Study of
Energy Use in Eastern Refinery Limited
and
Conservation Possibilities

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Study of Energy Use in Eastern Refinery Limited and Conservation Possibilities

by

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This thesis is respectfully dedicated
to my parents
to whom I owe so much

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ABSTRACT

This thesis discusses the energy use in Eastern Refinery Limited and the possibilities of energy conservation. It illustrates various processes, equipment and system oriented approach of energy conservation The objective of energy conservation can be achieved by auditing the energy consumed in every process unit and then to save energy, various steps are to be taken including designing more efficient and compact process units like heat exchanger, furnace etc, revamping existing equipment or overall BMRE (Balancing, Modernization and Rehabilitation)

This study on energy use and conservation possibilities was carried out in Eastern Refinery Ltd. (ERL), the only petroleum refinery plant in Bangladesh. ERL has five processing unit; the study was kept limited to Crude Distillation Unit, the main unit of the refinery plant. Eastern Refinery Ltd. was designed in the mid-60's and went on stream in 1968. At its early stage energy conservation was not so stringent as it is now in a refinery of 2001. ERL has taken necessary steps to improve the plant's efficiency and to conserve energy. ERL is now processing 1.3 – 1.4 million MT of crude oil per annum, as against its installed capacity of 1 5 million tons per year.

Furnace is the major consumer of energy, usually obtained by burning hydrocarbon fuel. It is estimated that total energy consumption of ERL is about 60.54 MMkcal per hour of which distillation furnace alone is consuming near about 28.62 MMkcal/hr (47.27 %). It has been found that, the furnace is running with a thermal efficiency about 65 % and operating with about 77 % excess air which is much higher than the designed value(40%). 1 – 2 % improvement in thermal efficiency is possible by reducing excess air to the furnace only and through proper monitoring. It has been found that 1% improvement in thermal efficiency in the furnace will result in energy saving of around 3256 MMkcal per year and savings in terms of money around Tk.22,34,104.00 per year.

Energy consumption in ERL per barrel of Crude oil processing is about 0.05 MM kcal. The overall processing expense of Crude oil is increasing every year. In 1990-91 financial years it was Tk.32.88 per Barrel and in the year 1999-2000 it increased

to Tk 58 70 per Barrel. About 4.99 MMkcal heat is wasted to cooling water during cooling products from process unit before transferring to storage tank. Heat exchanger can be placed before LGO and Residue water cooler, to save heat energy of about 3.99MM kcal /hr. A reasonable amount of energy can be saved by increasing flashing rate of crude oil in flashing dram before feeding to furnace, to avoid heating the same in the furnace.

TABLE OF CONTENTS

LIST OF FIGURES

LIST OF TABLES

NOMENCLATURE

A	Heat-transfer surface area, m^2
Cp	Specific heat, MMkcal / kg. °C
Cv	Valve flow Coefficient
Fc	Temperature-difference factor, ΔT = Fc x LMTD, dimensionless
G_f	Constant (Orifice)
h_1	Enthalpy at inlet, MMkcal/kg
H_1	Heat content at inlet , MMkcal/hr
h_2	Enthalpy at outlet, MMkcal/kg
H_2	Heat content at outlet, MMkcal/hr
K	Characterization factor
LMTD	Log Mean Temperature Difference
m	Moles of carbon
n	Moles of Hydrogen
P	Pressure
P_1	Pressure at inlet
P_2	Pressure at outlet
Q	Heat flow rate
R	Dimensionless temperature group
S	Dimensionless temperature group
s	Shell
t	Tube
t_1	Inlet temperature of cold fluid, °C
t_2	Outlet temperature of cold fluid, °C
T_1	Inlet temperature of hot fluid, °C
T_2	Outlet temperature of hot fluid, °C
t_{air}	Temperature of air, °C
T_{avg}	Average temperature, °C
$t_{flue\ gas}$	Temperature of flue gas, °C
U	Overall heat transfer coefficient

W	Flow rate (MT/h)
Tcm	Correct mean temperature
P	Electrical load, Kwatt
V	Volt
I	Current, Ampere
ΔT or Del T	Temperature difference
η	Thermal efficiency of furnace
θ	Angle

CHAPTER 1

Introduction

Energy is the prime mover of modern civilization. Its importance in our livelihoods is unquestionable. At the primitive stage, human muscle was the main source of energy. In course of time, men succeeded in building up mechanisms by which they could do work with greater energy and efficiency. They continued their effort to search for energies hidden in the extracted materials of earth. At the initial stage naturally occurring wood and coal were the main source of energy. With the advancement of technology, natural gas, petroleum oil, chemicals, nuclear energy, etc. are now used as the main sources of energy.

Internal combustion engine has given a revolutionary development in industry, particularly in power generation and transportation. At present electrical energy is obtained mainly by the conversion of energy from fuel, wind, water, chemical, nuclear sources, etc. Most energy generation involves the conversion of some type of fuel, usually by combustion. Temperature is the driving force for energy conversion, so it is usually an important variable

It is estimated that in 1850, coal, oil and gas provided 5 percent of the world's energy and the muscles of humans and animals, 95 percent. Today, coal, oil, gas, and nuclear sources account for approximately 94 percent, water power about 1 percent and the muscle of humans and animals supply the remaining 5 percent.

Almost all the commercial liquid fuels are derived from natural petroleum (crude oil). Crude oil is obtained from underground reservoirs in certain parts of the world. The liquid fuel consists of hydrocarbons. The natural petroleum may be separated in petrol or gasoline, paraffin oil or kerosene, fuel oils and lubricating oils by boiling the crude at different temperatures and by subsequent fractional distillation.

Currently the annual production of oil in the world is about 2300 MMta and natural gas over one trillion cubic meters and is expected to reach a maximum production of 2700 - 2800 MMta of oil during this decade (Rao, 1997).

The importance of petroleum in present day civilization is ever increasing due to its unmatched contribution for energy requirement in power generation, transportation, lnbrication and petrochemical fields. Sixty percent of the energy needs of the world are met by petroleum The frequent hikes in prices of petroleum and its products both national and international scenes is for two reasons; the mounting demand and fast depletion of reserves. To meet future demand of energy scientists and engineers are very busy to find out the new source of energy alternatives to the petroleum fuels and at the same time they are trying to find out the ways, to run the petroleum industry in a cost effective manner i.e with the highest degree of energy conservation.

Energy cost has increased dramatically over the past 30 years primarily due to OPEC regulation of crude oil pricing, since the Arab oil embargo in 1973, energy conservation has been a popular subject of discussion. Consequently, refiners are finding it necessary to make energy audit to find where energy is being wasted, develop ways to rednce or eliminate the wastage, and to monitor their operations to maintain an energy efficient operation. Energy efficiency is one of the keys to refinery profitability. Energy cost is the single largest component of variable operating expenses. Petroleum operating companies constantly seek opportunities to minimize manufacturing cost. It is clear that improving energy efficiency can benefit profitability. Improving efficiency by only a few percent, therefore, has obvious financial rewards.

To keep pace with the global market, Eastern Refinery Limited should take necessary steps to minimize production cost by reducing energy waste, improving equipment efficiency aud ensuring highest degree of conservation of energy.

CHAPTER 2

Literature Review

2.1 Introduction

A general literature survey has been carried out with major emphasis on energy conservation in crude petroleum processing industries. The purpose of literature survey usually is, to find the extent and amount of research work done on a particular topic and on related fields. This particular literature search has provided much vital information about latest development in energy sector, its demand, reserve, supply and energy consumption particularly petroleum products. Modern concept of energy conservation in crude oil processing industry particularly in designing equipment, techniques and operations in a cost effective manner is being implemented in many advanced country.

2.2 Plant Energy Conservation

Following discussion is based on study report on "*Technical Presentation for Implementation of Plant Energy conservation*" prepared by Technip, an organization expert in Petroleum industry.

The objective of the energy conservation is to do something to significantly reduce energy consumption and cost, thereby increase profits. There is nothing magic or unique about conserving energy. It simply gets down to assigning competent people to study the plants to determine where the energy can be saved, and then applying well proven techniques to recover energy which are economically attractive. Energy is consumed in many places and in many forms (fuel, steam, electricity, etc.). Attention to details is important to maximize energy consumption reductions.

Discussion of energy conservation may be divided into the following areas.

- The problem or looking at the opportunity to conserve energy and increase profits.

- The methodology employed to expedite an energy conservation program and then monitor operations to retain the improvements.
- The discrete steps in the execution of an energy conservation program from the auditing of performance through to the installation of energy saving equipment.

This opportunity is the ability to do something to significantly reduce energy consumption and cost, thereby increasing profits. Energy conservation can be very profitable Refiners have used several approaches to effect energy conservation. Most refiners have accomplished significant energy reductions in the "easy to do" areas such as reducing excess air in furnace stacks, minimizing steam leaks and installing insulation on high heat loss areas. Most of these techniques measure consumption and not efficiency. Refiners stress that their efforts are aimed at improving efficiency not reducing raw consumption. Minimum energy consumption is not necessarily optimum consumption. The system used to measure efficiency of individual process units provides the tool to measure and monitor progress needed to effectively motivate the organization.

The methodology measures excess energy loss on individual refinery units (as well as on overall refinery basis) as compared to a standard highly efficient process unit. The technique helps one to identify and quantify the specific energy loss areas and then analyze the losses for potential recovery. Most of the energy saved today has been achieved by fine-tuning operations and good housekeeping measures requiring little or no investment.

The next steps of energy saving moves will likely require investments. During the 1972 to 1990 period, fuel cost has increased about ten-fold, and equipment costs have increased by about a factor of three. The relatively high increase in fuel costs compared to equipment costs suggests that the plants built prior to 1973 can probably be economically revamped for better energy utilization. Of course, the high interest rates on capital investment will have some moderating influence.

Under conservation of energy programme many industries have achieved 15% - 20% reduction in energy consumption per barrel of crude processed, now they are targeting for another 10 % - 20 % reduction. Others indicate that they are considering some significant investments for energy conservation projects.

2.2.1 Energy conservation program

The six basic steps in a preferred programme towards conservation of energy are:

I. Start by measuring the actual energy consumption in each discrete operating unit such as Boiler, Furnace, etc. reporting all energy in an equivalent measuring unit. Steam and electrical power are also converted to same unit by using simplified factors.

II. The measured energy consumption is then compared to a high efficiency baseline unit. These baseline units are designed employing selected criteria intended to establish minimum energy usage in unit by using proven technology. Where baseline units are not available, they can be developed using practical energy conservation measures with respect to previous condition.

III. The next step is to identify and quantify the energy losses to permit analyzing the losses for potential recovery.

IV. In studying the specific energy loss areas, it will become evident that some energy reductions may be easily made whereas others may be difficult or not economical to recover. Energy conservation measure are often separated into three categories:

- Easy to do with little effort and little or no investment. These items are usually operating and maintenance improvements.
- Investment projects that are usually high-energy recovery projects, but that must be further developed to justify the investment.
- Low level heat sink energy items that one usually sets aside to accumulate until sufficient energy is available to merit serious consideration.

V The implementation of energy loss recovery simply gets down to doing the easy to do items as quickly as possible and then developing a programmed approach to implement those projects requiring significant investment

VI Probably one of the most important targets of an effective program is the monitoring of performance on a continuous basis to insure that any gains made are maintained and that the energy savings program does not regress but progresses.

Steps in a preferred energy program have been discussed; now the feature of a program in more depth will be considered to see how it works. An effective energy management system should have:

- A reliable yardstick to measure energy performance
- A means to define where and how much energy is wasted
- A means to monitor energy performance progress towards eliminating the wastage.

The baseline high energy efficient units provide the yardstick needed to measure energy efficiency and monitor performance. It is customary to convert all forms of energy usage into a fuel oil equivalent for easy comparisons. The high efficiency target units should be attainable designs employing high energy efficiency design criteria. The criteria used to develop these units will form a handy and valuable check list to help identify and quantify energy losses. Provisions should be made to adjust energy consumption on a given unit for factors that can have significant impact on the unit's energy consumption.

The bottom line economics of any energy conservation program depends on day-to-day, week-to-week, monitoring and stewardship of performance against an established goal. Monitoring should be done at each processing unit. In some instance, it may be desirable to group several units together within one identifiable processing facility. Examples of such are the combining of the atmospheric and vacuum distillation units, the naphtha hydrotreater and catalytic reformer, and the complete fractionation facilities within a gas plant. In monitoring towards a goal, a useful indicator of performance is to report efficiency as percent of target. This approach can be used as a management information tool providing management at any level with a performance indicator.

The major components evaluated in process units.

- **Furnaces and Boilers:**

  Of the energy used in a typical refinery, 75-80 percent is hydrocarbon fuel that is burned in process heaters or boilers. Therefore, improvements in the efficiency of furnaces and boilers can make a major contribution in reduction of energy consumption. Emphasis should be given to excess air control, recovering heat from stack gases and eliminating air leakage into the firebox.

- **Heat exchangers:**

  The primary function of heat exchange is to reuse process heat, thus reducing the amount of fuel supplied. Due to the current need to conserve increasingly scarce and costly fuel, there has been a drastic change in economic flux limits. Consequently, there is a need to reevaluate systems designed on past economic flux criteria.

- **Insulation:**

  Today's high energy costs can justify increased thickness of insulation over that considered economical only a few years ago.

- **Fractionation:**

  The minimization of fractionation energy requirements is concerned with such things as multi-unit integration, heat pump applications, lower pressure operation and reflux optimization. Extensive process instrumentation control and the use of non-conventional distillation schemes merit consideration.

- **Steam-power system:**

  A rational evaluation of possible energy savings steps requires knowledge of the plant steam system balance and the effect of proposed changes on the balance. Another area of consideration is the use of synchronous motors to help improve power factor

- **Steam traps:**

  The proper type of steam trap, the sizing and most importantly the maintenance of steam traps is very critical in an energy conservation program. Individual steam traps appear so insignificant; however, the large number of potentially leaking steam traps in a typical plant can result in a large energy loss. One leaking trap can be wasting $1000 per year of steam.

- **Flares and off sites:**

  The incentive to reduce or eliminate flaring has increased substantially with using fuel costs. Flare gas compression and recovery is becoming a money-making proposition in most plants.

- **Miscellaneous:**

  In auditing the energy consumption of any plant, one should look for steam leaks from flanges, valves, broken steam tracing, etc. Another useful tool is infrared temperature surveys used to locate hot spots and heat leaks from equipment. Leaking safety valves, poorly insulated pipe and tanks, and other equipment requiring insulation have been spotted using infrared equipment surveying from a plane.

2.3 Energy Conservation and overall Loss Control in Eastern Refinery Ltd (ERL).

Crude Distillation and Reforming Unit of ERL were designed in the mid-60 and require relatively high input of energy and various utilities. Loss control system is also not as stringent as it is now in a refinery of 2002. ERL can accomplish some energy reduction in the "easy to do" area such as frequently controlling and monitoring excess air in the furnace Total energy consumption in the entire refinery is to be measured and relate this to crude thru put, feed rates and processing severities i.e. standardizing energy consumption and loss and monitoring against these standards

The Distillation (topping) furnace can operate at 77% efficiency which would improve further with the present NFK burners which can operate at 20% excess air (Technip, 1991).

Large portion of heat liberated in the furnace is also wasted with hot flue gases leaving around 430 °C. Heat recovery from flue gas would require converting this furnace to balanced draft furnace which requires major investment.

2.3.1 Energy conservation

At present furnace inlet temperature can be increased by rearrangement of existing heat train and also additional heat transfer surface. An increase of 20 °C is very much achievable i.e. from 210 °C to 230 °C which would result in energy saving in the order 3.5 MM Kcal/hr or equal energy can be utilized for boiler feed water heating or steam generation. BFW (Boiler feed water) preheater or a deminaralized water preheater can be placed on the atmospheric residue between exchanger E1108 and E1115 (Technip, 1991).

Heat recovery scheme:

- Lowering the product outlet temperature of preheat train reduced the load on the trim cooler and improve furnace inlet temperature and decrease evaporation loss.
- Export of energy for BFW heating / steam generating.

At present, at low thruput operation, preheat temperature of Distillation furnace is around 230 °C. Old refineries like ERL should revamped the preheat train to increase the furnace inlet temperature to 230 °C - 240 °C range.

I. Rearranging present exchangers:

The easiest way with minimum capital investment is by rearranging the present exchanger. Exchanger E1108 is almost ineffective because of very low approach temperature. If E1108 is placed after E1105, the heat exchanged will be raised by about 1 0 MMkcal/hr, but this would adversely effect in E1106, E1107 and E1109 (Technip, 1991).

II. Rearranging with modification of flow pattern:

As in (1), placing E1108 after E1105 and splitting up the Crude in two streams. One stream is to pass through E1106 & E1107 and other stream through E1109. To maintain the same shell side flow rate, number of tubes is to be doubled and flow type in the shell of E1106, E1107 and E1109 is to be changed from mixed type to cross type. This leads to a saving of 1.0 MMkcal/hr (Technip, 1991)

III. Adding new Exchanger:

To reduce the evaporation loss and train cooler load, extra heat transfer surface is to be added i.e. additional exchangers are to be installed.

