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# ENERGY AND EXERGY ANALYSES OF CRUDE OIL DISTILLATION PLANTS

BY

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A Thesis Presented to the  
DEANSHIP OF GRADUATE STUDIES

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DHAHRAN, SAUDI ARABIA

In Partial Fulfillment of the  
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**MASTER OF SCIENCE**

In

**MECHANICAL ENGINEERING**

RAMADHAN 1423 H  
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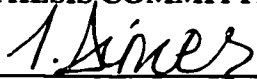
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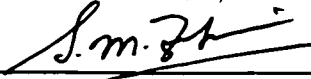
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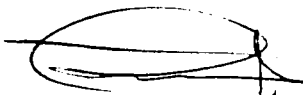
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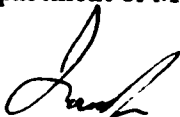
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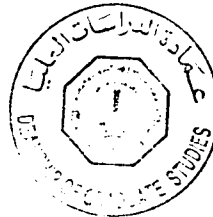
  
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*Dedicated to my wife*

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**“In the name of Allah (God), the Most Gracious, the Most Merciful. My blessing is only from Allah (God).” (The Noble Qur’an)**

**First and above all, all praise to Allah, Who gave us the gift of mind and inspired me with the will to accomplish this work. And second, prayers and peace upon His Prophet, Mohammed. I ask Allah to accept my efforts and to make this work beneficial to Islam and the Muslims.**

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## **THESIS ABSTRACT**

**NAME OF STUDENT: AL-MUSLIM, HUSAIN MUHAMMAD**

**TITLE OF STUDY: ENERGY AND EXERGY ANALYSES OF  
CRUDE OIL DISTILLATION PLANTS**

**MAJOR FIELD: MECHANICAL ENGINEERING**

**DATE OF DEGREE: NOVEMBER 2002**

**This study deals with energy and exergy analyses of crude oil distillation plants. In these plants, high degree of irreversibilities (exergy losses) occur. These losses cannot be identified by the first law of thermodynamics (energy analysis) only as it is a measure of quantity only whereas the second law of thermodynamics (exergy analysis) is a measure of quality as well as quantity. Exergy analysis appears to be a potential tool to determine the losses. In refinery distillation plants, there are many components of interest to be analyzed thermodynamically, e.g. the crude oil heating furnace, the distillation column and a network of heat exchangers. The previous literature works were mainly concerned with the yield percentage of the distillation plant and to some extent the energy efficiency, no detailed work was done using exergy analysis. In this study, crude oil distillation plants are modeled and analyzed thermodynamically with the aid of simulation software (SimSci/PRO II) to determine the energy efficiency, exergy efficiency and irreversibility rate of the individual components and the overall system. Effects of temperature profile and pressure profile of the distillation columns on the efficiencies and irreversibility rates are illustrated. Also, effect of modifying the conventional crude oil distillation plant is studied. Those modifications are pump-back circuits instead of pump-around circuits, adding reboiler to the distillation column and two-stage distillation columns. Results showed that operating conditions are the optimum conditions yielding the highest efficiencies and lowest irreversibility rates. Modifying the conventional model lead to an increase up to 58% for the system overall efficiency and decreases down to 40% system overall irreversibilities.**

**MASTER OF SCIENCE DEGREE  
KING FAHD UNIVERSITY OF PETROLUUM AND MINERALS  
Dhahran, Saudi Arabia  
November 2002**

## خلاصة الرسالة

الاسم : حسين محمد المسلم  
عنوان الرسالة : تحليل الطاقة و تحليل الإكسيري جي لمعامل تقطير الزيت  
التخصص : هندسة ميكانيكية  
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هذه الرسالة تطبق تحليل الطاقة و تحليل الإكسيري جي لمعامل تقطير الزيت. في هذه المعامل، هناك كميات ضائعة كبيرة من الإكسيري جي. هذه الكميات لا يمكن التعرف عليها من خلال تطبيق تحليل الطاقة فقط لأن الطاقة تقيس الكم فقط بينما الإكسيري جي تقيس الكم و الكيف. تحليل الإكسيري جي هو الأداة الفعالة لتحديد هذه الكميات الضائعة. في معمل تقطير الزيت، يوجد العديد من المعدات التي يهمننا تطبيق تحليل الطاقة و تحليل الإكسيري جي عليها مثل المسخنات الحرارية، عوالميد تقطير الزيت و لمبادلات الحرارية. معظم الدراسات السابقة ركزت على تحسين نسبة المشتقات المنتجة من معامل تقطير الزيت و إلى حد ما كفاءة الطاقة. ليس هناك دراسة مفصلة باستخدام التحليل الإكسيري جي. في هذه الرسالة، تحليل الطاقة و تحليل الإكسيري جي سيطبق على نموذج لمعمل تقطير الزيت بمساعدة برنامج محاكاة لتحديد كفاءة الطاقة، كفاءة الإكسيري جي و كميات الإكسيري جي الضائعة لمكونات المعمل منفصلة و للمعمل ككل. سيتم أيضا دراسة تأثير الحرارة و الضغط الخاصين بعوالميد للتقطير. أيضا، سيتم دراسة نماذج معدلة من معامل تقطير الزيت مثل الدوائر المرجعة بدل الدوائر المعقدة، و إضافة معبد تسخين لعوالميد التقطير، و فصل عامود التقطير إلى عامودين. تظهر النتائج أن حالات التشغيل الحالية تعطي النتائج الأفضل. تعديل نموذج معمل التقطير يعطي زيادة في الكفاءة الكلية تصل إلى 58% و تقليل من الكميات الضائعة يصل إلى 40%.

ماجستير في العلوم

جامعة الملك فهد للبترول و المعادن

الظهران - المملكة العربية السعودية

رمضان 1423 هـ (نوفمبر 2002م)

# CHAPTER 1

## INTRODUCTION

### 1.1 REFINERY COMPLEX AND CRUDE OIL DISTILLATION PLANT

Crude oil has low value in its original state and no practical applications. Even after separation of gas, water, H<sub>2</sub>S and other components, crude oil is still a mixture of thousands of hydrocarbons ranging from very light to very heavy components. The objective of a refinery complex is to produce the desired final practical products, e.g. liquefied petroleum gas (LPG), kerosene, diesel, gasoline, jet fuels, asphalt, etc. In order to do this the crude oil has to go through successive chemical and physical processes in different plants. As the refinery user wishes to increase yield of lighter better quality products, more processes have to be added. Figure 1.1 shows a schematic diagram of a modern refinery complex.

The crude oil distillation plant is the first step in a refinery complex. The process here is purely physical to separate the crude oil into different fractions depending on the difference of boiling temperatures. In most distillation plants, the crude oil is processed into two towers. The first is the atmospheric tower where light hydrocarbons are separated, and the second is the vacuum tower where heavier hydrocarbons are separated. The products of the crude oil distillation plant can be either final products or feedstock to other plants for further processing. Figure 1.2 shows a schematic diagram of a crude oil distillation plant. (Watkins, 1973; Ras Tanura Refinery, 2002).

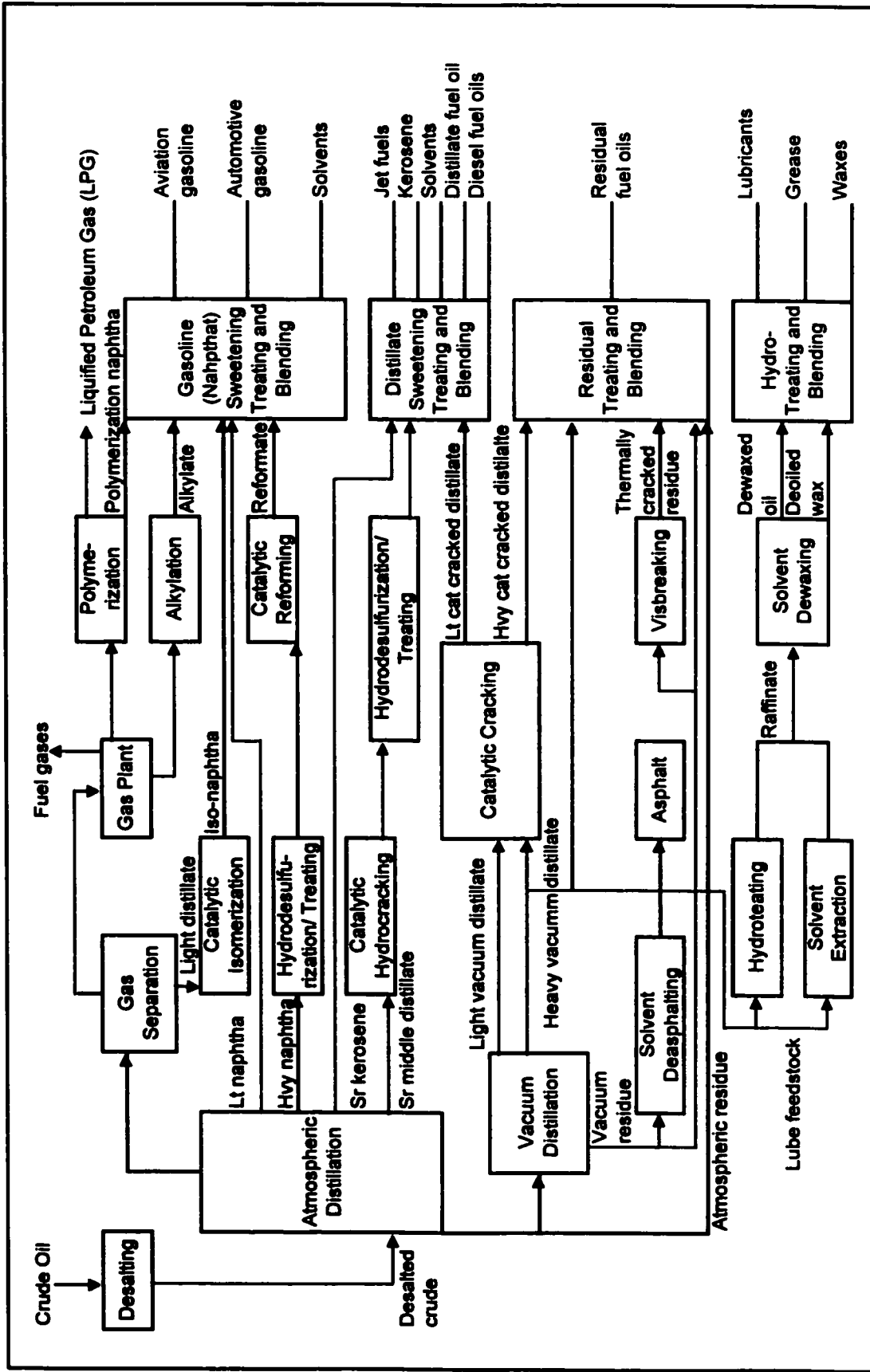
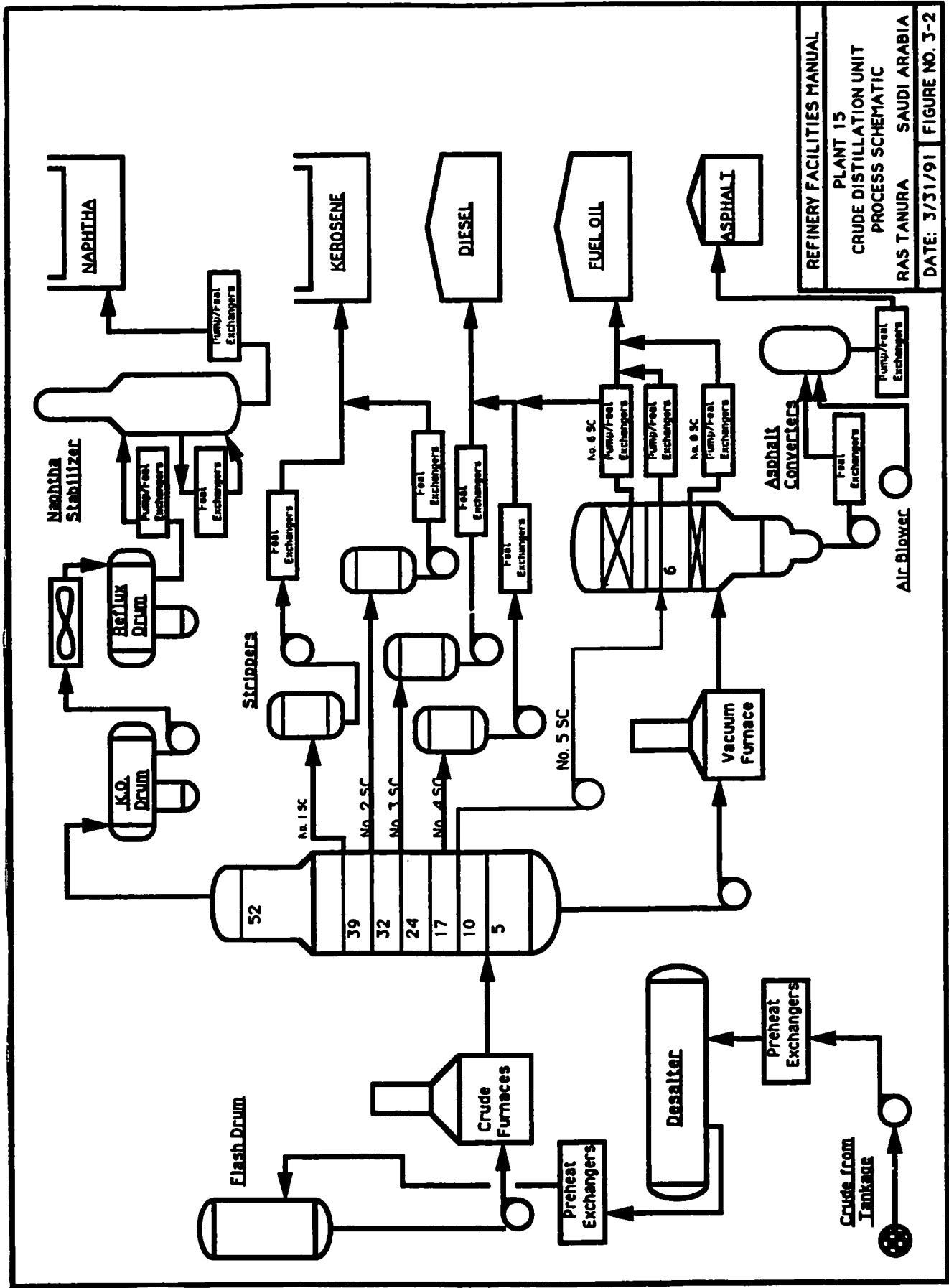


Figure 1.1 Schematic diagram of a refinery complex. (Ras Tanura Refinery, 2002)



REFINERY FACILITIES MANUAL  
 PLANT 15  
 CRUDE DISTILLATION UNIT  
 PROCESS SCHEMATIC  
 RAS TANURA SAUDI ARABIA  
 DATE: 3/31/91 FIGURE NO. 3-2

Figure 1.2 Schematic diagram of a crude oil distillation plant. (Ras Tanura Refinery, 2002)

The crude oil distillation plant has many components of interest to be analyzed thermodynamically, i.e. the crude oil heat furnace, the distillation towers and the heat exchangers network. The primary objective of conventional energy analysis of a crude oil distillation plant is to maximize the yield of the desired products. However, for economical and environmental reasons and the limited resources of energy, we have to maximize utilization of those energy resources. The first law (Energy) and the second law (exergy) analyses are the two basic tools to do so.

## **1.2 THERMODYNAMIC (ENERGY AND EXERGY) ANALYSIS**

We can generally view thermodynamics as the science of energy and entropy. It includes all aspects of energy and energy transformations, including power production, refrigeration, and relationships among the properties of matter. The main concern of thermal engineering is to make better use of available energy resources. (Dincer and Cengel, 2001)

There are two essential tools for the thermodynamic analysis. The first is the energy analysis, referred to as the first law of thermodynamics or the conservation of energy equation. The second is the exergy analysis, referred to as the second law of thermodynamic analysis or the entropy generation equation. These two equations in addition to the mass conservation equation represent the three governing equations in the thermodynamic analysis.

Energy is a scalar quantity, but it is difficult to directly measure. Therefore, interest is always in finding change of energy levels rather than finding absolute energy values. Energy cannot be destructed in a process, i.e. energy is always conserved for real processes and only converted from one form to another. Energy in a system includes internal energy, potential energy and kinetic energy. Energy analysis

is only a measure of quantity, and therefore, it does not give any information about the direction in which processes can spontaneously occur.

Exergy is the maximum shaft power that could be produced by a system when brought to equilibrium with its reference surroundings including temperature, pressure and chemical composition. A system in complete equilibrium with its surroundings does not have exergy and the more the system deviates from the surroundings the higher its exergy. Exergy is a measure of both quality and quantity. Actual processes occur in direction of decreasing exergy. Exergy is only conserved in reversible processes. Therefore, in real process, the exergy input is always greater than the exergy output due to exergy destruction or irreversibilities in the process. Exergy efficiency is the ratio of the exergy out to the exergy in. Temperature is a direct measure of exergy, and therefore, high exergy losses occur in processes where there is a high temperature difference.

The exergy analysis identifies the type, magnitude and location of exergy losses or irreversibilities. This helps to optimize processes and identify sustainable development to enhance efficiencies. The thermodynamic system is optimized by minimizing the irreversibilities. (Dincer and Al-Muslim, 2001)

The utilization of thermodynamic analysis has grown in the industrial world due to two reasons: the limited resources of energy and the strict policies of environmental issues. The increase in the oil prices in the seventies and adverse environmental impact in the eighties and nineties have gained considerable attention to optimize efficiencies of existing systems and future designs using the laws of thermodynamics. Also, energy use is one of the major sources of environmental problems, especially air pollution (Dincer, 2000).

### **1.3 PROCESS SIMULATION USING COMPUTER SOFTWARE**

The use of computer software is essential to be competitive in process simulation for easy manipulation of the process parameters to give clearer insight of their importance and effect. Here, it is even more critical because the crude oil and the products of crude oil distillation plants do not have established thermodynamic tables because they are mixtures of many pure components, and the composition of each varies enormously depending on many factors.

The distillation process is very dynamic and it is almost impossible to conduct manual simulation to it. It is also costly and not practical to conduct experimental runs. Most refineries will first run software simulation first before implementing any experimental test runs.



## CHAPTER 2

### LITERATURE REVIEW

#### 2.1 PREVIOUS WORKS

Exergy analysis has been utilized to optimize use of energy resources for economical and environmental aspects (Gaggioli et al., 1991; Gong and Wall, 1997; Rosen and Dincer, 1997; Dincer and Rosen, 1999; Dincer, 2000; Dincer, 2001a; Dincer, 2001b; Dincer, 2001c). Efforts have been made to achieve sustainable developments in thermodynamic processes and systems (Norton, 1991; MacRae, 1992; Rosen, 1996; Dincer, 1998; Dincer and Rosen, 1998; Rosen and Dincer, 2001). Bejan (1994) studied the application of entropy generation minimization principles to formulate energy policy aspects which was first introduced by Bejan (1982). Dincer (2002) utilized exergy as an efficient tool for energy policy making applications since exergy is a measure of quantity and quality of the energy sources.

Extensive amount of exergy analysis was carried out on standard thermodynamic cycles and their application in industry such as power plants, refrigeration systems and to some extent use of solar energy. However attention is focused mostly on energy analysis for industrial processes rather than exergy analysis (Cornelissen, 1994).

Abdi and Meisen (1999) used Aspen Plus simulation program to simulate purification of contaminated amine solution using distillation column. They showed that by modifying the process into two-stage distillation and varying the process parameters, full recovery of pure substance can be achieved.

Doldersum (1998) used exergy analysis to prove viability of distillation process modification. Doldersum used Aspen Plus program with ExerCom subroutine for process simulation and showed that by decreasing operating pressure and thus lowering operating temperature, using high pressure steam reboilers instead of furnaces and splitting feed stream; total exergy losses maybe reduced by 70% that directly results in a primary fuel reduction of almost 40% and energy saving of 10% due to splitting of the feed stream.

Exergy analysis of the column predicts that having an infinite number of stages (N) with a heat exchanger on each stage would result in zero entropy generation in the distillation column. For a 19-tray distillation column, introducing four heat exchangers would result in 25% reduction in entropy production, 6 heat exchangers would result in 37% reduction and 19 heat exchangers (i.e. heat exchanger for each tray) would result in 70% reduction. (Andersen et al., 2000)

Anaya et al. (1990) showed that efficiency of the atmospheric and vacuum distillation units is 8.6% for a refinery in Mexico. Cornelissen (1997) used Aspen Plus to simulate the distillation process and Advent to simulate the heat exchanger network in a refinery crude oil distillation plant. This study showed that the rational efficiency of the Crude oil Distillation Unit is 0.270, of the Vacuum Distillation Unit is 0.373, of the Heat Exchanger Network is 0.541 and of the furnace is 0.541. The overall rational efficiency is 0.0518. Cornelissen found the biggest causes of irreversibility are the use of high-quality fuel for the production of middle-quality heat in the furnace and the dissipation of low-quality energy in the atmosphere when cooling the products. Cornelissen proposed that utilizing the low-quality energy dissipated in the atmosphere could result in 16% reduction in irreversibilities, and also proposed to use combined heat and power cycle to reduce the irreversibility by utilizing useful work

lost in the distillation plant. Cornelissen anticipated that this reduction in entropy production would be between 19 and 60%.

Ji (2001) conducted detailed analysis on crude oil distillation plant using a rigorous targeting procedure that utilizes the heat demand-supply diagrams along with commercial software for process simulation. This study was only involved in energy in terms of fuel consumption ratio to crude oil processed, which was about 2%. Ji did not consider exergy analysis. Ji optimized processing of two types of crude oil, and also considered two alternatives to conventional design of crude oil distillation plant: the pre-flash design and the stripping-type design. The pre-flash design was more efficient than the conventional design on the expense of the products yield. The stripping-type distillation design was less competitive than the conventional design in terms of energy consumption and investment cost. The problem of designing a multi-period heat exchanger network for a complete distillation plant was also addressed.

## **2.2 OBJECTIVES OF THE STUDY**

The objective of this thesis is to conduct energy and exergy analyses to determine location, type and true magnitude of exergy loss (or exergy destruction) in a crude oil distillation plant with the aid of SimSci/PRO II program that is used to simulate the process and obtain the performance results.

Crude oil distillation plants are modeled and analyzed thermodynamically to determine the energy efficiency, exergy efficiency and irreversibility rate of the individual components and the overall system. Effects of temperature profile and pressure profile of the distillation columns on the efficiencies and irreversibility rates are illustrated. Also, effect of modifying the conventional crude oil distillation plant is

studied. Those modifications are pump-back circuits instead of pump-around circuits, adding reboiler to the distillation column and two-stage distillation columns.

## CHAPTER 3

### MODELLING

#### 3.1 DESCRIPTION OF EQUIPMENT AND PROCESS PARAMETERS

Figure 3.1 illustrates the model under study. The model consists of two crude oil distillation units: the Atmospheric Distillation Unit (ADU) and the Vacuum Distillation Unit (VDU), two crude oil furnaces and a network of heat exchangers (HEN). HEN is not shown, but its effect on utilizing the high temperature of product streams to preheat the crude oil is considered. In the model, a charge crude oil mass flow rate of 507 kg/s (300,000 barrels per day) is considered. Below is a description of the components and parameters of the model:

##### 3.1.1 Conventional Crude Oil Distillation Plant

###### 3.1.1.1 Crude Oil Furnace 1 – Heater 1

The crude oil is heated in this furnace by convection with hot exhaust of fuel combustion. The crude oil starts vaporizing as heat input increases. However, at the outlet the crude oil is not totally vaporized. It is typically 75% vapor. Typical outlet temperature values are 350°C-380°C. The charge (crude oil) can be directly from storage tanks at atmospheric temperatures, but in most crude oil distillation plants, the crude oil passes through a series of preheat exchangers that utilize the high temperatures of the distillation columns product streams. The outlet temperature of the preheat exchangers could reach up to 280°C.

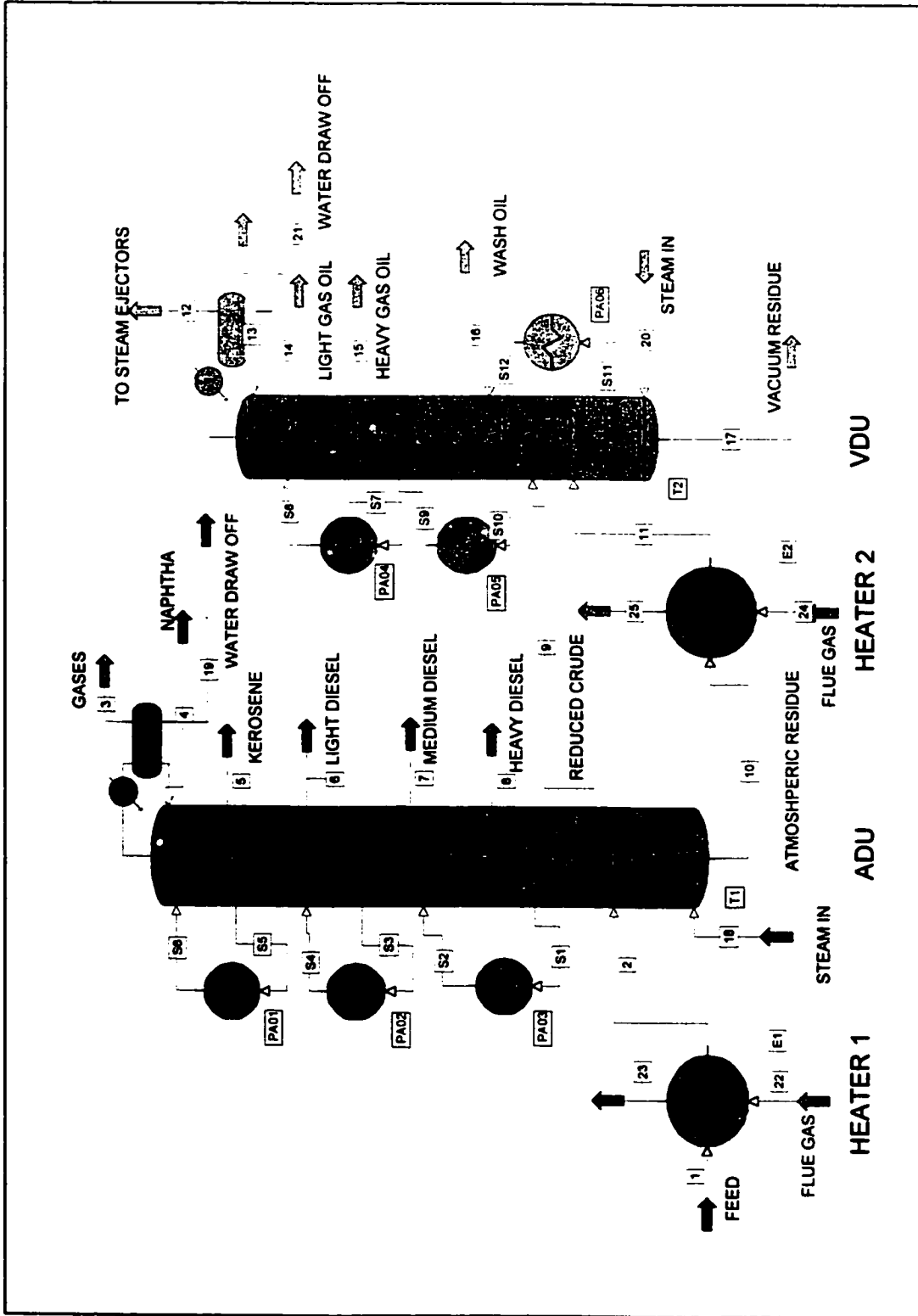


Figure 3.1 Model of a crude oil distillation plant under study.

### 3.1.1.2 Atmospheric Distillation Unit (ADU)

The Atmospheric Distillation Column operates at or above atmospheric pressure (typically 200 kPa). It is a long column that consists of many trays. These trays have bubbles or holes to allow vapor to pass through them. They are either half-cross trays or quarter-cross trays. The crude oil enters the Atmospheric Distillation Unit in a tray above the bottom in the so-called flash zone. The crude oil vaporizes more as it enters and rises up. The liquid is drawn from the bottom tray and is called the atmospheric residue. As the vapor rises, it cools as it passes through the trays and comes in contact with the liquid. Crude fractions settle in trays in the rectifying section and are drawn off at five liquid side cuts depending on their average boiling point.

The light products which have a low boiling point will be at the top and the heavier products with higher average boiling point will be at the bottom. The side cuts are from heavy to light: heavy diesel, medium diesel, light diesel and kerosene. The very light products, butane and lighter in addition to light naphtha, leave as vapor at the top of the column. The atmospheric overhead is partially condensed in heat exchangers. When they are cooled, the naphtha condenses and leaves as liquid while the lighter products remain as gases. Uncondensed vapor flows to the fuel gas system, which is used as fuel for the furnaces. In our model, a 27-tray column is assumed. The crude oil is introduced in tray no. 5. The side cut trays are no. 7 for residue, no. 10 for heavy diesel, no. 14 for medium diesel, no. 19 for light diesel and no. 24 for kerosene. The very light material leave from the overhead tray no. 1 and the very heavy material leave from the bottom tray no. 27.

Heat is removed from the column through the overhead condenser and a number of pump-around circuits, typically three. The column overhead vapor passes through

heat exchanger to cool and condense it. A portion of the condensed liquid is returned to the top of the column at a lower temperature. The pump-around circuit draws liquid from a certain tray, cool the liquid in heat exchangers and return the liquid to a tray above the original tray at a lower temperature. The pump-around circuit is defined by the withdraw tray, the return tray, the mass flow rate, and the return temperature. The ADU has three pump-around circuits: PA01, PA02, and PA03 which are presented in Table 3.1.

Superheated steam is introduced from the bottom of the column at about 350°C, which is used to reduce the partial pressure in the column and thus enhance vaporization and separation of the crude oil. The temperature profile must be designed so that the lowest temperature is well above boiling temperature of the steam and thus it will not condense in the column. Typical flow rate of steam is about 1 kg steam/ 100 kg of crude oil (10 lb steam/ barrel of crude oil).

The important parameters of the ADU are: no. of trays, crude oil entrance (flash zone), side cuts location and side cuts draw off rates as inputs. However, all those mentioned should be considered in design stage. Parameters that are frequently changed during plant operations are: temperature profile and pressure profile of the column. Temperature profile can be controlled by varying pump-around circuits flow rate and return temperature. Pressure profile can be controlled by varying the column overhead pressure. Those profiles are important indication of column operation as to produce the desired products.

### 3.1.1.3 Crude Oil Furnace 2 – Heater 2

The purpose of this furnace is to heat the bottom residue that comes out of the ADU from 350°C to 400°C. Its operation is similar to Heater 1.



#### 3.1.1.4 Vacuum Distillation Unit (VDU)

The VDU operates at vacuum pressure to help in separating of heavy hydrocarbons and allow lower temperatures for distillation than are possible in the ADU. The vacuum pressure is typically 18 kPa (2.0 psia) at the vacuum column flash zone and 10 kPa (1.1 psia) at the column top. Vacuum is created by placing a series of ejectors in service.

It works with the same principles as the ADU but with fewer trays and side cuts. In our model, a 12-tray column is assumed. Part of the residue is charged directly from the ADU at tray no. 7 and the other part is charged from the furnace at tray no. 2. The three side cuts are from heavy to light: wash oil at tray no. 4, heavy vacuum gas oil at tray no. 7 and light vacuum gas oil at tray no. 10. The vacuum residue leaves from the bottom tray no. 12 and usually goes to asphalt plant.

Superheated steam is also introduced at the bottom of the column. The function is exactly identical to that in the ADU. Heat from the VDU is removed mainly by the pump-around circuits. The VDU has three pump-around circuits: PA04, PA05, and PA06 which are presented in Table 3.1. Again, the temperature and pressure profile of the column are the key parameters to control the operation of the unit.

#### 3.1.1.5 Heat Exchangers Network (HEN)

HEN has two purposes: (i) they cool the product streams from their boiling point temperatures to about 60°C and (ii) they are used to preheat the crude oil from atmospheric temperatures to about 128°C. All those heat exchangers are normally shell and tube type.

