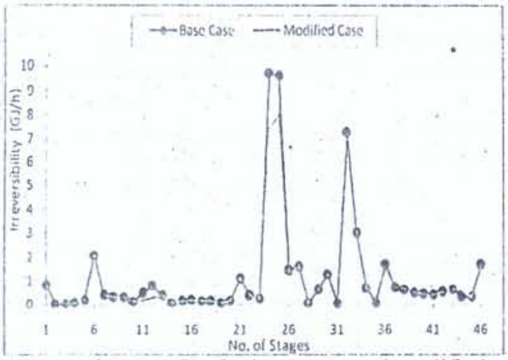

Fig 7 Exergy loss profile for base case and modified case in the Fractionator

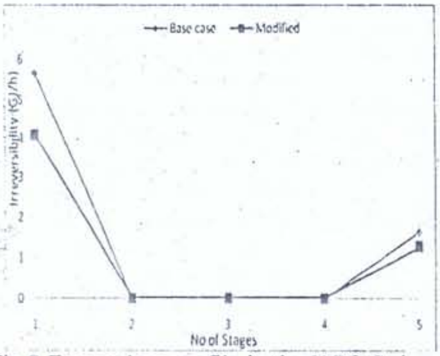

Fig 8 Exergy loss profile in the HGO Stripper

Fig 9 Exergy loss profile in the LGO Stripper

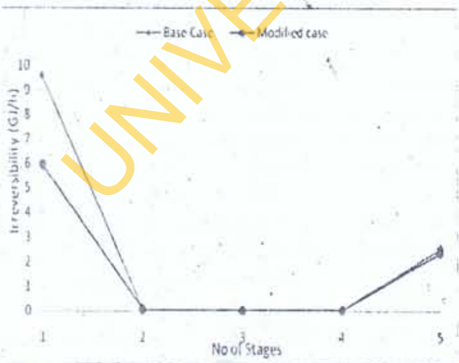

Fig 10 Exergy loss profile in the KERO Stripper

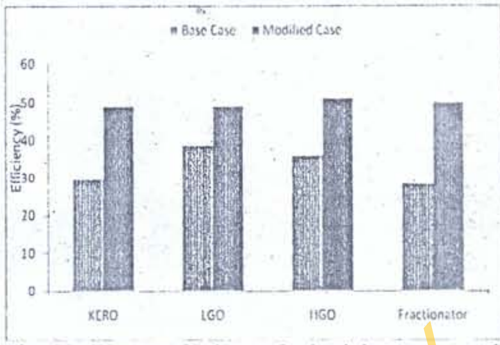

Fig. 11: Exergy efficiency for both base case and modified case.

5. CONCLUSIONS

In this work, exergy method was conducted on the atmospheric distillation unit of WRPC refinery using HYSYS® 2003 process simulator. The main focus was to identify sources of thermodynamic inefficiency. The application of exergy method to stage by stage clearly shows large irreversibilities. It was found that the highest irreversibility occurred in the main fractionator followed by TPA, IPA, kerosene stripper, BPA, LGO stripper and HGO strippers and BPA. Base on the result from exergy loss profile changes according to information gained was applied leading to improved distillation column.

ABBREVIATIONS:

PA:	Pump-around
BPA:	Bottom Pump-around
IPA:	Intermediate Pump-around
TPA:	Top Pump-around
HGO:	Heavy gas oil
LGO:	Light gas oil
KERO	Kerosene
10-CO-01:	Main Fractionator
10-CO-02:	Kerosene stripper
10-CO-03:	Light gas oil stripper
10-CO-04:	Heavy gas oil Stripper

6. REFERENCES

[1] I Dincer, & M Rosen, Exergy: Energy, Environment and Sustainable development. Elsevier, Oxford, UK, 2007.

[2] T. J. Kotas, The Exergy Method of thermal Plant Analysis, Reprint Edition, Krieger, Malabar, FL., 1995.

[3] H Al–Muslim. and I. Dincer, Thermodynamic analysis of crude oil distillation systems, Int. J. Energy Res; 29, 2005, 637–655.

[4] H. Al–Muslim, I. Dincer & S.M. Zubair,. Effect of reference state on exergy efficiencies of one and two stage crude oil distillation plant. International Journal of thermal science, 44, 2005, 65-73.

[5] M Errico, G. Tola & M Mascia, Energy saving in a crude distillation unit by a preflash implementation, Appl. Therm. Eng. doi:10.1016

[6] C. Koroneos, G. Roumbas & N. Moussiopoulos, Exergy analysis of cement production, Int. J. Exergy, 2, 2005, 1. 2005.