IV. Boiler feed water (BFW) heating for Steam generation:

Amount of heat energy available from the product stream is much more than the amounts of heat recovered by the present crude preheat train. By installing a steam generator in place of first LGO Exchanger (E1106), about 3.0 MMkcal/hr of heat can be recovered. A considerable amount of heat is rejected to cooling water in E115 that can also be recovered by modifying the flow scheme. Since a considerable amount of heat is available within the temperature range of 90 - 180 °C from different stream, a split up scheme may be developed. In this scheme, the energy level of product stream K II, LGO and Residue (in the 5^{th}, 6^{th} exchanger), the crude stream is also split up in three parts (Technip, 1991).

2.3.2 Loss control in Eastern Refinery Ltd.

ERL has taken necessary measures, to control losses in different discrete units are discussed below.

I. Flare loss:

Flare loss can be minimize by:

- Good operation practice
- Efficient overhead cooling
- Increase in LPG production.

In 1983 the flare loss was around 24 MT/day. At present feed rate, the loss is about 6 MT/day which is 0.20% of the feed (Report, ERL, 1990)

II. Refinery gas:

In Refinery the term Gas + Loss is generally referred to as Refinery gas and used as fuel gas. About 1.2% of the Refinery gas (Gas + Loss) is consumed in the furnaces. Reforming unit produces about 700 Kg/hour gas at the present feed and as such has a big effect on Gas + Loss which increases by 0.40 %. Flare gas compression and recovery is not attractive at present (Report, ERL, 1990).

III. Tank farm:

Losses that could be incurred in tank farm particularly for light product depend on:

- Volatility of the product
- Storage temperature
- Type of tank
- Inter tank transfer of products and blending operation.

Losses from tanks can be categorized as follows:

- Breathing loss
- Filling loss

- Standing loss
- Withdrawal loss

A rough estimate of ERL tank losses to be about 2000 MT per year which is about 0.20 % of crude thruput (Study Report, ERL, 1990).

Steps that could be considered for reduction of this loss are (Report, ERL, 1990):

- Lower storage temperature
- Minimum operation of tanks for a certain thruput
- To avoid storing volatile product in fixed roof tanks. A close monitor could bring down this loss.

IV. Sludge and water:

There is a considerable amount of sludge deposited in crude oil tanks and ponds. At present there is no way to accurately quantify the amount of sludge Some of crude that comes with water during water purging from storage tank, can be reduced further with more settling time. Periodical quantification of sludge and oil from the separator and monitoring could be beneficial and lead to more recovery of oil and cut down loss (Study Report, ERL, 1990).

V. Transfer loss from ERL to marketing companies:

To reduce loss during product transfer from ERL to marketing companies, following steps should be considered (Study Report, ERL, 1990)

- The tank should be recalibrated by reputed company at regular interval of time.
- As far as possible the tanks should be filled up before pumping schedule is given.
- For better monitoring pumping should commence as far as possible during office hours.
- The parcel size should be increased as far as practicable.

- The valves both at ERL and marketing companies should be checked and replaced at a certain interval.
- Wherever possible accurate dynamic quantification system is to be introduced.

VI. Process:

The Energy conservation can be made more meaningful by close monitoring of the process units performance. The prime objectives are:

- To identify and document operating and maintenance improvement opportunities (which is reasonably done).
- To identify and document energy losses (which can be more accurately done in addition of necessary measuring instruments).
- To establish targets/base lines for the development of an energy management reporting system. It should be done collectively by operation, planning and may be research & development (R and D).

All the above energy conservation task may be undertaken in phased manner for overall improvement. An organized program should accomplish increased energy savings in shorter time than simply assigning people to the job to develop their own approach (Study Report, ERL, 1990)

Probably one of the important facets of an effective program is the monitoring of performance on a continuous basis to ensure that any gain made are maintained and the energy saving program does not regress but progress.

2.4 Refining into the Next Decade

An ultra modern deep-conversion refinery of the twenty-first century, producing high quality products having low emission characteristics, could be up to four times more costly in capital investment than a simple vintage grass root refinery of 1970 and between two and two and a half times more expensive than a conventional 1980's complex refinery according to *Daniel Morel* (1998), executive vice president of Institute François du Petrol (IFP).

At the IFP's (Institute François du pétrole) recent 'Upgrading Heavy Ends' symposium *Daniel Morel* forecast that with in the next 50 years, emerging countries will account for more than 60 percent of the world's primary energy consumption. America will maintain a constant global market; Europe will have a decline and Asia will maintain its percentage at around 60 percent, while Africa will increase significantly.

2.5 Energy Demand

During the period 1996 – 2010, oil consumption will increase from 3.3 to 4.3 billion tons/year of oil equivalent (toe), according to the International Energy Agency (IEA).This extra increase in demand generates questions about future crude supplies. Long term oil potential will remain in the following OPEC countries: Saudi Arabia, Kuwait, Iran, Iraq, The United Arab Emirates and Venezuela. Crude from these regions can be exploited at a relatively low cost. The drawback is in their relatively higher densities and lower product qualities, i.e. heavier crude with higher sulphur and nitrogen content. OPEC oil supply should be between 1995 and 2010, increase from 28 to about 47-49 million bpd, representing an increase in OPEC's market share from 40 % in 1995 to 50 % in 2010" (PTQ, 1998).

Almost 50% of world oil consumption is used for transportation fuels. This trend will intensify as demand increases at a rate of 2.9 % /year, to boost oil demand from the 1995 level of 1.5 billion toes to 2.3 billion toes by 2010. Fossil fuels like coal, oil and natural gas will continue to provide 90 % of world's primary energy; although gas will displace coal in some regions. Petroleum will remain the dominant fuel, meeting 40% of world energy needs. Nuclear power output will remain constant in absolute terms but decline as older nuclear reactor in Europe and North America are retired. New renewable energy sources will increase rapidly, from 2% to 3% of total demand (Wendy, 2001).

2.6 Recent developments in Distillation

Within the past decade several significant developments have occurred in distillation technology applied to petroleum and chemical processing. Most of them can be grouped in to three general categories:

- New design practices
- New contacting devices and new applications for existing contacting devices.
- New troubleshooting techniques.

The developments have originated and been perfected both in academic and industrial Research and Development (R&D) centers. Notable among the former are three groups, those headed by Professor J R.Fair at the separations research program, University of Texas at Austin; by Professor R. Billet at the Institute for thermal separation processes, Ruhr University, Bo chum; and by Professor F. J. Zeiderweg, consultant of fractionation research, Inc., at the Technical University, Delft. In the Industrial sector many firms have made contributions to design, manufacture and use of distillation equipment in recent years, the most prominent being Glitsch, Koch, Norton, Nutter and Sulzer (Chen, 1989).

2.7 Heat Exchange System (Pinch Analysis)

The recent rise in energy costs, offers an increased incentive to develop any process design with a particular emphasis on energy recovery. Method proposed in recent year for designing optimal heat exchanger net works, generally, all have a similar structure.

- An analysis procedure to identify the maximum possible energy recovery for a given approach temperature
- A synthesis step which locates an "Optimal" heat exchanger network.

The analysis procedure proposed by Linhoff and Flower uses a "Problem Table" to calculate the minimum heating and cooling utility load to achieve the specified temperature changes in the hot and cold stream. It is also identified as a thermodynamic bottle neck called the "pinch", which has been one of the main concepts contributing to recent progress in heat transfer network design (Wadud, 1986).

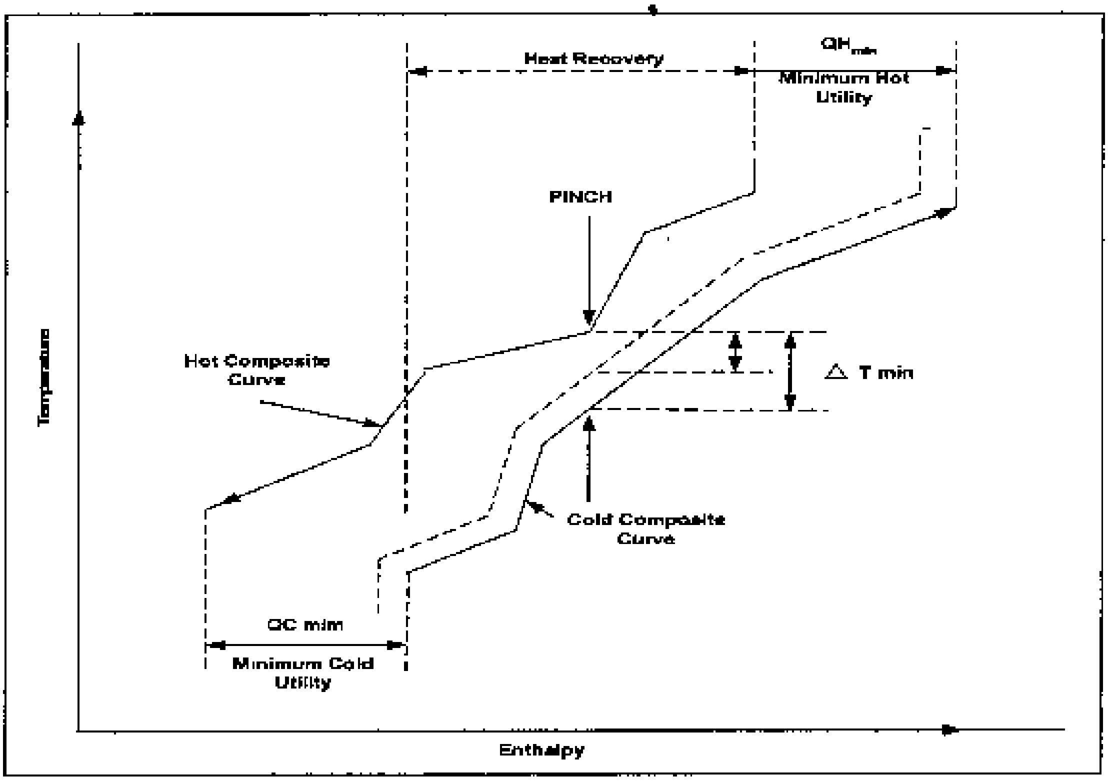

Figure 2.1 Energy targets and "Pinch" with composite curve (ICE, 1983).

Figure 2 1 shows a typical pair of composite curves. Shifting of the curves lead to behavior similar to that shown by the two-stream problem. Now, though, the "kinked" nature of the composites means that ΔT_{min} can occur anywhere in the interchange region Note that although there are many streams in the problem, a general ΔT_{min} occurs at only one point, termed the "pinch" (ICE, 1983).

At this point, some explanation of 'pinch' is deemed necessary. Petroleum refinery process can be considered as a series of stream requiring either heating or cooling. This heating and cooling is achieved by a combination of process interchanges (matches between process streams) and heaters and coolers supplied with utilities. Heat from the external hot utility is supplied to the first interval and any surplus heat from this interval is passed on or "cascaded" to the second interval. Surplus heat is cascaded through the other intervals in the same way until finally heat is rejected to the cold utility from the last interval. Within the overall cascade, heat flows from high to low temperatures. Therefore it can be use the cascade model to determine the lowest possible value for the input. This value will be minimum when it produces one point in the cascade of zero heat flow. This particular point has been termed as "Pinch" by Linhoff et al (ICE, 1983). The pinch divides process into two parts. Above the pinch temperature, heat is taken in and no heat is rejected; this part of the process acts as a 'heat sink'. Below the pinch the reverse occurs and the process acts as a 'heat source'.

The basis of all Pinch Analysis is the set of composite curves (Figure 2.1) that can be drawn for any process to represent all the heating and cooling duties in the process. The composition curve allow the designer to calculate hot and cold utility targets ahead of design, to understand driving forces for heat transfer and to locate the heat recovery "pinch".

The utility target depends on the value of ΔT_{min}. A small ΔT_{min} brings the curves close together, reducing hot and cold utility demands and giving lower operating costs. This is at the expense of larger heat exchanger area and hence greater capital cost. The optimum choice of ΔT_{min} depends on a trade-off between capital and energy (Brown, 1999).

Linhoff sums up 'Pinch Technology' in optimum heat exchanger network design as follows (ICE, 1983):

I. **Drawing of composite curves:** The hot and cold stream in a process can be represented on a temperature - heat content (enthalpy) graph once their input and output temperatures (or "supply" and "target" temperatures) and their flow rates and physical properties are known. Starting from the individual streams, it is possible to construct one "composite curve" of all heat streams in the process and another of all cold streams by simple addition of heat contents over the temperature ranges in the problem.

II. **Targets:** Once the composite curve is known, we know exactly how much external heating is required Near-optimal processes are confirmed as such and non-optimal processes are identified with great speed and confidence.

III. **The pinch:** Above the pinch, the process needs external heating and below the pinch, it needs external cooling. This tells us where to place furnaces, steam heaters, coolers etc

IV. **More in, more out:** An inefficient process requires more than the minimum external heating and therefore more than the minimum external cooling.

V. **Freedom of choice:** The "heat sink" and the "heat source" in the network are separate. As long as the designer obeys this constraint, he has the freedom in choosing plant layouts, control arrangements, etc.

VI. **Trade offs:** A simple relationship exists between the number of streams (process streams plus utilities) in a problem and the minimum number of heat exchange units (i.e. heaters, coolers and interchangers). Thus if designer goes for the best energy recovery, designing the "heat source" and "heat sink" sections separately, he or she will incur the need for more units than if the pinch division is ignored hence a new type of trade-off has been identified, between energy recovery and number of units

In ERL in crude preheat train; there are 6 hot streams and one cold stream. Hot stream are at different temperatures. There is no utility stream (that is noheateror no cooler is utilized). Only heat is added to the crude oil by thermal exchange with hot product stream from Distillation column

2.8 Consumption of Petroleum Products in Bangladesh

Bangladesh is a poor country of South East Asia. A reasonable amount of gas reserve has been discovered but no oil field has yet been discovered except Haripur oil field, production from which is remained suspended. At present annual consumption of petroleum products is more than 34 lac M.Ton To meet the national demand of petroleum products Bangladesh has to import crude oil and refined products from Middle Eastern countries at cost of more than Tk. 2,500 Crore (C & F Cost) every year.

Consumption of petroleum products in Bangladesh is increasing every year. It is observed that annual increment of demand of petroleum products in Bangladesh is about 6.6 %. To meet the increasing demand of petroleum products, a lion portion of our national budget is expended in importing Petroleum products. So conservation of energy should be considered at every step of energy use. Emphasis should be given on natural gas. The use of natural gas alternative to petroleum oil in every sector may relieve the dependence on imported oil. Consumption of petroleum products in Bangladesh during last 10 years is shown in Table 2.1 (MIS, Dec-2001).