**Table 3.1 Pump-around circuits of the atmospheric distillation unit and vacuum distillation unit.**

<b>Pump-Around Circuit</b>	<b>Withdraw Tray</b>	<b>Return Tray</b>	<b>Mass Flow Rate (kg/s)</b>	<b>Return Temperature (°C)</b>
PA01	5	3	0.1	50
PA02	10	7	0.4	125
PA03	17	13	0.3	220
PA04	3	2	0.2	75
PA05	6	4	0.08	245
PA06	9	7	0.03	200

### 3.1.1.6 Parameters

In the model, 17 state points are identified for the distillation process itself, 4 state points for steaming and 4 state points for heating. We have three pump-around circuits for each of the ADU and VDU. Important parameters for the study are temperature, pressure and flow rates as inputs for each stream including pump-around circuits. By defining those parameters, enthalpies and entropies are found to carry out the thermodynamic analysis. The first simulation is for plant operating conditions. Then, the input parameters are varied to see how they affect energy and exergy efficiencies and irreversibility rates of individual components and the overall system. Table 3.2 summarizes how these parameters for the first simulation case.

## 3.1.2 Modified Crude Oil Distillation Plants

### 3.1.2.1 Changing Pump-Around Circuits To Pump-Back Circuits

The model is illustrated in Figure 3.2. In this modification, the pump-around circuits are replaced by pump-back circuits. This is not common in refineries, but interest is in the effect on exergy efficiencies and irreversibility rate.

Table 3.2 Parameter presentation for initial simulation- operating conditions.  
(L: liquid, G: gas)

State	Stream	Phase	T (°C)	P (kPa)	Flow (kg/s)
1	Crude feed	L	25	101	507
2	Crude heated	Mixed	350	101	507
3	Off gas	G	70	205	2
4	Naphtha	L	70	205	5
5	Kerosene	L	90	210	15
6	Light diesel	L	110	215	35
7	Medium diesel	L	130	220	75
8	Heavy diesel	L	160	225	55
9	Reduced oil	L	260	230	60
10	Atmospheric residue	L	350	235	260
11	Atm. residue heated	Mixed	400	235	260
12	Vacuum gas	G	200	10	10
13	Vacuum condense	L	200	10	2
14	Light gas oil	L	220	12	25
15	Heavy gas oil	L	290	14	25
16	Wash oil	L	310	16	50
17	Vacuum residue	L	380	18	140
18	Steam	G	350	500	5
19	Water	L	70	101	2
20	Steam	G	400	205	2
21	Water	L	70	101	0.5
22	Flue gas in	G	1100	101	600
23	Flue gas out	G	350	101	600
24	Flue gas in	G	1100	101	60
25	Flue gas out	G	350	101	60

Reference: Ras Tanura Refinery, 2002

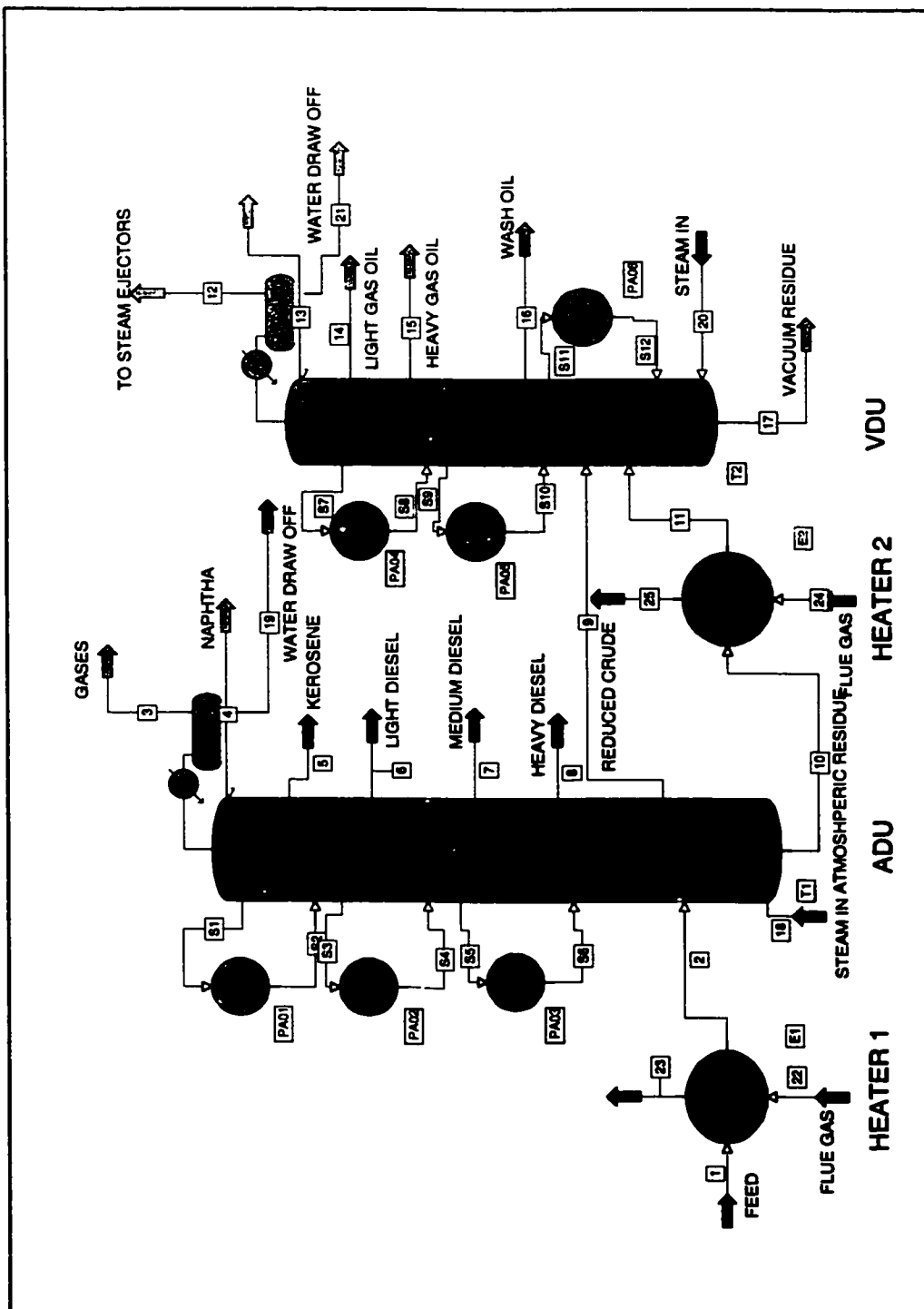


Figure 3.2 Model of crude oil distillation plant with pump-back circuits.

In the pump-around circuit, liquid is taken hot from a certain tray, cooled and returned to a tray above the original tray. This serves a dual purpose of cooling the distillation columns and pre-heating the crude oil feed.

In the pump-back circuit, liquid is taken from a certain tray, heated and returned to a tray below the original tray. This enhances separation of the light components from the heavy components. However, additional heaters are needed and the crude feed cannot be re-heated using this circuit.

### 3.1.2.2 Adding Reboiler to Distillation Columns

The model is illustrated in Figure 3.3. In this modification, a reboiler is added to the bottom tray of the distillation column. This is not common in refineries, but interest is in the effect on exergy efficiencies and irreversibility rate.

The objective of the reboiler is to heat the bottom products and return it to the distillation column. This enhances separation of the lighter products from the heavy products, but there is addition of components to the system.

### 3.1.2.3 Two-Stage Distillation Column

This modification is only applied to the ADU but not the VDU. This is because the VDU has less number of trays of 12 compared to 27 trays for the ADU. Moreover, the VDU has only two yield products while the ADU has six yield products. Therefore, it is not practical to split the vacuum distillation column into two stages.

The model is shown in Figure 3.4. The ADU column is split into two distillation columns, the first one with 13 trays and the second one with 14 trays, with a heater between the two columns. The lighter components, LPG, naphtha, kerosene and light

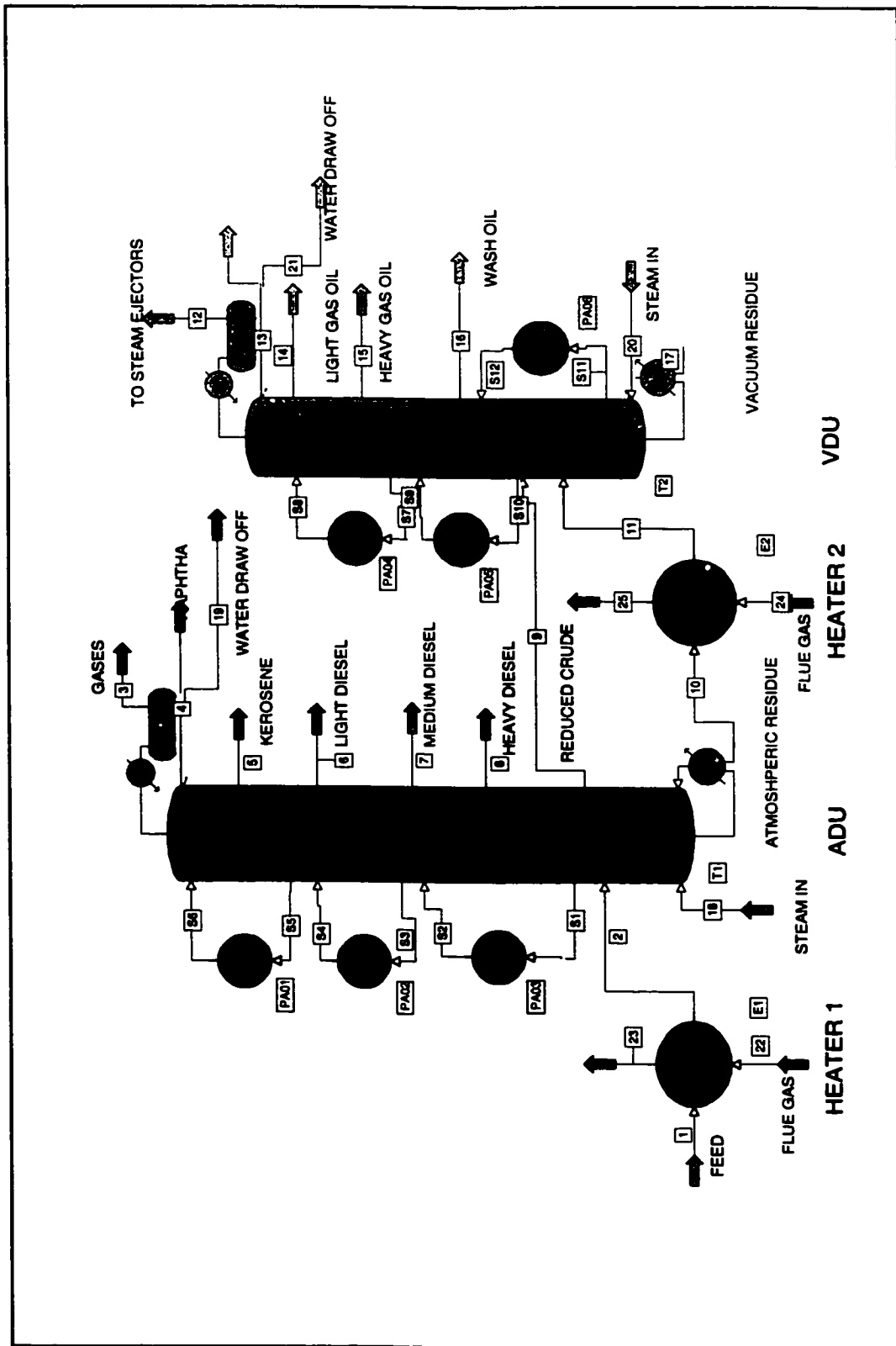


Figure 3.3 Model of crude oil distillation plant with reboiler



diesel, are extracted from the first stage. The heavier components, medium diesel, heavy diesel, and VDU partial feed, are extracted from the second stage. The second stage bottom product is the feed to the VDU.

## **3.2 DESCRIPTION OF SOFTWARE AND SIMULATION PROCEDURE**

### **3.2.1 Description of Software**

SimSci/PROII software is used to simulate the model proposed to find the exact side cuts, temperature profiles, pressure profiles, enthalpies and entropies of all state points. The program has the power and flexibility to model refinery processes. Its modeling capabilities address a wide range of applications from crude oil characterization and preheating to complex reaction and separation units. Applications of the software include: (Simulation Sciences Inc, 2001)

- Designing new processes.
- Evaluating alternative plant configurations.
- Modernizing and revamping existing plants.

The software include the following processes

- Crude preheating.
- Atmospheric crude distillation.
- Vacuum distillation.

### **3.2.2 Simulation Procedure**

After developing the model which is described in section 3.1, seven steps are required to get the results as described below. Figure 3.5 illustrates the simulation flow chart.



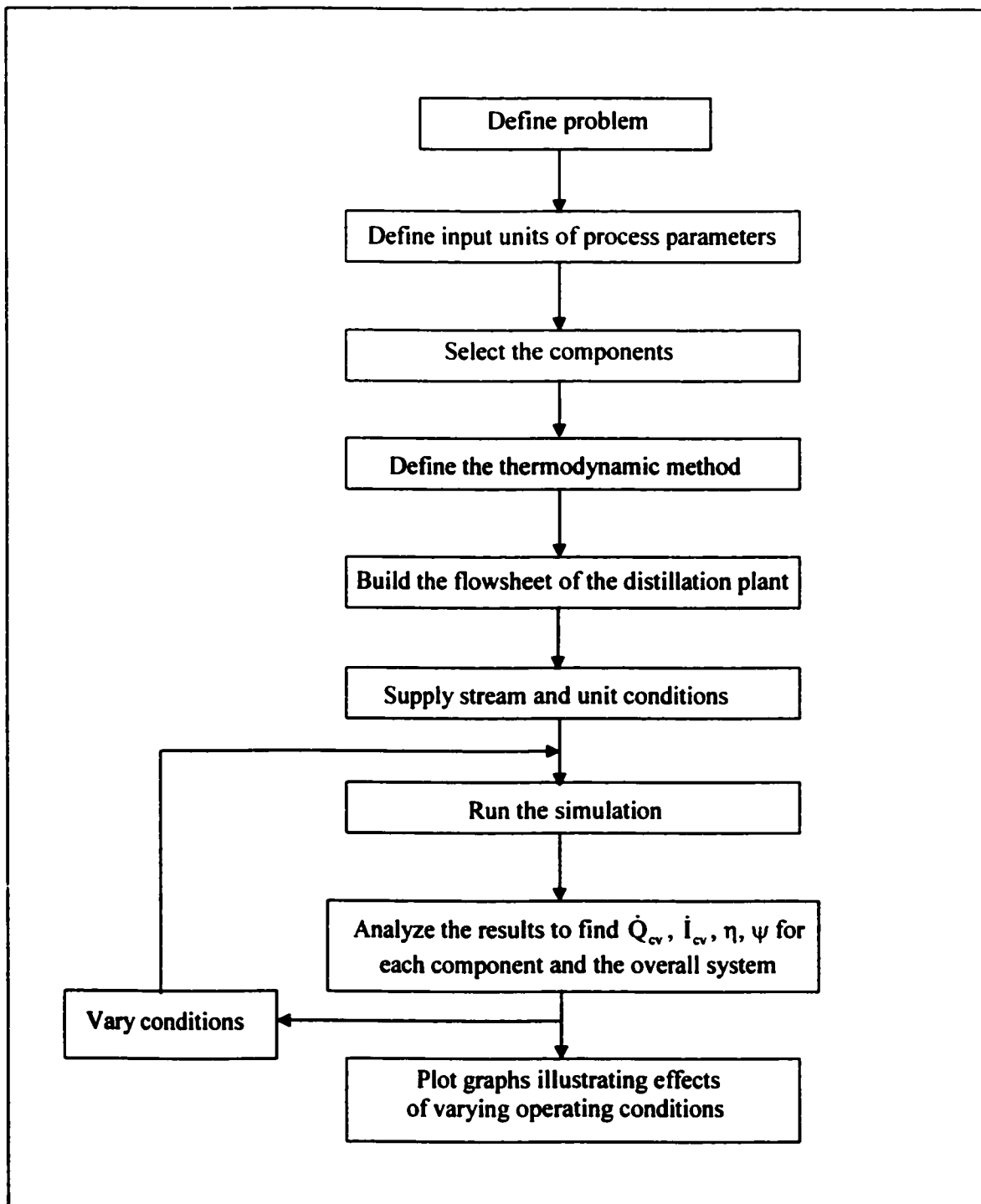


Figure 3.5 Simulation flow chart.

### **3.2.2.1 Defining the Input Units of Measure**

The units of temperature, pressure, flow rate, enthalpy, entropy, etc. should be defined. A set of units such as the SI or English units can be selected. Customized definition is also allowed. In this study, the SI units are used.

### **3.2.2.2 Selecting the Components**

PRO/II component databanks contains wide range of components such as water, air, nitrogen, pure hydrocarbons. New components can be added such as crude oil and refined products by defining their properties including average boiling point, API gravity, molecular weight, etc.

### **3.2.2.3 Defining the Thermodynamic Method**

PRO/II provides variety of thermodynamic methods for solving the problem. The most commonly used for refinery and crude oil distillation processes is the Peng-Robinson which utilizes the Peng-Robinson equation of state to calculate the liquid and vapor phase enthalpies as automatically used in the software.

### **3.2.2.4 Building the Flowsheet**

PRO/II has many default components that can be selected to build the flowsheet including distillation columns and heaters. Then, the appropriate stream should be added to the flowsheet. Streams can be external or internal.

### 3.2.2.5 Specifying the Process Stream and Unit Data

The distillation column is defined by its number of trays including location of feed stream and product streams. Also, defining the pump-around circuits and overhead pressure is essential. The heater is defined by the number of tube passes.

The stream is defined by its temperature, pressure, flow rate and composition. Defining the external streams is a must but defining the internal streams is optional as initial estimate which help the simulation to converge faster.

### 3.2.2.6 Running the Simulation

Once the flowsheet is built, and all required are entered, it is possible to run the simulation.

### 3.2.2.7 Analyzing the Results

There are a number of different ways to review the results such as reports, stream properties table, charts, etc. These properties include flow rate, enthalpy and entropy which are the key properties for energy and exergy analyses.

Then, the mathematical formulas developed in Section 3.3 are used to conduct the energy and exergy analyses. By doing so, areas of high exergy losses can be found. Finally, by varying the input data, effects of varying operating conditions on our results will be illustrated.

### 3.3 MATHEMATICAL FORMULATION

Mass balance equation, energy balance equation and exergy balance equation are applied to each component and the overall system to find heat added, energy efficiency, exergy efficiency and irreversibility rate.

#### 3.3.1 Assumptions

- Steady-state steady-flow conditions.
- Kinetic energy and potential energy are negligible.
- Reference state is:
  - $T_0 = 25^\circ \text{C} = 298.15 \text{ K}$ .
  - $P_0 = 101 \text{ kPa}$ .
  - $N_{0,i} = 1.0$ .

#### 3.3.2 Governing Equations of Thermodynamics

In general, the mass balance equation, the energy balance equation and the exergy balance equation are derived for an example open system shown in Figure 3.6. (Bejan, 1997; Van Wylen, 1994)

For the open system (control volume) shown, there is mass, energy and exergy of the control volume. There are also mass, energy and exergy flowing into the open system and mass, energy and exergy flowing out of the open system. Heat is added to the open system and work is done by the system.

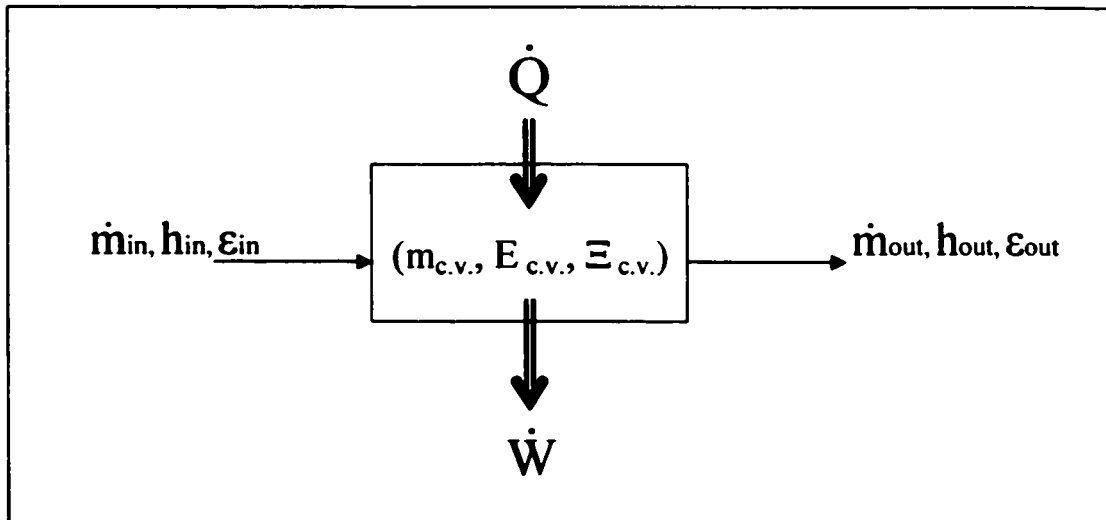


Figure 3.6 Derivation of governing equations of thermodynamics for an open system.

### 3.3.2.1 Mass Balance Equation

Mass cannot be created nor destroyed in a system. Therefore, mass is always conserved. The rate of increase of mass within a system is equal to the total mass flow rates in minus the total of mass flow rates out.

$$\frac{dm_{c.v.}}{dt} = \sum \dot{m}_{in} - \sum \dot{m}_{out} \quad (3.1)$$

### 3.3.2.2 Energy Balance Equation

Energy is always conserved in a system. Energy is just converted from one form to another like from heat to work or from heat to increase in internal energy of the system. Energy balance equation does not tell in which direction should the process go.

$$\frac{dE_{c.v.}}{dt} = \dot{Q}_{c.v.} - \dot{W}_{c.v.} + \sum_{in} \dot{m}h - \sum_{out} \dot{m}h \quad (3.2)$$

Energy efficiency  $\eta$  is defined as ratio of the output desired on the input given. For the system presented in Section 3.1, there is no work interaction. Therefore, the energy efficiency is defined as the ratio of the energy output to the energy input.

$$\eta = \frac{\dot{E}_{out}}{\dot{E}_{in}} \quad (3.3)$$

### 3.3.2.3 Exergy Balance equation

Exergy is only conserved for reversible systems. For real systems, there are always exergy losses or irreversibilities associated, and actual systems always move in the direction of decreasing exergy. Exergy loss is directly proportional to the entropy generation in the system.

$$\dot{I} = \dot{\Xi}_{loss} = \sum_{in} \dot{m}\epsilon - \sum_{out} \dot{m}\epsilon + \sum (1 - \frac{T_0}{T_i})\dot{Q}_i - \dot{W}_{c.v.} = T_0 \cdot \dot{S}_{gen} \quad (3.4)$$

where

$$\dot{S}_{gen} = \frac{dS_{c.v.}}{dt} - \sum \frac{\dot{Q}_i}{T_i} - \sum_{in} \dot{m}s + \sum_{out} \dot{m}s \quad (3.5)$$

The exergy is composed of physical exergy, which is the maximum shaft power a system can produce when it is brought to the reference temperature and pressure. This exergy is applicable to all thermodynamic systems. However, dealing mixtures, another exergy term comes into the picture, the chemical exergy. It is defined as the maximum shaft power a system of mixtures can produce when the composition of the components in the mixture are brought to the reference state. In the system studied, the reference molar composition of the components is assumed to be 1.0. The derivation of physical exergy and chemical exergy terms is illustrated below:

$$\varepsilon = \varepsilon_{ph} + \varepsilon_{ch} \quad (3.6)$$

where

$$\varepsilon_{ph} = u + p_0 v - T_0 s = h - T_0 s \quad (3.7a)$$

$$\varepsilon_{ch} = - \sum \mu_{o,i} N_i \quad (3.7b)$$

$$\mu_{o,i} = h_{o,i} - T_0 s_{o,i} \quad (3.8)$$

The exergy efficiency is defined as the ratio of the exergy output to the exergy input:

$$\Psi = \sum \dot{\varepsilon}_{out} / \sum \dot{\varepsilon}_{in} \quad (3.9)$$

### 3.3.3 Conventional Crude Oil Distillation Plant

The equations developed in Section 3.3.2 are applied to the conventional crude oil plant shown in Figure 3.1.

#### 3.3.3.1 Crude Oil Furnace 1 – Heater 1

Applying energy balance equation (3.2), we find the required  $\dot{m}_{22}$ :

$$\dot{m}_{22} = \dot{m}_1 (h_2 - h_1) / (h_{22} - h_{23}) \quad (3.10)$$

The energy efficiency is not defined for the heater as the process here is adiabatic, and therefore, the heat input is equal to the heat output.

Applying exergy balance equation (3.4), we find irreversibility rate. We note that no chemical exergy change occurs here, as no separation takes place in the furnace. The composition of the crude oil entering the furnace is the same as the composition of the crude oil leaving the furnace:

$$\dot{I}_{HEATER1} = \dot{m}_1 (\varepsilon_1 - \varepsilon_2) + \dot{m}_{22} (\varepsilon_{22} - \varepsilon_{23}) \quad (3.11)$$

The exergy efficiency is obtained from equation (3.9):

$$\Psi_{\text{HEATER1}} = (\Sigma \dot{\Xi}_{\text{in}} - \dot{I}_{\text{HEATER1}}) / \Sigma \dot{\Xi}_{\text{in}} \quad (3.12)$$

where

$$\Sigma \dot{\Xi}_{\text{in}} = \dot{m}_1 \varepsilon_1 + \dot{m}_{22} \varepsilon_{22} \quad (3.13)$$

### 3.3.3.2 Atmospheric Distillation Unit (ADU)

Applying energy balance equation (3.2), we find heat transfer from the column:

$$\begin{aligned} \dot{Q}_{\text{ADU}} = & \dot{m}_3 h_3 + \dot{m}_4 h_4 + \dot{m}_5 h_5 + \dot{m}_6 h_6 + \dot{m}_7 h_7 + \dot{m}_8 h_8 + \dot{m}_9 h_9 + \dot{m}_{10} h_{10} \\ & + \dot{m}_{19} h_{19} - \dot{m}_2 h_2 - \dot{m}_{18} h_{18} \end{aligned} \quad (3.14)$$

The energy efficiency is obtained by applying equation (3.3)

$$\eta_{\text{ADU}} = \frac{\dot{m}_3 h_3 + \dot{m}_4 h_4 + \dot{m}_5 h_5 + \dot{m}_6 h_6 + \dot{m}_7 h_7 + \dot{m}_8 h_8 + \dot{m}_9 h_9 + \dot{m}_{10} h_{10} + \dot{m}_{19} h_{19}}{\dot{m}_2 h_2 + \dot{m}_{18} h_{18}} \quad (3.15)$$

Applying exergy balance equation (3.4), we find irreversibility rate, which includes chemical exergy losses because of the separation process:

$$\begin{aligned} \dot{I}_{\text{ADU}} = & \dot{m}_2 \varepsilon + \dot{m}_{18} \varepsilon_{18} - \dot{m}_3 \varepsilon_3 - \dot{m}_4 \varepsilon_4 - \dot{m}_5 \varepsilon_5 - \dot{m}_6 \varepsilon_6 - \dot{m}_7 \varepsilon_7 - \dot{m}_8 \varepsilon_8 - \dot{m}_9 \varepsilon_9 \\ & - \dot{m}_{10} \varepsilon_{10} - \dot{m}_{19} \varepsilon_{19} + \sum (1 - T_0/T) \dot{Q} \end{aligned} \quad (3.16)$$

The exergy efficiency is obtained by applying equation (3.9):

$$\Psi_{\text{ADU}} = [\Sigma \dot{\Xi}_{\text{in}} + \sum (1 - T_0/T) \dot{Q} - \dot{I}_{\text{ADU}}] / \Sigma \dot{\Xi}_{\text{in}} \quad (3.17)$$

where

$$\Sigma_i \Xi_i = \dot{m}_2 \varepsilon_2 + \dot{m}_{18} \varepsilon_{18} \quad (3.18)$$

### 3.3.3.3 Crude Oil Furnace 2 – Heater 2

Applying energy balance equation (3.2), we find the required  $\dot{m}_{24}$ :

$$\dot{m}_{24} = \dot{m}_{10} (h_{11} - h_{10}) / (h_{24} - h_{25}) \quad (3.19)$$



The energy efficiency is not defined for the heater as the process here is adiabatic, and therefore, the heat input is equal to the heat output.

Applying exergy balance equation (3.4), we find irreversibility rate. Note that no chemical exergy change occurs here, as no separation takes place in the furnace. The composition of the crude oil entering the furnace is the same as the composition of the crude oil leaving the furnace:

$$\dot{I}_{\text{HEATER2}} = \dot{m}_{10}(\varepsilon_{10} - \varepsilon_{11}) + \dot{m}_{24}(\varepsilon_{24} - \varepsilon_{25}) \quad (3.20)$$

The exergy efficiency is obtained from equation (3.9):

$$\Psi_{\text{HEATER2}} = (\Sigma \dot{\Xi}_{\text{in}} - \dot{I}_{\text{HEATER2}}) / \Sigma \dot{\Xi}_{\text{in}} \quad (3.21)$$

where

$$\Sigma \dot{\Xi}_{\text{in}} = \dot{m}_{10}\varepsilon_{10} + \dot{m}_{24}\varepsilon_{24} \quad (3.22)$$

### 3.3.3.4 Vacuum Distillation Unit (VDU)

Applying energy balance equation (3.2), we find heat transfer from the column:

$$\dot{Q}_{\text{VDU}} = \dot{m}_{12}h_{12} + \dot{m}_{13}h_{13} + \dot{m}_{14}h_{14} + \dot{m}_{15}h_{15} + \dot{m}_{16}h_{16} + \dot{m}_{17}h_{17} + \dot{m}_{21}h_{21} - \dot{m}_{11}h_{11} - \dot{m}_{20}h_{20} \quad (3.23)$$

The energy efficiency is obtained from equation (3.3)

$$\eta_{\text{VDU}} = \frac{\dot{m}_{12}h_{12} + \dot{m}_{13}h_{13} + \dot{m}_{14}h_{14} + \dot{m}_{15}h_{15} + \dot{m}_{16}h_{16} + \dot{m}_{17}h_{17} + \dot{m}_{21}h_{21}}{\dot{m}_{11}h_{11} + \dot{m}_{20}h_{20}} \quad (3.24)$$

Applying exergy balance equation (3.4), we find irreversibility rate, which includes chemical exergy losses because of the separation process:

$$\dot{I}_{\text{VDU}} = \dot{m}_{11}\varepsilon_{11} + \dot{m}_{20}\varepsilon_{20} - \dot{m}_{12}\varepsilon_{12} - \dot{m}_{13}\varepsilon_{13} - \dot{m}_{14}\varepsilon_{14} - \dot{m}_{15}\varepsilon_{15} - \dot{m}_{16}\varepsilon_{16} - \dot{m}_{17}\varepsilon_{17} - \dot{m}_{21}\varepsilon_{21} + \sum (1 - T_0/T) \dot{Q} \quad (3.25)$$

The exergy efficiency is obtained by applying equation (3.9):

$$\Psi_{VDU} = \left[ \sum \dot{\Xi}_{in} + \sum (1 - T_0/T) \dot{Q} - \dot{i}_{VDU} \right] / \sum \dot{\Xi}_{in} \quad (3.26)$$

where

$$\sum_i \Xi_i = \dot{m}_{11} \epsilon_{11} + \dot{m}_{21} \epsilon_{21} \quad (3.27)$$

### 3.3.3.5 Overall System

Taking the whole unit as our control volume, we notice that heat transfer occurs at ADU and VDU. The overall efficiency is obtained from equation (3.3) where:

$$\dot{E}_{in} = \dot{m}_1 h_1 + \dot{m}_{18} h_{18} + \dot{m}_{20} h_{20} + \dot{m}_{22} h_{22} + \dot{m}_{24} h_{24} \quad (3.28)$$

$$\begin{aligned} \dot{E}_{out} = & \dot{m}_3 h_3 + \dot{m}_4 h_4 + \dot{m}_5 h_5 + \dot{m}_6 h_6 + \dot{m}_7 h_7 + \dot{m}_8 h_8 + \dot{m}_{19} h_{19} + \dot{m}_{12} h_{12} \\ & + \dot{m}_{13} h_{13} + \dot{m}_{14} h_{14} + \dot{m}_{15} h_{15} + \dot{m}_{16} h_{16} + \dot{m}_{17} h_{17} + \dot{m}_{21} h_{21} + \dot{m}_{23} h_{23} + \dot{m}_{25} h_{25} \end{aligned} \quad (3.29)$$

The total irreversibility rate is the summation of the irreversibility rates of the individual components.

$$\dot{I}_O = \dot{I}_{HEATER1} + \dot{I}_{ADU} + \dot{I}_{HEATER2} + \dot{I}_{VDU2} \quad (3.30)$$

The exergy efficiency is obtained by applying equation (3.9)

$$\Psi_O = \left[ \sum \dot{\Xi}_{in} + \sum (1 - T_0/T) \dot{Q} - \dot{I}_O \right] / \sum \dot{\Xi}_{in} \quad (3.31)$$

where

$$\sum_i \Xi_i = \dot{m}_1 \epsilon_1 + \dot{m}_{18} \epsilon_{18} + \dot{m}_{20} \epsilon_{20} + \dot{m}_{22} \epsilon_{22} + \dot{m}_{24} \epsilon_{24} \quad (3.32)$$

### 3.3.4 Two-Stage Atmospheric Distillation Unit (ADU)

The governing equations developed in Section 3.3.2 are applied to the proposed two – stage distillation model shown in Figure 3.2. The two-stage ADU efficiency is obtained by applying equation (3.3) where:

$$\dot{E}_{in} = \dot{m}_1 h_1 + \dot{m}_2 h_2 + \dot{m}_{18} h_{18} + \dot{m}_{30} h_{30} + \dot{m}_{32} h_{32} \quad (3.33)$$

$$\begin{aligned} \dot{E}_{out} = & \dot{m}_3 h_3 + \dot{m}_4 h_4 + \dot{m}_5 h_5 + \dot{m}_6 h_6 + \dot{m}_7 h_7 + \dot{m}_8 h_8 + \dot{m}_9 h_9 + \dot{m}_{10} h_{10} \\ & + \dot{m}_{28} h_{28} + \dot{m}_{29} h_{29} + \dot{m}_{31} h_{31} + \dot{m}_{33} h_{33} \end{aligned} \quad (3.34)$$

Applying exergy balance equation (3.4), we find irreversibility rate:

$$\begin{aligned} \dot{I}_{ADU} = & \dot{m}_1 \varepsilon_1 + \dot{m}_2 \varepsilon_2 + \dot{m}_{18} \varepsilon_{18} + \dot{m}_{30} \varepsilon_{30} + \dot{m}_{32} \varepsilon_{32} - \dot{m}_3 \varepsilon_3 - \dot{m}_4 \varepsilon_4 - \dot{m}_5 \varepsilon_5 \\ & - \dot{m}_6 \varepsilon_6 - \dot{m}_7 \varepsilon_7 - \dot{m}_8 \varepsilon_8 - \dot{m}_9 \varepsilon_9 - \dot{m}_{10} \varepsilon_{10} - \dot{m}_{28} \varepsilon_{28} - \dot{m}_{29} \varepsilon_{29} - \dot{m}_{31} \varepsilon_{31} \\ & - \dot{m}_{33} \varepsilon_{33} + \sum (1 - T_0/T) \dot{Q} \end{aligned} \quad (3.35)$$

The exergy efficiency is obtained by applying equation (3.9):

$$\Psi_{ADU} = \left[ \sum \dot{\Xi}_{in} + \sum (1 - T_0/T) \dot{Q} - \dot{I}_{ADU} \right] / \sum \dot{\Xi}_{in} \quad (3.36)$$

where

$$\sum \dot{\Xi}_{in} = \dot{m}_1 \varepsilon_1 + \dot{m}_2 \varepsilon_2 + \dot{m}_{18} \varepsilon_{18} + \dot{m}_{30} \varepsilon_{30} + \dot{m}_{32} \varepsilon_{32} \quad (3.37)$$

## CHAPTER 4

### RESULTS AND DISCUSSOIN

The first simulation is run for the actual operating conditions of Ras Tanura Refinery crude oil distillation plant. The input data are presented in Table 3.2. The results of the simulation are summarized in Table 4.1.