[7] M.A., Waheed, S.O., Jekayinfa, J. O, Ojediran. & O. E. Imeokparia, Energetic analysis of fruit juice processing operations in Nigeria, Energy, 33, 2008, 35-45.

[8] H. Dalsgard, Simplification of process integration in medium-size industry, Unpublished Ph.D thesis, Department of Mechanical and Energy Engineering, University of Denmark, Denmark, 2002.

[9] E. Rotstein, Exergy analysis: a diagnosis and heat integration tool. In: Singh RP, editor. Energy in food processing. Elsevier; 1986.

[10] R.L. Cornelissen, Thermodynamics and sustainable development: the use of exergy analysis and the reduction of irreversibility. Ph.D. Thesis, University of Twente, The Netherlands, 1997.

[11] H. Al–Muslim, I. Dincer & S.M. Zubair, Exergy Analysis of Single- and Two-Stage Crude Oil Distillation Units, Journal of Energy Resources Technology, 125, 2003, 199-207.

[12] R. Rivero, R. Consuelo, L. Montroy, The Exergy of Crude Oil Mixtures and Petroleum Fractions: Calculation and Application, Int.J. Applied Thermodynamics, .2, (No.3), 1999, 115-123.

[13] R. Rivero, R. Consuelo, G. Salvador, Exergy and exergoeconomics analysis of a crude oil combined distillation unit. Energy 29, 2004, 1909–1927.

[14] A. Anaya, L.M. Caraveo, V.H. Cacia, A. Mendosa, Energetic optimization of a petroleum refinery applying second law of thermodynamics. Proceedings of the computer-aided energy analysis, ASME AES, 21, 1990, 49-54.

[15] S Ji, Optimal design of crude oil distillation plants. Shungchen Ji. Norman, 2001.

[16] E. O. Okeke & A. A. Osakwe-Akofe, Optimization of a refinery crude distillation unit in the context of total energy requirement, NNPC R&D Division, Port Harcourt, Nigeria, 2003.

[17] J. Szargut Exergy analysis: technical and ecological applications. WIT Press, Southampton, UK, 2005.

[18] N. Nguyen &Y. Demirel, Retrofit of distillation columns in biodiesel production plants, Energy 35 (2010) 1625–1632

[19] A.N. Anozie, O.J. Odejobi and O.P. Ayoola, Relevance of exergy analysis in the assessment of thermal power plant performance. Proceedings of the first national conference of the Faculty of Technology, Obafemi Awolowo University , Ile-Ife, Nigeria, (2009) pp. 193-202.

[20] A Rosjorde., S Kjelstrup. The second law optimal state of diabatic tray distillation column, Chemical Engineering Science, 60, 2005, 1199

[21] D. Mustapha, T. Sabria, O. Fatima. Distillation of a complex mixture. Part II: Performance Analysis of a Distillation Column Using Exergy. Entropy, 9, 2007, 137-151.

BIOGRAPHY OF AUTHORS

Dr. David Abimbola FADARE (*MNSE, Reg COREN, MIMechE*), is currently a lecturer at Mechanical Engineering Department, University of Ibadan, Nigeria. He holds a B.Sc. degree in Mechanical Engineering in 1992, M.Sc. degree in Mechanical Engineering in 1995, and Ph.D. in Mechanical Engineering in 2003 from the University of Ibadan. His research interests include Engineering systems modeling, Renewable energy applications, Digital image analysis, Optimization of metal cutting operations and energy audit of manufacturing process.

Mr. A. O. ONI has his first degree in Chemical Engineering from Ladoke Akintola University of Technology (LAUTECH), Ogbomoso in the year 2002 and M.Sc in Mechanical Engineering from University of Ibadan in 2007. His research interest is in digital image processing, modeling and energy audit of manufacturing process.

Prof. Mufutau Adekojo WAHEED (*MNSE, Reg COREN*), is a lecturer in Department of Mechanical Engineering, University of Agriculture, Abeokuta, Nigeria. He's currently the Deputy Dean of the College Engineering. He has first and second degrees in Mechanical Engineering from University of Ilorin, Ilorin, Nigeria in 1990 and 1995, respectively and Ph.D. from Aachen University of Technology (RWTH) Aachen, Germany in 2001. His research work is in Fluid Dynamics, Flow Stability and Control, Heat and Mass Transfer, Computational Fluid Dynamics, Solar Energy.