Table 2.1 Consumption of petroleum products in Bangladesh during last 10 year

(Quantity in M.Ton)

Products	1991-92	1992-93	1993-94	1994-95	1995-96	1996-97	1997-98	1998-99	1999-00	2000-01
JP-1	74198	77181	90344	95339	97087	107563	127847	120983	142070	169119
HOBC	16909	20835	24971	31184	35285	53960	59080	64115	75808	98458
MS	102950	116518	12851	156062	174390	224461	212694	197609	194376	197046
SKO	358478	374797	391651	434468	451645	539318	585846	568145	609402	636099
HSD	839204	930671	952055	1274616	1302787	1476198	1542004	1738933	1743251	1846239
LDO	9364	11691	10906	10062	9940	5987	2284	1209	1041	1809
JBO	20968	21005	19478	21531	19626	20921	22087	18564	17393	16855
FO	155949	226694	213228	240511	211470	272270	221908	168031	338916	306665
LUB	31946	35143	36456	44195	44552	47307	44360	45186	44596	37443
SBP/MTT	2666	3074	3628	4043	4392	5264	5524	5367	5650	5841
LPG	8015	3779	12549	15743	13446	15856	13539	11724	14845	20432
BITUMNE	49034	72897	75356	86602	119132	64979	71042	63008	44170	63113
TOTAL	**16,69,691**	**19,00,295**	**19,59,139**	**24,16,356**	**24,83,652**	**28,34,084**	**29,08,215**	**3,002,874**	**3231752**	**3403116**
(-)/+	+78,313	+2,30,604	+58,854	+4,57,217	+67,296	+3,50,432	+74,131	+94,659	+228878	+171364
%	+4.92	+13.81	+3.1	+23.34	+2.79	+14.11	+2.62	+3.25	7.08	+5.04

To meet the increasing demand of petroleum products, a lion portion of our national budget is expended in importing petroleum products. So conservation of energy should be considered at every step of energy use. Emphasis should be given on natural gas. The use of natural gas alternative to petroleum oil in every sector may relieve the dependence on imported oil. A sale forecast of petroleum products in Bangladesh up to 2010 is shown in Table 2.2 (MIS, Feb-2001)

Table 2.2 Sale forecasts of petroleum products in Bangladesh up to 2010

Quantity in Lac M.Ton

Product/year	2001-02	2002-03	2003-04	2004-05	2005-06	2006-07	2007-08	2008-09	2009-10
HOBC	0.88	0.98	1.07	1.18	1.31	1.44	1.57	1.73	1.9
MS	3.13	3.44	3.78	4.16	4.58	5.04	5.54	6.1	6.71
JP-1	1.44	1.46	1.47	1.49	1.5	1.52	1.53	1.55	1.58
SKO	6.58	6.77	6.98	7.19	7.4	7.63	7.85	8.09	8.33
HSD	22.67	24.49	26.45	28.56	30.85	33.32	35.98	38.86	41.98
LDO	0.05	0.05	0.05	0.05	0.05	0.05	0.05	0.05	0.05
FO	2.6	2.6	2.6	2.65	2.6	2.6	2.6	2.6	2.6
LPG	0.27	0.29	0.32	0.35	0.39	0.43	0.47	0.52	0.57
JBO	0.25	0.25	0.25	0.25	0.25	0.25	0.25	0.25	0.25
LUB	0.63	0.68	0.73	0.79	0.86	0.93	1	1.08	1.17
BITUMEN	1.27	1.34	1.38	1.42	1.49	1.53	1.61	1.69	1.77
TOTAL	39.77	42.35	45.08	48.04	51.28	54.74	58.45	62.52	66.88
% Increment		6.49%	6.45%	6.57%	6.74%	6.75%	6.78%	6.96%	6.97%

CHAPTER 3

Objective

3.1 Objective

The main objectives of the study based on energy use in Eastern Refinery Ltd. Conservation possibilities are as follows:

- To study energy use in Eastern refinery limited.
- To study the performance of Crude Distillation Furnace
- To study crude preheat train which utilize heat from the hot product streams from distillation column to increase furnace inlet temperature of feed stock.
- To find out the possible ways of energy conservation in ERL

3.2 Methodology

In calculating heating value of fuel, lower heating value (LHV) has been considered. Heat content in petroleum fraction has been determined by considering the characterization factor 'K'. Heat rejected by hot fluids and gained by the cold fluid (Crude) is calculated by heat balance ($Q = M\ Cp\ \Delta T$). Layout and design of heat exchangers were checked by "Pinch Technology". Based on above mentioned methods following calculations has been undertaken.

- Calculation of total energy consumed in ERL
- Calculation of thermal efficiency of distillation furnace, its relation with excess oxygen in flue gas and flue gas temperature.
- Application of "Pinch Technology" in identifying the bottlenecks in heat exchanger net work.
- Energy saving by crude oil flashing, by reducing excess air supply, increasing furnace inlet temperature of feed stock through preheating etc.

CHAPTER 4

Eastern Refinery Limited (ERL)

4.1 Introduction

Eastern Refinery Ltd is the lone petroleum refinery plant of Bangladesh. It is a subsidiary of Bangladesh Petroleum Corporation (BPC) and is a public limited company incorporated under company act of 1913 (amended in 1994). It owns and operates the sole petroleum refinery complex of Bangladesh The company installed the refining complex on picturesque expanse of 196 acres on the bank of river Karnaphuli at North Patenga area of Chittagong city. The project was conceived and initiated in 1963 by a group of the then Pakistani entrepreneurs. Under a turn-key contract with French firm ENSA / TECHNIP/ COFRI, ERL has completed the installation of refinery complex that went on stream commercially from 8th of May, 1968.

Operational activities of ERL are mostly limited to reception of crude oil, processing of crude oil into finished products and transferring these products to marketing companies for distribution through out the country. ERL processes low sulphur light crude oil namely Murban Crude (of Abu Dhabi) and Arabian Light Crude (of KSA)

This refinery is basically a fuel refinery and produces 16 petroleum fuel products including 4 nonfuel products. ERL in its first year of operation treated only 0.25 million MT of crude; now it processes 1.3 to 1.4 million MT of crude annually against the installed capacity of 1.5 million MT per year. ERL meets the demand of almost all of the petroleum products of the country. Only middle distillate e.g. kerosene and diesel oil are the major deficit products.

4.2 Physical Facilities of Eastern Refinery Ltd.

At present, ERL has the following main physical facilities:

Table 4.1 Physical facilities of ERL

PROCESS UNITS	INSTALLED CAPACITY
1. Crude distillation Unit	1.5 million MT/Year
2. Catalytic Reforming Unit	70,000 MT/Year
3 Asphaltic Bitumen Unit i) Vacuum Distillation Unit ii) Bitumen Blowing Unit	 2,00,000 MT/Year 70,000 MT/Year
4. Visbreaker Unit	5,22,000 MT/Year
5. Mild Hydro cracking Unit	57,000 MT/Year
ANCILLARY UNITS	
1. Drum Manufacturing & Drum Filling Unit	1,100 Drum /Day (8 hours)
2. Hydrogen Plant	790 MT/Year
TREATMENT UNITS	
1. LPG Merox Unit	24,000 MT/Year
2. Gasoline Merox Unit	65,000 MT/Year
3. Kerosene Merox Unit	1,25,000 MT/Year
UTILITIES UNIT	
1. Steam Generation Unit (3 Boilers)	60 MT/hour
2. Power Generation Units i) 2 Steam Turbine Generators ii) 3 Diesel Generators	 3,000 KW (each) 1,400 KW (each)
STORAGE	
1. Crude Tankage	3,28,000 M^3
2. LPG Tankage	2,200 M^3
3. Other product Tankage (Including Bitumen)	1,68,000 M^3

4.3 Process Units of ERL

ERL has the following five main processing units:

- Crude Distillation Unit
- Catalytic Reforming Unit
- Asphaltic Bitumen Unit
- Visbreaker Unit
- Mild Hydrocracking (MHC) Unit

Short description of each processing unit is given below:

4.3.1 Crude Distillation Unit

This is the primary crude oil processing unit and is the starting point for the petroleum refinery. In this unit, crude oil is preheated with hot product streams desalted and a part of it is then flash vaporized for introducing in the distillation column at low temperature. The rest is finally heated up to around 370 °C in a furnace

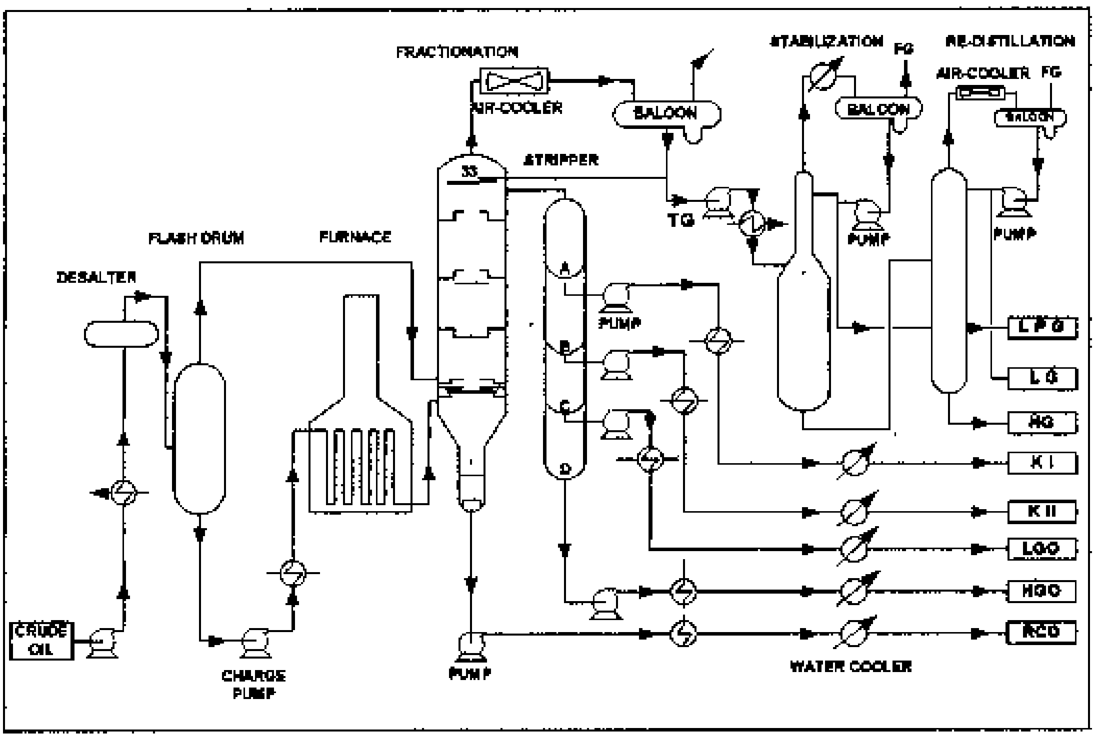

Figure 4.1 Crude Distillation Units

The heated crude oil is fed into the distillation column for fractionation. The distillation column is operated at slightly higher than atmospheric pressure and it is generally known as Atmospheric Distillation Unit (ADU) or Topping Unit. Figure 4.1, shows, the flow diagram of Crude Distillation Unit.

The straight-run fraction obtained from the column is: Total gasoline (TG), Light kerosene (K I), Heavy kerosene (K II), Light gas oil (LGO), Heavy gas oil (HGO) and Atmospheric residue (RCO) Lighter compounds (Propane and Butane) present in total gasoline are debutanise in a Stabilization unit and a part of it is recovered as LPG. Rest of the Propane and Butane goes to refinery fuel gas system and is used in furnaces and boilers. Debutanised gasoline from Stabilization is then separated into a Light gasoline and a Heavy gasoline stream in the Redistillation unit. The straight-run fractions need further processing and blending to make marketable products

4.3.2 Catalytic Reforming Unit

This unit is preceded by Pre-treatment unit where heavy gasoline (HG) from Atmospheric Distillation unit (Re-distillation unit) is desulfurized and hydrogenated to remove/ reduce catalyst poisons e g. Sulfur, Nitrogen, Oxygen, Arsenic, etc. from the feed. Flow diagram of Catalytic Reforming Unit with Pretreatment is shown in Figure 4 2.

The charge is mixed with recycle gas rich in hydrogen from reforming section, following preheating the mixture is heated in furnace and then fed into fixed bed reactor, where the charge is catalytically reformed over a catalyst containing Co-Mo on alumina carrier. The catalytic reaction is performed at high temperature (330-350°C) and high pressure (18 - 35 bars). After cooling, the reaction products are flashed in to separation drum. Flashed gases containing hydrogen sulfide along with mostly unreacted hydrogen are sent to the fuel gas collection header. Flashed liquid are sent to Stripper column (C1201) where all the water, ammonia and hydrogen sulfide are removed from the liquid. Gaseous distillate flowing out of the top of the tower also contains the remaining light hydrocarbon gases from the separation product. This stream is partially condensed and provides a liquid reflux to the top of the column.

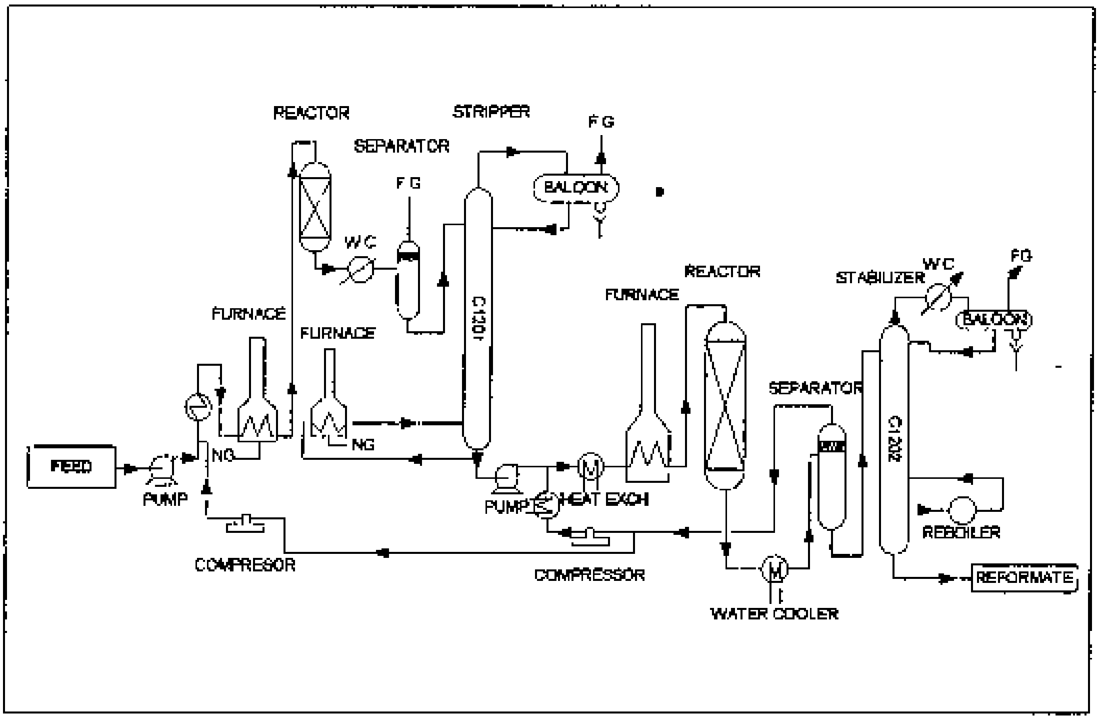

Figure 4.2 Catalytic Reforming Units with Pretreatment

Condensed water including the evolved ammonia in solution is separated in the reflux drum and sent to water decantation pot. Uncondensed hydrogen sulfide and light hydrocarbon gases supply to the fuel gas system. Stripping is provided with reboiler furnace which vaporizes part of the bottom liquid and sufficient vapor flow is provided to ensure efficient stripping action. Stripped liquid from the bottom of the column is charged to reforming section.

The charge at 480-500°C is mixed with hydrogen recycle, preheated in exchanger and heated in furnace to about 510-530°C, is passed through reactor containing a catalytic bed of precious metal Platinum Most of this reaction is endothermic. Therefore the charge is heated between the different reactors by furnace. The catalytic reforming reaction takes place at temperature of 425-510°C under pressure of 20 bar and subsequent cooling, separation and stabilization of the reactor effluent produce reformate of 86 to 89 RON (Research octane number). Increase of octane number from around 52 to 89 is mostly due to aromatization naphthenic hydrocarbons present in the feed Reformat is used for production of HOBC (100 octane) and is also blended with LG (Sweet) for augmenting octane number of Motor Spirit (MS).

4.3.3 Asphaltic Bitumen Unit

In this unit, atmospheric residue named as RCO (Reduced Crude Oil) of Arabian light crude (ALC) is further distilled under high vacuum, to recover light and heavy vacuum gas oils. Vacuum residue is then oxidized by air blowing in the Bitumen Blowing Unit to produce Bitumen of various grades. Flow diagram of Asphaltic Bitumen Unit is shown in Figure 4.3.

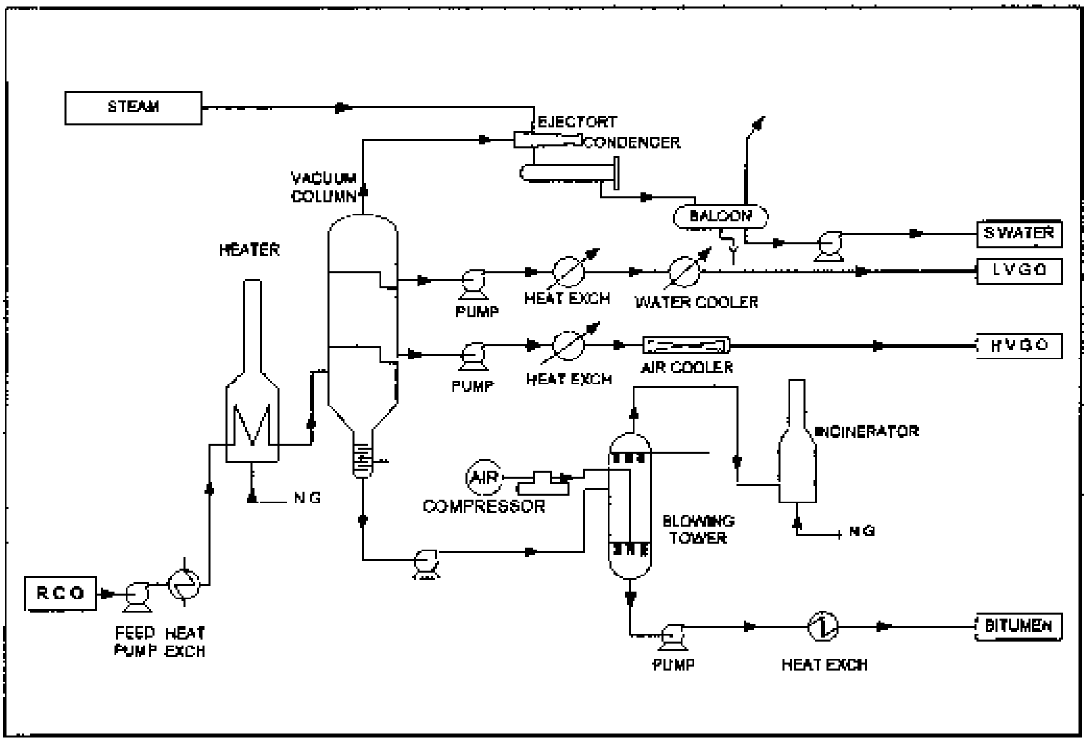

Figure 4.3 Asphaltic Bitumen Unit

RCO from storage is pumped by the feed pump and is preheated by heat exchanger. The preheated RCO is sent to fired heater and heated to about 390 °C. Heater effluent flashed in the Vacuum Distillation column. The flash zone temperature and pressure determine the quality of the vacuum residue Pressure is maintained at 51 mmHg by steam ejector control.