By applying the equations developed in Section 3.3.3, we can find the heat balance, energy efficiency, exergy efficiency and irreversibility rate for individual components and the overall system. The calculations for exergy are made twice: one with neglecting the chemical exergy term in Equation (3.7) and the other including the chemical exergy loss inherent to the separation process. This is done to see how much is the contribution of chemical exergy loss to the total exergy loss. The results are shown in Table 4.2 and Figure 4.1.

The energy efficiency of the ADU is 0.497 while the energy efficiency of the VDU is 0.579. The ADU energy efficiency is lower because the main separation takes place there. The energy efficiency of the overall system is 0.519. The energy efficiencies of the heaters are not included as we are assuming adiabatic heat transfer in the process.

The highest irreversibilities (exergy losses) occur in the ADU with 56% of the total irreversibility losses. This is again because the main separation takes place in the ADU. Those losses are composed of physical and chemical exergy losses. The chemical exergy losses are 6.8% of the total exergy losses.

Table 4.1 Simulation results for operating conditions. (l: light, m: medium, h: heavy, LVGO: light vacuum gas oil, HVGO: heavy vacuum gas oil, ex: exhaust)

<b>Stream Name</b>	<b>1</b>	<b>2</b>	<b>3</b>	<b>4</b>	<b>5</b>	<b>6</b>	<b>7</b>
<b>Stream Description</b>	feed	feed	offgas	naphtha	kerosene	l. diesel	m. diesel
<b>Phase</b>	Liquid	Vapor	Vapor	Mixed	Mixed	Mixed	Mixed
<b>Temperature (°C)</b>	25	352.3746	68.00522	68.00522	68.91907	68.86612	91.4024
<b>Pressure (kPa)</b>	102	102	206.6044	206.8148	208.4696	211.7791	215.9159
<b>Mass Flow Rate (kg/s)</b>	507.6221	507.6221	1.09E-12	0.566983	15.56693	32.14957	75.81233
<b>Sp. Entropy (kJ/kg-K)</b>	4.440448	6.615505	4.216201	2.509089	3.039461	2.323762	4.266169
<b>Sp. Enthalpy (kJ/kg)</b>	34.39055	1040.125	760.3895	164.5537	348.8646	182.8285	186.9895
<b>Stream Name</b>	<b>8</b>	<b>9</b>	<b>10</b>	<b>11</b>	<b>12</b>	<b>13</b>	<b>14</b>
<b>Stream Description</b>	h. diesel	reduced	adu bottom	adu bottom	offgas	condensed	LVGO
<b>Phase</b>	Liquid	Liquid	Liquid	Liquid	Vapor	Liquid	Liquid
<b>Temperature (°C)</b>	157.1815	261.4419	342.2222	408.8889	203.2595	203.2595	219.9582
<b>Pressure (kPa)</b>	219.2254	223.3622	228.8781	228.8781	11.10056	11.10056	11.10056
<b>Mass Flow Rate (kg/s)</b>	53.9682	58.0393	260.1879	260.1879	10.21269	0.834219	22.61737
<b>Sp. Entropy (kJ/kg-K)</b>	5.015905	5.878275	6.557474	6.858651	6.830466	5.768256	5.860224
<b>Sp. Enthalpy (kJ/kg)</b>	323.7495	555.9205	750.3401	945.6614	1125.868	398.4854	438.4018
<b>Stream Name</b>	<b>15</b>	<b>16</b>	<b>17</b>	<b>18</b>	<b>19</b>	<b>20</b>	<b>21</b>
<b>Stream Description</b>	HVGO	Wash oil	vd bottom	steam	water	steam	water
<b>Phase</b>	Liquid	Liquid	Liquid	Vapor	Water	Vapor	Water
<b>Temperature (°C)</b>	286.3149	312.649	299.5011	353.3333	68.00522	768	n/a
<b>Pressure (kPa)</b>	11.92794	13.16899	15.23742	515.0106	206.8148	413.6856	n/a
<b>Mass Flow Rate (kg/s)</b>	39.7135	46.74861	142.2034	5.415384	0.517354	2.141962	0
<b>Sp. Entropy (kJ/kg-K)</b>	6.238811	6.426736	6.386489	7.651651	0.96201	8.831165	n/a
<b>Sp. Enthalpy (kJ/kg)</b>	602.1979	666.1357	628.2944	3172.316	284.4449	4075.276	n/a
<b>Stream Name</b>	<b>22</b>	<b>23</b>	<b>24</b>	<b>25</b>			
<b>Stream Description</b>	FLUE	FLUE EX	FLUE2	FLUE2 EX			
<b>Phase</b>	Vapor	Vapor	Vapor	Vapor			
<b>Temperature (°C)</b>	1100	353.3333	1100	357			
<b>Pressure (kPa)</b>	102	102	102	102			
<b>Mass Flow Rate (kg/s)</b>	600	600	60	60			
<b>Sp. Entropy (kJ/kg-K)</b>	7.986952	7.100531	7.986952	7.106715			
<b>Sp. Enthalpy (kJ/kg)</b>	1180.016	329.1276	1180.016	333.0128			

Table 4.2 Model results for operating conditions.

	HEATER1	ADU	HEATER2	VDU	ALL
$\Sigma \dot{H}_{in}$ (kW)	725467	545170	266030	287044	822177
$\Sigma \dot{H}_{out}$ (kW)	725467	270692	266030	166148	426803
$\dot{Q}_{c.v.}$ (kW)	0	274478	0	120896	395374
$\eta$	-	0.497	-	0.579	0.519
$\Sigma \dot{E}_{ph.in}$ (kW)	549311	460180	211396	233428	632531
$\Sigma \dot{E}_{ph.out}$ (kW)	444036	201557	201436	124561	149806
$I_{ph}$ (kW)	105275	341129	9960	151841	608204
$\Sigma \dot{E}_{in}$ (kW)	589496	500365	223817	248621	672716
$\Sigma \dot{E}_{out}$ (kW)	484221	216749	213858	124560.5	149806
$I$ (kW)	105275	366121	9960	167033	648389
$\psi$	0.821	0.433	0.956	0.501	0.223
$I_{ch}$ (kW)	0	24993	0	15192.3	40185
% $I_{ch}$	0	6.8	0	9.1	6.2

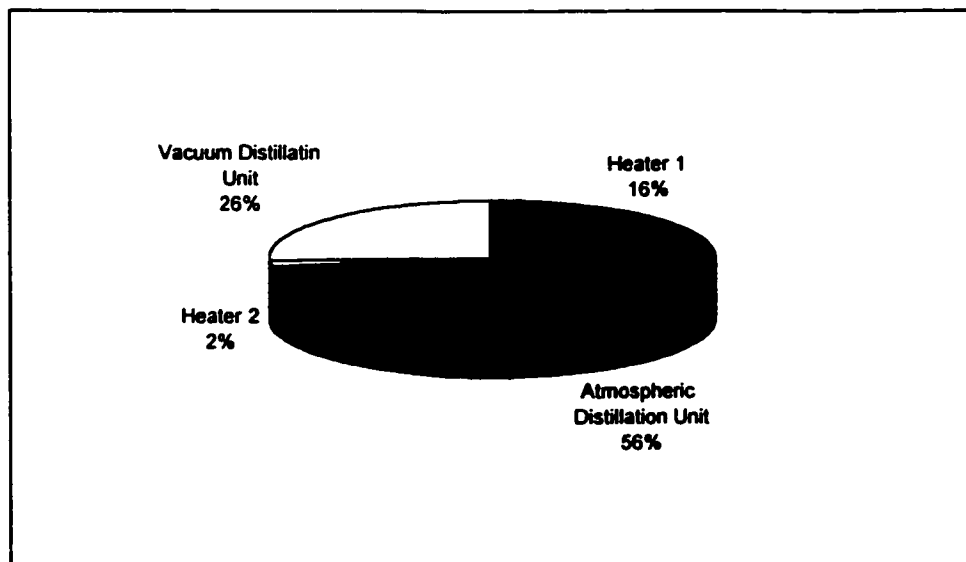


Figure 4.1 Distribution of irreversibilities.

The irreversibility losses in the VDU are still high, 26 % of the total irreversibility losses, but lower than those of the ADU. This is because less separation is involved. The chemical exergy losses represent 9.1% of the total VDU exergy losses.

In the two heaters, the irreversibility losses are 16% for Heater 1 and 2% for Heater 2. This is because most of the heating load is carried in Heater 1. In Heater 2 the flow is almost half of that in Heater 1 and temperature rise required is less. Only physical exergy losses are present here as there is no separation process in the heater.

The components that contribute most to the irreversibility losses have the least exergy efficiency. The exergy efficiencies are 0.433 for ADU, 0.501 for VDU, 0.821 for Heater 1 0.956 for Heater 2 and 0.233 for the overall system. The overall exergy efficiency is the lowest amongst the components efficiency as the more components are added to the system, the lower overall exergy efficiency. This is because exergy losses add up as more components are added, and thus, the exergy output of the system gets lower while the exergy input to the system remains constant.

The detailed simulation report is given in Appendix A. The report first gives the names of different streams and their basic input properties including temperature, pressure and composition. It then describes the input of unit operations including the distillations columns and furnaces (heat exchangers). After that, the report gives the output of the simulation including component data, calculation sequence, heat exchanger summary, column summary, stream molar component rates and stream summary.

## **4.1 EFFECT OF OPERATING CONDITIONS**

### **4.1.1 Effect Of Distillation Column Temperature Profile**

The temperature profile of the distillation column is controlled through the pump-around circuits. The two parameters are the pump-around flow rate and the pump-around return temperature. Effect of varying those parameters on energy efficiencies exergy efficiencies and irreversibility rates of both distillation columns and the overall system is illustrated in this section.

#### **4.1.1.1 Effect of Pump-Around Mass Flow Rate**

The ADU has three pump-around circuits PA01, PA02 and PA03. Figures 4.2 through 4.7 illustrate the effect of changing the mass flow rate of each pump-around on the energy efficiency, exergy efficiency and the irreversibility rate of the ADU and the overall system.

For the ADU, the operating condition, the middle point, is the optimum condition. The energy efficiency at this point is 4% to 9% higher than the adjacent points, the exergy efficiency is 6% to 10% higher than the adjacent points, and the irreversibility rate is 3 to 12% lower than the adjacent points. This is because as we departure from the operating condition, more load is required by the overhead condenser to give the same yield products.

For the overall system, there is no significant change with regards to the energy and exergy efficiencies. However the irreversibly rate follows the same trend as that of the ADU with minimum irreversibilities at the operating condition which is 5% to 10% lower than the adjacent points.



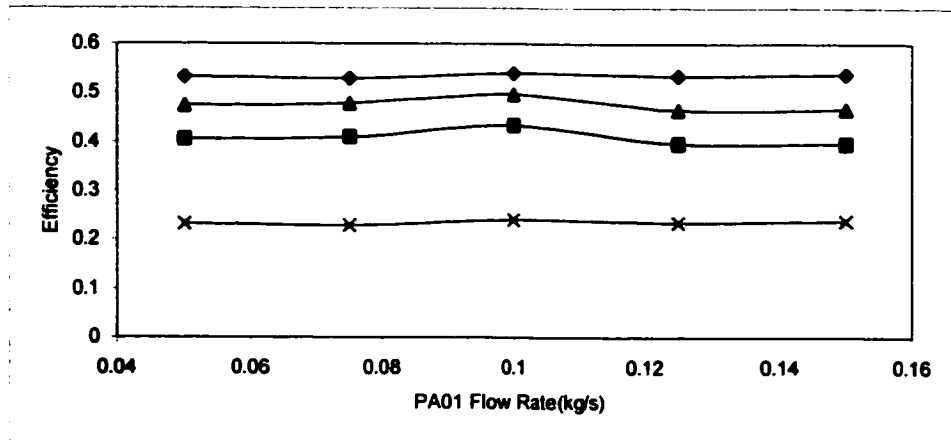


Figure 4.2 Energy and exergy efficiencies vs. pump-around O1 mass flow rate (◆ overall energy efficiency, ▲ atmospheric distillation unit energy efficiency, ■ atmospheric distillation unit exergy efficiency, × overall exergy efficiency).

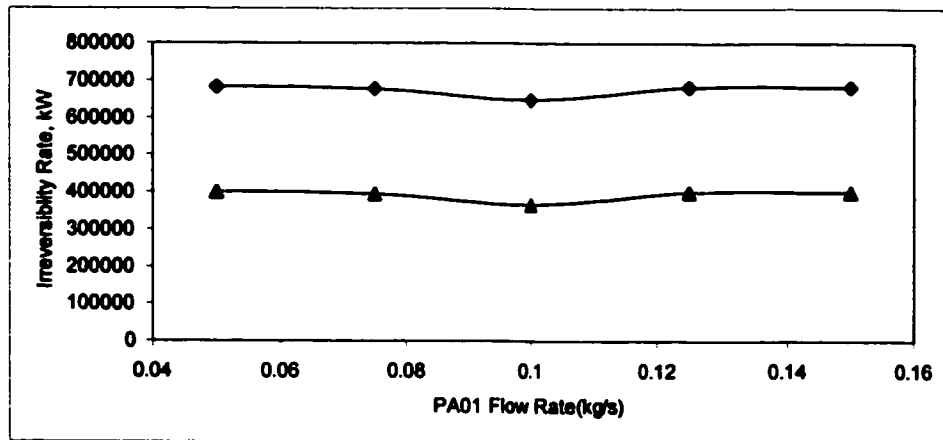


Figure 4.3 Irreversibility rate vs. pump-around O1 mass flow rate (◆ overall, ▲ atmospheric distillation unit).

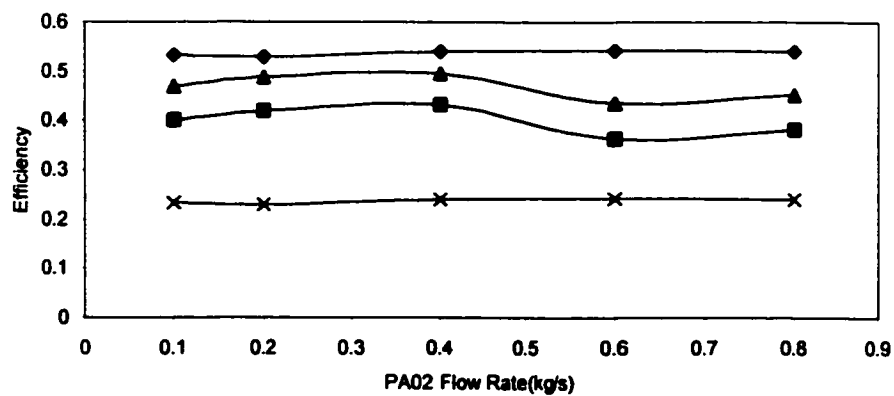


Figure 4.4 Energy and exergy efficiencies vs. pump-around O<sub>2</sub> mass flow rate (◆ overall energy efficiency, ▲ atmospheric distillation unit energy efficiency, ■ atmospheric distillation unit exergy efficiency, × overall exergy efficiency).

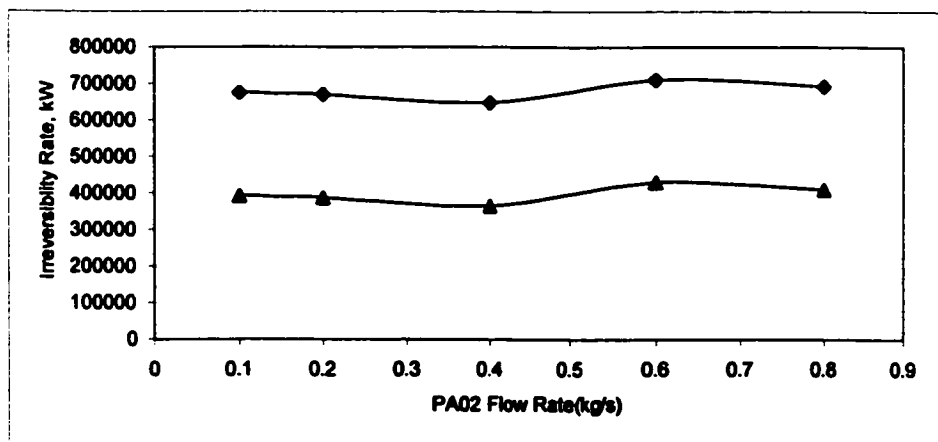


Figure 4.5 Irreversibility rate vs. pump-around O<sub>2</sub> mass flow rate (◆ overall, ▲ atmospheric distillation unit).

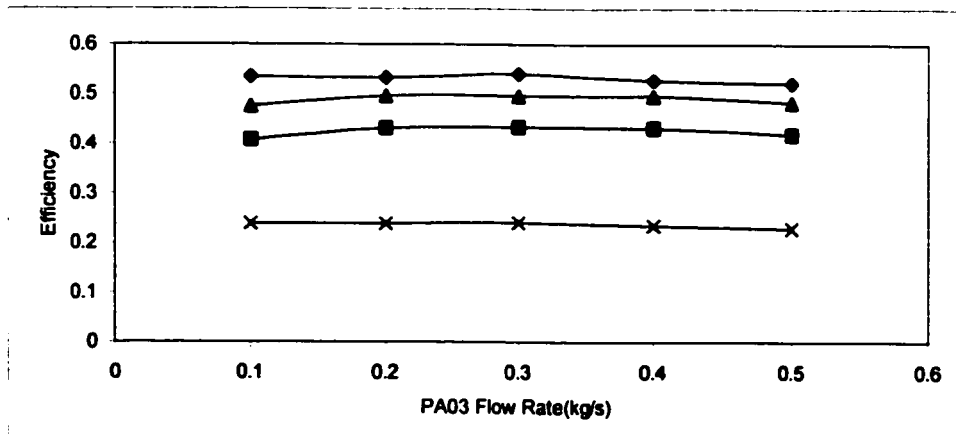


Figure 4.6 Energy and exergy efficiencies vs. pump-around 03 mass flow rate (◆ overall energy efficiency, ▲ atmospheric distillation unit energy efficiency, ■ atmospheric distillation unit exergy efficiency, × overall exergy efficiency).

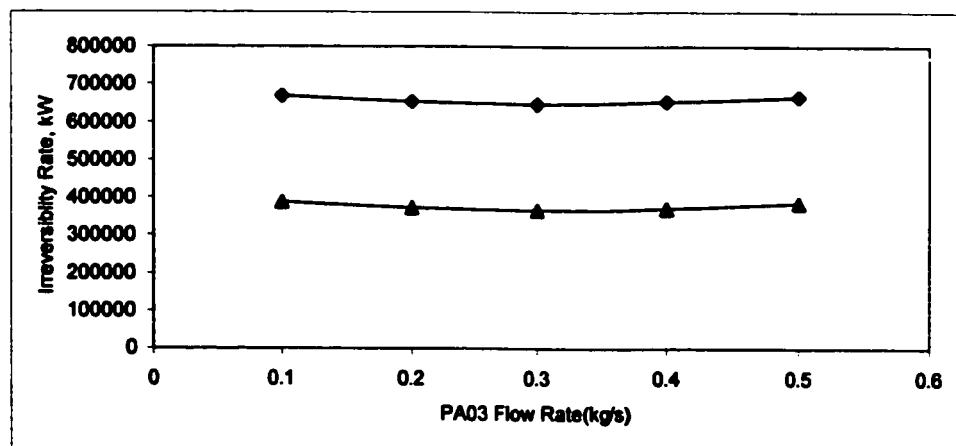


Figure 4.7 Irreversibility rate vs. pump-around 03 mass flow rate (◆ overall, ▲ atmospheric distillation unit).

The VDU has three pump-around circuits PA04, PA05 and PA06. Figures 4.8 through 4.13 illustrate the effect of changing the mass flow rate of each pump-around on the energy efficiency, exergy efficiency and the irreversibility rate of the VDU and the overall system.

As for the VDU, the energy efficiency variation is within 2%, the exergy efficiency variation is within 5% and the irreversibility rate variation is within 2%. The variation for the overall system energy efficiency, exergy efficiency and irreversibility rate are within 2%. This is because heat duty on the VDU overhead condenser is negligible and most of the heat duty is carried by the pump-around circuits.

#### 4.1.1.2 Effect of Pump-Around Return Temperature

Figures 4.14 through 4.19 illustrate the effect of varying PA01, PA02 and PA03 return temperature on the energy efficiency, exergy efficiency and the irreversibility rate of the ADU and the overall system.

The operating condition is again the optimum condition. The energy efficiency at this point is 4% to 11% higher than the adjacent points, the exergy efficiency is 5% to 14% higher than the adjacent points, and the irreversibility rate is 8% to 15% lower than the adjacent points. This is because as we departure from the operating condition, more load is required by the overhead condenser to give the same yield products.

For the overall system, there is no significant change with regards to the energy and exergy efficiencies. However the irreversibility rate follows the same trend as that of the ADU with minimum irreversibilities at the operating condition which is 4% to 9% lower than the adjacent points.

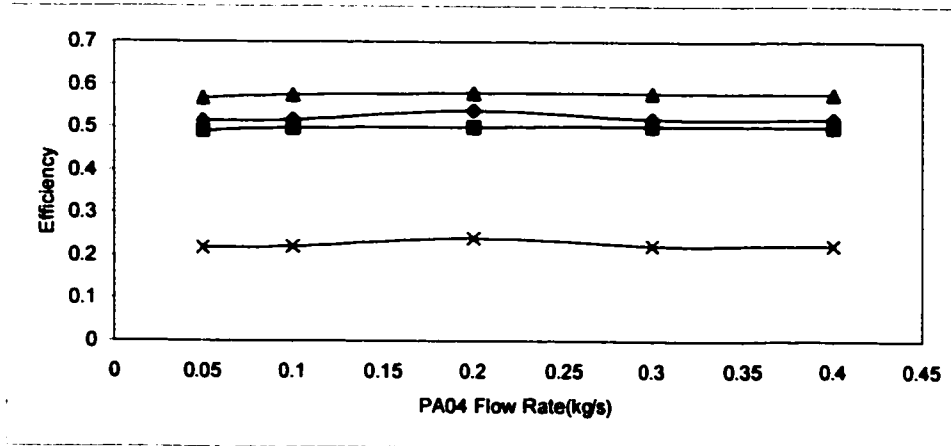


Figure 4.8 Energy and exergy efficiencies vs. pump-around 04 mass flow rate (▲ vacuum distillation unit energy efficiency, ◆ overall energy efficiency, ■ vacuum distillation unit exergy efficiency, × overall exergy efficiency).

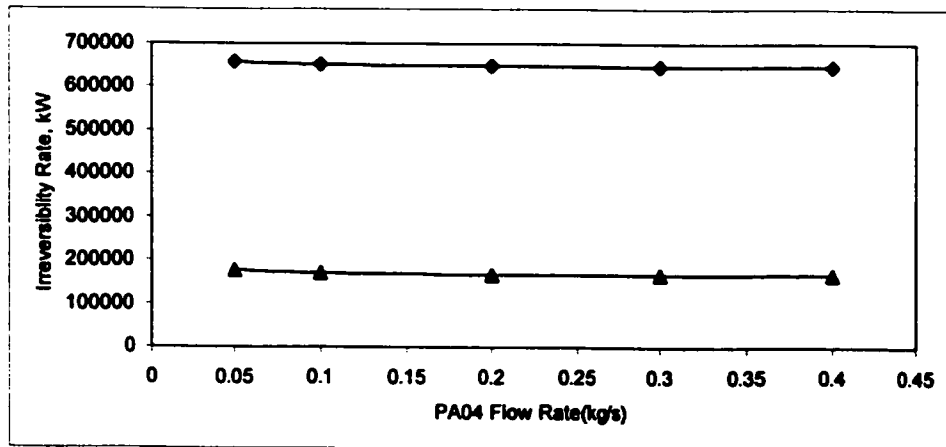


Figure 4.9 Irreversibility rate vs. pump-around 04 mass flow rate (◆ overall, ▲ vacuum distillation unit).

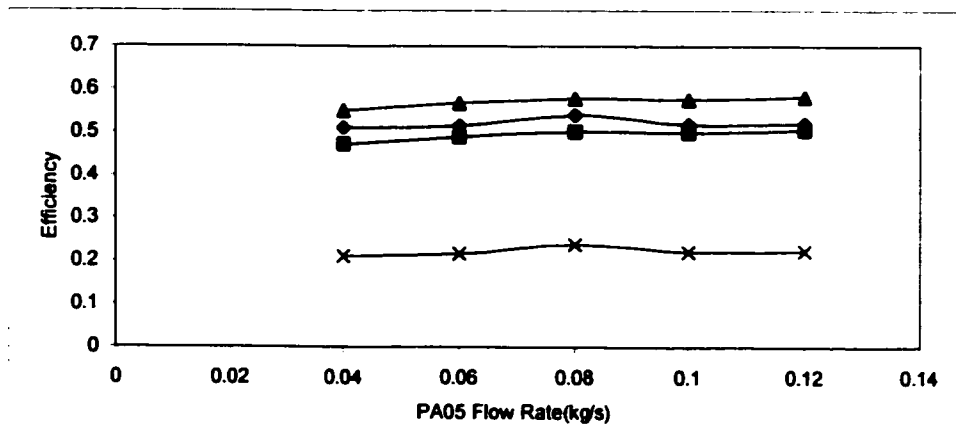


Figure 4.10 Energy and exergy efficiencies vs. pump-around 05 mass flow rate (▲ vacuum distillation unit energy efficiency, ◆ overall energy efficiency, ■ vacuum distillation unit exergy efficiency, × overall exergy efficiency).

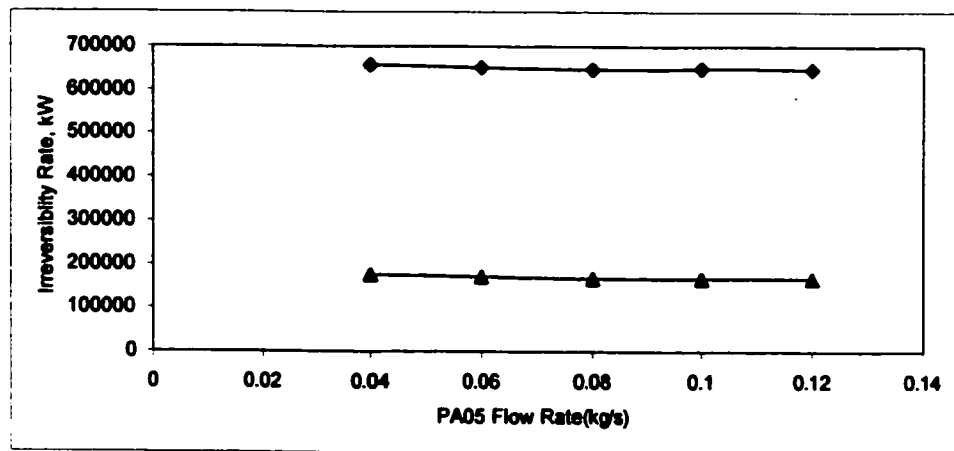


Figure 4.11 Irreversibility rate vs. pump-around 05 mass flow rate (◆ overall, ▲ vacuum distillation unit).

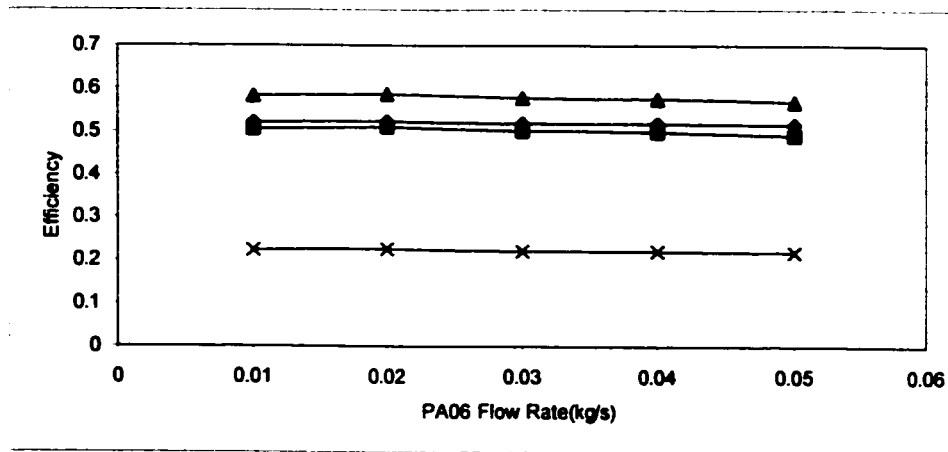


Figure 4.12 Energy and exergy efficiencies vs. pump-around 06 mass flow rate (▲ vacuum distillation unit energy efficiency, ◆ overall energy efficiency, ■ vacuum distillation unit exergy efficiency, × overall exergy efficiency).