Vacuum column is consists of three packed bed From the top bed LVGO (Light vacuum Gas oil) is obtained, after cooling it is blended with gas oils to increase production of HSD. From the intermediate bed HVGO (Heavy vacuum gas oil) is obtained, after cooling through exchanger and air cooler it is used as a useful stock for furnace oil production. Vacuum residue is pumped to the Blowing tower for production of Bitumen. Air is blown to the tower by air compressor. Top out let of this tower is burn in Incinerator as wastage. HVGO is now being used as a feed for Mild Hydrocracking (MHC) Unit.

4.3.4 Visbreaker Unit

Visbreaker is a mild thermal cracking process where heavy hydrocarbons (Residue) are converted into comparatively lighter hydrocarbons and in this process viscosity of the residue is reduced considerably (Figure 4.4).

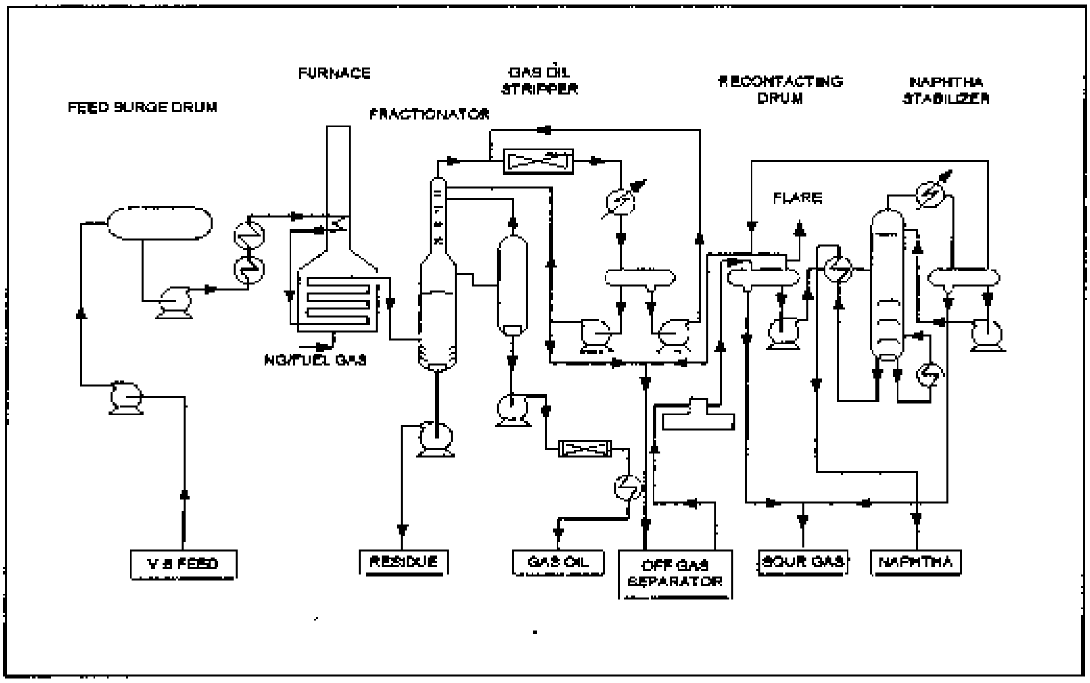

Figure-4.4 Visbreaker Unit

Atmospheric Residue from storage passes through surge drum is preheated with hot product-streams and finally heated in the Visbreaker furnace. Under the effect of temperature (480 °C), pressure (20 bars) and residence time, thermal creaking takes place inside the heater coil. Cracked products are passes to the fractionators for the separation of Naphtha minus, Gas oil and Residue fraction. The over head vapor from the fractionator are mixed with recycle sour water then condensed to yield a reflux back to the top of the fractionator.

After recompression the unstabilized naphtha, the compressed gases are recontacted at a high pressure to improve the recovery of Naphtha. The vapors from the recontacting systems are sent as light gas to the fuel gas header while the liquid is debutanized in a Naphtha stabilizer. Fuel gas and sour gas removed from the process are burnt as fuel in the furnace. Gas oil and visbroken residue are taken into diesel and furnace oil pool respectively.

4.3.5 Mild Hydrocracking (MHC) Unit

Heavy vacuum gas oil (HVGO) from storage is fed by pump, preheated in heat exchanger train, is mixed with compressed recycle gas rich in hydrogen and before entering the Reactor feed heater it is again mixed with Make-up hydrogen. The Make-up hydrogen come from the Hydrogen plant and is compressed by the compressor (Figure 4.5).

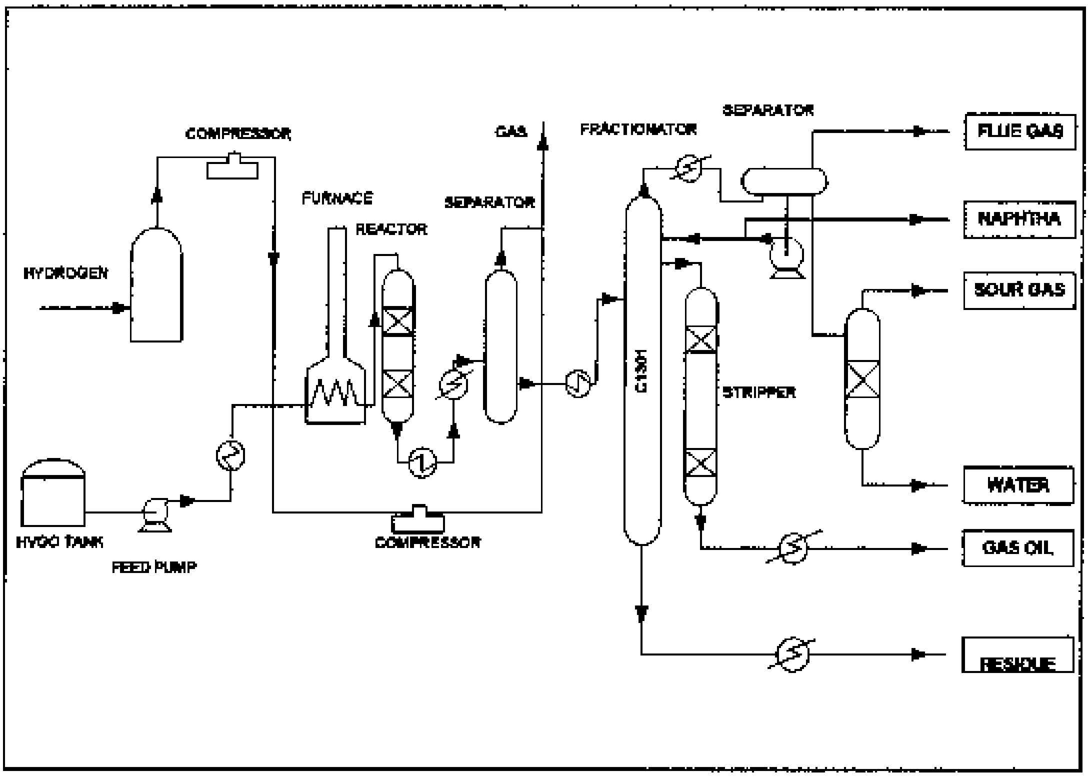

Figure.4.5 Mild Hydrocracking Unit

In the reactor feed heater the mixture feed, recycle gas and make-up hydrogen is heated in the furnace to about 405 °C to obtain the required conversion. In the MHC reactor, reaction of mild hydrocracking, desulphurization and other auxiliary reaction takes place Since the reaction, which occurs in the reactor is exothermic, the temperature of the effluent leaving the reactor is increased to about 415 °C. The effluent leaving the reactor is cooled and partially condensed in the exchanger and further cooled in water cooler and then flushed into separator.

The effluent is separated into following two streams:

- The gas leaving the separator at the top are mainly compressed by the recycle compressor
- The hydrocarbon phase which is fed to the Fractionator for fractionation.

The feed enters the Fractionator (C1301) at a temperature of 376 °C. Resulting products are separated in the fractionators as Fuel gas, Naphtha, Gas oil, and Residue.

4.4 Hydrogen Unit

This is an auxiliary unit of Mild Hydrocracking Unit The unit produces Hydrogen of 99.9% purity from steam reforming of natural gas. Flow diagram of Hydrogen Unit is shown in Figure 4.6.

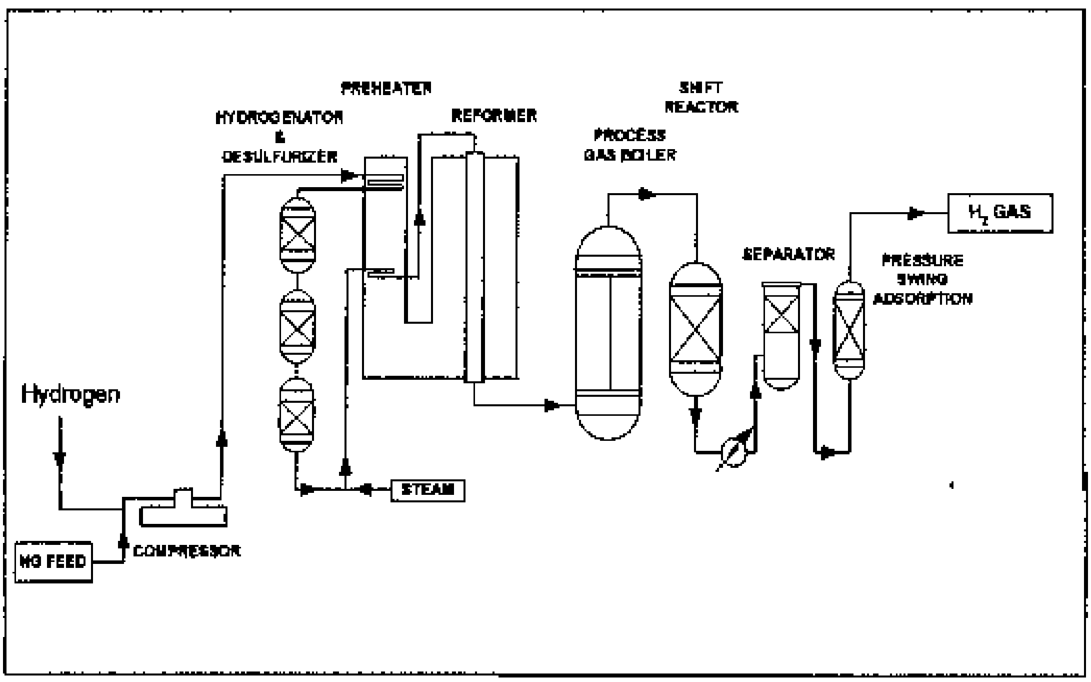

Figure-2.6 Hydrogen unit

Natural gas feed from battery limit at 4.1 bar is mixed with hydrogen recycle (5% of feed stock by volume) are than compressed through a compressor to 26 bar. The feed is then preheated to 400 °C in the Reformer connection section. The gas is then passes through hydrogenator and desulfurizer where sulfur compounds are converted to H_2S. The outlet temperature is 380 °C due to heat loss from the vessel. The hydrocarbon fed is mixed with saturated steam. The ratio between the two flows is 3.31 kg steam/kg feed which corresponds to a steam/carbon ratio of 3.0 on molar basis. The mixed feed /steam (250°C and 24.6 bars) is sent to the reformer feed preheater to be superheated to about 540°C before entering the reforming reactor. To obtain desired conversion of steam reforming, the temperature of the reactor maintained at about 865°C. The reformer effluent is cooled down in the process gas boiler and is fed to the high temperature shift reactor. The converted effluent is cooled to 41 °C in cooler and sent to a separator, where condensate is separated from the gaseous phase (raw hydrogen). Raw hydrogen from separator is fed to the PSA (Pressure swing adsorption) system where high purity hydrogen is produced by adsorption of the impurities on active carbon and molecular sieve. The purity of hydrogen here is 99.9 %.

4.5 Finished Products

ERL produces 16 petroleum products including four non-fuel products, namely, SBP, JBO, MTT, and BITUMEN. Some of these products are produced only on demand. A brief description of products with few specifications (Product Specification, 1991) follows:

- **RG** (Refinery Gas): It is not a marketable product. It contains mostly C_1-C_2 hydrocarbons and is used as fuel in heaters of refinery.

- **LPG** (Liquefied Petroleum Gas): It contains mostly Propane & Butane and is marketed in steel cylinders by a separate company named LP Gas Ltd. It is very popular as cooking fuel and is also used in motor vehicles. Compositions are Ethane, Propane, Butane and Pentane

Test	Method	Limit
Density at 15 °C, kg/Liter	ASTM D 657	Min. 0.55 Max. 0.60
Vapor pressure at 40 °C, Psi	ASTM D 1267	Max. 213
Volatility. Evaporation temp. for 95 % vol. °C	ASTM D 1837	Max. 2
Copper strip corrosion (1 hr. at 37 8 °C)	ASTM D 1835	Max. no. 1
Total volatile Sulphur, % mass	ASTM D 2784	Max 0 02
Hydrogen sulphide	UOP-212	Nil

- **SBP** (Special Boiling Point Solvent): It is essentially a sweetened light gasoline (LG) and is used as industrial solvent and also in dry-cleaning.

Test	Method	Limit
Sp. gr @60°F/60°F	ASTM D 1298	Min. 0.670
Colour, say bolt	ASTM D 156	Min. +25
Doctor Test	ASTM D 235	Negative
Aromatic content, % vol.	ASTM D 1319	Max. 5.0
Sulphur content, % wt.	ASTM D 1266	Max. 0.05
Distillation: Initial boiling point °C Final boiling point, °C Residue , % vol.	ASTM D 86	 Min. 30 Max. 115 Max. 1.5

- **MS** (Motor Spirit): Popularly known as petrol, is used as fuel in petrol engines. It is a light petroleum distillate dyed orange and has an octane rating of 80 RON (Research Octane Number).

Test	Method	Limit
Density at 15°C	ASTM D 1298	Min. 0.670
Colour, visual		Orange
Reid vapor pressure at 38 °C , Psi	ASTM D 235	Max. 10
Octane number (research method)	ASTM D 2699	Min. 80
Lead content(as pb), g/L	ASTM D 3341	Max. 0.5
Residue on evaporation, mg/100 ml	ASTM D 381	Max. 4
Sulphur , total, % mass	ASTM D 1266	Max. 0.1
Doctor test	ASTM D 235	Negative
Distillation . Initial boiling point, °C Final boiling point, °C Residue, % vol.	ASTM D 86	 To be reported Max. 210 Max. 2

- **HOBC** (High Octane Blending Component): Popularly known as 100 Octane (because, its octane number is close to 100) is used in combination with MS in petrol engines. The base stock is reformate, produced in Catalytic Reforming Unit. It is dyed red for marketing.

Tests	Method	Limit
Density at 15 °C, kg/L	ASTM D 1298	To be reported
Colour, visual		Red
Reid vapor pressure at 38°C, Psi	ASTM D 323	Max. 10
Octane number(Research method)	ASTM D 2699	Min. 96
Lead content (as pb), g/L	ASTM D 3341	Max. 0.84
Residue on evaporation mg/100ml	ASTM D 381	Max. 4
Sulphur, total, % mass	ASTM D 1266	Max 0.1
Doctor test	ASTM D 235	Negative
Distillation Initial Boiling , °C Final Boiling Point, °C	ASTM D 86	 To be reported Max. 210

- **NAPHTHA:** This is essentially untreated light and heavy gasoline that remains as excess. It is mainly used as feed for petrochemicals and also as solvents. Naphtha is mostly an exportable product.

Tests	Method	Limit
Sp. gr. @60°F/60°F	ASTM D 1298	Min. 0.670 Max. 0 740
Colour, say bolt	ASTM D 156	Min. +20
Reid vapor pressure at 38°C, Psi	ASTM D 323	Max. 12
Lead content (as pb), ppb	IP 224	Max. 50
Residue on evaporation mg/100ml	ASTM D 381	Max. 4
Sulphur, total, % mass	ASTM D 1266	Max. 0.075
Doctor test	ASTM D 235	Positive
Distillation Final Boiling Point, °C	ASTM D 86	 Max. 180
Paraffin content, % vol	ASTM D 1319	Max. 2
Aromatic content , % vol.	ASTM D 1319	Max. 15

- **SKO** (Superior Kerosene Oil): It is commonly known as Kerosene and mostly used for illumination and as a cooking fuel.

Tests	Method	Limit
Density at 15 °C, kg/L	ASTM D 1298	To be reported
Colour, ASTM	ASTM D 1500	Max. 3.0
Flash point °C	IP 170	Min. 43
Smoke point, mm	ASTM D 1322	Min. 10
Sulphur , total, % mass	ASTM D 1266	Max. 1.0
Copper strip corrosion	ASTM D 130	Max. no. 2

- **MTT** (Mineral Turpentine): It is the lighter kerosene specially treated for use as thinner and solvent in paints and varnishes. It is known as Turpentine oil or White sprit and is produced only on demand.

Tests	**Method**	**Limit**
Sp. gr. @60°F/60°F	ASTM D 1298	Min. 0.770
Colour, say bolt	ASTM D 156	Min. +18
Flash point °F	IP 170	Min. 85
Aromatic contents, % vol.	ASTM D 1319	Min. 14
Existent Gum, mg/100ml	ASTM D 381	Max. 5
Distillation	ASTM D 86	
Initial boiling point °C		Min. 135
Final Boiling Point, °C		Max. 210
Residue, % vol.		Max. 1.5

- **JP-1** (Jet Propulsion-1): It is specially treated kerosene with additives for improving certain product properties e g. very low freezing point (minus 47 deg. Celsius). This product is used as fuel of aviation turbine engine. JP-1 is now produced only on demand

Tests	**Method**	**Limit**
Sp. gr. @60°F/60°F	ASTM D 1298	Min. 0.775 Max. 0.830
Colour,		Clear bright
Flash point °F	IP 170	Min. 85
Freezing point °C	IP 16	Max. -47
Aromatic contents, % vol.	ASTM D 1319	Max 22
Sulphur, total, % mass	ASTM D 1266	Max 0.3
Smoke point. mm	ASTM D 1322	Min 25
Electrical conductivity Ps/m	IP 274	Max. 450 Min 50

Antioxidant, mg/L ,(Topanol)		Max 24
Distillation Initial boiling point °C Final Boiling Point, °C Residue, % vol.	ASTM D 86	 Report, Max. 300 Max. 1.5

- **JBO** (Jute batching oil): This is a brown colored straight-run petroleum distillate used in the form of emulsion (water in oil) for softening, for removal of dust and other impurities, before spinning of jute. It is an important nonfuel petroleum product.

Tests	Method	Limit
Sp. gr. @60°F/60°F	ASTM D 1298	Min. 0.825
Colour, ASTM	ASTM D 1500	Max 7 0
Flash point °F	ASTM D 93	Min. 100
Pour point, °C.	ASTM D 97	Max 30
Emulsification test	BDS 1448	Shall pass the test
Doctor test	ASTM D 235	Positive
Distillation Initial boiling point °C Final Boiling Point, °C Residue, % vol.	ASTM D 86	 Min. 250 Max. 371 Max. 2

- **HSD** (High Speed Diesel): It is popularly known as Diesel and is used as fuel in high speed diesel engines of automobiles and irrigation pumps It is also used in diesel generators for power generation. It is a light yellow colored distillate fuel.

Tests	Method	Limit
Sp. gr. @60°F/60°F	ASTM D 1298	Min. 0.820 Max 0.870
Colour, ASTM	ASTM D 1500	Min. 3.0
Flash point °F	IP 170	Min 32
Pour point, °C	ASTM D 97	Max.9 (for winter) Max.15(for summer)
Cetane no.	ASTM D 613	Min 45
Sulphur, total, % mass	ASTM D 2622	Max. 1.0
Ash, % mass	ASTM D 482	Max. 0.01
Kinematic viscosity at 38 °C, cst	ASTM D 445	Max. 9

- **LSDO** (Low Sulfur Diesel Oil): It is a special diesel oil of lower sulfur content and is used as fuel in medium speed marine engine of Bangladesh Navy's Vessels.

Tests	Method	Limit
Sp. gr. @60°F/60°F	ASTM D 1298	Report
Carbon residue, % mass	ASTM D 189	Max. 0.5
Flash point °F	ASTM	Min. 80
Pour point, °C	ASTM D 97	Max -10
Cetane index.	ASTM D 976	Min. 50
Sulphur, total, % mass	ASTM D 2622	Max. 0.5
Ash, % mass	ASTM D 482	Max. 0.02
Kinematic viscosity at 40 °C, cst	ASTM D 445	Max. 4.1 Min. 2.3

- **LDO** (Light Diesel Oil): It is a dark coloured diesel fuel used in large stationary or low speed (less than 600 rpm) marine diesel engines.

Tests	Method	Limit
Sp. gr. @60°F/60°F	ASTM D 1298	Min. 0.890
Carbon residue, % mass	ASTM D 189	Max. 2.0
Flash point °F	ASTM D 93	Min. 66
Pour point, °C	ASTM D 97	Max. 12(for winter) Max.18(for summer)
Sediment, % mass	ASTM D 473	Max. 0.1
Sulphur, total, % mass	ASTM D 2622	Max. 1.8
Ash, % mass	ASTM D 482	Max. 0.02
Kinematic viscosity at 38 °C, cst	ASTM D 445	Max. 16

- **HSFO** (High Sulfur Furnace Oil): It is commonly known as FO (Furnace Oil). This high sulfur residual fuel is now produced by blending straight run residue with visbroken residue and MHC unit residue. It is used solely as fuel for burning in furnaces and boilers.

Tests	Method	Limit
Sp. gr. @60°F/60°F	ASTM D 1298	Min. 0.890 Max. 0.96
Carbon residue, % mass	ASTM D 189	Max. 10
Flash point °F	ASTM D 93	Min. 66
Pour point, °C	ASTM D 97	Max. 33
Sediment, % mass	ASTM D 473	Max 0.1

Sulphur, total, % mass	ASTM D 2622	Max. 3 5
Carbon residue , % mass	ASTM D 189	Max. 10
Kinematic viscosity at 38 °C, cst	ASTM D 445	Max. 16
Calorific value, kcal/kg	ASTM D 240	Min. 10250

- **LSFO** (Low Sulfur Furnace Oil): This is the straight- run Murban residue. It is so named as its sulfur content is much lower (1.5% wt). Production of LSFO has been discontinued.

Tests	Method	Limit
Sp. gr. @60°F/60°F	ASTM D 1298	Min. 0.890 Max. 0 950
Carbon residue, % mass	ASTM D 189	Max. 10
Flash point PM (c.c)°F	ASTM D 93	Min. 150
Pour point, °C	ASTM D 97	Max. 90
Sediment, % mass	ASTM D 473	Max. 0.25
Sulphur, total, % mass	ASTM D 2622	Max. 1.5
Carbon residue , % mass	ASTM D 189	Max. 10
Kinematic viscosity at 100 °F, cst	ASTM D 445	Max. 150 Min. 25
Calorific value, kcal/kg	ASTM D 240	Min. 10230

- **BITUMEN:** Bitumen is the desired product of Asphaltic Bitumen Plant. This is a very important non-fuel petroleum product. ERL produces three grades of Bitumen namely 80/100, 60/70 and 10/20.The 80/100 and 60/70 grades are used for construction of bitumen-carpeted roads. The 10/20 grade Bitumen is used mostly as insulation material in cold storages. Specifications of different grades of Bitumen are given below.

Tests	Method	Limit		
		Grade:10-20	Grade:60-70	Grade:80-100
Sp.gr.@25°C/25°C	ASTM D 70	Min. 1.02 Max. 1.07	Min. 1.01 Max 1.06	Min. 1.01 Max. 1.06
Penetration at 25°C (100g,5 secs),	ASTM D 5	Min. 10 Max. 20	Min. 60 Max 70	Min. 80 Max. 100
Softening point, °C (Ring & Bell)	ASTM D 36	Min. 80 Max. 95	Min. 45 Max 56	Min. 44 Max. 55
Ductility at 25°C	ASTM D 113	Min. 0	Min. 100	Min. 100
Flash point °C	ASTM D 92	Min 250	Min. 230	Min. 230
Loss on heating,% wt	ASTM D 6		Max. 0.5	Max. 0.5

4.6 Crude Oil Processing

ERL processes low sulfur light crude of Middle Eastern countries namely Murban Crude (of Abu Dhabi) and Arabian Light Crude (of K.S.A.). Of these two, only ALC is suitable for bitumen production. On the other hand, Murban crude contains more valuable Light ends (LPG, Gasoline) and Middistillates (Kerosene & Diesel).

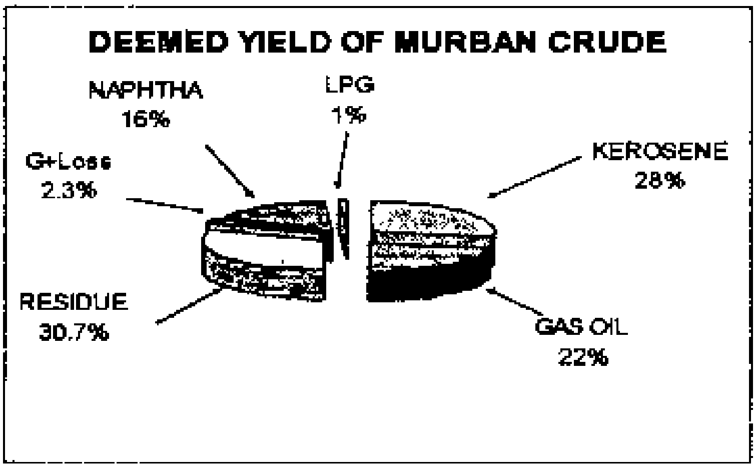

Figure 4.7 Deemed yield of Murban Crude

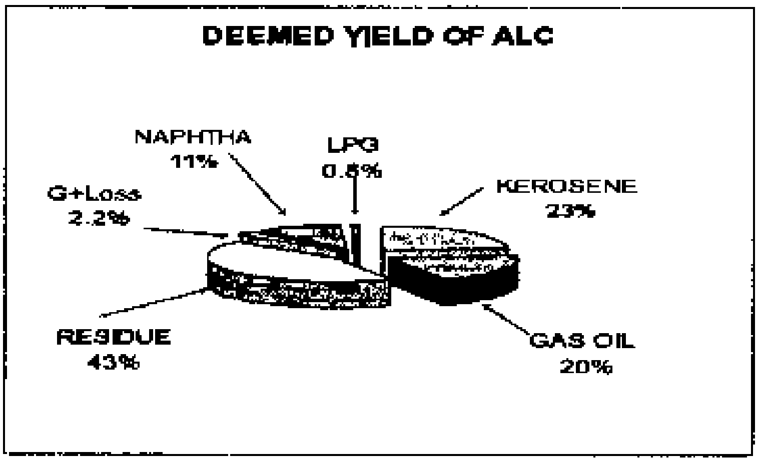

Figure 4.8 Deemed yield of Arabian Light Crude

Figure-4.7 and 4.8 shows deemed yield of general categories of products from processing of ALC and Murban Crude. ERL's processing strategy therefore, has been processing maximum quantity of Murban crude for maximizing light ends and middistillates production while processing just enough quantity of ALC required for meeting country's Bitumen demand. Recent addition of residue upgrading units will only slightly change this strategy. ERL also processes whatever crude oil it receives from Haripur field, the only oil field in Bangladesh and condensates received from gas fields.

With the addition of standby equipment and tankages, and with the improvement of utilities over the years, the operational reliabilities have improved remarkably. For the last 8 years or so, ERL has maintained on stream factor of around 330 days or 8000 hours per year and during this period ERL has treated 1.0-1.3 million MT of crude oil annually. It needs to be mentioned that such treatment target is set solely on overall economic consideration. In coming years, subject to economic viability, ERL plans to treat 1.3 to 1 4 million MT of crude annually. The histogram (Figure 4.9) shown below gives last 10 year's crude processing status.

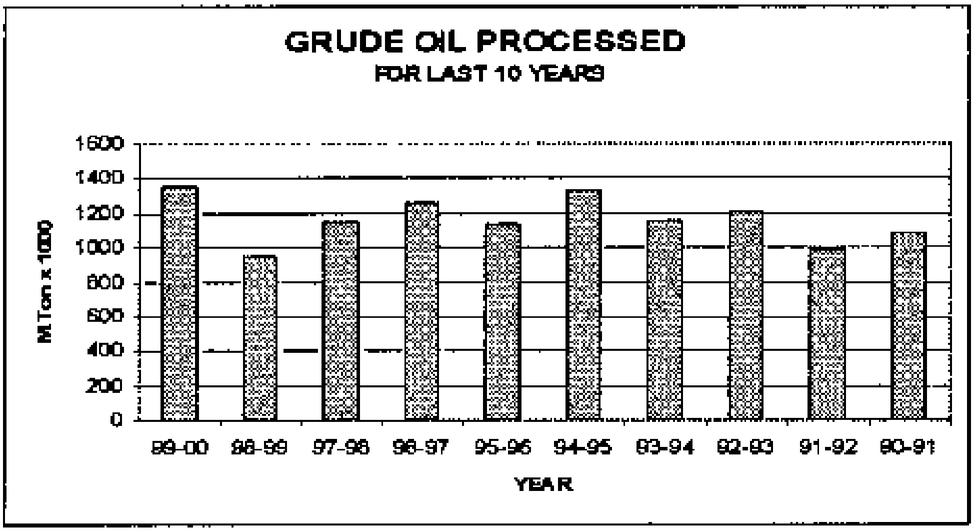

Figure 4.9 Crude oil processed for the last 10years

4.6.1 Processing expense of crude oil

The processing expense of ERL is increasing day by day. It includes the overhead expenditure related to crude oil processing. Processing expenses can be kept at minimum level by processing maximum amount of crude oil with proper utilization of existing facilities. Processing expense of Crude oil in Taka per Barrel for the last ten years with the amount of crude oil processed is shown in Table 4.2.

Table-4.2 Processing expenses of ERL for the last 10 years (ERL, 1999-2000)

Year	Crude oil Processed per year M.Ton	Processing Expense per year Taka in Lacs	Processing Expense Per M.Ton (Tk.)	Processing Expense Per Barrel (Tk.)
1990-91	1109690.00	2554.24	230.18	32.88
1991-92	1017315.00	3146.45	309.29	44.18
1992-93	1240740.00	3771.51	303.97	43.42
1993-94	1194723.00	3859.11	323.01	46.14
1994-95	1373598.00	5756.84	419.11	59.87
1995-95	1159930.00	6163.96	531.41	75.92
1996-97	1323765.00	6256.29	472.61	67.52
1997-98	1176320.00	5806.68	493.63	70.52
1998-99	996240.00	6146.81	617.00	88.14
1999-2000	1377600.00	5660.44	410.89	58.70
Total	11969921.00	49122.33	4111.10	587.30
Average	1196992.10	4912.23	411.11	58.73

Graphical representation of processing expense in Taka per Barrel for the last ten years is shown in Figure 4.10.

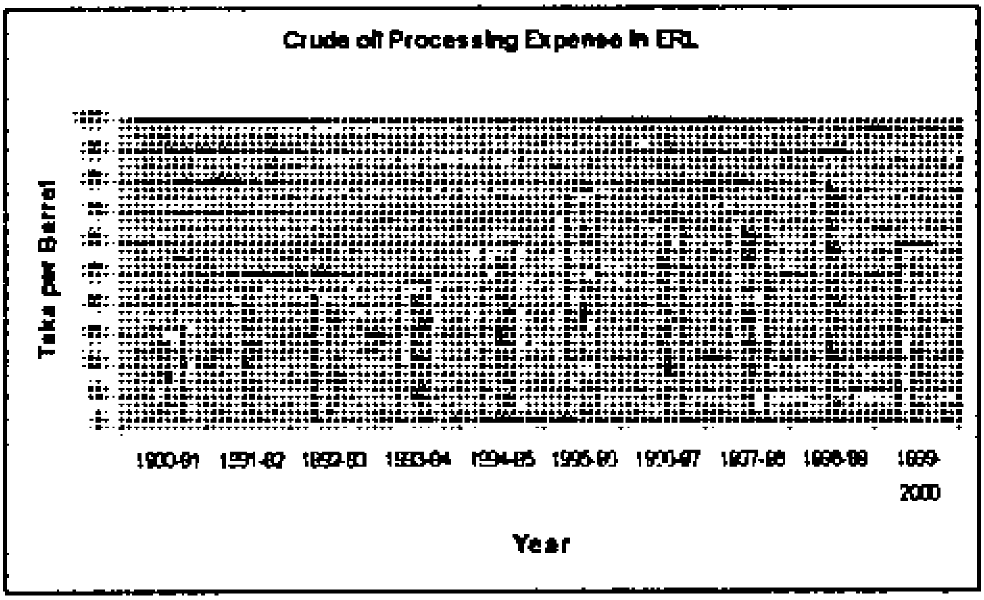

Figure 4.10 Processing expenses of ERL for the last 10 years

4.6.2 Cost of energy in processing crude oil

The overhead processing expenses of crude oil include the cost of energy. In ERL fuel, power, Bakhrabad gas and water are taken as the sources of energy. The cost of energy consumed in processing crude oil for the last five year is shown in Table 4.3.