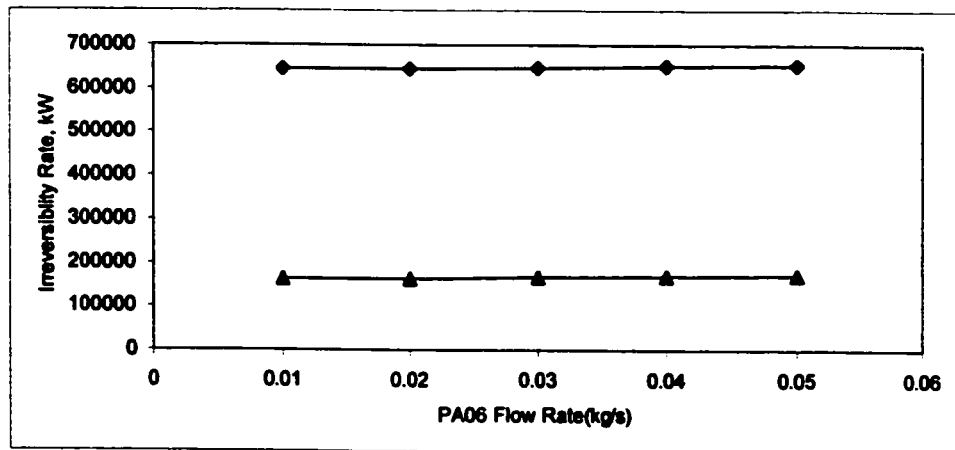


Figure 4.13 Irreversibility rate vs. pump-around 06 mass flow rate (◆ overall, ▲ vacuum distillation unit).

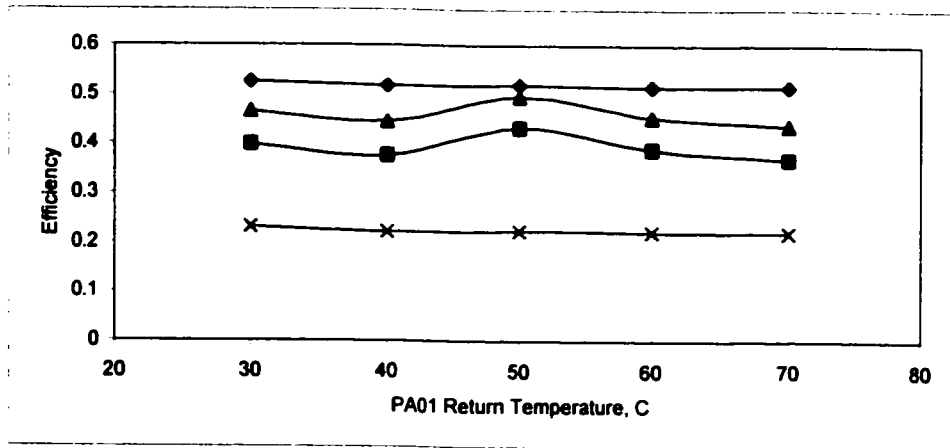


Figure 4.14 Energy and exergy efficiencies vs. pump-around 01 return temperature (◆ overall energy efficiency, ▲ atmospheric distillation unit energy efficiency, ■ atmospheric distillation unit exergy efficiency, × overall exergy efficiency).

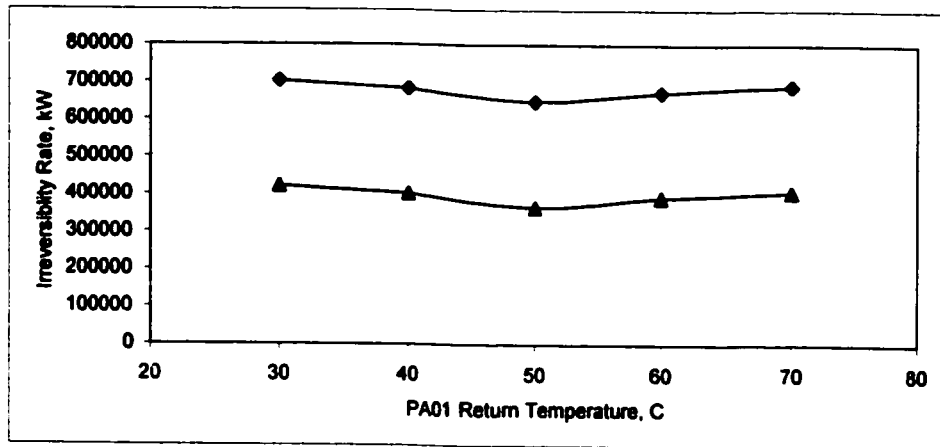


Figure 4.15 Irreversibility rate vs. pump-around 01 return temperature (◆ overall, ▲ atmospheric distillation unit).



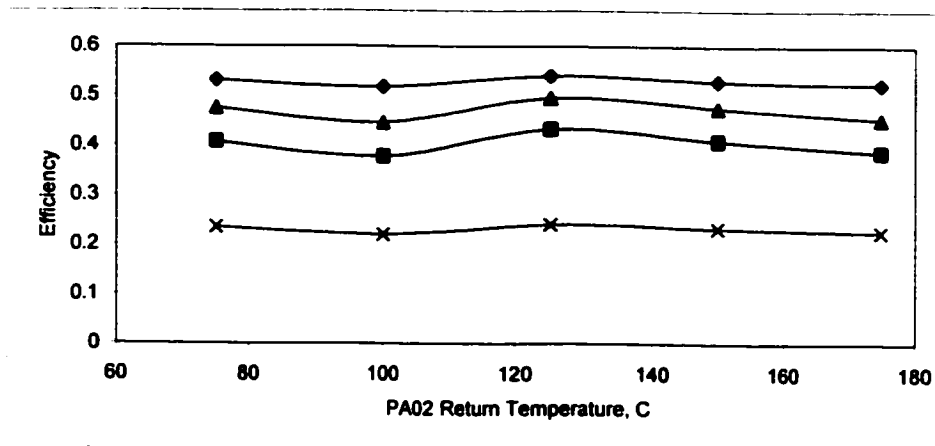


Figure 4.16 Energy and exergy efficiencies vs. pump-around O2 return temperature (◆ overall energy efficiency, ▲ atmospheric distillation unit energy efficiency, ■ atmospheric distillation unit exergy efficiency, × overall exergy efficiency).

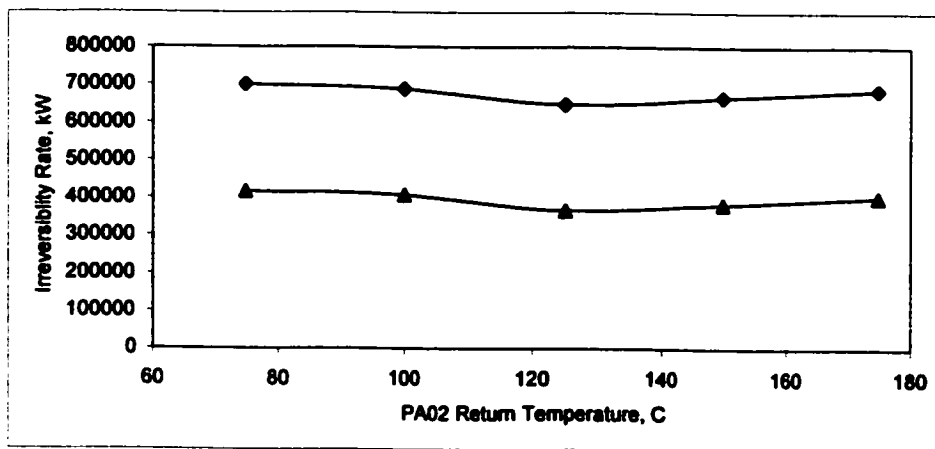


Figure 4.17 Irreversibility rate vs. pump-around O2 return temperature (◆ overall, ▲ atmospheric distillation unit)

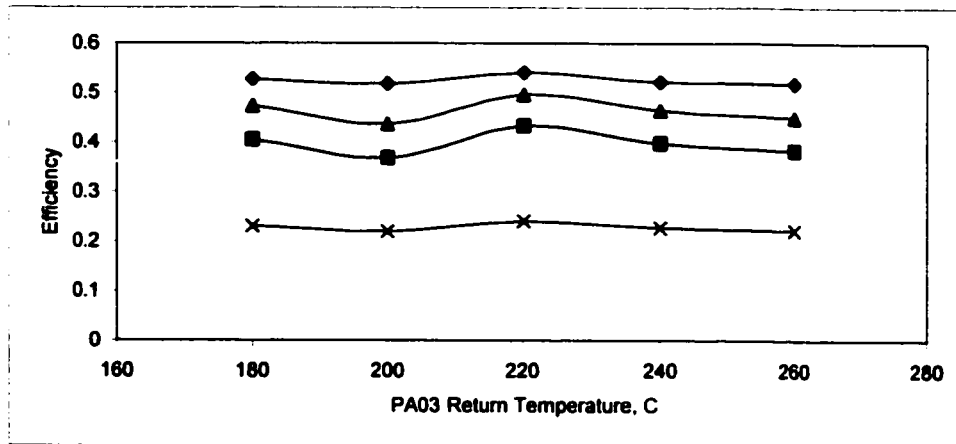


Figure 4.18 Energy and exergy efficiencies vs. pump-around 03 return temperature (◆ overall energy efficiency, ▲ atmospheric distillation unit energy efficiency, ■ atmospheric distillation unit exergy efficiency, × overall exergy efficiency).

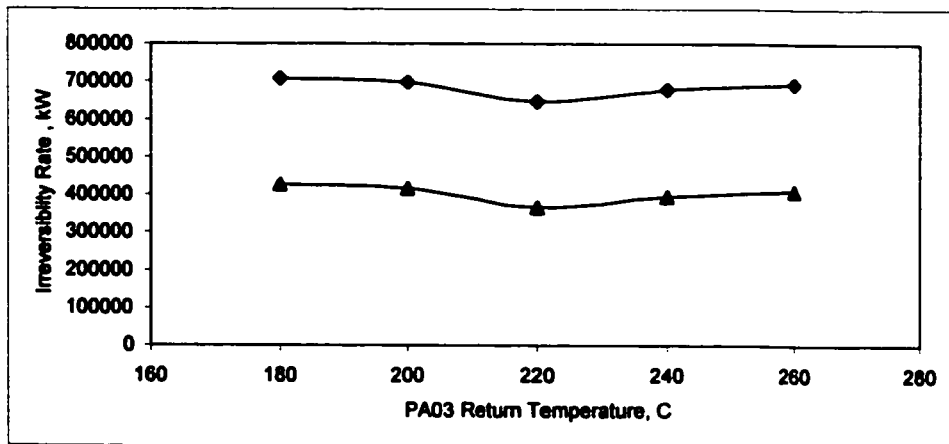


Figure 4.19 Irreversibility rate vs. pump-around 03 return temperature (◆ overall, ▲ atmospheric distillation unit).

Figures 4.20 through 4.25 illustrate the effect of changing the return temperature of PA04, PA05 and PA06 on the energy efficiency, exergy efficiency and the irreversibility rate of the VDU and the overall system.

Variation the return temperature of PA04, PA05 and PA06 has no significant effect on the energy efficiency, exergy efficiency and irreversibility rate of the VDU and the overall system. As for the VDU, the energy efficiency, exergy efficiency and the irreversibility rate variations are within 2%. The variation for the overall system energy efficiency, exergy efficiency and irreversibility rate are within 1%. This is because heat duty on the VDU overhead condenser is negligible and most of the heat duty is carried by the pump-around circuits.

#### **4.1.2 Effect of Distillation Column Pressure Profile**

The pressure profile of the distillation column is controlled through the overhead pressure of the column. The vapor flow to the condenser is restricted to increase the pressure. As the column overhead pressure increases, the yield of the heavy hydrocarbons increase and the yield of the light hydrocarbons decrease. Effect of varying the overhead pressure on the energy efficiency, exergy efficiency and irreversibility rate of both distillation columns and the overall system is illustrated in this section.

##### **4.1.2.1 Atmospheric Distillation Unit (ADU) Overhead Pressure**

Effect of varying the ADU overhead pressure on the exergy efficiency is shown on Figure 4.26 while its effect on the irreversibility rate is shown on Figure 4.27.

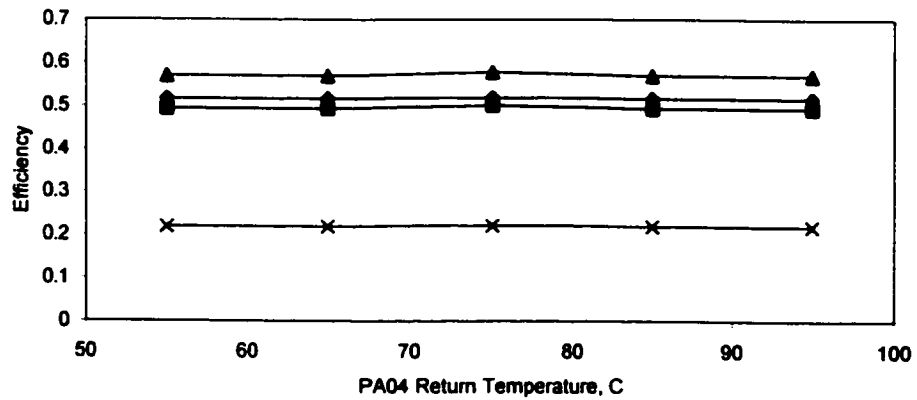


Figure 4.20 Energy and exergy efficiencies vs. pump-around 04 return temperature (▲ vacuum distillation unit energy efficiency, ◆ overall energy efficiency, ■ vacuum distillation unit exergy efficiency, × overall exergy efficiency).

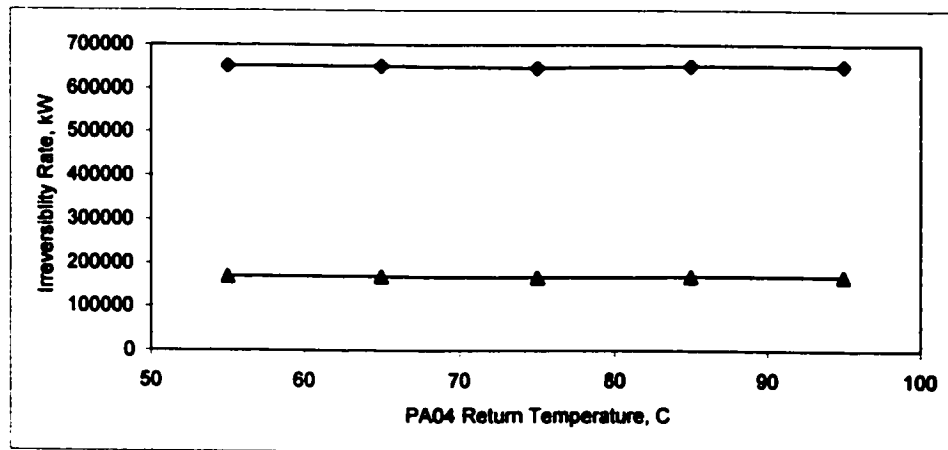


Figure 4.21 Irreversibility rate vs. pump-around 04 return temperature (◆ overall, ▲ vacuum distillation unit).

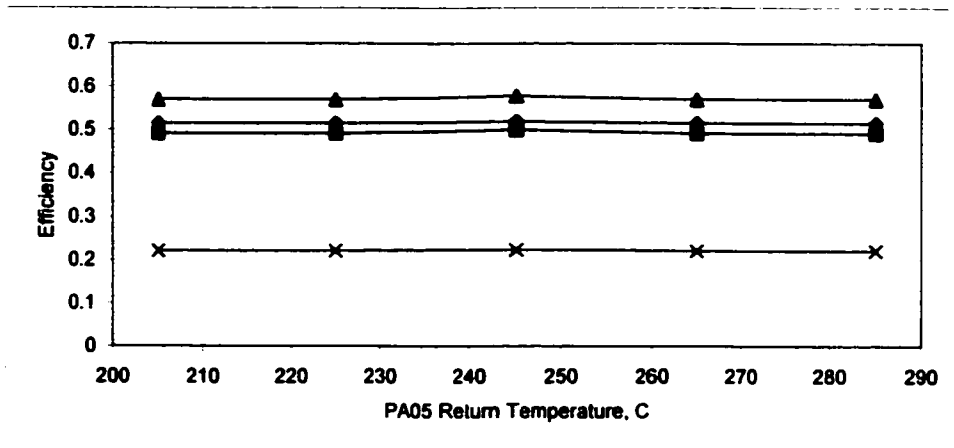


Figure 4.22 Energy and exergy efficiencies vs. pump-around 05 return temperature (▲ vacuum distillation unit energy efficiency, ◆ overall energy efficiency, ■ vacuum distillation unit exergy efficiency, × overall exergy efficiency).

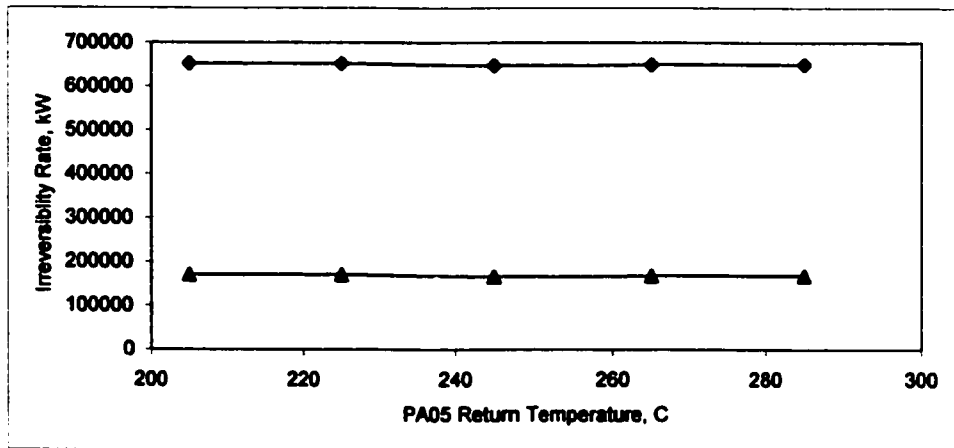


Figure 4.23 Irreversibility rate vs. pump-around 05 return temperature (◆ overall, ▲ vacuum distillation unit).

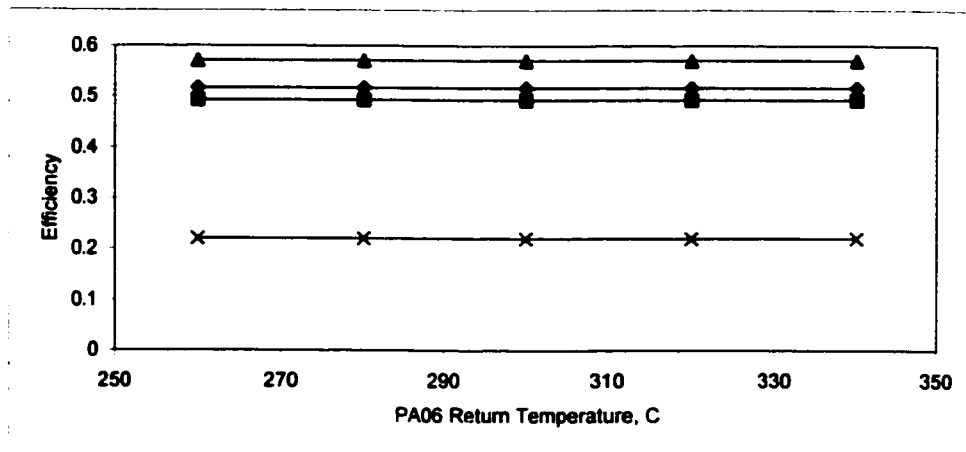


Figure 4.24 Energy and exergy efficiencies vs. pump-around 06 return temperature (▲ vacuum distillation unit energy efficiency, ◆ overall energy efficiency, ■ vacuum distillation unit exergy efficiency, × overall exergy efficiency).

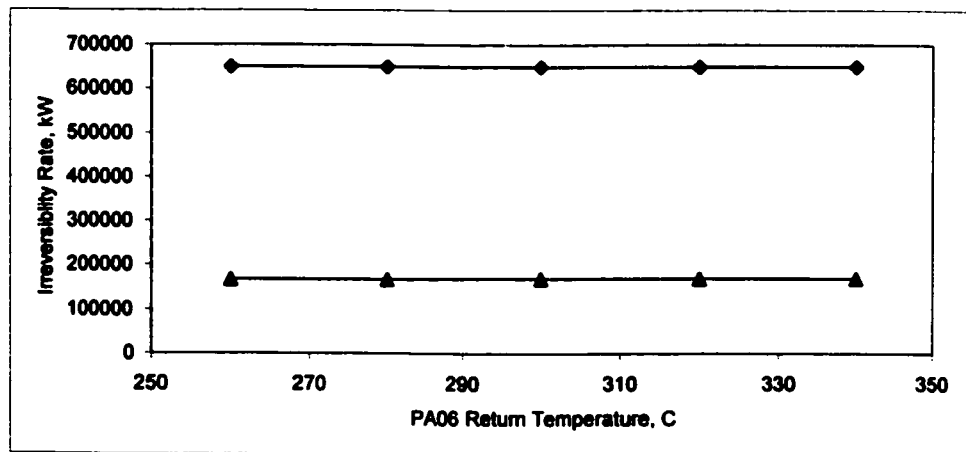


Figure 4.25 Irreversibility rate vs. pump-around 06 return temperature (◆ overall, ▲ vacuum distillation unit).

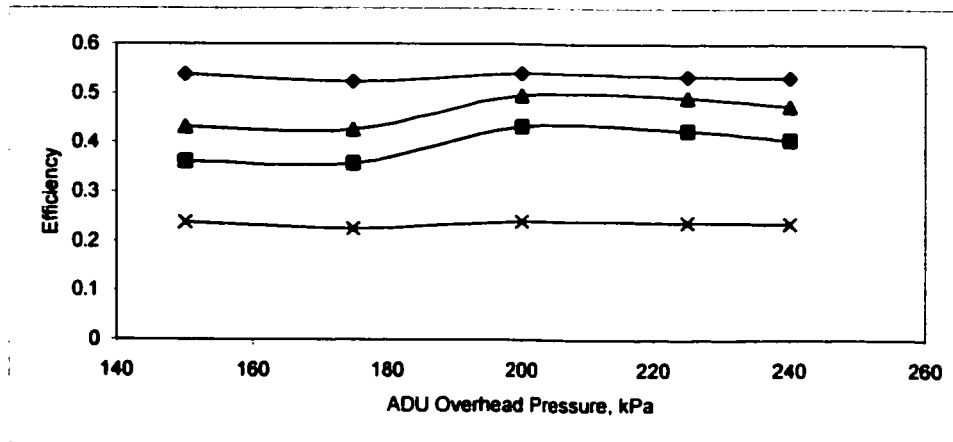


Figure 4.26 Energy and exergy efficiencies vs. atmospheric distillation unit overhead pressure (◆ overall energy efficiency, ▲ atmospheric distillation unit energy efficiency, ■ atmospheric distillation unit exergy efficiency, × overall exergy efficiency).

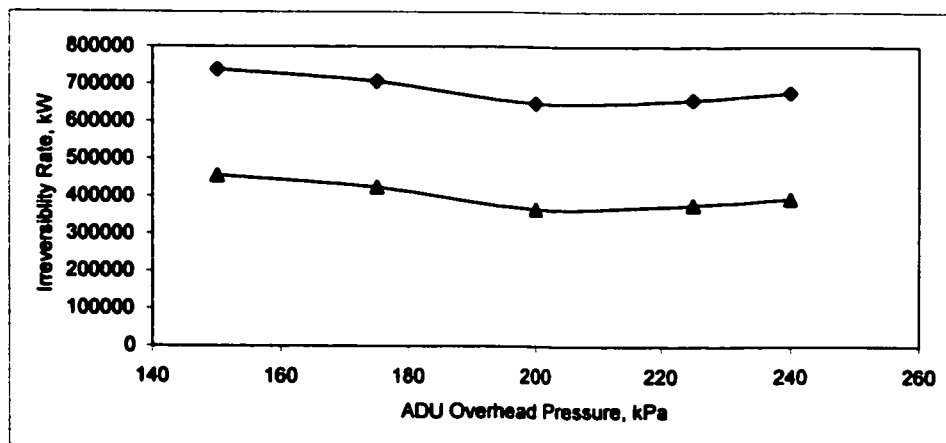


Figure 4.27 Irreversibility rate vs. atmospheric distillation unit overhead pressure (◆ overall, ▲ atmospheric distillation unit).

The operating overhead pressure is the optimum pressure. The energy efficiency at this pressure is 4% to 13% higher than the adjacent points, the exergy efficiency is 6% to 16% higher than the adjacent points, and the irreversibility rate is 5 to 21% lower than the adjacent points. This is because as we depart from the operating condition, more load is required by the overhead condenser to give the same yield products.

For the overall system, there is no significant change with regards to the energy and exergy efficiencies. However the irreversibility rate follows the same trend as that of the ADU with minimum irreversibilities at the operating condition which is 4% to 13% lower than the adjacent points.

#### 4.1.2.2 Vacuum Distillation Unit (VDU) Overhead Pressure

Effect of varying the VDU overhead pressure on the exergy efficiency is shown on Figure 4.28 while its effect on the irreversibility rate is shown on Figure 4.29.

The VDU overhead pressure has little effect on the energy efficiency, exergy efficiency and irreversibility rate of the VDU and the overall system. As for the VDU, the energy efficiency, exergy efficiency and the irreversibility rate variations are within 5%. The variation for the overall system energy efficiency, exergy efficiency and irreversibility rate are within 2%. This is because heat duty on the VDU overhead condenser is negligible and most of the heat duty is carried by the pump-around circuits.



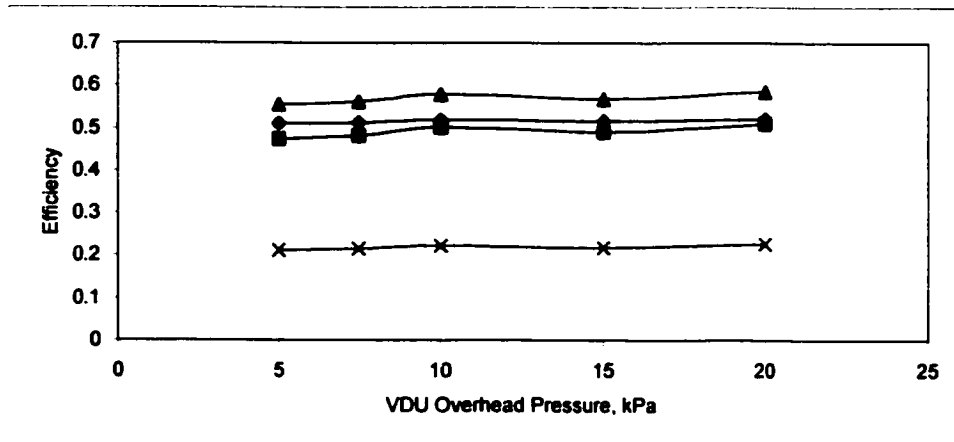


Figure 4.28 Energy and exergy efficiencies vs. vacuum distillation unit overhead pressure (▲ vacuum distillation unit energy efficiency, ◆ overall energy efficiency, ■ vacuum distillation unit exergy efficiency, × overall exergy efficiency).

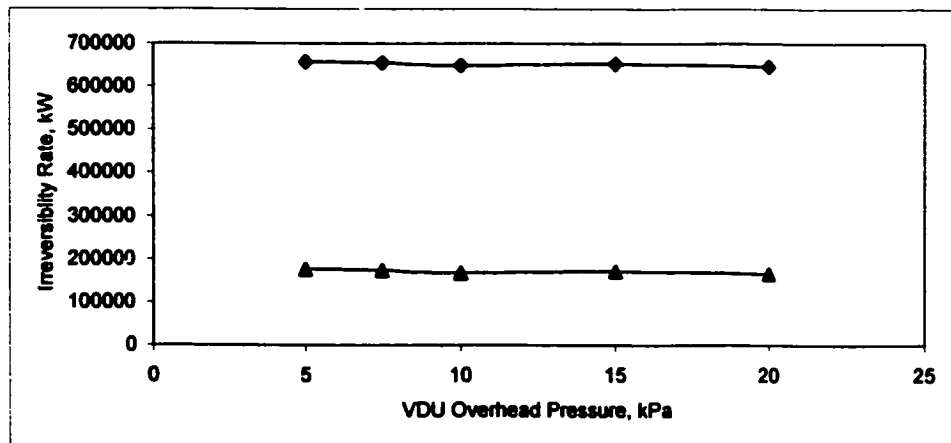


Figure 4.29 Irreversibility rate vs. vacuum distillation unit overhead pressure (◆ overall, ▲ vacuum distillation unit).

## **4.2 EFFECT OF MODIFYING DISTILLATION PLANT MODEL**

### **4.2.1 Changing Pump-Around Circuits To Pump-Back Circuits**

#### **4.2.1.1 Atmospheric Distillation Unit (ADU)**

Figures 4.30 through 4.38 illustrate varying the flow rates of the pump-back circuits on the energy efficiency, exergy efficiency and irreversibility rate of the ADU and overall system. It also compares the results to that of the pump-around circuits.

By changing the pump-around circuits to pump-back circuits, The ADU energy efficiency increases from 15% to 30%, the exergy efficiency increases from 19% to 32%. However, the side effect of that is the overall energy efficiency decreases by about 9% and the exergy efficiency decreases by 10% to 16%. The reason is that looking into the ADU individually the energy output and exergy output increase while the energy input and exergy input remain constant leading to a higher efficiency. The energy output and exergy output are considered inputs to HEATER 2 and the VDU. Because their energy and exergy outputs remain constant, their efficiencies decrease. The overall effect of this is a decrease in the system efficiency.

As for the irreversibility rate, there is a decrease of about 16% to 28% for the ADU and 9% to 17% for the overall system. The amount of decrease in irreversibility is equal but the original irreversibility of the ADU is less than that of the overall system and thus the percentage decrease is higher.

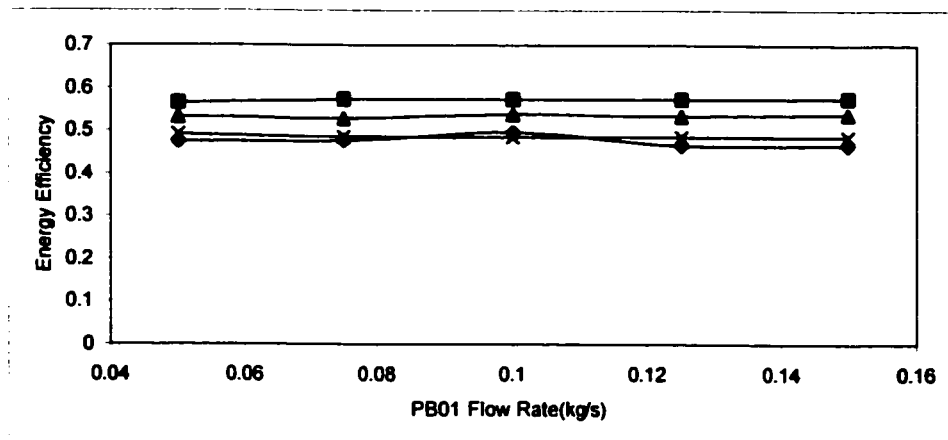


Figure 4.30 Energy efficiency vs. pump-back O1 mass flow rate (■ modified plant: atmospheric distillation unit, ▲ conventional plant: overall, × modified plant: overall, ◆ conventional plant: atmospheric distillation unit).

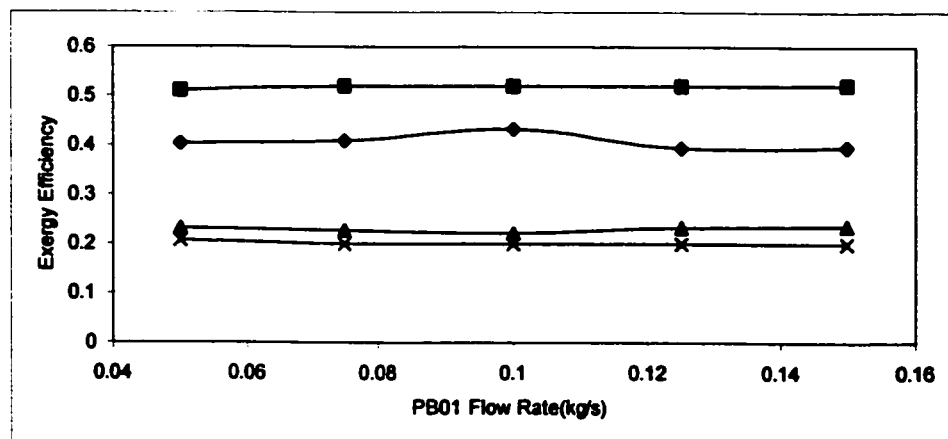


Figure 4.31 Exergy efficiency vs. pump-back O1 mass flow rate (■ modified plant: atmospheric distillation unit, ◆ conventional plant: atmospheric distillation unit, ▲ conventional plant: overall, × modified plant: overall).

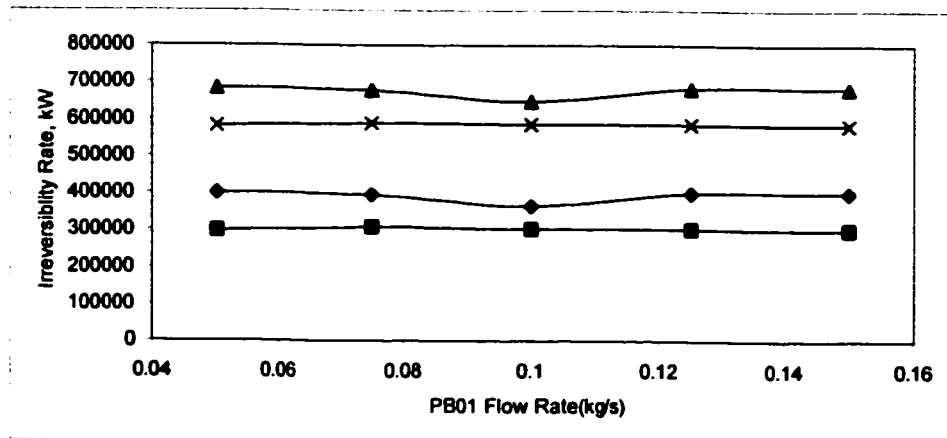


Figure 4.32 Irreversibility rate vs. pump-back 01 mass flow rate (▲ conventional plant: overall, × modified plant: overall, ◆ conventional plant: atmospheric distillation unit, ■ modified plant: atmospheric distillation unit).

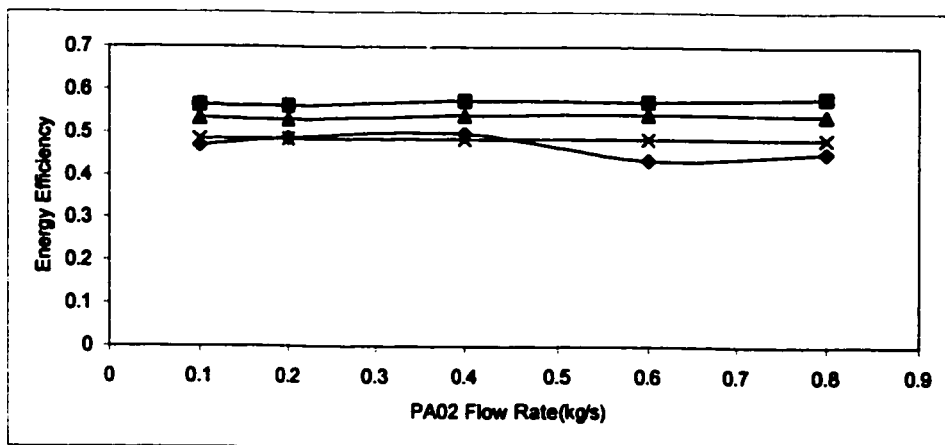


Figure 4.33 Energy efficiency vs. pump-back 02 mass flow rate (■ modified plant: atmospheric distillation unit, ▲ conventional plant: overall, × modified plant: overall, ◆ conventional plant: atmospheric distillation unit).

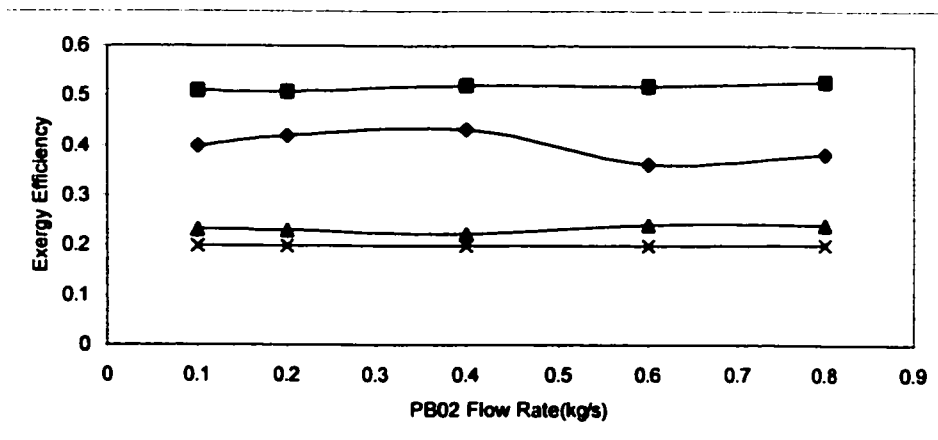


Figure 4.34 Exergy efficiency vs. pump-back O<sub>2</sub> mass flow rate (■ modified plant: atmospheric distillation unit, ◆ conventional plant: atmospheric distillation unit, ▲ conventional plant: overall, × modified plant: overall).

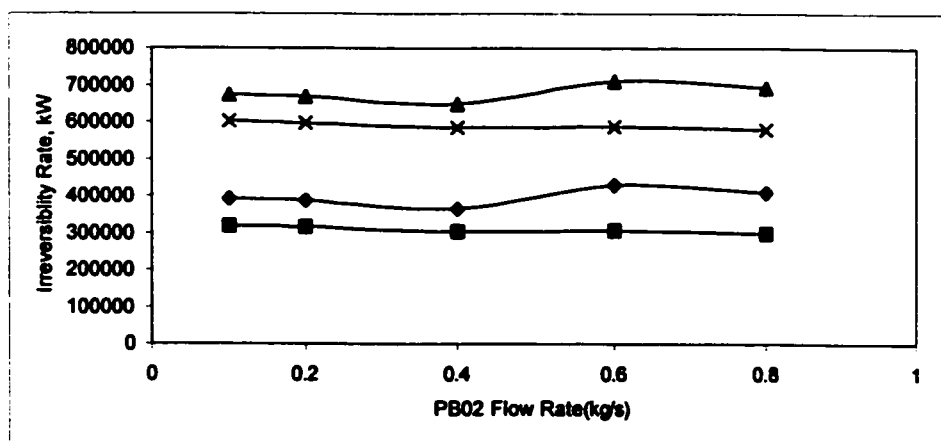


Figure 4.35 Irreversibility rate vs. pump-back O<sub>2</sub> mass flow rate (▲ conventional plant: overall, × modified plant: overall, ◆ conventional plant: atmospheric distillation unit, ■ modified plant: atmospheric distillation unit).

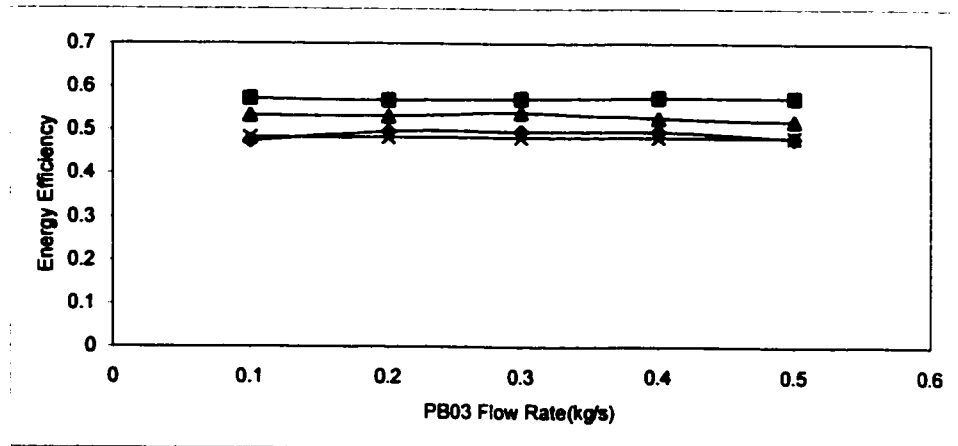


Figure 4.36 Energy efficiency vs. pump-back 03 mass flow rate (■ modified plant: atmospheric distillation unit, ▲ conventional plant: overall, × modified plant: overall, ◆ conventional plant: atmospheric distillation unit).

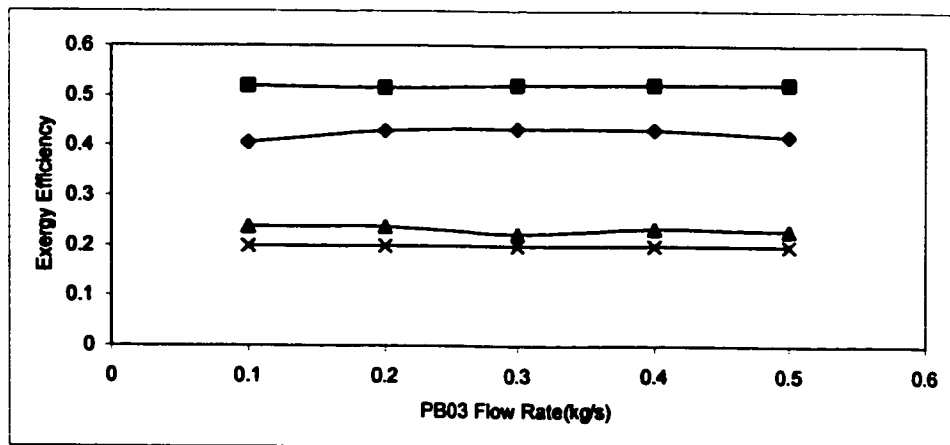


Figure 4.37 Exergy efficiency vs. pump-back 03 mass flow rate (■ modified plant: atmospheric distillation unit, ◆ conventional plant: atmospheric distillation unit, ▲ conventional plant: overall, × modified plant: overall).

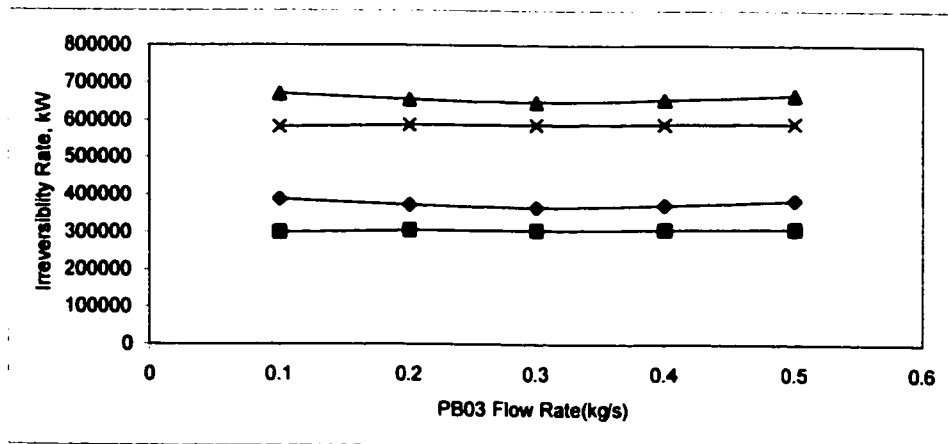


Figure 4.38 Irreversibility rate vs. pump-back 03 mass flow rate (▲ conventional plant: overall, × modified plant: overall, ◆ conventional plant: atmospheric distillation unit, ■ modified plant: atmospheric distillation unit).

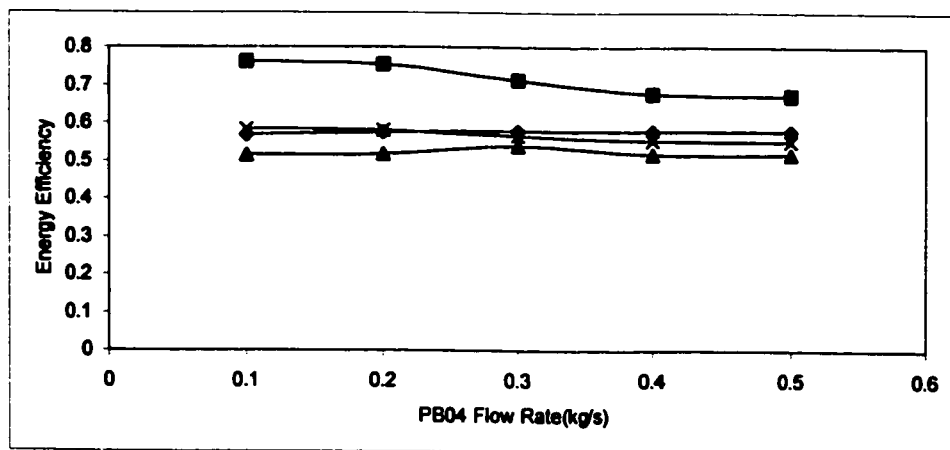


Figure 4.39 Energy efficiency vs. pump-back 04 mass flow rate (■ modified plant: vacuum distillation unit, ◆ conventional plant: vacuum distillation unit, × modified plant: overall, ▲ conventional plant: overall).

#### **4.2.1.2 Vacuum Distillations Unit (VDU)**

Figures 4.39 through 4.47 illustrate varying the flow rates of the pump-back circuits on the energy efficiency, exergy efficiency and irreversibility rate of the VDU and overall system. It also compares the results to that of the pump-around circuits.

The energy efficiency of the VDU increases between 16% and 33% and the exergy efficiency of the VDU increases between 18% and 40%. Moreover, the energy efficiency of the overall system increases from 2% to 12% and the exergy efficiency of the overall system increases from 14% to 35%. This is because increase in the energy and exergy outputs of the VDU means increase of energy and exergy outputs of the overall system.

The irreversibility rate of the VDU decreases between 29% and 47%. This high percentage is due to small amount of irreversibility in the original condition leading that small amount of changes results in high percentage change. This is more seen by noting that the irreversibility rate of the overall system decreases only by 6% to 12%.

### **4.2.2 Distillation Column With Reboiler**

#### **4.2.2.1 Atmospheric Distillation Unit (ADU)**

Figures 4.48 through Figure 4.56 illustrate varying the flow rates of the pump-around circuits on the energy efficiency, exergy efficiency and irreversibility rate of the ADU with reboiler and overall system. It also compares the results to that of the distillation column without reboiler.



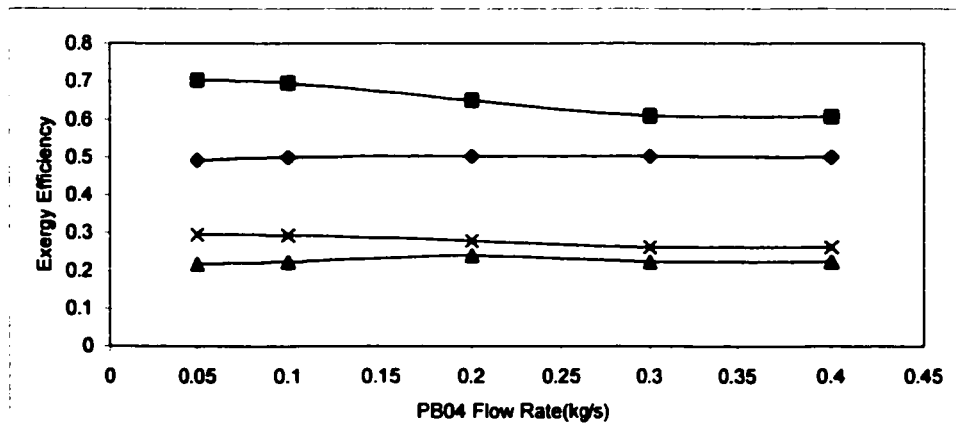


Figure 4.40 Exergy efficiency vs. pump-back 04 mass flow rate (■ modified plant: vacuum distillation unit, ◆ conventional plant: vacuum distillation unit, × modified plant: overall, ▲ conventional plant: overall).

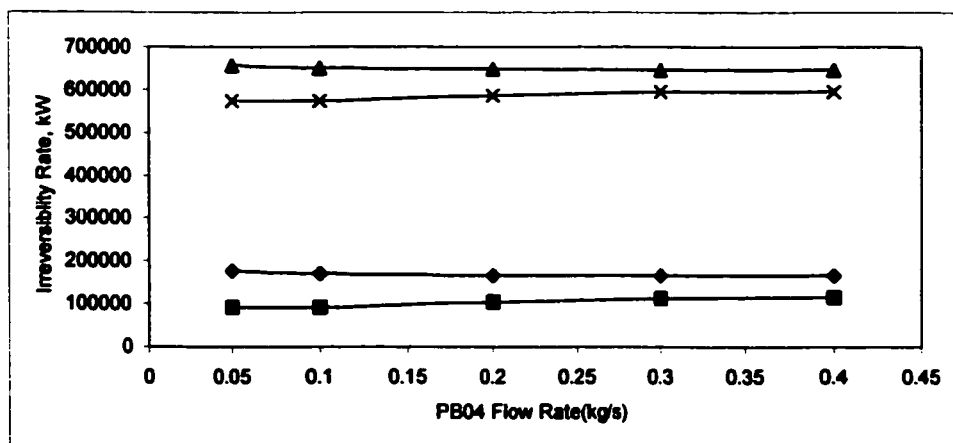


Figure 4.41 Irreversibility rate vs. pump-back 04 mass flow rate (▲ conventional plant: overall, × modified plant: overall, ◆ conventional plant: vacuum distillation unit, ■ modified plant: vacuum distillation unit).

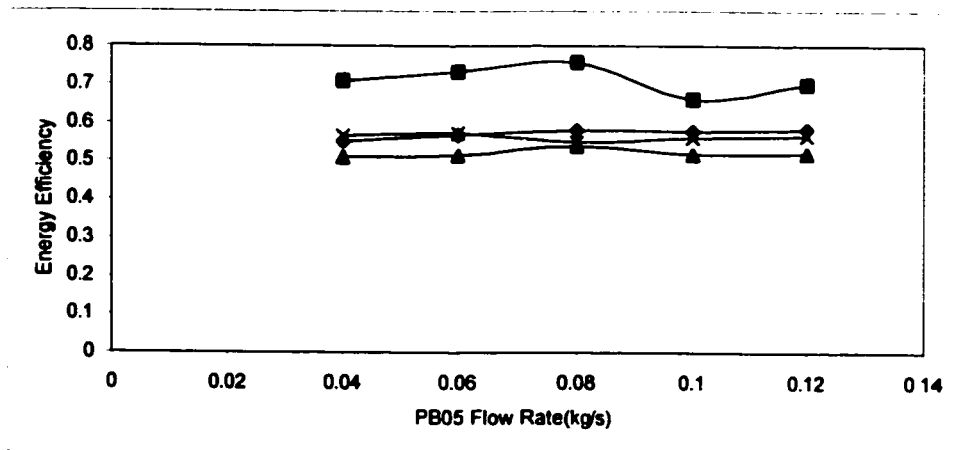


Figure 4.42 Energy efficiency vs. pump-back 05 mass flow rate (■ modified plant: vacuum distillation unit, ◆ conventional plant: vacuum distillation unit, × modified plant: overall, ▲ conventional plant: overall).

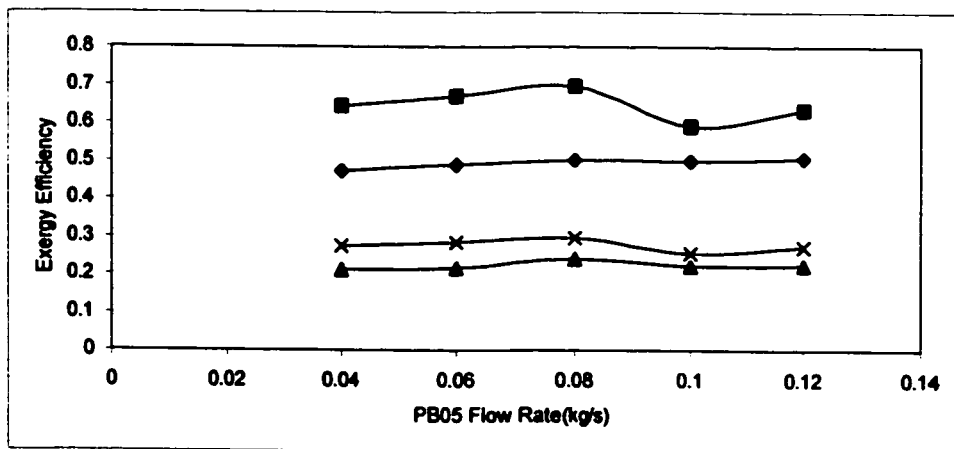


Figure 4.43 Exergy efficiency vs. pump-back 05 mass flow rate (■ modified plant: vacuum distillation unit, ◆ conventional plant: vacuum distillation unit, × modified plant: overall, ▲ conventional plant: overall).

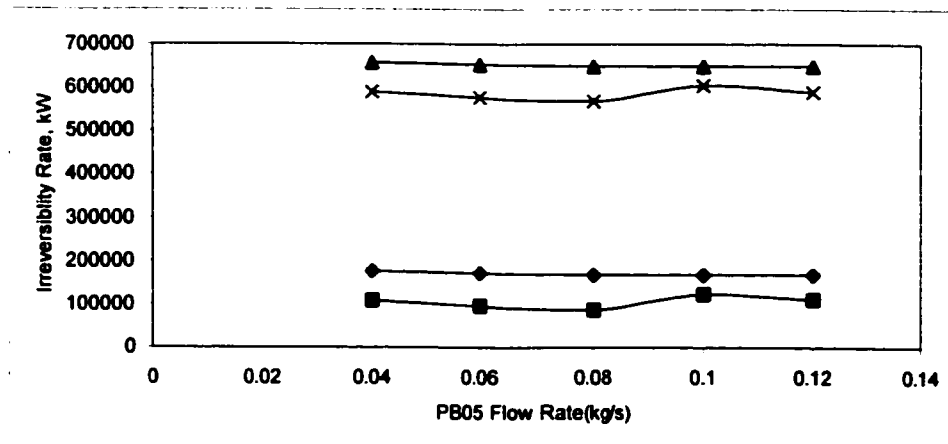


Figure 4.44 Irreversibility rate vs. pump-back 05 mass flow rate (▲ conventional plant: overall, × modified plant: overall, ◆ conventional plant: vacuum distillation unit, ■ modified plant: vacuum distillation unit).

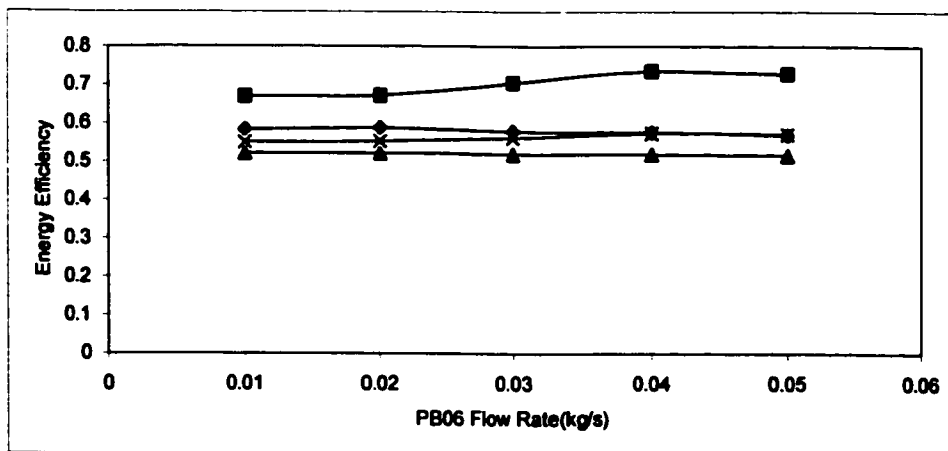


Figure 4.45 Energy efficiency vs. pump-back 06 mass flow rate (■ modified plant: vacuum distillation unit, ◆ conventional plant: vacuum distillation unit, × modified plant: overall, ▲ conventional plant: overall).

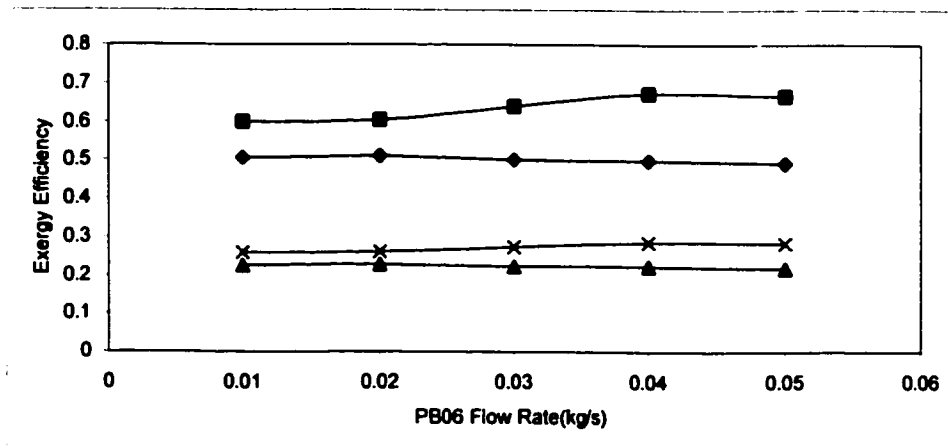


Figure 4.46 Exergy efficiency vs. pump-back 06 mass flow rate (■ modified plant: vacuum distillation unit, ◆ conventional plant: vacuum distillation unit, × modified plant: overall, ▲ conventional plant: overall).

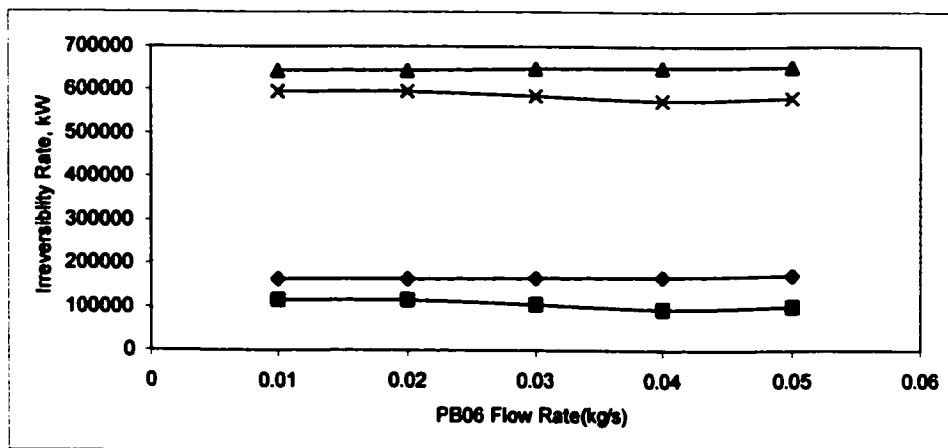


Figure 4.47 Irreversibility rate vs. pump-back 06 mass flow rate (▲ conventional plant: overall, × modified plant: overall, ◆ conventional plant: vacuum distillation unit, ■ modified plant: vacuum distillation unit).

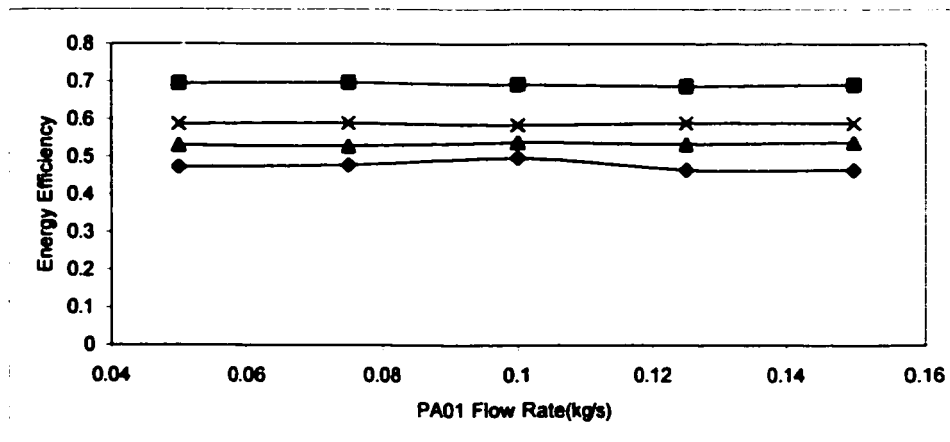


Figure 4.48 Vs. Energy efficiency vs. pump-around O1 mass flow rate of atmospheric distillation unit with reboiler (■ modified plant: atmospheric distillation unit, × modified plant: overall, ▲ conventional plant: overall, ◆ conventional plant: atmospheric distillation unit)

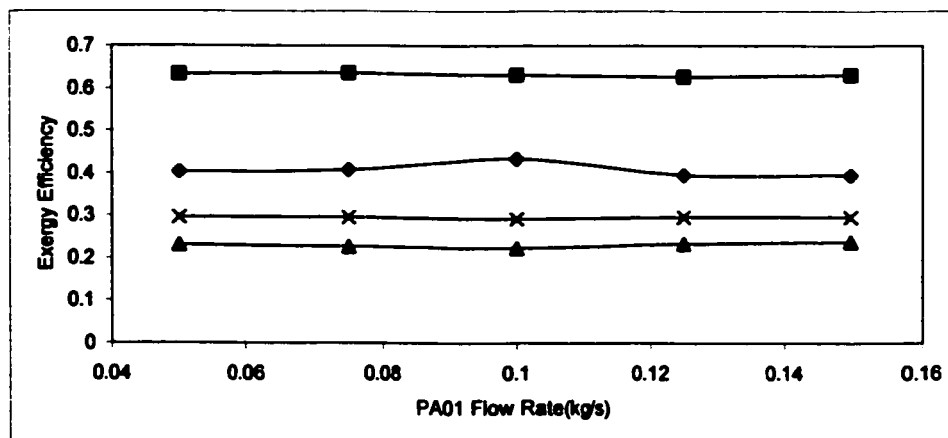


Figure 4.49 Exergy efficiency vs. pump-around O1 mass flow rate of atmospheric distillation unit with reboiler (■ modified plant: atmospheric distillation unit, ◆ conventional plant: atmospheric distillation unit, × modified plant: overall, ▲ conventional plant: overall)

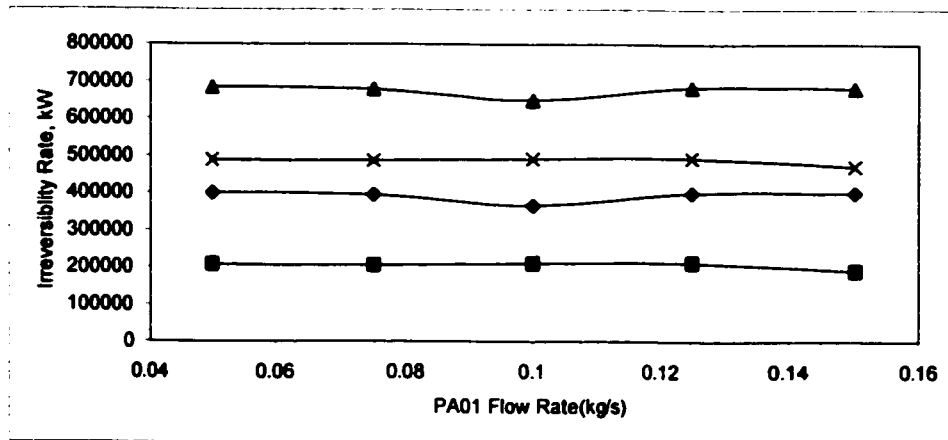


Figure 4.50 Irreversibility rate vs. pump-around 01 mass flow rate of atmospheric distillation unit with reboiler (▲ conventional plant: overall, × modified plant: overall, ◆ conventional plant: atmospheric distillation unit, ■ modified plant: atmospheric distillation unit).

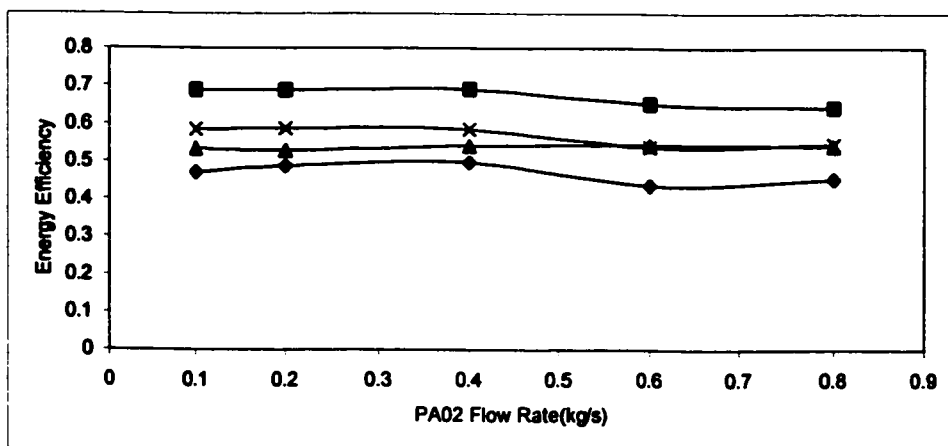


Figure 4.51 Energy efficiency vs. pump-around 02 mass flow rate of atmospheric distillation unit with reboiler (■ modified plant: atmospheric distillation unit, × modified plant: overall, ▲ conventional plant: overall, ◆ conventional plant: atmospheric distillation unit).