Table 4.3 Cost of energy in processing crude oil for the last five years

Particulars	1995-96	1996-97	1997-98	1998-99	1999-2000
Fuel, Power & Water (Tk. in lac)	68.02	102.09	65.64	45 46	45 33
Bakhrabad Gas (Tk. in lac)	923.48	683.82	749 44	695.47	921 94
Total cost of energy (Tk in lac)	991.5	785.91	815.08	740.93	967.27
Crude oil Processed(M.Ton)	1159930	1323765	1176320	996240	1377600
Cost of Energy per M.Ton(Tk.)	85.48	59.37	69.29	74 37	70.21
Cost of Energy per Barrel(Tk)	12.21	8.48	9.90	10.62	10.03

Graphical representation of cost of energy in processing crude oil in Taka per barrel for the last five years is shown in Figure 4.11.

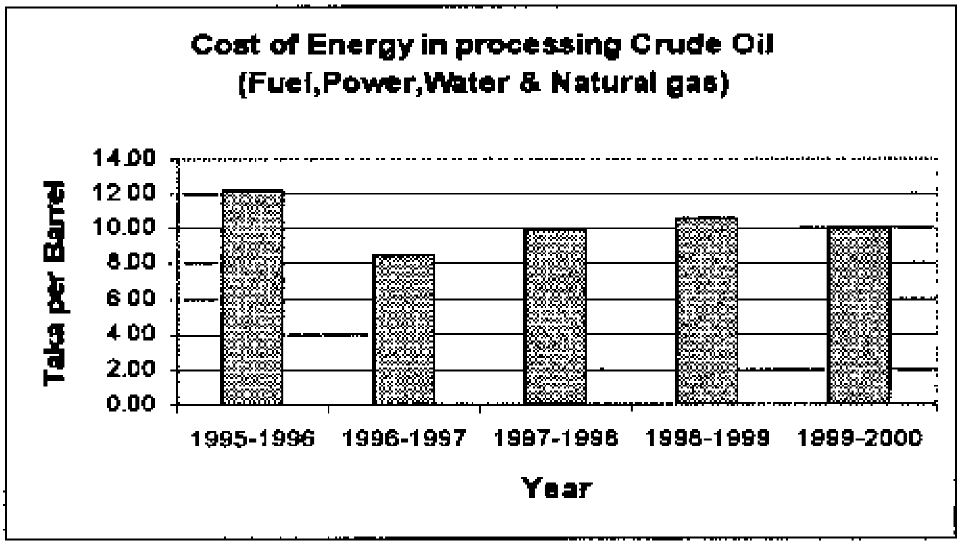

Figure 4.11 Cost of energy in processing crude oil for the last five years

CHAPTER 5

Details Study of Crude Distillation Unit

5.1 Introduction

Crude distillation is usually referred to as complete operation in which heating, vaporization, fractionation, condensation, and cooling of crude oil are practiced. Crude Distillation Unit is the primary crude oil processing unit and is the starting point for the petroleum refinery In this unit, crude oil is preheated with hot product streams, desalted, and a part of it is then flash vaporized for introducing in the distillation column. The rest is finally heated up to around 370 °C in a furnace and fed into the Distillation column for fractionation

5.2 Description of Crude Distillation Unit

The detailed flow diagram of Crude Distillation Unit of ERL is shown in Figure.5 1. The crude oil is pumped directly from storage tanks by feed Pump (P1101A/B) and sent to the Flash drum B1101 after desalting and thermal exchange with the following products:

- Kerosene I (K I) in the Exchanger E1101
- Kerosene II (K II) in the Exchanger E1102
- Cold light gas-oil (LGO) in the Exchanger E1103
- Cold circulating reflux in the Exchanger E1104
- Hot circulating reflux in the Exchanger E1105.

Here 'E' stands for exchanger, '11' stands for crude distillation unit and 01, 02, 03.... is the serial no of the exchanger. After preheating the crude through the following exchangers the feed is introduced in the Furnace F1101 by four passes.

- Hot light gas oil (LGO) in the Exchanger E1106
- Heavy gas oil (HGO) in the Exchanger E1107
- Cold Residue in the Exchanger E1108
- Hot Residue in the Exchanger E1109A/B.

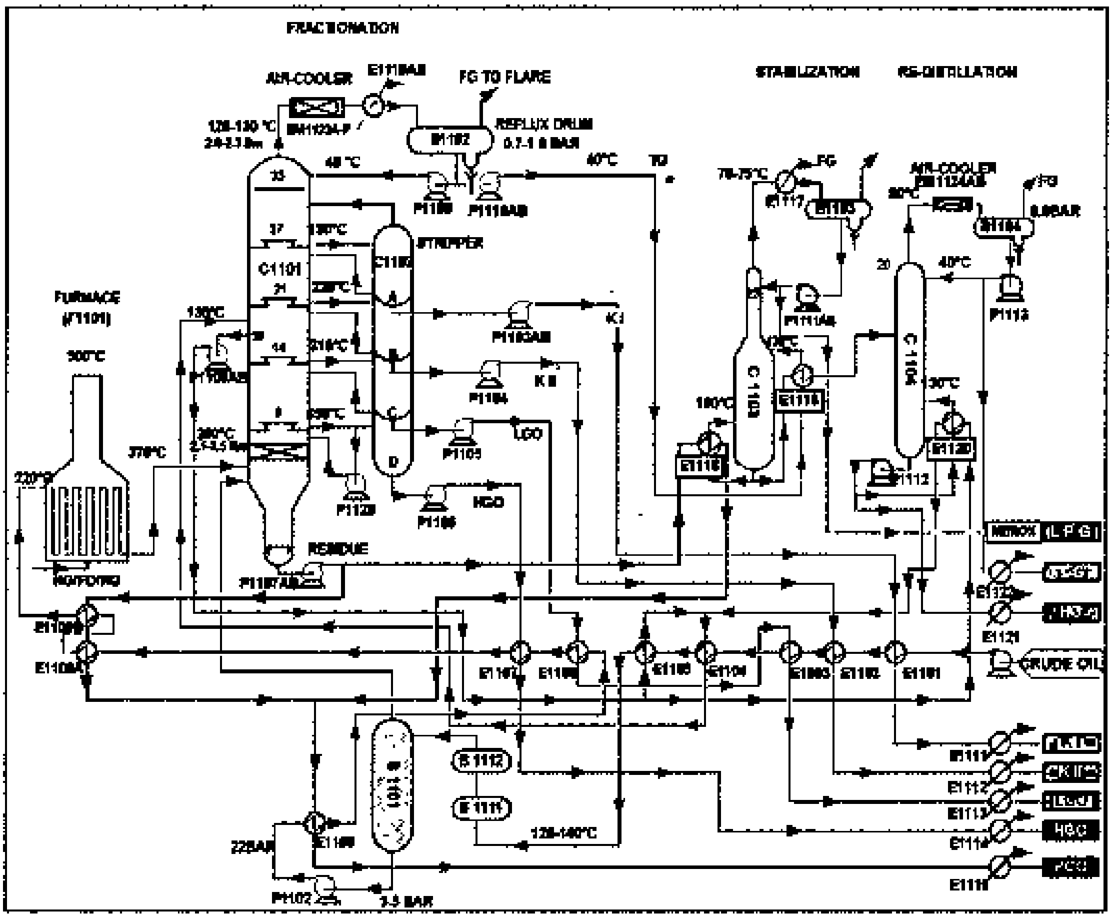

Figure 5.1 Flow diagram of Crude Distillation Unit

The flow rate is controlled by the flow rate control valves feeding each one through two passes; the adjustment of the passes is done manually by acting on the glove valve, which is placed on each of them The outlet temperature of the furnace is automatically controlled by the flow of fuel oil or fuel-gas. The furnace has a steam overheating section, which is used for the stripping of the atmospheric section. At the out let of the Furnace (F1101) the partially vaporized crude is then sent in the Atmospheric Distillation column (C1101) which is equipped with 33 trays. In this distillation column the crude oil is fractionated to give the following products:

- Total gasoline (TG)
- Kerosene I (K I)
- Kerosene II (K II)
- Light gas-oil (LGO)

- Heavy gas-oil (HGO)
- Atmospheric residue (Residue)

The Total Gasoline and the steam used for different stripping are the overhead products of the column. The overhead is condensed in the Aero-condenser E1123 then in the Water condenser E1110A/B. The pressure in the Reflux drum B1102 is alternately controlled by departure of gas to the flare or by admission of fuel gas. In this reflux drum, gasoline and water are separated by decantation Part of the gasoline is taken out by the Pump P1109 to be refluxed under temperature control at the top of the column. The other part is taken out by Pump P1110A/B and sent as feed for the stabilization column under the level control valve of B 1102. Water is evacuated by gravity to the sewer under liquid control of the Drum B 1102.

Kerosene I is withdrawn from Tray 27 and flows by gravity into the Stripper C1102A under the level controller of the stripper. The stripping of Kerosene- I is done by the superheated steam. Stripping is required to reduce the partial pressure of hydrocarbon mixture which gives fractionation at low temperature The stripped Kerosene- I is taken out by the Pump P1103A/B. After exchanging heat with the crude in the Exchanger E1101, it is cooled in the Water cooler E1111 and sent under the flow control, partly to the Kerosene- I sweetening unit (Merox), rest to the storage.

Kerosene II is withdrawn from Tray 21 and flows by gravity into the Stripper C1102B under the level controller of the stripper. The stripping of Kerosene II is done by superheated steam. The stripped product is taken out by the Pump P1104 to be sent into the Exchanger E1102 in which there is a heat exchange with the crude. It is then cooled in the Water cooler E1112 and sent under flow rate control, partially to the storage, remainder to the sweetening unit (Merox).

Part of the heat of the atmospheric column is recovered below the withdrawal tray of Kerosene II by circulating reflux. The circulating reflux is withdrawn from Tray 18, taken out by the Pump P1108A/B and partially sent to the Exchanger E1105 where it exchanges heat with the crude, remainder to the Reboiler E1120 of the redistillation column. The outlet temperature of the reboiler is controlled by the circulating reflux flow-rate directed to this one through the temperature control valve located at the outlet of the Exchanger E1105. The whole circulating reflux, then, exchanges heat with the crude in the Exchanger E1104 and is reintroduced in the column under Tray 21. The temperature control valve acting on the by-pass of the Exchanger E1104 controls the back temperature of the circulating reflux.

Light gas oil is withdrawn from Tray 14 and flows by gravity into the Stripper C1102C under the level controller of the stripper. The stripping of the light gas-oil is done by super heated steam The light gas-oil is taken out by the Pump P1105 to be sent in the Exchanger E1106 and E1103, in which it exchanges heat with the crude. It is then cooled in a Water cooler E1113 to be sent to the storage under flow-rate control.

Heavy gas-oil is withdrawn from Tray 9 and flows by gravity to the Stripper C1102D under the level controller of the stripper. The stripping of the heavy gas-oil is done by super heated steam. Then heavy gas oil is taken out by the Pump P1106. After exchanging heat with the crude, in the Exchanger E1107, it is cooled in a Water cooler E1114 to be sent to the storage under flow-rate control.

Atmospheric residue is stripped by superheated steam in the bottom of the atmospheric column. The stripped residue is taken out by the Pump P1107A/B and partially sent to the Exchanger E1109 A and B, where it exchanges heat with the crude, partially to the Reboiler E1118 of the stabilization column. The outlet temperature of the reboiler is controlled by the flow-rate of the residue by temperature control valve .Then, the entire residue exchanges heat with the crude in the Exchanger E1108 and is cooled in the Water cooler E1115 to be sent to the storage under flow-rate control.

To prevent corrosion in the top of the column and its equipment, the following is provided.

- A continuous ammonia injection under Tray 31,
- A corrosion inhibitor injection in the headline.

5.3 Stabilization of Gasoline

The total gasoline from top Reflux drum B1102 is pumped by Pump P1110A/B is preheated by stabilized gasoline in the Exchanger E1116A and B, before being introduced into the Stabilization column C1103 which is equipped with 25 trays (Figure 5.2). The inlet temperature of the column is controlled by the by-pass of stabilized gasoline around E1116.

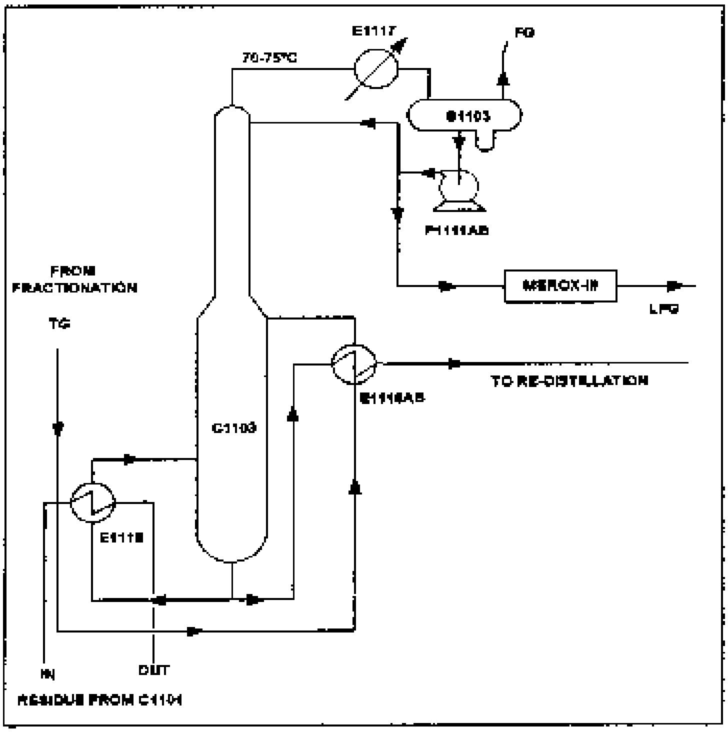

Figure 5.2 Stabilization

<u>**The overhead vapors**</u> are partially condensed in the Water condenser E1117. The vapor distillate is evacuated to the fuel gas system. The evacuation of gases controlling the pressure in the Reflux drum B1103 is done by the pressure control valve A part of the condensed vapor is taken out as reflux by the Pump P111A/B and reintroduced in the column above Tray 25 under temperature control flow and rest is feed to merox unit to sweeten to produce LPG.

<u>**The stabilized gasoline**</u> going out from the bottom of the Column C1103, feeds the Redistillation column C1104 under bottom level control valve, after exchanging heat with the feed in the Exchanger E1116. The reboiling of the column is done by

the Reboiler E1118 which is heated by a part of the atmospheric residue. The outlet temperature of the reboiler is controlled by the temperature control valve acting on the flow-rate of atmospheric residue going through E1118.

5.4 Redistillation of Gasoline

The Redistillation column (C 1104) is equipped with 20 trays (shown in Figure 5.3). Stabilized gasoline feed to redistillation column for further fractionation.

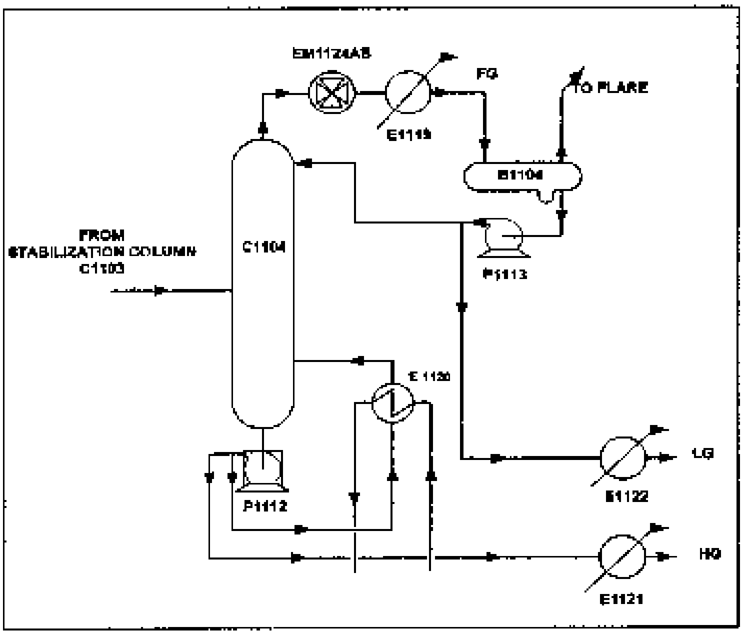

Figure 5.3 Redistillation Unit

<u>The overhead vapors</u> of redistillation column are cooled through Aero condensers EM 1124A and B then condensed in the Water cooler E1119 and constitute the Light gasoline. The pressure in the reflux drum is alternately controlled by departure of gas to flare and by admission of fuel gas.