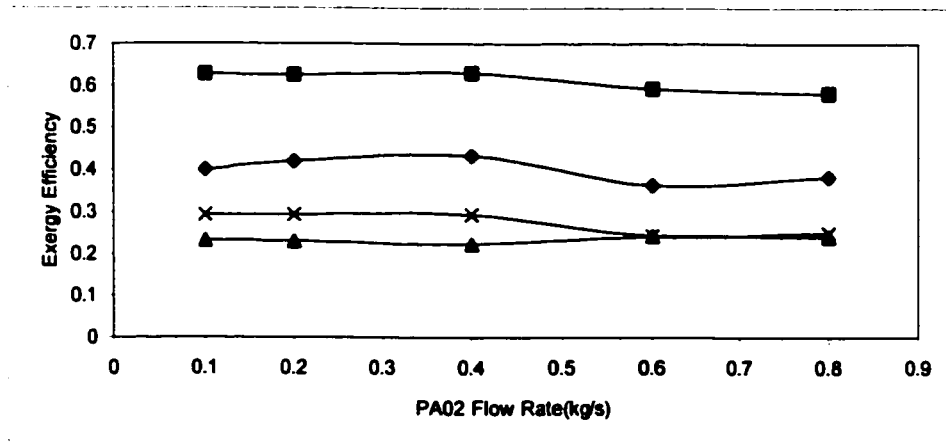


Figure 4.52 Exergy efficiency vs. pump-around O<sub>2</sub> mass flow rate of atmospheric distillation unit with reboiler (■ modified plant: atmospheric distillation unit, ◆ conventional plant: atmospheric distillation unit, × modified plant: overall, ▲ conventional plant: overall).

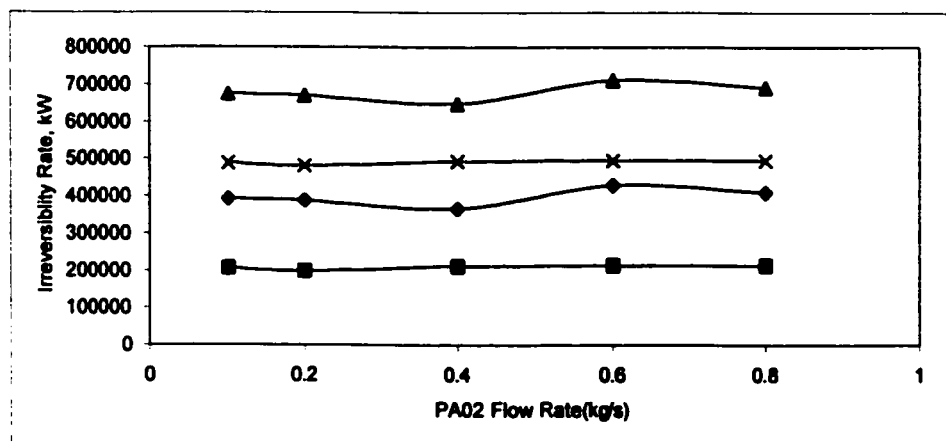


Figure 4.53 Irreversibility rate vs. pump-around O<sub>2</sub> mass flow rate of atmospheric distillation unit with reboiler (▲ conventional plant: overall, × modified plant: overall, ◆ conventional plant: atmospheric distillation unit, ■ modified plant: atmospheric distillation unit).

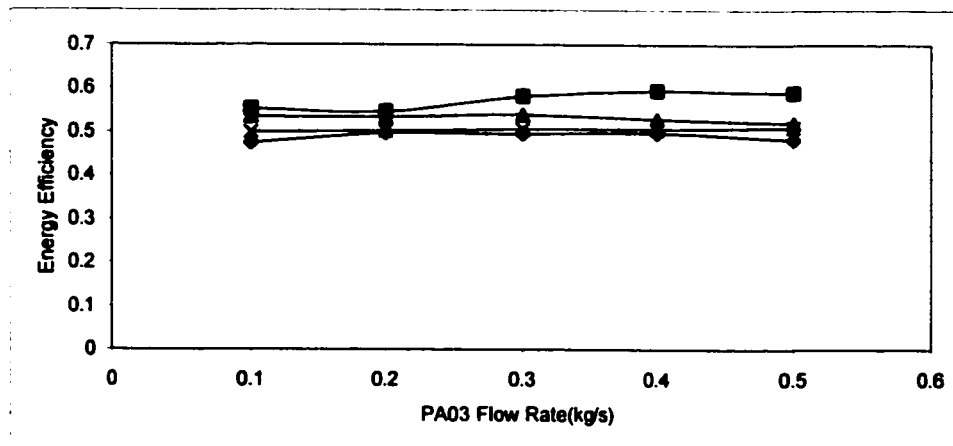


Figure 4.54 Energy efficiency vs. pump-around O<sub>3</sub> mass flow rate of atmospheric distillation unit with reboiler (■ modified plant: atmospheric distillation unit, × modified plant: overall, ▲ conventional plant: overall, ◆ conventional plant: atmospheric distillation unit).

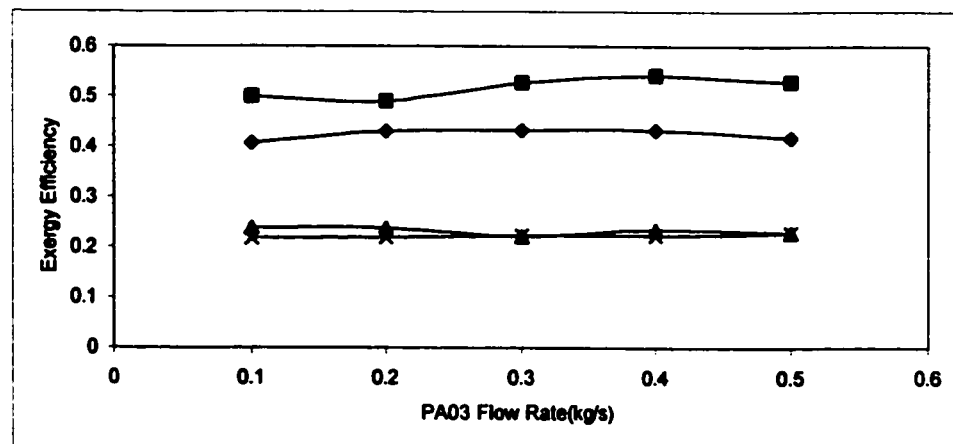


Figure 4.55 Exergy efficiency vs. pump-around O<sub>3</sub> mass flow rate of atmospheric distillation unit with reboiler (■ modified plant: atmospheric distillation unit, ◆ conventional plant: atmospheric distillation unit, × modified plant: overall, ▲ conventional plant: overall).



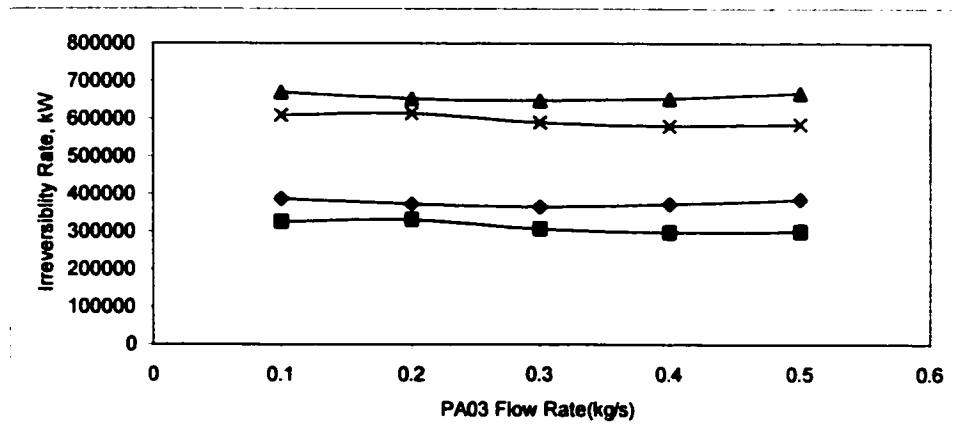


Figure 4.56 Irreversibility rate vs. pump-around 03 mass flow rate of atmospheric distillation unit with reboiler (▲ conventional plant: overall, × modified plant: overall, ◆ conventional plant: atmospheric distillation unit, ■ modified plant: atmospheric distillation unit).

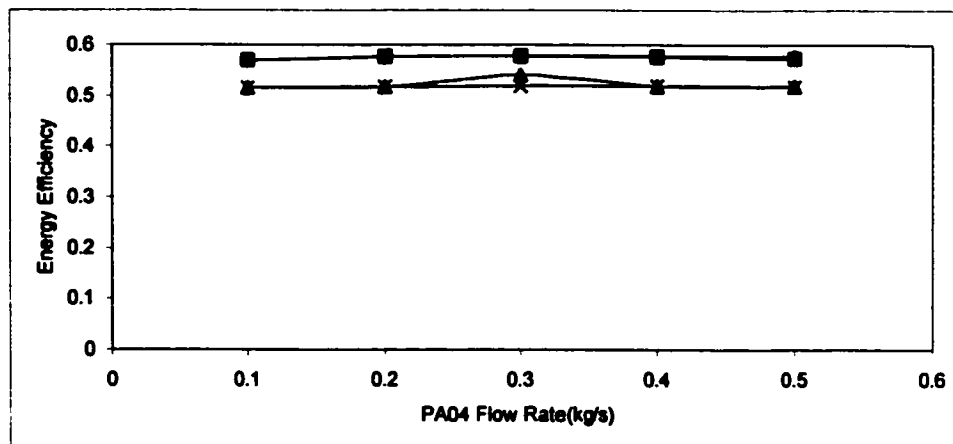


Figure 4.57 Energy efficiency vs. pump-around 04 mass flow rate of vacuum distillation unit with reboiler (■ modified plant: vacuum distillation unit, × modified plant: overall, ▲ conventional plant: overall, ◆ conventional plant: vacuum distillation unit)

The energy efficiency of the ADU with reboiler increases between 9% and 22% while for the overall system it increases up to 10%. The exergy efficiency increases between 14% and 46% while for the overall system it increases up to 31%. This is due to the energy and exergy outputs increase of the ADU and thus of the overall system.

The irreversibility rate of the ADU with reboiler decreases between 10% and 52% while the irreversibility rate of the overall system decreases between 8% and 31%. The amount of decrease is the same for the ADU and the overall system. However, in terms of percentage difference, it is higher for the ADU because we are dividing by the ADU irreversibility rate (315 MW) while for the overall system we are dividing by the overall irreversibility rate (608 MW).

#### 4.2.2.2 Vacuum Distillation Unit (VDU)

Figures 4.57 through 4.65 illustrate the effect of varying the flow rates of the pump-around circuits on the energy efficiency, exergy efficiency and irreversibility rate of the VDU with reboiler and overall system. It also compares the results to that of the distillation column without reboiler.

There is not significant change to the energy efficiency and exergy efficiency of the VDU and the overall system after adding the reboiler. The variations are within  $\pm 5\%$ . This is because the components of the VDU are mostly heavy and there is minimum duty on the reboiler to strip the light components.

The irreversibility rate of the VDU decreases slightly between 8% and 11% while the irreversibility rate decrease for the overall system is within 4% which is not significant.

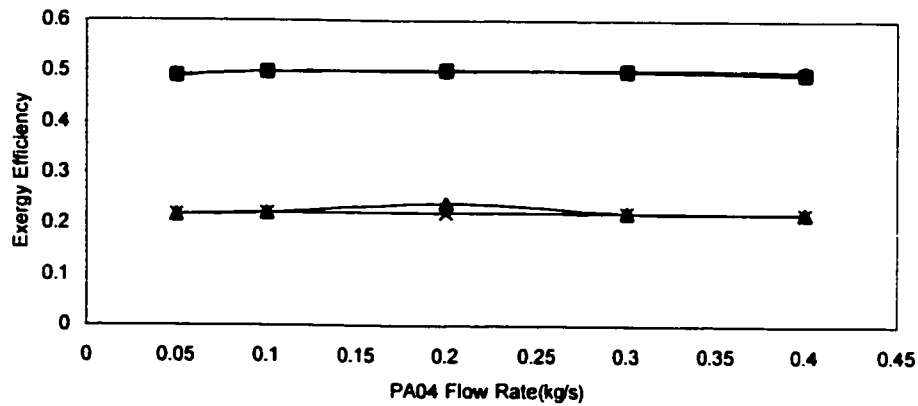


Figure 4.58 Exergy efficiency vs. pump-around 04 mass flow rate of vacuum distillation unit with reboiler (■ modified plant: vacuum distillation unit, ◆ conventional plant: vacuum distillation unit, × modified plant: overall, ▲ conventional plant: overall).

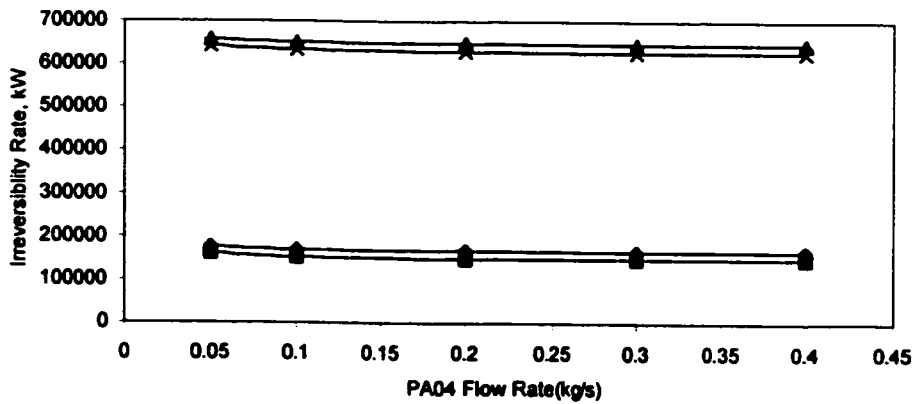


Figure 4.59 Irreversibility rate vs. pump-around 04 mass flow rate of vacuum distillation unit with reboiler (▲ conventional plant: overall, × modified plant: overall, ◆ conventional plant: vacuum distillation unit, ■ modified plant: vacuum distillation unit).

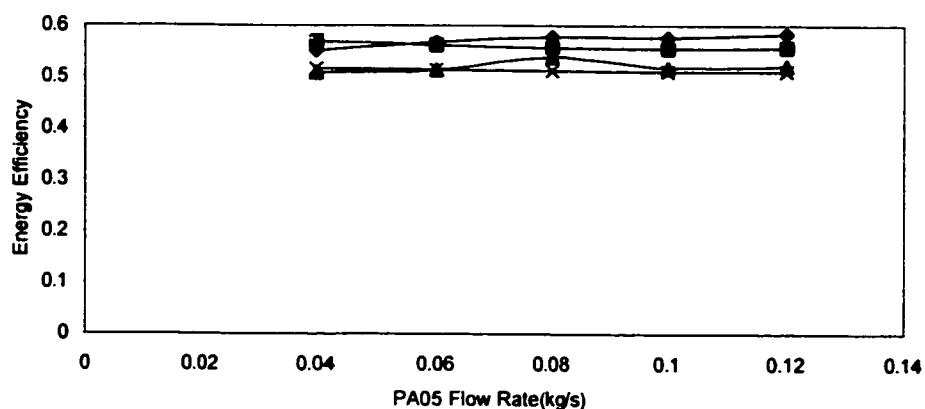


Figure 4.60 Energy efficiency vs. pump-around 05 mass flow rate of vacuum distillation unit with reboiler (■ modified plant: vacuum distillation unit, × modified plant: overall, ▲ conventional plant: overall, ◆ conventional plant: vacuum distillation unit).

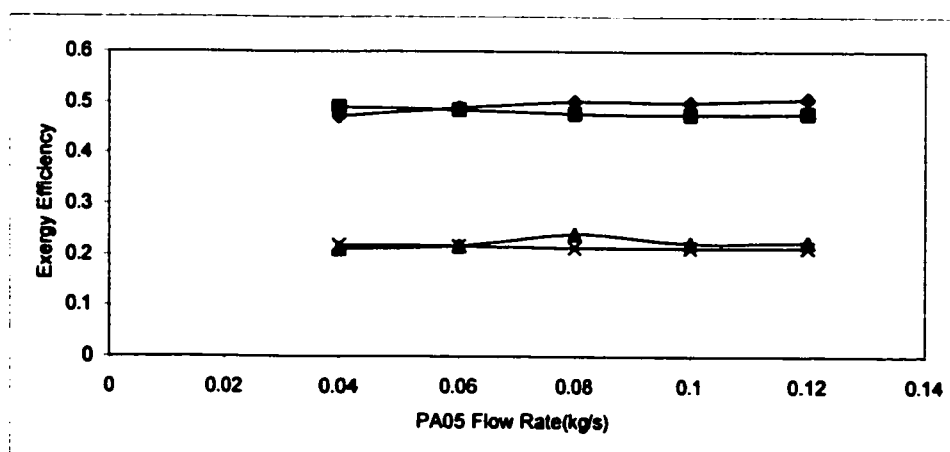


Figure 4.61 Exergy efficiency vs. pump-around 05 mass flow rate of vacuum distillation unit with reboiler (■ modified plant: vacuum distillation unit, ◆ conventional plant: vacuum distillation unit, × modified plant: overall, ▲ conventional plant: overall).

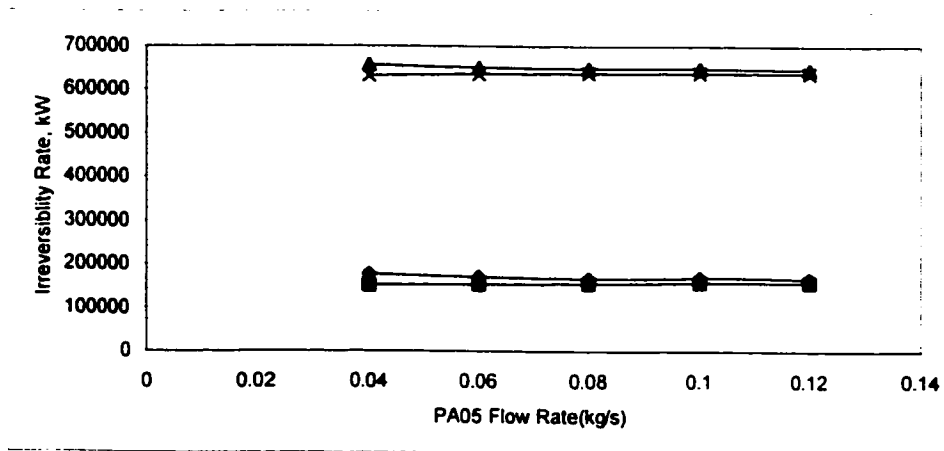


Figure 4.62 Irreversibility rate vs. pump-around 05 mass flow rate of vacuum distillation unit with reboiler (▲ conventional plant: overall, × modified plant: overall, ◆ conventional plant: vacuum distillation unit, ■ modified plant: vacuum distillation unit).

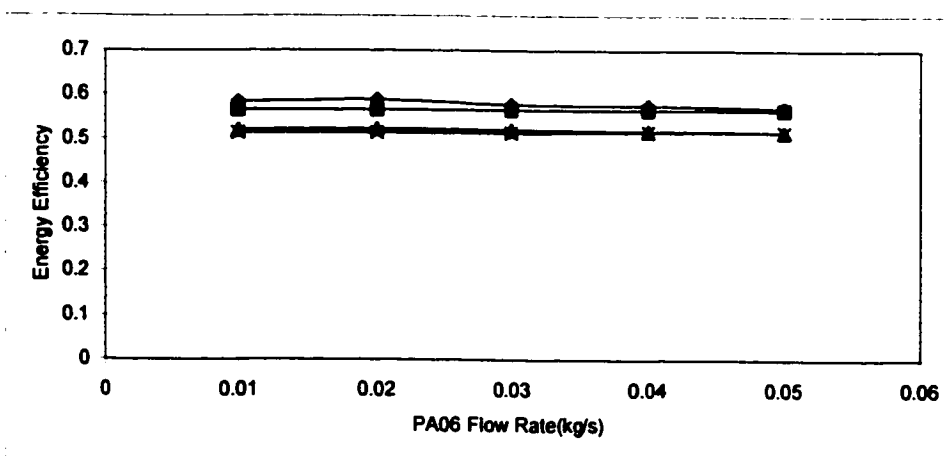


Figure 4.63 Energy efficiency vs. pump-around 06 mass flow rate of vacuum distillation unit with reboiler (■ modified plant: vacuum distillation unit, × modified plant: overall, ▲ conventional plant: overall, ◆ conventional plant: vacuum distillation unit).

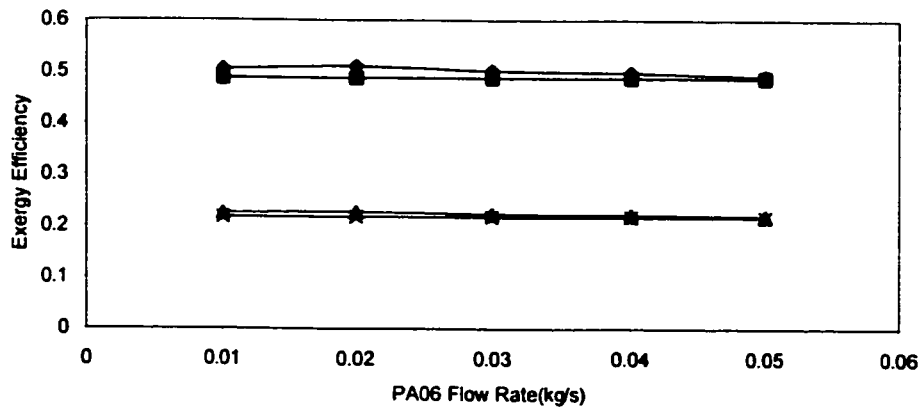


Figure 4.64 Exergy efficiency vs. pump-around 06 mass flow rate of vacuum distillation unit with reboiler (■ modified plant: vacuum distillation unit, ◆ conventional plant: vacuum distillation unit, × modified plant: overall, ▲ conventional plant: overall).

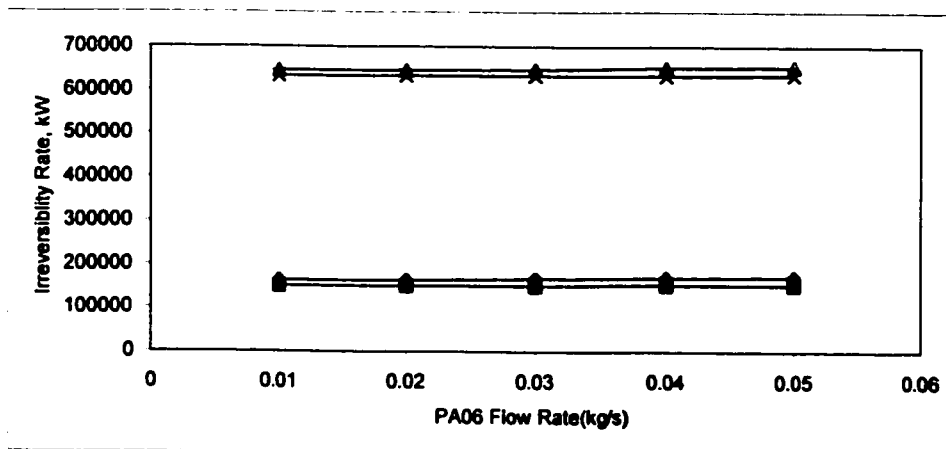


Figure 4.65 Irreversibility rate vs. pump-around 06 mass flow rate of vacuum distillation unit with reboiler (▲ conventional plant: overall, × modified plant: overall, ◆ conventional plant: vacuum distillation unit, ■ modified plant: vacuum distillation unit).

### **4.2.3 Two-Stage Atmospheric Distillation Unit (ADU)**

Figures 4.66 through 4.74 illustrate the effect of varying the flow rate of the pump-around circuits on the energy efficiency, exergy efficiency and irreversibility rate of the two-stage ADU and compare it to that of the single-stage ADU.

The energy efficiency of the two-stage ADU is 53% to 68% higher than that of the single-stage ADU. This also results in 9% to 13% increase in the overall energy efficiency.

The exergy efficiency of the two-stage ADU is 46% to 66% higher than that of the single-stage ADU. This also results in 44% to 58% increase in the overall exergy efficiency.

The irreversibility rate of the two-stage ADU decreases by about 45% to 50% compared to that of the single-stage ADU. The decrease of irreversibility rate of the overall system is about 35% to 40%.

This enhancement is due to two reasons. The first is that the lighter components are only heated to 250° C not to 350° C as in the case for the single-stage ADU. This results in less heater duty and thus less fuel consumption. The second reason is that the temperature difference in the two-stage distillation ADU units is less than that of the single-stage distillation. It is important to note that temperature difference is a direct measure of exergy. The higher the temperature difference in a unit the more exergy losses are generated.

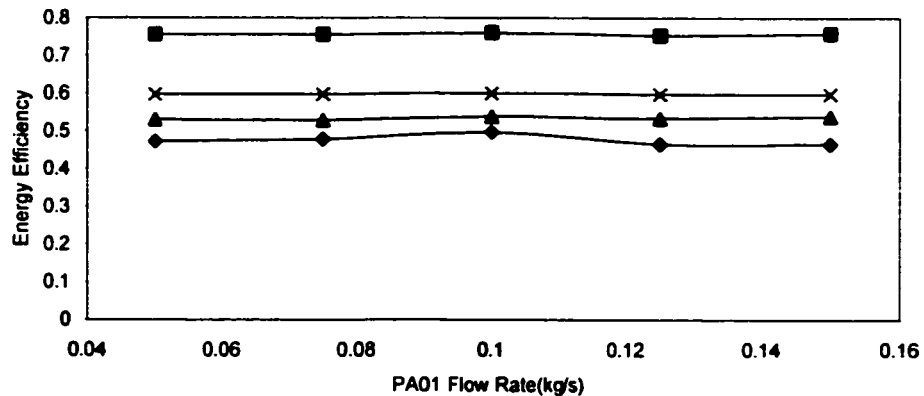


Figure 4.66 Energy efficiency vs. pump-around O1 mass flow rate of two-stage atmospheric distillation unit (■ modified plant: atmospheric distillation unit, × modified plant: overall, ▲ conventional plant: overall, ◆ conventional plant: atmospheric distillation unit).

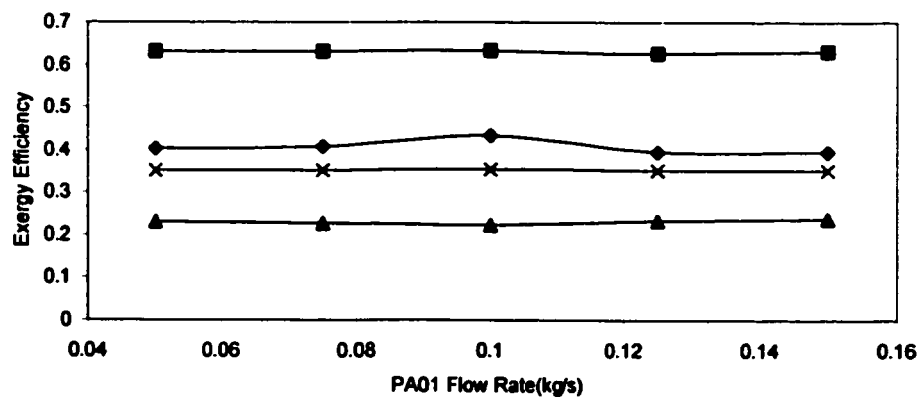


Figure 4.67 Exergy efficiency vs. pump-around O1 mass flow rate of two-stage atmospheric distillation unit (■ modified plant: atmospheric distillation unit, ◆ conventional plant: atmospheric distillation unit, × modified plant: overall, ▲ conventional plant: overall).



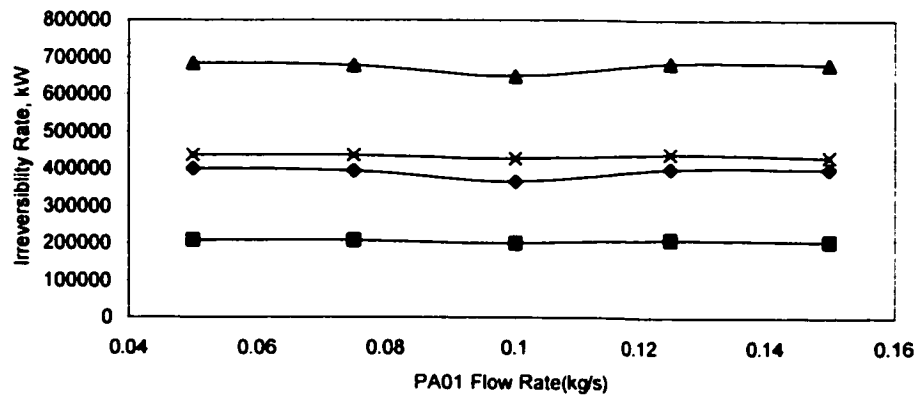


Figure 4.68 Irreversibility rate vs. pump-around 01 mass flow rate of two-stage atmospheric distillation unit (▲ conventional plant: overall, × modified plant: overall, ◆ conventional plant: atmospheric distillation unit, ■ modified plant: atmospheric distillation unit).

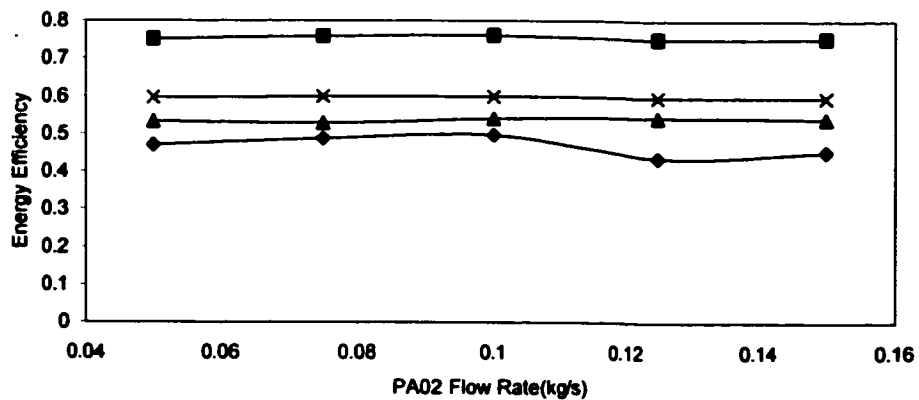


Figure 4.69 Energy efficiency vs. pump-around 02 mass flow rate of two-stage atmospheric distillation unit (■ modified plant: atmospheric distillation unit, × modified plant: overall, ▲ conventional plant: overall, ◆ conventional plant: atmospheric distillation unit).