<u>The light gasoline</u> is pumped out by the Pump P1113 and sent, partly as reflux under flow-rate control and rest to the Water cooler E1122. This cooled light gasoline is sent, partially to the sweetening of gasoline, remainder to the storage.

<u>The heavy gasoline</u> going out at the bottom of the column is pumped by the Pump P1112 and after cooling in the Water cooler E1121, is sent to the storage under level control of the C1104. The Reboiler E1120 heated by part of the circulating reflux provides reboiling of the column

5.5 Auxiliary Unit of Crude Distillation Unit

5.5.1 Distillation furnace

A fired heater or furnace is the backbone of any refinery. A fired heater consists of three major components, heating coil, enclosure and combustion equipment. The heating coil consists of tubes connected together in series that carry the charge being heated. Heat is transferred to the material passing through the tubes. Figure 5.4 provides a cross-sectional view of a furnace. The enclosure consists of a firebox. It is a steel structure lined with refractory material that holds the generated heat. Burners generate the heat by burning fuel, either oil or gas.

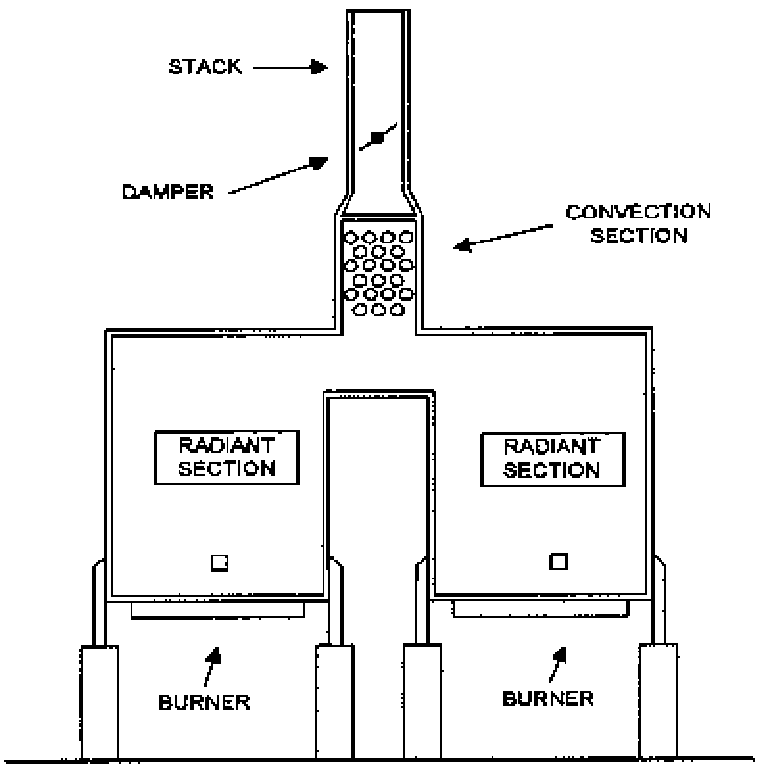

Figure 5.4 Side view of a twin-cell, horizontal-tube cabin heater

The heating coil absorbs the heat mostly by radiant heat transfer and convective heat transfer from flue gases, which are vented to the atmosphere through the stack. Burners are located on the floor or sidewalls. Combustion air is drawn from the atmosphere. For increased heat recovery, an air preheater is installed downstream of the convection section. Instruments are provided to control the fuel-firing rate and flow through the coils to maintain desired operating conditions. The convection section is mounted on top of the radiant section. It consists of bare and extended surface tubes. The stack is mounted on the top of the convection section to provide draught and to dispose of the flue gas safely.

Finned tubes are used in the gas-fired heater and studded tubes are used in the oil-fired heaters. The convection section absorbs 20-30 percent of the total heat duty. A number of tubes are provided in the convection section for superheating steam, use as stripping steam. Hot flue gases inside the firebox and stack are lighter than the cold ambient air. This results in a slightly negative pressure inside the furnace. Combustion air is drawn into the burners and hot gas flows out the stack due to this pressure differential. In natural draft furnace air is drawn into the furnace by a draft created by the stack. The taller the stack, the more draft available.

Furnace of distillation unit in ERL is a twin-cell cabin heater with vertical tubes in the radiant section and horizontal tubes in the convection section. Six numbers of burners are placed on the floor of each cabin and are so spaced as to provide uniform heat distribution and air is drawn in to the furnace by natural draft.

5.5.2 Dehydration and desalting of crude oil

All crude oil contains moisture and salts to varying degrees. When the water content is high, the pressure in the distillation column grows, its productivity decreases and excessive heat is spent to heat and evaporate the water.

Salts, mainly chlorides, have an adverse effect. They are deposited in the tubes of the heat exchangers, furnaces, that require frequent cleaning of the tubes and lower the heat transfer coefficient. Any crude that contains more than 5 kgs of total salts expressed in terms of sodium chloride per thousand barrels may be regarded as salty crude (Rao, 1997). Water being good solvent for these salts, the removal is very much effective in the form of brine. The first stage of water and salt removal is performed at the oil field At refineries, the second stage of dehydration and desalting of crude is performed.

The following general method is versatile for dehydration of crude (Rao,1997):

- Chemical treatment
- Gravity settling
- Centrifugal separation
- Electric desalting.

At the oil field, salt is removed by settling or by adding chemicals or by combination of these. The settling techniques are not effective and are time consuming. Continuous operation is not possible, as large amount of space and equipment is to be isolated for this purpose Similarly centrifugal separation is also not economical due to the huge energy requirements and less quantity handled. All these have given the way to electric desalting.

Electric desalting:

Simultaneous desalting and dehydration is achieved in this unit with a spectacular removal of more than 90% salt in just less than half an hour. Figure 5.5 shows the schematic diagram of Electric Desalter.

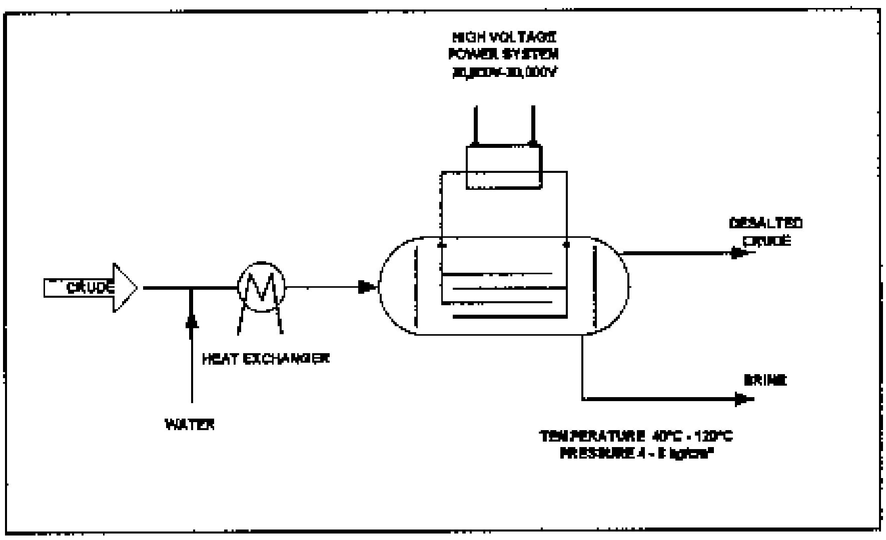

Figure 5.5 Electric desalting methods.

The principle of separation is very simple, under a charged electric field the polar molecules orient. A potential of 20,000 to 30,000 volts is applied between electrodes through which crude is passed. Water present in the form of emulsion also coalesces and agglomerates in to a stream entrapping all the salts in this process, brine collects at the bottom of the desalter, while crude floats above and forms a separate stream Almost all refineries have adopted electric desalting technique due to compactness, efficiency and ease of operation.

Eastern Refinery Ltd. is equipped with electric desalting facilities. Temperature during electric desalting is maintained around 140 to 150°C and a pressure of 14 to 16 bars is maintained Power consumption is very small, frequently of the order 0.01 kWh per barrel. 'Separol' is used as a demulsifying agent and dosed at the rate of 10 ppm.

5.5.3 Flashing of Crude oil

Every kind of crude oil contains gas, liquid and solid as a mixture of hydrocarbons; Crude has a boiling range of –160 °C (methane) to + 1000 °C or more (pitch) (Rao, 1997). When the crude contains a good amount of soluble gases, to avoid heating in the furnace, a preflashing is employed, which is directly fed to the distillation column for fractionation. Preflashing is also useful when crude has to be transported to a long distance. Crude containing less than 6% light ends (gases) usually offers no problem in transportation

Preflashing is conducted at 100 °C under pressure of 3-5 atmospheres to remove these light ends In ERL the feed is preheated through a series of heat exchangers E1101, E1102, E1103, E1104, and E1105 (Figure 5.1). Temperature of the oil rises to 120 - 130 °C, after desalting and passing through Drums B1111 and B1112 the stock is flashed in to the Flash drum B1101. The liquid level in side the Flash drum B1101 is control by valve Vapor evolved by oil flashing in B1101 directly flows to the Distillation column C1101. The flow rate of the vapor from B1101 to C1101 is controlled by the flow control valve.

CHAPTER 6

Energy use

6.1 Introduction

To refine crude petroleum, huge amount of energy is required to operate different process unit ofRefinery. Energy use in refinery is in different forms like, fuel, steam and electricity. Energy cost is the single largest component of variable operating expenses of crude oil processing. To take necessary measure in favor of energy conservation, it is required to find out the actual energy consumption in each discrete operating unit. Fuel is the main source of energy. Crude oil is heated in furnace by combustion of fuel for fractionation in distillation column. In distillation unit steam is used as stripping steam to reduce the partial pressure of hydrocarbons for fractionation at low temperature. Electricity is used in refinery to operate pumps, compressor, control valve etc. of the process unit. Electricity generated in ERL by Steam or Diesel generator. Certain amount of electrical energy is received from Power Development Board (PDB) through national grid.

6.2 Energy Use in Eastern Refinery Ltd.

The overall energy consumption in ERL can be easily found out by calculating the actual fuel consumption. Average fuel consumption per day in ERL is shown in Table 6.1.

Table 6 1 Average fuel consumption per day:

Consumption of Fuel	
NG(MSCF)	1842.67
RG(MT)	31.67
HSFO(MT)	54.33
LDO(MT)	4.02

Average PDB power consumption is 5935 Kwh Power generated by Diesel and Steam Generator of ERL is not included (considered fuel consumption). Average crude oil (ALC) processing in ERL is about 4064.33 M.Ton per day.

Sample Calculation

Energy consumption per day.

Natural gas	= 1842.67 Mscf
	= 1000 x 1842.67 x 0.028317 x 8889.71/10^6
	= 463.86 MM kcal (1 ft^3 = 0.028317 m^3)
Refinery gas	= 31.67MT
	= 31.67 x 1000 x 12211/10^6 = 386.72 MM kcal
HSFO	= 54.33 MT
	= 54.33x 1000 x 10230 /10^6 = 555.80 MM kcal
LDO	= 4.02 MT
	= 4.02 x 1000 x 10340 /10^6 = 41.57 MM kcal
Power (PDB)	= 5935 Kwh
	= 5935 x 859.6 / 10^6 = 5.10 MM kcal

Total Energy consumption = 1453.05 MM kcal per day

= 60.54 MM kcal per hr.

In processing crude oil energy consumption $= \frac{1453.05 MMkcal/day}{4064.33 MT/day}$

= 0.36MM kcal per MT

= 0.05 MM kcal per Barrel

(Taken, 1 MT = 7 Bbl)

Overall energy consumption in processing crude oil in ERL is about 0.05 MMkcal per Barrel

Heating value of fuel use in ERL (LHV)

Natural gas	8889.71 kcal/m^3
Refinery gas	12211 kcal/kg
HSFO	10230 kcal/kg
LDO	10340 kcal/kg

6.3 Energy Use in Crude Distillation Unit

Energy is consumed in following discrete components of Crude Distillation Unit.

(1) Distillation Furnace : Liquid petroleum (heavy) oil - HSFO
Gaseous fuel –Natural gas, Refinery gas

(2) Distillation Column : Stripping Steam

(3) Utility : Electrical Energy

6.3.1 Energy consumption in distillation furnace

Furnace is an essential component of petroleum processing plant. It is used to heat crude oil for fractionation, generation of steam etc. Furnace is the major consumer of energy. It consume about 70-80 percent of the total energy consumption of refinery. To reduce the production cost it is generally operated with residual fuel, HSFO (High sulfur furnace oil) along with certain amount of refinery gas. Saturated steam is used in furnace to atomize the liquid fuel.

Average consumption of energy in distillation furnace with equivalent energy is as follows:

A) High Sulphur Furnace Oil (HSFO) : 2292 kg /hr = 23.49 MMkcal/hr

B) Refinery gas (RG) : 420 kg /hr = 5.13 MMkcal/hr

C) Atomizing Steam (0.3 kg /kg of fuel): 0.05 MMkcal/hr

Total average energy consumption in distillation furnace is around 28.67 MMkcal/hr

6.3.2 Energy consumption in distillation column (Topping Unit)

Steam is a form of energy. In refinery steam is used in different unit for different purposes. A significant amount of steam usage in distillation unit as stripping steam. Steam being inert causes the reduction in partial pressure of hydrocarbons in the distillation column to fractionate at low temperature. Reduction of partial pressure of hydrocarbons may contribute to a temperature drop of 10 – 20 °C in distillation column and side stream strippers. Certain amount of steam is used to atomize liquid fuel used in distillation furnace for proper combustion.

For a maximum feed rate of crude oil, at about 203.22 m^3/hr, the consumption of stripping steam in distillation unit is shown in Table 6.2

Table 6.2 Steam used in Distillation Unit

Steam	kg/hr	kg/ m^3
Stripping steam for K I	280	1.66
Stripping steam for K II	210	1 24
Stripping steam for LGO	146	0.86
Stripping steam for HGO	110	0.65
Stripping steam for column	1140	6.74
Fuel atomizing Steam for furnace (0.3kg/kg of fuel)	687.6	3.38
Total	1647.60	14.53

It is observed that, distillation unit consumes about 1647.60 kg/hr or 14.53 kg steam per cubic meter of crude oil processing.

6.3.3 Energy consumption in utility unit:

Electricity is another form of energy plays a vital role in any petroleum refinery. Equipment like pumps, compressors, agitators etc is operated by electric power. More over proper illumination is essential during night. Uninterrupted power supply in refinery is very essential, break down even for a few seconds, plays adverse effect on process flow system. For uninterrupted supply of power, ERL has its own power generation units with the following capacities:

- Steam Turbine Generator (Unit-1) 3,000 kw
- Steam Turbine Generator (Unit-2) 3,000 kwh
- Diesel Generator 3 nos. 2,500 kwh (Total)

Power consumptions in distillation unit

Consumption of electrical energy by different equipments used in distillation unit is listed in Table 6.3.

Table 6.3 Power consumptions in Distillation Unit

Equipment name	Rated load kw	Volts	Current Amp.	RPM	Available load kw,(P)	Function
PM 1101AB	210	550	20.25	3000	154.33	Crude feed to Distillation unit
PM 1102	210	550	20.25	3000	154.33	Crude feed to Furnace
PM 1103AB	18.5	380	18.75	3000	9.87	Stripping product K I
PM 1104	18 5	380	18.75	3000	9.87	Stripping product K II
PM 1105AB	18.5	380	18.75	3000	9.87	Stripping product LGO
PM 1106	7.5	380	11.025	3000	5.81	Stripping product HGO
PM 1107AB	75	380	99 75	3000	52.52	Atmospheric residue
PM 1108AB	37	380	50.25	3000	26.46	Internal reflux
PM 1109	30	380	42.75	3000	22.51	Top reflux
PM 1110A	75	380	99.75	3000	52.52	Stabilization charge
PM 1110B	45	380	61.5	3000	32.38	Stabilization charge
PM 1111AB	7.5	380	11.025	3000	5.81	Stabilization head reflux
PM 1112	18.5	380	26.25	3000	13.82	HG product pump
PM 1113	22	380	30.75	3000	16.19	LG product pump
PM 1115	0.25	380	0.375	965	0.20	Corrosion inhibitor to column
PM 1121AB	18.5	380	26.25	3000	13.82	Water to B1112
PM 1122AB	30	380	42.75	3000	22.51	Water to B1111
PM 1123	0.25	380	0.375	965	0 20	Separol to crude oil
PM 1124	2.2	380	3 45	1500	1.82	Caustic to crude
EM 1123A	30	380	42.75	1500	22.51	TG aero condenser
EM 1123B	30	380	42.75	1500	22.51	TG aero condenser
EM 1123C	30	380	42.75	1500	22.51	TG aero condenser
EM 1123D	30	380	42.75	1500	22.51	TG aero condenser
EM 1123E	30	380	42.75	1500	22.51	TG aero condenser
EM 1123F	30	380	42.75	1500	22 51	TG aero condenser
EM 1124A	30	380	42.75	1500	22.51	Redistillation aero condenser
EM 1124B	30	380	42 75	1500	22.51	Redistillation aero condenser
	1084.2				784.91	

Available load for the equipment are obtained by the following formula:

$$P = \sqrt{3}\, V\, I \cos\theta \text{ watt}$$

where, V = Volt

I = Current

$Cos\ \theta$ = Power factor

= 0.8 (Available in ERL)

Total available load for pumps & lightening used in distillation Unit is 784.91 kw

Average plant running time of the plant per year is 8000 hr.