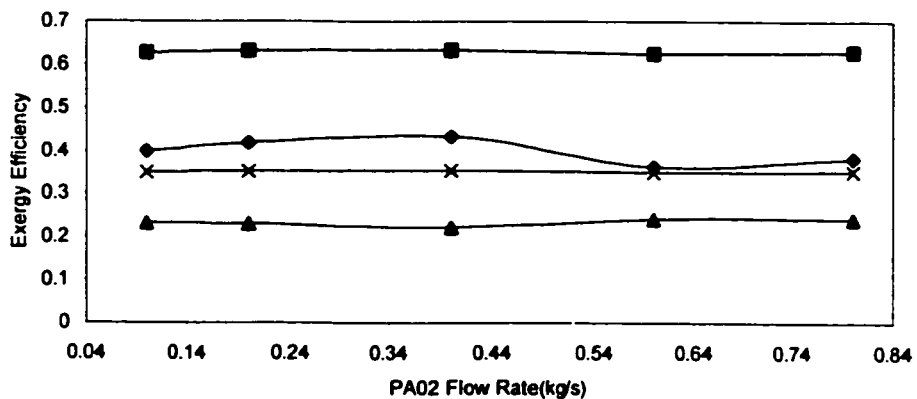


Figure 4.70 Exergy efficiency vs. pump-around O<sub>2</sub> mass flow rate of two-stage atmospheric distillation unit (■ modified plant: atmospheric distillation unit, ◆ conventional plant: atmospheric distillation unit, × modified plant: overall, ▲ conventional plant: overall)

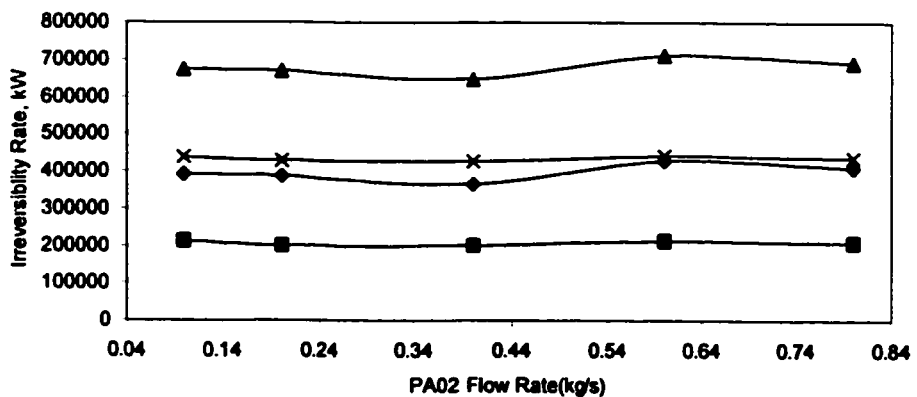


Figure 4.71 Irreversibility rate vs. pump-around O<sub>2</sub> mass flow rate of two-stage atmospheric distillation unit (▲ conventional plant: overall, × modified plant: overall, ◆ conventional plant: atmospheric distillation unit, ■ modified plant: atmospheric distillation unit).

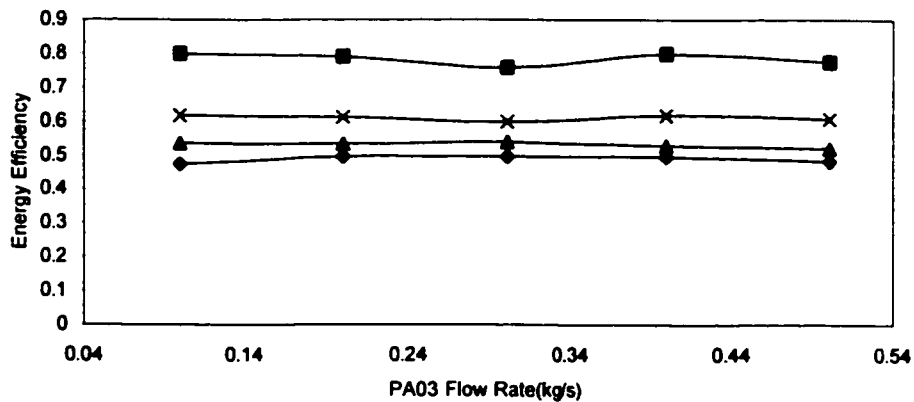


Figure 4.72 Energy efficiency vs. pump-around O3 mass flow rate of two-stage atmospheric distillation unit (■ modified plant: atmospheric distillation unit, × modified plant: overall, ▲ conventional plant: overall, ◆ conventional plant: atmospheric distillation unit).

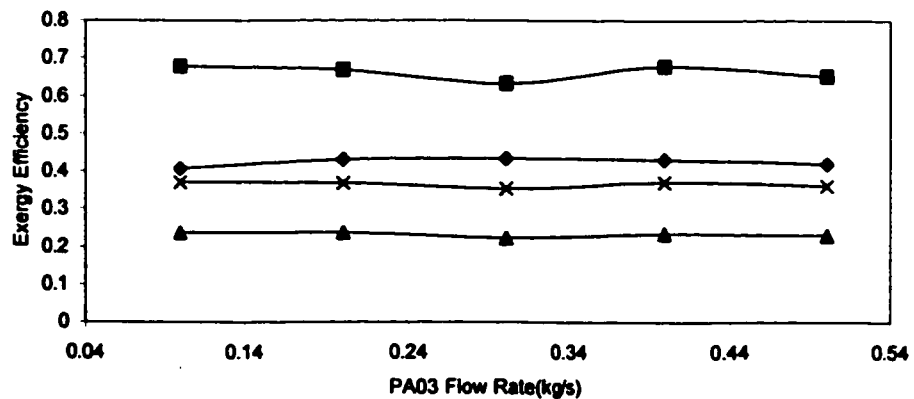


Figure 4.73 Exergy efficiency vs. pump-around O3 mass flow rate of two-stage atmospheric distillation unit (■ modified plant: atmospheric distillation unit, ◆ conventional plant: atmospheric distillation unit, × modified plant: overall, ▲ conventional plant: overall)

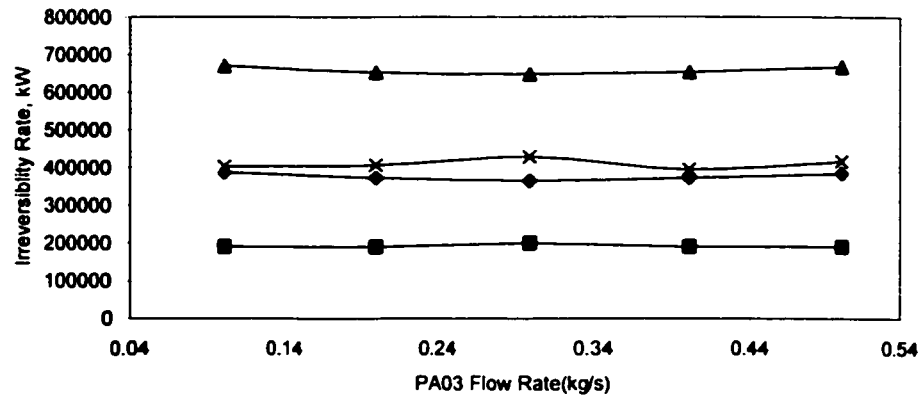


Figure 4.74 Irreversibility rate vs. pump-around O3 mass flow rate of two-stage atmospheric distillation unit ( $\blacktriangle$  conventional plant: overall,  $\times$  modified plant: overall,  $\blacklozenge$  conventional plant: atmospheric distillation unit,  $\blacksquare$  modified plant: atmospheric distillation unit).

## CHAPTER 5

# CONCLUSIONS AND RECOMMENDATIONS

### 5.1 CONCLUSIONS

In this thesis, energy and exergy analyses are conducted on crude oil distillation plant model. The model consists of two heaters, atmospheric distillation column (ADU) and vacuum distillation column (VDU). 144 simulations are run to find the effect of the operating conditions as well as modifying the plant model on the energy efficiencies, exergy efficiencies and irreversibility rate of the individual components and the overall system.

The first simulation is carried for the operating conditions of the model. The total irreversibility losses are 608 MW. 56% are contributed by the ADU, 26% by the VDU and 18% by the heaters. The highest irreversibility losses occur in the ADU as the main separation takes place there. 6.2% of the losses are due to chemical exergy losses associated with the separation process itself. The rest of the losses are due to the physical exergy losses mainly because of the temperature difference. The energy efficiencies are 0.497 for the ADU, 0.579 for the VDU and 0.519 for the overall system. The exergy efficiencies are 0.433 for the ADU, 0.501 for the VDU, 0.821 for Heater 1, 0.956 for Heater 2 and 0.233 for the overall system. The overall efficiency is not equal to the product of the components efficiencies.

For the operating conditions, two cases are studied: varying the temperature profile of the columns and varying the pressure profile of the columns. The temperature profile and pressure profile of the ADU at operating condition gives optimum energy

efficiency, exergy efficiencies and irreversibility rate. Changing the temperature profile and pressure profile of the VDU does not give significant changes in the energy efficiencies, exergy efficiencies and irreversibility rate.

The first modification on the model is to change the pump-around circuits to pump-back circuits. For this case, heaters should be added to the pump-back circuit whereas this is not needed in the pump-around circuits, as the stream should be cooled using the crude oil feed. This also serves for preheating the crude oil feed and minimizing the heaters duty. The ADU energy efficiency increases from 15% to 30%, the exergy efficiency increases from 19% to 32%. However, the side effect of that is the overall energy efficiency decreases by about 9% and the exergy efficiency decreases by 10% to 16%. The irreversibility rate decreases by about 16% to 28% for the ADU and 9% to 17% for the overall system.

Changing the pump-around circuits to pump-back circuits in the VDU increases the energy efficiency between 16% and 33% and the exergy efficiency between 18% and 40%. Moreover, the energy efficiency of the overall system increases from 2% to 12% and the exergy efficiency of the overall system increases from 14% to 35%. The irreversibility rate of the VDU decreases between 29% and 47% while for the overall system it decreases only by 6% to 12%.

The second modification is to add a reboiler on the bottom tray of the distillation columns. The energy efficiency of the ADU with reboiler increases between 9% and 22% while for the overall system it increases up to 10%. The exergy efficiency increases between 14% and 46% while for the overall system it increases up to 31%. The irreversibility rate decreases between 10% and 52% while the irreversibility rate of the overall system decreases between 8% and 31%.

There is not significant change to the energy efficiency and exergy efficiency of the VDU and the overall system after adding the reboiler. The variations are within  $\pm 5\%$ . The irreversibility rate of the VDU decreases slightly between 8% and 11% while the irreversibility rate decrease for the overall system is within 4% which is not significant.

The final modification is to split the separation occurring in the ADU into two stages. The energy efficiency of the two-stage ADU is 53% to 68% higher than that of the single-stage ADU. The overall energy efficiency of the distillation plant increases from 9% to 13%. The exergy efficiency of the two-stage ADU is 46% to 66% higher than that of the single-stage ADU. This also results in 44% to 58% increase in the overall exergy efficiency. The irreversibility rate of the two-stage ADU decreases by about 45% to 50% compared to that of the single-stage ADU. The decrease of irreversibility rate of the overall system is about 35% to 40%.

## **5.2 RECOMMENDATIONS**

1. Optimization should be focused on the ADU as it represents 56% of the total irreversibility losses of the plant mode. Moreover, The enhancements are generally better and more significant in the ADU compared to the VDU.
2. Operating conditions should be kept the same as they give the optimum energy efficiencies, exergy efficiencies and irreversibility rates.
3. Pump-back circuits should not be added to the ADU as they result in a decrease of the overall plant energy and exergy efficiencies.
4. Adding pump-back circuits to the VDU should be considered by conducting thermo-economic analysis to compare the added cost of the heaters and the effect

on the product yield (plant profit) to the enhancement of the energy and exergy efficiencies.

5. Adding reboiler to the ADU should be considered by conducting thermo-economic analysis to compare the added cost of reboiler and the effect on the product yield (plant profit) to the enhancement of the energy and exergy efficiencies.
6. Reboiler should not be added to the VDU as it does not give significant changes on the energy and exergy efficiencies and irreversibility rates.
7. In future designs, having two-stage ADU rather than single-stage ADU should be considered by conducting thermo-economic analysis to compare the added cost of two smaller columns and two smaller heaters instead of one larger column and one larger heater, and the effect on the product yield (plant profit) to the enhancement of the exergy efficiency.

The cost of a recent built distillation plant is \$ 125,000,000. Note that this cost may vary from one company to another and from one location to another. Changing pump-around circuits to pump-back circuits may add 2% to 5% of the initial cost. Adding a reboiler to the distillation columns could add up to 2% of the initial cost. Building a two-stage atmospheric distillation column is expected to increase the initial cost 5% to 10%.



## **Appendix A**

### **Simulation Report of Operating Conditions**

§ Generated by PRO/II Keyword Generation System <version 5.6>  
 \$ Generated on: Tue Sep 17 19:38:41 2002

## TITLE

PRINT PWRATE, STREAM=ALL, RATE=M  
 DIMENSION SI, TEMP=C, TIME=SEC, LDENSITY=KG/M3, VDENSITY=KG/M3  
 SEQUENCE SIMSCI  
 CALCULATION RVPBASIS=APIN, TVP=37.778

## COMPONENT DATA

LIBID 11,AIR/12,H2O, BANK=PROCESS,SIMSCI  
 PETRO 1,LPG,52,701.541,48.889  
 PETRO 2,NAPHTAS,114,720.859,115  
 PETRO 3,KEROSENE,170,791.049,198.889  
 PETRO 4,DIESEL2,254,824.259,246.111  
 PETRO 5,DIESEL3,310,849.012,268.333  
 PETRO 6,DIESEL4,,875.297,300  
 PETRO 7,VLGO,,897.527,348.333  
 PETRO 8,VHGO,,920.915,376.111  
 PETRO 9,WASHOIL,,942.403,407.222  
 PETRO 10,RESIDUE,,951.922,426.667

## THERMODYNAMIC DATA

METHOD SYSTEM=PR, SET=PR01, DEFAULT

## STREAM DATA

PROPERTY STREAM=S1, TEMPERATURE=25, PRESSURE=102, PHASE=M, &  
 COMPOSITION(LV,MBBL/D)=1,49.39/2,5.26/3,40.91/4,22.27/ &  
 5,24.46/6,29.99/7,11.62/8,23.67/9,27.39/10,85.77  
 PROPERTY STREAM=S3, TEMPERATURE=150, PHASE=M, LFRACTION=0.0001, &  
 COMPOSITION(WT,KG/S)=1,8.4  
 PROPERTY STREAM=S4, TEMPERATURE=191.67, PRESSURE=211.64, PHASE=M, &  
 RATE(LV)=0.0752796, COMPOSITION(M)=2,0.05/3,0.85/4,0.05/ &  
 5,0.05  
 PROPERTY STREAM=S5, TEMPERATURE=245.56, PRESSURE=215.09, PHASE=M, &  
 RATE(LV)=0.0409796, COMPOSITION(M)=3,0.05/4,0.85/5,0.05/ &  
 6,0.05  
 PROPERTY STREAM=S6, TEMPERATURE=267.78, PRESSURE=218.54, PHASE=M, &  
 RATE(LV)=4.50095E-5, COMPOSITION(M)=4,0.05/5,0.85/6,0.05/ &  
 7,0.05  
 PROPERTY STREAM=S7, TEMPERATURE=572, PRESSURE=221.98, PHASE=M, &  
 RATE(LV)=0.0551854, COMPOSITION(M)=5,0.05/6,0.85/7,0.05/ &  
 8,0.05  
 PROPERTY STREAM=S8, TEMPERATURE=342.22, PRESSURE=228.88, PHASE=M, &  
 COMPOSITION(LV,MBBL/D)=7,11.61/8,23.48/9,29.4/10,85.77  
 PROPERTY STREAM=S9, TEMPERATURE=400, PRESSURE=102, PHASE=M, &  
 COMPOSITION(WT,KG/S)=11,600  
 PROPERTY STREAM=S10, TEMPERATURE=1100, PRESSURE=102, PHASE=M, &  
 COMPOSITION(WT,KG/S)=11,600  
 PROPERTY STREAM=S11, TEMPERATURE=50, PRESSURE=206.82, PHASE=M, &  
 RATE(LV)=0.0956866, COMPOSITION(M)=1,0.05/2,0.85/3,0.05/ &  
 4,0.05  
 PROPERTY STREAM=S12, TEMPERATURE=339.44, PRESSURE=225.43, PHASE=M, &  
 RATE(LV)=0.00362505, COMPOSITION(M)=7,0.25/8,0.25/9,0.25/ &  
 10,0.25  
 PROPERTY STREAM=S13, TEMPERATURE=353.33, PRESSURE=515.01, PHASE=M, &  
 RATE(WT)=5.41538, COMPOSITION(M)=12,1  
 PROPERTY STREAM=S16, TEMPERATURE=1100, PRESSURE=102, PHASE=M, &  
 RATE(WT)=60, COMPOSITION(M)=11,1  
 PROPERTY STREAM=S18, TEMPERATURE=162.22, PRESSURE=317.06, PHASE=M, &  
 RATE(LV)=0.0213639, COMPOSITION(M)=7,0.85/8,0.1/9,0.05  
 PROPERTY STREAM=S19, TEMPERATURE=307.22, PRESSURE=254.66, PHASE=M, &  
 RATE(LV)=0.0435558, COMPOSITION(M)=7,0.05/8,0.85/9,0.05/ &  
 10,0.05  
 PROPERTY STREAM=S20, TEMPERATURE=358.89, PHASE=M, LFRACTION=0.99, &

RATE(LV)=0.0504011, COMPOSITION(M)=8,0.05/9,0.85/10,0.1  
 PROPERTY STREAM=S22, TEMPERATURE=362.78, PHASE=M, LFRACTION=0.99, &  
 RATE(LV)=0.157828, COMPOSITION(M)=9,0.1/10,0.9  
 PROPERTY STREAM=S26, TEMPERATURE=768, PRESSURE=413.69, PHASE=M, &  
 RATE(WT)=2.14196, COMPOSITION(M)=12,1  
 NAME S1,FEED/S10,FLUE/S16,FLUE2  
 UNIT OPERATIONS  
 HX UID=E1, NAME=HEATER1  
 HOT FEED=S10, M=S9  
 COLD FEED=S1, M=S2  
 CONFIGURE COUNTER, TPASS=8, SPASS=4  
 OPER HTEMP=353.33, TOL=0.01  
 COLUMN UID=T1, NAME=ADU  
 PARAMETER TRAY=27,IO  
 FEED S2,23/S13,27  
 PRODUCT OVHD(WT)=S3, WATER(M)=S14,1, LDRAW(LV)=S4,4,0.0752796, &  
 LDRAW(LV)=S5,8,0.0409796, LDRAW(LV)=S6,13,0.0450095, &  
 LDRAW(LV)=S7,17,0.0551854, LDRAW(LV)=S11,1,0.0962571, &  
 LDRAW(LV)=S12,22,0.00362505, BTMS(LV)=S21,0.236438, &  
 SUPERSEDE=ON  
 CONDENSER TYPE=PART  
 DUTY 1,1/2,3/3,7/4,13  
 PA FROM=5, TO=3, PHASE=L, RATE(LV)=0.0958706, TEMP=50.556  
 PA FROM=10, TO=7, PHASE=L, RATE(LV)=0.397467, TEMP=125  
 PA FROM=17, TO=13, PHASE=L, RATE(LV)=0.320182, TEMP=219.444  
 PSPEC PTOP=206.82, DPCOLUMN=20.684  
 PRINT PROPTABLE=PART  
 ESTIMATE MODEL=REFINING  
 SPEC STREAM=S21, RATE(LV,MBBL/D), COMP=7,10,WET, VALUE=128.49  
 VARY DUTY=1  
 TOLERANCE KVALUE=0.01, EQUILIBRIUM=0.01, ENTHALPY=0.01  
 COLUMN UID=T2, NAME=VDU  
 PARAMETER TRAY=12,IO  
 FEED S15,10/S26,12  
 PRODUCT BTMS(LV)=S22,0.157828, OVHD(M)=S23, WATER(M)=S24,1, &  
 LDRAW(LV)=S18,2,0.0213823, LDRAW(LV)=S19,4,0.0435558, &  
 LDRAW(LV)=S20,7,0.0504011, LDRAW(LV)=S25,1, &  
 0.00184013  
 CONDENSER TYPE=PART  
 DUTY 1,1/2,2/3,4/4,7  
 PA FROM=3, TO=2, PHASE=L, RATE(LV)=0.204254, TEMP=73.333  
 PA FROM=6, TO=4, PHASE=L, RATE(LV)=0.0759421, TEMP=245  
 PA FROM=9, TO=7, PHASE=L, RATE(LV)=0.0320182, TEMP=198.333  
 PSPEC PTOP=11.101, DPCOLUMN=4.1369  
 PRINT PROPTABLE=PART  
 TOLERANCE KVALUE=0.1, EQUILIBRIUM=0.1, ENTHALPY=0.1  
 HX UID=E2, NAME=HEATER2  
 HOT FEED=S16, M=S17  
 COLD FEED=S8, M=S15  
 CONFIGURE COUNTER  
 OPER HTEMP=408.89, TOL=0.01  
 END

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COMPONENT	COMP. TYPE	PHASE	MOL. WEIGHT	DENSITY KG/M3
1	LPG	PETRO CUT	52.000	701.541
2	NAPHTAS	PETRO CUT	114.000	720.859
3	KEROSENE	PETRO CUT	170.000	791.049
4	DIESEL2	PETRO CUT	254.000	824.259
5	DIESEL3	PETRO CUT	310.000	849.012
6	DIESEL4	PETRO CUT	236.461	875.297
7	VLGO	PETRO CUT	284.046	897.527
8	VHGO	PETRO CUT	305.832	920.915
9	WASHOIL	PETRO CUT	334.575	942.403
10	RESIDUE	PETRO CUT	357.964	951.922
11	AIR	LIBRARY	28.972	871.096
12	H2O	LIBRARY	18.015	998.566

COMPONENT	NBP C	CRIT. TEMP. C	CRIT. PRES. KPA	CRIT. VOLM. M3/KG-MOL	
1	LPG	48.889	226.017	3677.118	0.3107
2	NAPHTAS	115.000	291.692	2710.259	0.4484
3	KEROSENE	198.889	384.823	2185.737	0.6197
4	DIESEL2	246.111	434.094	1953.064	0.7264
5	DIESEL3	268.333	460.248	1926.365	0.7575
6	DIESEL4	300.000	494.325	1846.643	0.8144
7	VLGO	348.333	539.350	1637.034	0.9420
8	VHGO	376.111	569.182	1604.767	0.9833
9	WASHOIL	407.222	600.761	1547.550	1.0397
10	RESIDUE	426.667	618.903	1491.725	1.0864
11	AIR	-194.360	-140.700	3779.420	0.0883
12	H2O	100.000	374.200	22119.199	0.0554

COMPONENT	ACEN. FACT.	HEAT FORM. KJ/KG-MOL	G FORM. KJ/KG-MOL	
1	LPG	0.21205	-35007.18	MISSING
2	NAPHTAS	0.34167	-86127.63	MISSING
3	KEROSENE	0.45568	-197287.65	MISSING
4	DIESEL2	0.53186	-227545.65	MISSING
5	DIESEL3	0.55943	-146610.66	MISSING
6	DIESEL4	0.61124	-106437.02	MISSING
7	VLGO	0.71467	-111036.96	MISSING
8	VHGO	0.76457	-84984.61	MISSING
9	WASHOIL	0.82734	-108083.25	MISSING
10	RESIDUE	0.87223	-145718.96	MISSING
11	AIR	0.04000	0.00	0.00
12	H2O	0.34800	-241997.00	-228796.00

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HEAT EXCHANGER SUMMARY

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UNIT 1, 'E1', 'HEATER1'

OPERATING CONDITIONS

DUTY, M*KJ/SEC	0.511
LMTD, C	509.547
F FACTOR (FT)	0.990
MTD, C	504.499
U*A, KW/K	1011.961

HOT SIDE CONDITIONS

	INLET	OUTLET
	-----	-----
FEED	S10	
VAPOR PRODUCT		S9
VAPOR, KG-MOL/SEC	20.710	20.710
K*KG/SEC	0.600	0.600
CP, KJ/KG-C	1.210	1.059
TOTAL, KG-MOL/SEC	20.710	20.710
K*KG/SEC	0.600	0.600
CONDENSATION, KG-MOL/SEC		0.000
TEMPERATURE, C	1100.000	353.333
PRESSURE, KPA	102.000	102.000

COLD SIDE CONDITIONS

	INLET	OUTLET
	-----	-----
FEED	S1	
VAPOR PRODUCT		S2
VAPOR, KG-MOL/SEC		2.859
K*KG/SEC		0.508
CP, KJ/KG-C		2.617
LIQUID, KG-MOL/SEC	2.859	
K*KG/SEC	0.508	
CP, KJ/KG-C	1.802	
TOTAL, KG-MOL/SEC	2.859	2.859
K*KG/SEC	0.508	0.508
VAPORIZATION, KG-MOL/SEC		2.859
TEMPERATURE, C	25.000	352.375
PRESSURE, KPA	102.000	102.000

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UNIT 2, 'T1', 'ADU'

TOTAL NUMBER OF ITERATIONS

IN/OUT METHOD 2

COLUMN SUMMARY

TRAY	TEMP DEG C	PRESSURE KPA	NET FLOW RATES				HEATER DUTIES M*KJ/SEC
			LIQUID	VAPOR	FEED	PRODUCT	
			KG-MOL/SEC				
1C	89.5	206.81	1.9			1.2V 0.7L 0.0W	-0.1876
2	211.0	206.81	2.0	3.8			
3	226.9	207.64	2.5	3.9	0.3P		-0.0309
4	239.8	208.47	2.9	4.1		0.2L	
5	255.9	209.30	3.3	4.7		0.3P	
6	269.9	210.12	3.7	5.4			
7	280.8	210.95	5.3	5.8	1.0P		-0.1115
8	287.8	211.78	4.9	6.4		0.1L	
9	292.0	212.61	4.4	6.1			
10	294.9	213.43	2.7	5.6		1.0P	
11	295.8	214.26	2.1	4.9			
12	294.1	215.09	1.6	4.3			
13	290.0	215.92	1.6	3.8	0.3P	0.1L	-0.0122
14	286.3	216.74	1.4	3.5			
15	282.5	217.57	1.2	3.3			
16	279.5	218.40	1.1	3.1			
17	278.2	219.23	0.7	3.0		0.3P 0.1L	
18	280.5	220.05	0.7	3.0			
19	285.1	220.88	0.7	3.0			
20	289.9	221.71	0.7	3.0			
21	293.5	222.53	0.7	3.0			
22	301.3	223.36	0.8	3.1		0.0L	
23	329.1	224.19	1.3	3.1	2.9M		
24	359.4	225.02	1.3	0.8			
25	370.1	225.84	1.4	0.7			
26	394.9	226.67	0.9	0.8			
27	346.6	227.50		0.4	0.3V	0.8L	

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UNIT 2, 'T1', 'ADU' (CONT)

FEED AND PRODUCT STREAMS

TYPE	STREAM	PHASE	FROM TRAY	TO TRAY	LIQUID FRAC	FLOW RATES KG-MOL/SEC	HEAT RATES M*KJ/SEC
FEED	S2	MIXED		23	0.1080	2.86	0.5280
FEED	S13	VAPOR		27	0.0000	0.30	0.0172
PROD	S3	VAPOR	1			1.16	0.0458
PROD	S11	LIQUID	1			0.70	0.0142
PROD	S14	WATER	1				0.0000
PROD	S4	LIQUID	4			0.21	0.0254
PROD	S5	LIQUID	8			0.13	0.0213
PROD	S6	LIQUID	13			0.06	0.0088
PROD	S7	LIQUID	17			0.05	0.0073
PROD	S12	LIQUID	22			0.00	0.0007
PROD	S21	LIQUID	27			0.85	0.2110
OVERALL MOLE BALANCE, (FEEDS - PRODUCTS)						-1.1102E-16	
OVERALL HEAT BALANCE, (H(IN) - H(OUT) )							-0.1316

PUMPAROUNDS

TRAY FROM	TRAY TO	TEMP, DEG C FROM	TEMP, DEG C TO	LIQUID FRACTION FROM	LIQUID FRACTION TO	RATES KG-MOL/SEC	RATES K*KG/SEC	RATES STD M3/SEC
5	3	255.9	50.6	1.0000	1.0000	0.26	0.067	0.08
10	7	294.9	125.0	1.0000	1.0000	1.05	0.269	0.31
17	13	278.2	219.4	1.0000	1.0000	0.34	0.080	0.09

SPECIFICATIONS

PARAMETER TYPE	TRAY NO	COMP NO	SPECIFICATION TYPE	SPECIFIED VALUE	CALCULATED VALUE
STRM S21	27	7- 10	LV RATE	1.285E+02	1.482E+02
STRM S11	1	1- 12	LV RATE	9.626E-02	9.881E-02
STRM S4	4	1- 12	LV RATE	7.528E-02	5.982E-02
STRM S5	8	1- 12	LV RATE	4.098E-02	3.978E-02
STRM S6	13	1- 12	LV RATE	4.501E-02	1.600E-02
STRM S7	17	1- 12	LV RATE	5.519E-02	1.388E-02
STRM S12	22	1- 12	LV RATE	3.625E-03	1.262E-03



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UNIT 2, 'T1', 'ADU' (CONT)

REFLUX RATIOS

	----- REFLUX RATIOS -----		
	MOLAR	WEIGHT	STD L VOL
	-----	-----	-----
REFLUX / FEED STREAM S2	0.6665	0.4040	0.4548
REFLUX / VAPOR DISTILLATE	1.6406	3.9561	3.7513

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STREAM MOLAR COMPONENT RATES

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STREAM ID	S1	S2	S3	S4	
NAME	FEED				
PHASE	DRY LIQUID	DRY VAPOR	WET VAPOR	WET LIQUID	
FLUID RATES, KG-MOL/SEC					
1	LPG	1.2261	1.2261	0.8566	4.9770E-03
2	NAPHTAS	0.0612	0.0612	0.0135	7.3498E-04
3	KEROSENE	0.3503	0.3503	3.3530E-03	0.0656
4	DIESEL2	0.1330	0.1330	2.8572E-05	0.0844
5	DIESEL3	0.1233	0.1233	1.2933E-06	0.0487
6	DIESEL4	0.2043	0.2043	3.6378E-09	5.2871E-03
7	VLGO	0.0676	0.0676	1.1016E-17	4.6339E-09
8	VHGO	0.1312	0.1312	0.0000	2.5890E-12
9	WASHOIL	0.1420	0.1420	0.0000	2.5470E-16
10	RESIDUE	0.4197	0.4197	0.0000	1.6947E-18
11	AIR	0.0000	0.0000	0.0000	0.0000
12	H2O	0.0000	0.0000	0.2878	2.1477E-04
TOTAL RATE, KG-MOL/SEC					
		2.8586	2.8586	1.1613	0.2099
TEMPERATURE, C					
		25.0000	352.3746	89.4643	239.8062
PRESSURE, KPA					
		102.0000	102.0000	206.8148	208.4696
ENTHALPY, M*KJ/SEC					
		0.0175	0.5280	0.0458	0.0254
MOLECULAR WEIGHT					
		177.5801	177.5801	44.6425	234.7974
MOLE FRAC VAPOR					
		0.0000	1.0000	1.0000	0.0000
MOLE FRAC TOTAL LIQUID					
		1.0000	0.0000	0.0000	1.0000
MOLE FRAC H/C LIQUID					
		1.0000	0.0000	0.0000	1.0000
MOLE FRAC FREE WATER					
		0.0000	0.0000	0.0000	0.0000

SIMULATION SCIENCES INC.  
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STREAM MOLAR COMPONENT RATES

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STREAM ID	S5	S6	S7	S8
NAME				
PHASE	WET LIQUID	WET LIQUID	WET LIQUID	DRY LIQUID
FLUID RATES, KG-MOL/SEC				
1	1.5040E-03	1.1323E-03	1.2177E-03	0.0000
2	1.6893E-04	1.2141E-04	1.3917E-04	0.0000
3	4.9330E-03	2.5367E-03	3.1573E-03	0.0000
4	0.0228	2.7379E-03	3.2054E-03	0.0000
5	0.0539	7.4889E-03	5.5562E-03	0.0000
6	0.0440	0.0436	0.0372	0.0000
7	6.5018E-07	2.4031E-05	6.6792E-05	0.0675
8	1.7461E-09	4.9022E-07	1.7025E-06	0.1301
9	1.0245E-12	2.9013E-09	1.1273E-08	0.1524
10	2.1570E-14	2.7939E-10	1.1261E-09	0.4197
11	0.0000	0.0000	0.0000	0.0000
12	9.8398E-05	7.9326E-05	7.6643E-05	0.0000
TOTAL RATE, KG-MOL/SEC	0.1274	0.0577	0.0506	0.7697
TEMPERATURE, C	287.7628	289.9721	278.2172	342.2222
PRESSURE, KPA	211.7790	215.9159	219.2254	228.8781
ENTHALPY, M*KJ/SEC	0.0213	8.8452E-03	7.2843E-03	0.1952
MOLECULAR WEIGHT	265.6227	239.7597	236.4584	338.0387
MOLE FRAC VAPOR	0.0000	0.0000	0.0000	0.0000
MOLE FRAC TOTAL LIQUID	1.0000	1.0000	1.0000	1.0000
MOLE FRAC H/C LIQUID	1.0000	1.0000	1.0000	1.0000
MOLE FRAC FREE WATER	0.0000	0.0000	0.0000	0.0000

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STREAM MOLAR COMPONENT RATES

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STREAM ID		S9	S10	S11	S12
NAME			FLUE		
PHASE		DRY VAPOR	DRY VAPOR	WET LIQUID	WET LIQUID
FLUID RATES, KG-MOL/SEC					
1	LPG	0.0000	0.0000	0.3607	7.6467E-05
2	NAPHTAS	0.0000	0.0000	0.0465	8.6798E-06
3	KEROSENE	0.0000	0.0000	0.2704	1.9698E-04
4	DIESEL2	0.0000	0.0000	0.0184	2.0391E-04
5	DIESEL3	0.0000	0.0000	2.3973E-03	3.5098E-04
6	DIESEL4	0.0000	0.0000	3.4314E-05	2.1747E-03
7	VLGO	0.0000	0.0000	1.7293E-12	4.2687E-04
8	VHGO	0.0000	0.0000	1.7736E-16	3.6773E-04
9	WASHOIL	0.0000	0.0000	0.0000	1.7353E-04
10	RESIDUE	0.0000	0.0000	0.0000	3.0703E-04
11	AIR	20.7097	20.7097	0.0000	0.0000
12	H2O	0.0000	0.0000	3.0152E-03	5.0948E-06
TOTAL RATE, KG-MOL/SEC		20.7097	20.7097	0.7014	4.2920E-03
TEMPERATURE, C		353.3333	1100.0000	89.4643	301.3017
PRESSURE, KPA		102.0000	102.0000	206.8148	223.3622
ENTHALPY, M*KJ/SEC		0.1975	0.7080	0.0142	7.3881E-04
MOLECULAR WEIGHT		28.9720	28.9720	107.6484	259.7991
MOLE FRAC VAPOR		1.0000	1.0000	0.0000	0.0000
MOLE FRAC TOTAL LIQUID		0.0000	0.0000	1.0000	1.0000
MOLE FRAC H/C LIQUID		0.0000	0.0000	1.0000	1.0000
MOLE FRAC FREE WATER		0.0000	0.0000	0.0000	0.0000

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STREAM MOLAR COMPONENT RATES

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STREAM ID		S13	S16	S18	S19
NAME			FLUE2		
PHASE		WATER VAPOR	DRY VAPOR	WET LIQUID	MIXED
FLUID RATES, KG-MOL/SEC					
1	LPG	0.0000	0.0000	0.0000	0.0000
2	NAPHTAS	0.0000	0.0000	0.0000	0.0000
3	KEROSENE	0.0000	0.0000	0.0000	0.0000
4	DIESEL2	0.0000	0.0000	0.0000	0.0000
5	DIESEL3	0.0000	0.0000	0.0000	0.0000
6	DIESEL4	0.0000	0.0000	0.0000	0.0000
7	VLGO	0.0000	0.0000	0.0264	0.0213
8	VHGO	0.0000	0.0000	0.0302	0.0579
9	WASHOIL	0.0000	0.0000	6.0116E-03	0.0302
10	RESIDUE	0.0000	0.0000	2.0398E-03	0.0188
11	AIR	0.0000	2.0710	0.0000	0.0000
12	H2O	0.3006	0.0000	6.3879E-05	1.7833E-05
TOTAL RATE, KG-MOL/SEC		0.3006	2.0710	0.0647	0.1283
TEMPERATURE, C		353.3333	1100.0000	190.3522	282.1823
PRESSURE, KPA		515.0106	102.0000	11.1006	11.9279
ENTHALPY, M*KJ/SEC		0.0172	0.0708	7.1546E-03	0.0241
MOLECULAR WEIGHT		18.0150	28.9720	300.9838	316.5947
MOLE FRAC VAPOR		1.0000	1.0000	0.0000	7.1548E-05
MOLE FRAC TOTAL LIQUID		0.0000	0.0000	1.0000	0.9999
MOLE FRAC H/C LIQUID		0.0000	0.0000	1.0000	0.9999
MOLE FRAC FREE WATER		0.0000	0.0000	0.0000	0.0000

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STREAM MOLAR COMPONENT RATES

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STREAM ID	S20	S21	S22	S26	
NAME					
PHASE	WET LIQUID	WET LIQUID	WET LIQUID	WATER VAPOR	
FLUID RATES, KG-MOL/SEC					
1	LPG	0.0000	2.8957E-07	0.0000	0.0000
2	NAPHTAS	0.0000	2.6987E-07	0.0000	0.0000
3	KEROSENE	0.0000	1.7761E-04	0.0000	0.0000
4	DIESEL2	0.0000	1.2133E-03	0.0000	0.0000
5	DIESEL3	0.0000	4.8466E-03	0.0000	0.0000
6	DIESEL4	0.0000	0.0720	0.0000	0.0000
7	VLGO	3.5504E-03	0.0670	2.2820E-03	0.0000
8	VHGO	0.0184	0.1308	0.0170	0.0000
9	WASHOIL	0.0406	0.1418	0.0750	0.0000
10	RESIDUE	0.0657	0.4194	0.3331	0.0000
11	AIR	0.0000	0.0000	0.0000	0.0000
12	H2O	1.5543E-05	9.2854E-03	2.3627E-04	0.1189
TOTAL RATE, KG-MOL/SEC					
		0.1283	0.8466	0.4277	0.1189
TEMPERATURE, C					
		310.5296	346.5667	300.8271	768.0000
PRESSURE, KPA					
		13.1497	227.4991	13.6615	413.6856
ENTHALPY, M*KJ/SEC					
		0.0289	0.2110	0.0949	8.7291E-03
MOLECULAR WEIGHT					
		340.9960	325.6115	351.2039	18.0150
MOLE FRAC VAPOR					
		0.0000	0.0000	0.0000	1.0000
MOLE FRAC TOTAL LIQUID					
		1.0000	1.0000	1.0000	0.0000
MOLE FRAC H/C LIQUID					
		1.0000	1.0000	1.0000	0.0000
MOLE FRAC FREE WATER					
		0.0000	0.0000	0.0000	0.0000

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STREAM ID	S1	S2	S3	S4
NAME	FEED			
PHASE	DRY LIQUID	DRY VAPOR	WET VAPOR	WET LIQUID
----- TOTAL STREAM -----				
RATE, KG-MOL/SEC	2.859	2.859	1.161	0.210
K*KG/SEC	0.508	0.508	5.184E-02	4.929E-02
TEMPERATURE, C	25.000	352.375	89.464	239.806
PRESSURE, KPA	102.000	102.000	206.815	208.470
MOLECULAR WEIGHT	177.580	177.580	44.643	234.797
ENTHALPY, M*KJ/SEC	1.746E-02	0.528	4.584E-02	2.538E-02
KJ/KG	34.391	1040.125	884.270	514.903
MOLE FRACTION LIQUID	1.00000	0.00000	0.00000	1.00000
MOLE FRACTION FREE WATER	0.00000	0.00000	0.00000	0.00000
----- TOTAL VAPOR -----				
RATE, KG-MOL/SEC	N/A	2.859	1.161	N/A
K*KG/SEC	N/A	0.508	5.184E-02	N/A
K*M3/SEC	N/A	0.142	1.636E-02	N/A
NORM VAP RATE(1), K*M3/SEC	N/A	6.407E-02	2.603E-02	N/A
MOLECULAR WEIGHT	N/A	177.580	44.643	N/A
ENTHALPY, KJ/KG	N/A	1040.125	884.270	N/A
CP, KJ/KG-C	N/A	2.617	1.817	N/A
DENSITY, KG/K*M3	N/A	3572.923	3169.749	N/A
Z (FROM DENSITY)	N/A	0.9747	0.9661	N/A
----- TOTAL LIQUID -----				
RATE, KG-MOL/SEC	2.859	N/A	N/A	0.210
K*KG/SEC	0.508	N/A	N/A	4.929E-02
M3/SEC	0.595	N/A	N/A	7.503E-02
GAL/MIN	9429.584	N/A	N/A	1189.294
STD LIQ RATE, M3/SEC	0.590	N/A	N/A	5.982E-02
MOLECULAR WEIGHT	177.580	N/A	N/A	234.797
ENTHALPY, KJ/KG	34.391	N/A	N/A	514.903
CP, KJ/KG-C	1.802	N/A	N/A	2.626
DENSITY, KG/M3	853.268	N/A	N/A	656.947
Z (FROM DENSITY)	8.5633E-03	N/A	N/A	0.0175

(1) NORMAL VAPOR VOLUME IS 22.414 M3/KG-MOLE (273.15 K AND 1 ATM)

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STREAM ID	S1	S2	S3	S4
NAME	FEED			
PHASE	DRY LIQUID	DRY VAPOR	WET VAPOR	WET LIQUID
----- DRY STREAM -----				
RATE, KG-MOL/SEC	2.859	2.859	0.873	0.210
K*KG/SEC	0.508	0.508	4.666E-02	4.929E-02
STD LIQ RATE, M3/SEC	0.590	0.590	6.635E-02	5.981E-02
MOLECULAR WEIGHT	177.580	177.580	53.417	235.019
MOLE FRACTION LIQUID	1.0000	0.0000	0.0000	1.0000
REDUCED TEMP (KAYS RULE)	0.4528	0.9500	0.7241	0.7391
PRES (KAYS RULE)	0.0390	0.0390	0.0566	0.1012
ACENTRIC FACTOR	0.4678	0.4678	0.2150	0.5082
WATSON K (UOPK)	11.583	11.583	11.889	11.812
STD LIQ DENSITY, KG/M3	860.107	860.107	703.151	824.059
SPECIFIC GRAVITY	0.8610	0.8610	0.7038	0.8249
API GRAVITY	32.852	32.852	69.539	40.042
----- DRY VAPOR -----				
RATE, KG-MOL/SEC	N/A	2.859	0.373	N/A
K*KG/SEC	N/A	0.508	4.666E-02	N/A
K*M3/SEC	N/A	0.142	1.220E-02	N/A
NORM VAP RATE(1), K*M3/SEC	N/A	6.407E-02	1.958E-02	N/A
SPECIFIC GRAVITY (AIR=1.0)	N/A	6.131	1.844	N/A
MOLECULAR WEIGHT	N/A	177.580	53.417	N/A
CP, KJ/KG-C	N/A	2.617	1.799	N/A
DENSITY, KG/K*M3	N/A	3572.923	3825.371	N/A
----- DRY LIQUID -----				
RATE, KG-MOL/SEC	2.859	N/A	N/A	0.210
K*KG/SEC	0.508	N/A	N/A	4.929E-02
M3/SEC	0.595	N/A	N/A	7.503E-02
GAL/MIN	9429.584	N/A	N/A	1189.219
STD LIQ RATE, M3/SEC	0.590	N/A	N/A	5.981E-02
SPECIFIC GRAVITY (H2O=1.0)	0.8610	N/A	N/A	0.8249
MOLECULAR WEIGHT	177.580	N/A	N/A	235.019
CP, KJ/KG-C	1.802	N/A	N/A	2.626
DENSITY, KG/M3	853.268	N/A	N/A	656.937

(1) NORMAL VAPOR VOLUME IS 22.414 M3/KG-MOLE (273.15 K AND 1 ATM)



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STREAM ID	S5	S6	S7	S8
NAME				
PHASE	WET LIQUID	WET LIQUID	WET LIQUID	DRY LIQUID
----- TOTAL STREAM -----				
RATE, KG-MOL/SEC	0.127	5.768E-02	5.061E-02	0.770
K*KG/SEC	3.384E-02	1.383E-02	1.197E-02	0.260
TEMPERATURE, C	287.763	289.972	278.217	342.222
PRESSURE, KPA	211.779	215.916	219.225	228.878
MOLECULAR WEIGHT	265.623	239.760	236.458	338.039
ENTHALPY, M*KJ/SEC	2.129E-02	8.845E-03	7.284E-03	0.195
KJ/KG	629.169	639.585	608.667	750.340
MOLE FRACTION LIQUID	1.00000	1.00000	1.00000	1.00000
MOLE FRACTION FREE WATER	0.00000	0.00000	0.00000	0.00000
----- TOTAL VAPOR -----				
RATE, KG-MOL/SEC	N/A	N/A	N/A	N/A
K*KG/SEC	N/A	N/A	N/A	N/A
K*M3/SEC	N/A	N/A	N/A	N/A
NORM VAP RATE(1), K*M3/SEC	N/A	N/A	N/A	N/A
MOLECULAR WEIGHT	N/A	N/A	N/A	N/A
ENTHALPY, KJ/KG	N/A	N/A	N/A	N/A
CP, KJ/KG-C	N/A	N/A	N/A	N/A
DENSITY, KG/K*M3	N/A	N/A	N/A	N/A
Z (FROM DENSITY)	N/A	N/A	N/A	N/A
----- TOTAL LIQUID -----				
RATE, KG-MOL/SEC	0.127	5.768E-02	5.061E-02	0.770
K*KG/SEC	3.384E-02	1.383E-02	1.197E-02	0.260
M3/SEC	5.188E-02	2.068E-02	1.768E-02	0.352
GAL/MIN	822.377	327.738	280.298	5572.958
STD LIQ RATE, M3/SEC	3.978E-02	1.600E-02	1.388E-02	0.276
MOLECULAR WEIGHT	265.623	239.760	236.458	338.039
ENTHALPY, KJ/KG	629.169	639.585	608.667	750.340
CP, KJ/KG-C	2.745	2.767	2.728	2.826
DENSITY, KG/M3	652.133	668.841	676.748	740.012
Z (FROM DENSITY)	0.0185	0.0165	0.0167	0.0204

(1) NORMAL VAPOR VOLUME IS 22.414 M3/KG-MOLE (273.15 K AND 1 ATM)

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STREAM ID	S5	S6	S7	S8
NAME				
PHASE	WET LIQUID	WET LIQUID	WET LIQUID	DRY LIQUID
----- DRY STREAM -----				
RATE, KG-MOL/SEC	0.127	5.760E-02	5.054E-02	0.770
K*KG/SEC	3.383E-02	1.383E-02	1.197E-02	0.260
STD LIQ RATE, M3/SEC	3.978E-02	1.600E-02	1.388E-02	0.276
MOLECULAR WEIGHT	265.814	240.065	236.790	338.039
MOLE FRACTION LIQUID	1.0000	1.0000	1.0000	1.0000
REDUCED TEMP (KAYS RULE)	0.7636	0.7512	0.7390	0.7048
PRES (KAYS RULE)	0.1094	0.1128	0.1136	0.1491
ACENTRIC FACTOR	0.5640	0.5855	0.5806	0.8313
WATSON K (UOPK)	11.665	11.587	11.596	11.366
STD LIQ DENSITY, KG/M3	850.461	864.159	862.362	941.012
SPECIFIC GRAVITY	0.8513	0.8650	0.8632	0.9419
API GRAVITY	34.716	32.082	32.422	18.722
----- DRY VAPOR -----				
RATE, KG-MOL/SEC	N/A	N/A	N/A	N/A
K*KG/SEC	N/A	N/A	N/A	N/A
K*M3/SEC	N/A	N/A	N/A	N/A
NORM VAP RATE(1), K*M3/SEC	N/A	N/A	N/A	N/A
SPECIFIC GRAVITY (AIR=1.0)	N/A	N/A	N/A	N/A
MOLECULAR WEIGHT	N/A	N/A	N/A	N/A
CP, KJ/KG-C	N/A	N/A	N/A	N/A
DENSITY, KG/K*M3	N/A	N/A	N/A	N/A
----- DRY LIQUID -----				
RATE, KG-MOL/SEC	0.127	5.760E-02	5.054E-02	0.770
K*KG/SEC	3.383E-02	1.383E-02	1.197E-02	0.260
M3/SEC	5.188E-02	2.068E-02	1.768E-02	0.352
GAL/MIN	822.339	327.707	280.269	5572.958
STD LIQ RATE, M3/SEC	3.978E-02	1.600E-02	1.388E-02	0.276
SPECIFIC GRAVITY (H2O=1.0)	0.8513	0.8650	0.8632	0.9419
MOLECULAR WEIGHT	265.814	240.065	236.790	338.039
CP, KJ/KG-C	2.744	2.767	2.728	2.826
DENSITY, KG/M3	652.129	668.835	676.740	740.012

(1) NORMAL VAPOR VOLUME IS 22.414 M3/KG-MOLE (273.15 K AND 1 ATM)

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STREAM ID	S9	S10	S11	S12
NAME		FLUE		
PHASE	DRY VAPOR	DRY VAPOR	WET LIQUID	WET LIQUID
----- TOTAL STREAM -----				
RATE, KG-MOL/SEC	20.710	20.710	0.701	4.292E-03
K*KG/SEC	0.600	0.600	7.551E-02	1.115E-03
TEMPERATURE, C	353.333	1100.000	89.464	301.302
PRESSURE, KPA	102.000	102.000	206.815	223.362
MOLECULAR WEIGHT	28.972	28.972	107.648	259.799
ENTHALPY, M*KJ/SEC	0.197	0.708	1.420E-02	7.388E-04
KJ/KG	329.128	1180.016	188.089	662.585
MOLE FRACTION LIQUID	0.00000	0.00000	1.00000	1.00000
MOLE FRACTION FREE WATER	0.00000	0.00000	0.00000	0.00000
----- TOTAL VAPOR -----				
RATE, KG-MOL/SEC	20.710	20.710	N/A	N/A
K*KG/SEC	0.600	0.600	N/A	N/A
K*M3/SEC	1.058	2.319	N/A	N/A
NORM VAP RATE(1), K*M3/SEC	0.464	0.464	N/A	N/A
MOLECULAR WEIGHT	28.972	28.972	N/A	N/A
ENTHALPY, KJ/KG	329.128	1180.016	N/A	N/A
CP, KJ/KG-C	1.059	1.210	N/A	N/A
DENSITY, KG/K*M3	567.151	258.784	N/A	N/A
Z (FROM DENSITY)	1.0003	1.0002	N/A	N/A
----- TOTAL LIQUID -----				
RATE, KG-MOL/SEC	N/A	N/A	0.701	4.292E-03
K*KG/SEC	N/A	N/A	7.551E-02	1.115E-03
M3/SEC	N/A	N/A	0.108	1.624E-03
GAL/MIN	N/A	N/A	1704.520	25.742
STD LIQ RATE, M3/SEC	N/A	N/A	9.881E-02	1.262E-03
MOLECULAR WEIGHT	N/A	N/A	107.648	259.799
ENTHALPY, KJ/KG	N/A	N/A	188.089	662.585
CP, KJ/KG-C	N/A	N/A	2.240	2.778
DENSITY, KG/M3	N/A	N/A	702.147	686.571
Z (FROM DENSITY)	N/A	N/A	0.0105	0.0177

(1) NORMAL VAPOR VOLUME IS 22.414 M3/KG-MOLE (273.15 K AND 1 ATM)

SIMULATION SCIENCES INC.  
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STREAM ID	S9	S10	S11	S12
NAME		FLUE		
PHASE	DRY VAPOR	DRY VAPOR	WET LIQUID	WET LIQUID
----- DRY STREAM -----				
RATE, KG-MOL/SEC	20.710	20.710	0.698	4.287E-03
K*KG/SEC	0.600	0.600	7.545E-02	1.115E-03
STD LIQ RATE, M3/SEC	0.689	0.689	9.876E-02	1.262E-03
MOLECULAR WEIGHT	28.972	28.972	108.035	260.086
MOLE FRACTION LIQUID	0.0000	0.0000	1.0000	1.0000
REDUCED TEMP (KAYS RULE)	4.7300	10.3673	0.6347	0.7405
PRES (KAYS RULE)	0.0270	0.0270	0.0693	0.1221
ACENTRIC FACTOR	0.0400	0.0400	0.3246	0.6393
WATSON K (UOPK)	5.981	5.981	11.953	11.537
STD LIQ DENSITY, KG/M3	871.096	871.096	764.021	883.392
SPECIFIC GRAVITY	0.8720	0.8720	0.7648	0.8843
API GRAVITY	30.779	30.779	53.522	28.520
----- DRY VAPOR -----				
RATE, KG-MOL/SEC	20.710	20.710	N/A	N/A
K*KG/SEC	0.600	0.600	N/A	N/A
K*M3/SEC	1.058	2.319	N/A	N/A
NORM VAP RATE(1), K*M3/SEC	0.464	0.464	N/A	N/A
SPECIFIC GRAVITY (AIR=1.0)	1.000	1.000	N/A	N/A
MOLECULAR WEIGHT	28.972	28.972	N/A	N/A
CP, KJ/KG-C	1.059	1.210	N/A	N/A
DENSITY, KG/K*M3	567.151	258.784	N/A	N/A
----- DRY LIQUID -----				
RATE, KG-MOL/SEC	N/A	N/A	0.698	4.287E-03
K*KG/SEC	N/A	N/A	7.545E-02	1.115E-03
M3/SEC	N/A	N/A	0.107	1.624E-03
GAL/MIN	N/A	N/A	1703.628	25.740
STD LIQ RATE, M3/SEC	N/A	N/A	9.876E-02	1.262E-03
SPECIFIC GRAVITY (H2O=1.0)	N/A	N/A	0.7648	0.8843
MOLECULAR WEIGHT	N/A	N/A	108.035	260.086
CP, KJ/KG-C	N/A	N/A	2.239	2.778
DENSITY, KG/M3	N/A	N/A	702.009	686.569

(1) NORMAL VAPOR VOLUME IS 22.414 M3/KG-MOLE (273.15 K AND 1 ATM)

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STREAM ID	S13	S16	S18	S19
NAME		FLUE2		
PHASE	WATER VAPOR	DRY VAPOR	WET LIQUID	MIXED
----- TOTAL STREAM -----				
RATE, KG-MOL/SEC	0.301	2.071	6.466E-02	0.128
K*KG/SEC	5.415E-03	6.000E-02	1.946E-02	4.061E-02
TEMPERATURE, C	353.333	1100.000	190.352	282.182
PRESSURE, KPA	515.011	102.000	11.101	11.928
MOLECULAR WEIGHT	18.015	28.972	300.984	316.595
ENTHALPY, M*KJ/SEC	1.718E-02	7.080E-02	7.155E-03	2.406E-02
KJ/KG	3172.316	1180.016	367.630	592.637
MOLE FRACTION LIQUID	0.00000	0.00000	1.00000	0.99993
MOLE FRACTION FREE WATER	0.00000	0.00000	0.00000	0.00000
----- TOTAL VAPOR -----				
RATE, KG-MOL/SEC	0.301	2.071	N/A	9.177E-06
K*KG/SEC	5.415E-03	6.000E-02	N/A	2.408E-06
K*M3/SEC	3.014E-03	0.232	N/A	3.522E-06
NORM VAP RATE(1), K*M3/SEC	6.738E-03	4.642E-02	N/A	2.057E-07
MOLECULAR WEIGHT	18.015	28.972	N/A	262.376
ENTHALPY, KJ/KG	3172.316	1180.016	N/A	856.909
CP, KJ/KG-C	2.064	1.210	N/A	2.368
DENSITY, KG/K*M3	1796.905	258.784	N/A	683.604
Z (FROM DENSITY)	0.9912	1.0002	N/A	0.9915
----- TOTAL LIQUID -----				
RATE, KG-MOL/SEC	N/A	N/A	6.466E-02	0.128
K*KG/SEC	N/A	N/A	1.946E-02	4.060E-02
M3/SEC	N/A	N/A	2.398E-02	5.302E-02
GAL/MIN	N/A	N/A	380.049	840.440
STD LIQ RATE, M3/SEC	N/A	N/A	2.127E-02	4.377E-02
MOLECULAR WEIGHT	N/A	N/A	300.984	316.599
ENTHALPY, KJ/KG	N/A	N/A	367.630	592.621
CP, KJ/KG-C	N/A	N/A	2.361	2.656
DENSITY, KG/M3	N/A	N/A	811.651	765.759
Z (FROM DENSITY)	N/A	N/A	1.0682E-03	1.0681E-03

(1) NORMAL VAPOR VOLUME IS 22.414 M3/KG-MOLE (273.15 K AND 1 ATM)

SIMULATION SCIENCES INC.  
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STREAM ID	S13	S16	S18	S19
NAME		FLUE2		
PHASE	WATER VAPOR	DRY VAPOR	WET LIQUID	MIXED
----- DRY STREAM -----				
RATE, KG-MOL/SEC	N/A	2.071	6.460E-02	0.128
K*KG/SEC	N/A	6.000E-02	1.946E-02	4.061E-02
STD LIQ RATE, M3/SEC	N/A	6.888E-02	2.127E-02	4.377E-02
MOLECULAR WEIGHT	N/A	28.972	301.264	316.636
MOLE FRACTION LIQUID	N/A	0.0000	1.0000	0.9999
REDUCED TEMP (KAYS RULE)	N/A	10.3673	0.5553	0.6517
PRES (KAYS RULE)	N/A	0.0270	6.8989E-03	7.5490E-03
ACENTRIC FACTOR	N/A	0.0400	0.7535	0.7869
WATSON K (UOPK)	N/A	5.981	11.464	11.409
STD LIQ DENSITY, KG/M3	N/A	871.096	915.017	927.594
SPECIFIC GRAVITY	N/A	0.8720	0.9159	0.9285
API GRAVITY	N/A	30.779	22.990	20.895
----- DRY VAPOR -----				
RATE, KG-MOL/SEC	N/A	2.071	N/A	7.866E-06
K*KG/SEC	N/A	6.000E-02	N/A	2.384E-06
K*M3/SEC	N/A	0.232	N/A	3.015E-06
NORM VAP RATE(1), K*M3/SEC	N/A	4.642E-02	N/A	1.763E-07
SPECIFIC GRAVITY (AIR=1.0)	N/A	1.000	N/A	10.465
MOLECULAR WEIGHT	N/A	28.972	N/A	303.100
CP, KJ/KG-C	N/A	1.210	N/A	2.372
DENSITY, KG/K*M3	N/A	258.784	N/A	790.840
----- DRY LIQUID -----				
RATE, KG-MOL/SEC	N/A	N/A	6.460E-02	0.128
K*KG/SEC	N/A	N/A	1.946E-02	4.060E-02
M3/SEC	N/A	N/A	2.398E-02	5.302E-02
GAL/MIN	N/A	N/A	380.028	840.434
STD LIQ RATE, M3/SEC	N/A	N/A	2.127E-02	4.377E-02
SPECIFIC GRAVITY (H2O=1.0)	N/A	N/A	0.9159	0.9285
MOLECULAR WEIGHT	N/A	N/A	301.264	316.637
CP, KJ/KG-C	N/A	N/A	2.361	2.656
DENSITY, KG/M3	N/A	N/A	811.648	765.759

(1) NORMAL VAPOR VOLUME IS 22.414 M3/KG-MOLE (273.15 K AND 1 ATM)

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STREAM ID	S20	S21	S22	S26
NAME				
PHASE	WET LIQUID	WET LIQUID	WET LIQUID	WATER VAPOR
----- TOTAL STREAM -----				
RATE, KG-MOL/SEC	0.128	0.847	0.428	0.119
K*KG/SEC	4.375E-02	0.276	0.150	2.142E-03
TEMPERATURE, C	310.530	346.567	300.827	768.000
PRESSURE, KPA	13.150	227.499	13.662	413.686
MOLECULAR WEIGHT	340.996	325.612	351.204	18.015
ENTHALPY, M*KJ/SEC	2.891E-02	0.211	9.491E-02	8.729E-03
KJ/KG	660.836	765.620	631.917	4075.276
MOLE FRACTION LIQUID	1.00000	1.00000	1.00000	0.00000
MOLE FRACTION FREE WATER	0.00000	0.00000	0.00000	0.00000
----- TOTAL VAPOR -----				
RATE, KG-MOL/SEC	N/A	N/A	N/A	0.119
K*KG/SEC	N/A	N/A	N/A	2.142E-03
K*M3/SEC	N/A	N/A	N/A	2.486E-03
NORM VAP RATE(1), K*M3/SEC	N/A	N/A	N/A	2.665E-03
MOLECULAR WEIGHT	N/A	N/A	N/A	18.015
ENTHALPY, KJ/KG	N/A	N/A	N/A	4075.276
CP, KJ/KG-C	N/A	N/A	N/A	2.309
DENSITY, KG/K*M3	N/A	N/A	N/A	861.701
Z (FROM DENSITY)	N/A	N/A	N/A	0.9991
----- TOTAL LIQUID -----				
RATE, KG-MOL/SEC	0.128	0.847	0.428	N/A
K*KG/SEC	4.375E-02	0.276	0.150	N/A
M3/SEC	5.711E-02	0.378	0.193	N/A
GAL/MIN	905.164	5998.785	3051.647	N/A
STD LIQ RATE, M3/SEC	4.636E-02	0.295	0.158	N/A
MOLECULAR WEIGHT	340.996	325.612	351.204	N/A
ENTHALPY, KJ/KG	660.836	765.620	631.917	N/A
CP, KJ/KG-C	2.724	2.851	2.687	N/A
DENSITY, KG/M3	766.062	728.330	780.144	N/A
Z (FROM DENSITY)	1.2061E-03	0.0197	1.2887E-03	N/A

(1) NORMAL VAPOR VOLUME IS 22.414 M3/KG-MOLE (273.15 K AND 1 ATM)

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STREAM ID	S20	S21	S22	S26
NAME				
PHASE	WET LIQUID	WET LIQUID	WET LIQUID	WATER VAPOR
----- DRY STREAM -----				
RATE, KG-MOL/SEC	0.128	0.837	0.427	N/A
K*KG/SEC	4.375E-02	0.275	0.150	N/A
STD LIQ RATE, M3/SEC	4.636E-02	0.294	0.158	N/A
MOLECULAR WEIGHT	341.035	329.023	351.388	N/A
MOLE FRACTION LIQUID	1.0000	1.0000	1.0000	N/A
REDUCED TEMP (KAYS RULE)	0.6656	0.7182	0.6475	N/A
PRES (KAYS RULE)	8.5966E-03	0.1454	9.0666E-03	N/A
ACENTRIC FACTOR	0.8382	0.8104	0.8592	N/A
WATSON K (UOPK)	11.353	11.379	11.340	N/A
STD LIQ DENSITY, KG/M3	943.558	935.932	948.965	N/A
SPECIFIC GRAVITY	0.9445	0.9369	0.9499	N/A
API GRAVITY	18.316	19.537	17.463	N/A
----- DRY VAPOR -----				
RATE, KG-MOL/SEC	N/A	N/A	N/A	N/A
K*KG/SEC	N/A	N/A	N/A	N/A
K*M3/SEC	N/A	N/A	N/A	N/A
NORM VAP RATE(1), K*M3/SEC	N/A	N/A	N/A	N/A
SPECIFIC GRAVITY (AIR=1.0)	N/A	N/A	N/A	N/A
MOLECULAR WEIGHT	N/A	N/A	N/A	N/A
CP, KJ/KG-C	N/A	N/A	N/A	N/A
DENSITY, KG/K*M3	N/A	N/A	N/A	N/A
----- DRY LIQUID -----				
RATE, KG-MOL/SEC	0.128	0.837	0.427	N/A
K*KG/SEC	4.375E-02	0.275	0.150	N/A
M3/SEC	5.711E-02	0.378	0.193	N/A
GAL/MIN	905.158	5994.261	3051.552	N/A
STD LIQ RATE, M3/SEC	4.636E-02	0.294	0.158	N/A
SPECIFIC GRAVITY (H2O=1.0)	0.9445	0.9369	0.9499	N/A
MOLECULAR WEIGHT	341.035	329.023	351.388	N/A
CP, KJ/KG-C	2.724	2.847	2.687	N/A
DENSITY, KG/M3	766.063	728.437	780.146	N/A

(1) NORMAL VAPOR VOLUME IS 22.414 M3/KG-MOLE (273.15 K AND 1 ATM)



## NOMENCLATURE

ADU	:	atmospheric distillation unit
$\dot{E}$	:	energy rate, kW
$h$	:	specific enthalpy, kJ/kg
$h^{\circ}$	:	specific enthalpy of formation, kJ/kg
$\dot{i}$	:	irreversibility rate, kW
$\dot{m}$	:	mass flow rate, kg/s
$N$	:	molar ratio
$P$	:	pressure, kPa
$\dot{Q}$	:	heat transfer rate, kW
$\dot{S}$	:	Entropy rate, kW/K
$s$	:	specific entropy, kJ/kg.K
$s^{\circ}$	:	specific entropy of formation, kJ/kg.K
$T$	:	temperature, K
$v$	:	specific volume, m <sup>3</sup> /kg
$\dot{W}$	:	Work rate, kW
VDU	:	vacuum distillation unit
$\dot{\Xi}$	:	exergy rate. kW
$\epsilon$	:	specific exergy, kJ/kg
$\eta$	:	energy efficiency
$\mu$	:	chemical potential, kJ/kg
$\psi$	:	exergy efficiency

**Subscripts**

**0** : **reference condition**

**ch** : **chemical**

**cv** : **control volume**

**gen** : **generation**

**ph** : **physical**

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