If the equipments listed in Table 6.3 is operated simultaneously through out the year then,

power consumption or available load in use per year	= 784 91 x 8000
	= 62, 79,280kwh
	= 62, 79,280x 859.6 kcal
	= 5397.67 MM kcal
	(1 kwh = 859.6 kcal)

Cost of 1 unit of power (1 kwh) = Tk.3 30

Total cost of power /energy used per year =62, 79,280 kwh x Tk. 3.30/kwh

= Tk. 2, 07, 21,624.00

It shows that huge amount of electrical energy is required in processing crude oil in ERL. Steps may be taken to cut down certain percentage of power consumption

6.4 Performance of Distillation Furnace

A detailed study has been made on distillation furnace Furnace consumes about 70-80 percent of the total energy consumption of refinery. Therefore, improvements in efficiency of furnaces can make a major contribution to reduce energy consumption. Distillation furnace of ERL is a natural draft furnace provided with dual fuel burning facilities (gaseous and liquid fuel). The design characteristics of distillation furnace of ERL are given below.

- Total heat absorbed : 25.44 MM Kcal/hr
- Heat absorption efficiency : 60%
- Excess air : 40%
- Heat transfer co-efficient (overall) : 58500 kcal/hr m^2 ^oC
- Radiation section height : 12.75 m
- Chimney height : 26.60 m
- Convection section height : 3.40 m
- Radiation section diameter : 4.30 m
- Number of tube in radiation zone : 52
- Radiation section tube thickness : 6 mm
- Radiation section tube material : High carbon alloy
- Radiation section tube length : 11.65 m
- Radiation section tube diameter : 125 mm

- Nos. of tube in convection section : 112
- Tube thickness in convection section : 6.03 mm
- Tube material in convection section · A 100 - Group
- Tube diameter in convection section : 125 mm
- Flow distribution . 4 nos. of pass
- Residence time 6 min.

To study the performance of distillation furnace, a sample calculation was made with the available data. Average fuel characteristics and operation conditions of distillation furnace are as follows:

(1)Fuel **High Sulfur Furnace Oil (HSFO)**

Consumption		: 2.292 MT/hr
Temperature of fuel		: 112 °C
Lower Heating value		· 10250 kcal/kg
Chemical composition		
	Carbon	: 85 %
	Hydrogen	· 13 %
	Sulfur	: 2 %
Atomizing steam		: 0.3 kg/kg of fuel
Temperature		: 180 °C (356 °F)
Pressure		: 6 bar

(2)Fuel: Refinery gas

Consumption	· 420 kg/hr
Lower heating value	: 12211 kcal/kg

Stack temperature	: 471 °C
Ambient temperature	: 30 °C
Datum temperature	: 15.6 °C
Average value of excess oxygen in dry flue gas	: 9.5 %
Feed rate (ALC)	: 4305 MT/day

Sample calculation

Fuel: High Sulfur Furnace Oil (HSFO)

Combustion reaction -

$$C + O_2 = CO_2$$

$$H_2 + \frac{1}{2} O_2 = H_2O$$

$$S + O_2 = SO_2$$

Basis 1 kg of fuel (HSFO

Moles of combustion products:

Carbon, C = 0.85/12 = 0.0708

Hydrogen, H = 0.13/2 = 0.065

Sulfur, S = 0.02/32 = 0.000625

Oxygen required to burn 1kg fuel (no excess):

		Mole
For carbon	: 0.0708 x 1	= 0 0708
For hydrogen	: 0 065 x ½	= 0 0325
For sulfur	: 0.000625 x 1	= 0.000625
Total oxygen required		= 0.103925
Total air required		= 100/21x0.103925
		= 0.4949 mole

Let supply of air be 'Y' mole, then flue gas contains:

Composition	Mole
CO_2	: 0.0708
H_2O	: 0.065
SO_2	: 0 000625
N_2	: 0.79Y
O_2	: (0.21Y - 0.103925)

Total flue gas : (Y + 0 0325) mole

Dry flue gas (excluding H_2O) = (Y + 0.0325 - 0 065) = Y - 0.0325 mole

Dry flue gas contains 9.5 % excess oxygen.

Then, excess O_2 in dry flue gas, $9.5 = \frac{0.21Y - 0.103925}{Y - 0.0325} \times 100$

Total air supply, Y = 0.8768 mole

Then composition of flue gas

Composition	Mole
CO_2	0.0708
H_2O	: 0.065
SO_2	. 0.000625
N_2	: 0.6927
O_2	: 0.0802
Total	0 9093 mole

Material balance of water vapor:

Water in	Mole
Atomizing steam	(2292 x 03 / 18) = 38 2
From hydrogen in fuel	= 0 065
From air (negligible)	= 0.00
Total	= 38.265

Fuel: Refinery gas (RG)

Refinery gas is a mixture of various refinery gaseous products as shown in Table 6.4.

Table 6.4 Typical refinery fuel gas mixture (Talavera, 2000)

	Mol. Wt mole	Mole %	Equivalent carbon	Equivalent Hydrogen
CH_4	16.04	60	0.6 x 1 = 0.6	0.6 x 4 = 2.4
C_2H_6	30 07	17	0.17 x 2 = 0 34	0.17 x 6 = 1.02
H_2	2.02	20	0	0.2 x 2 = 0.4
C_3H_8	44.0	2	0.02 x 3 = 0 06	0.02 x 8 = 0 16
C_4H_{10}	58.1	1	0.01 x 4 = 0.04	0.01 x 10 = 0.1
Total		100	1 04	4.08

Molecular formula of RG can be represented by: $C_{1.04}H_{4.08}$

Basis 1 kg of Refinery gas

Percentage composition of refinery gas.

$$\text{Carbon} = \frac{12 \times 1.04}{12 \times 1.04 + 1 \times 4.08} \times 100 = 75.36\%$$

$$\text{Hydrogen} = \frac{1 \times 4.08}{12 \times 1.04 + 1 \times 4.08} \times 100 = 24.64\%$$

Molecular weight of

Carbon	12
Hydrogen	1

Moles of combustion products

Carbon, C = 0.7536/12 = 0.0628

Hydrogen, H_2 = 0.2464/2 = 0.1232

Oxygen required per kg fuel gas burning (no excess):

Composition	Mole
For carbon	: 0.0628 x 1 = 0.0628
For hydrogen	: 0.1232 x ½ = 0.0616

Total oxygen required = 0.1244 mole

Air required = 0.5924 mole

Let supply of air (with excess O_2) be 'X' mole

Then flue gas contains

Composition	Mole
CO_2	: 0.0628
H_2O	: 0.1232
N_2	: $0.79X$
O_2	: (0.21X - 0.1244)

Total flue gas = (X + 0.0616)

Dry flue gas (excluding H_2O) = (X - 0.0616) mole

Average value of excess oxygen in dry flue gas is 9.5 %

We have excess O_2 in dry flue gas, $9.5 = \dfrac{0.21X - 0.1244}{X - 0.0616} \times 100$

Supply of air, X = 1.0308 mole

Then, flue gas composition;

Composition	Mole
CO_2	. 0.0628
H_2O	: 0.1232
N_2	. 0.8143
O_2	. 0.0921
Total	1 0924

Heat entering the system:

Fuel : HSFO	Consumption 2292 kg/hr	MM kcal/hr
Net heating value of fuel	2292x10250/ 10^6	23.4930
Heat in air	2292x0 8768x29x0.241x(30 - 15.5)/10^6	0.2037
Heat in steam (Nelson ,1958)	2292x0.3x2.2/18x(2800 - 260)x.252/10^6	0.0538
Sensible heat content of fuel oil	2292x0.49x(112 - 15.5)/10^6	0.1084
Total		23.8588
Fuel :RG	Consumption : 420 kg/hr	
Net heating value of fuel	420x12211/1E-6	5.1286
Heat in air	420x1.0308x29x0.241(30 - 15.5)/E-6	0.0439
Total		5.1725
Total heat in the system		**29.0313**

<u>Heat leaving the system</u>

Stack loss:

Fuel . HSFO (2292 kg/hr)		MMkcal/hr
Flue gas	0.9093x2292x30x.27x(471 - 15.5)/10^6	7.6894
Water vapor/steam (Nelson,1958)	2292x.3x2.2/18(7500 - 260)x0.252/10^6	0.1533
Total		7.8428
Fuel : RG (420kg/hr)		
Flue gas	1.0924x420x30x.27x(471 - 15.5)/10^6	1.6928
Total		1.6928
Total heat leaving		**9.5356**

Molecular Weight of air = 29

Molecular Weight of flue gas = 30

Molecular Weight of steam = 18

Sp. heat of air = 0.241 kcal/kg°C

Sp. heat of flue gas = 0 27 kcal/kg°C

Sp. heat of fuel oil = 0.49 kcal/kg°C

Sp. heat of feed stock = 0.753 kcal/kg°C

Useful heat		MMkcal/hr
Heat gain by crude oil	$179.375 \times 1000 \times 0.753 \times (372-233)/10^6$	18 7741
Heat gain by water to steam (Nelson ,1958)	$0.3 \times 2292 \times 2.2/18 \times (5500-480) \times 252/10^6$	0.1063
Total useful heat		**18.8804**

Summery of heat balance		%
Into system	**29.0313**	
Out of system		
Useful heat	18.8804	65.03
Stack loss	9.5356	32.85
Unaccounted for and loss	0.6153	2.12
	29.0313	100.00

Thermal efficiency of furnace:

Thermal efficiency of a furnace is defined as the heat absorbed in the furnace divided by the heat input.

$$\text{Thermal efficiency, } \eta = \frac{\text{Heat absorbed}}{\text{Heat input}} \times 100$$

$$\eta = \frac{\text{Heat Input - Heat loss}}{\text{Heat input}} \times 100$$

Thermal efficiency, $\eta = \frac{\text{Useful heat}}{\text{Heat input}} \times 100$

$$\eta = \frac{18.8804}{29.0313} \times 100 = 65.03$$

Heater efficiency is lower due to certain heat losses through heater wall. This loss is usually taken as 2% of total heat input

6.4.1 Study of performance of distillation furnace

Heater efficiency operating with stoichiometric amount of air and with 2% heat loss is given by a heat balance around the heater and expressed as_

$$\eta = 98 - \frac{\text{Heat out}}{\text{Heat input}} \times 100$$

Heat of flue gas:

For 1 kg HSFO fuel burning, the composition of flue gas is:

(Required amount of O_2 = 0.1039 mole)

Flue gas composition	Mole
CO_2	: 0.0708
H_2O	: 0.065
SO_2	: 0.000625
N_2	: 0.3909

For 1 kg refinery gas burning, composition of flue gas:

(Required O_2 = 0.1244 mole)

Flue gas composition	Mole
CO_2	: 0.0628
H_2O	: 0 1232
N_2	: 0.4680 mole

Since, HSFO and RG are burnt together, so flue gas contains the mixture of both products mole of flue gas per one kg HSFO and one kg RG fuel burning with stoichiometric amount of air.

Product	HSFO Mole/kg	RG Mole/.kg	Total Mole/2kg	Total mole/1kg
CO_2	0.0708	0.0628	0 1336	0 0668
H_2O	0.0650	0 1232	0.1882	0.0941
SO_2	0.000625	---	0.000625	0.00031
N_2	0.3909	0.4680	0.8589	0.4295
Total				0.591

For 1 kg. fuel (mixture) burnt with stoichiometric amount oxygen gives

Total flue gas =0.591 mole

Dry flue gas (except H_2O) = 0.497

		Mole
Stoichiometric amount oxygen for 1 kg, HSFO burning	=	0.1039
Stoichiometric amount oxygen for 1 kg, RG burning	=	0.1244
Stoichiometric amount oxygen for 2 kg fuel (HSFO + RG) burning	=	0 2283
Stoichiometric amount oxygen for 1 kg fuel (HSFO + RG) burning	=	0 1142

Stoichiometric amount air for 1 kg fuel (mixture) burning = (0.1142 x 100/21)

= 0.5436 mole

In order to complete combustion with excess O_2, let supply of air be "Y" mole

Then,

Amount of O_2 in air = 0.21Y

Amount of N_2 in air = 0.79Y

1 kg fuel mixture (HSFO + RG) burning, flue gas contents

Product	Mole /kg
CO_2	0 0668
H_2O	0.0941
SO_2	0.00031
N_2	0.79Y
O_2	(0.21Y – 0.1142) (excess)

Total amount of dry flue gas per kg of fuel burning =Y - 0.047

Total amount of flue gas (including H_2O) = Y + 0.047

Relation between excess O_2 (%) in dry flue gas and excess air (%) in supply

Excess O_2 in dry flue gas = O_2 in supply air - Required O_2

= (0.21Y - 0.1142)

$$\text{Excess Oxygen, } \%O_2 = \frac{\text{Excess } O_2 \text{in dry flue gas}}{\text{Total amount of dry flue gas}} \times 100 \qquad (6.1)$$

$$= \frac{0.212Y - 0.1142}{Y - 0.047} \times 100 \qquad (6.2)$$

$$Y = \frac{11.42 - 0.047\%O_2}{21 - \%O_2} \qquad (6.3)$$

Again,

$$\%\text{ excess air} = \frac{\text{Supply air - Required air}}{\text{Required air}} \times 100 \qquad (6.4)$$

$$= \frac{Y - 0.5426}{0.5426} \times 100 \qquad (6.5)$$

Putting value of Y in Equation, 6 5

$$\%\text{ excess air,} = 183.96 \times \left(\frac{11.42 - 0.047\%O_2}{21 - \%O_2} \right) \qquad (6.6)$$

$$= \frac{91.351\%O_2}{21 - \%O_2} \qquad (6.7)$$

Equation 6.7 gives the relation between excess Oxygen in dry flue gas and excess air supply for combustion of fuel in furnace.

Total heat into the system per kg. fuel burning:

kcal/kg

For 1 kg HSFO burning

Heat of combustion = 10250

Heat in steam [0.3 x 2.2/18 x (2800 - 260) x 0.252] = 23.4696

Sensible heat in fuel oil [0.49 x (112 - 15.5] = 47.285

For 1kg RG burning.

Heat of combustion = 12211

Total heat,

for 2kg. fuel (HSFO+ RG) burning: = 22531.7546

for 1kg. fuel (HSFO+ RG) burning: = 11265.8773

Total heat of air per kg. fuel (HSFO&RG) = Y x 29 x 0.24x (t_{air} - 15.5)

= 6.96 x Y x (t_{air} - 15.5)

Total heat in to the system per kg fuel = 11265.877 + 6.96 x Y (t_{air} - 5.5)

Total heat out of system per kg. fuel burning

Heat of flue gas = (Y + 0.047) x 30 x 0.27($t_{flue\ gas}$ - 15.5)

Heat out with steam [0.3 x 2.2/18(7500 - 260) x 0.252] = 66.8976

Total heat out of the system = 66.8976 + 8.1(Y + 0.047) x ($t_{flue\ gas}$-15.5)

Therefore, efficiency of furnace can be written as:

$$\eta = 98 - \frac{66.8976 + 8.1(Y + 0.047) \times (t_{flue\ gas} - 15.5)}{11265.877 + 6.96Y(t_{air} - 15.5)} \times 100 \quad (6.8)$$

Considering air temperature, t_{air} = 30 °C, equation (6.9) can be written as

$$\eta = 98 - \frac{66.8976 + 8.1(Y + 0.047) \times (t_{flue\ gas} - 15.5)}{11265.877 + 100.92Y} \times 100 \quad (6.9)$$

Substituting the value of Y (Equation 6.3) in Equation (6.9) one obtains;

$$\eta = 98 - \frac{66.8976 + 8.1\left(\frac{11.41 - 0.047\,\%O_2}{21 - \%O_2} + 0.047\right) \times (t_{flue\ gas} - 15.5)}{11265.877 + 100.92\left(\frac{11.42 - 0.047\,\%O_2}{21 - \%O_2}\right)} \times 100 \quad (6.10)$$

On simplifying above Equation (6.10), efficiency of the furnace obtained by:

$$\eta = 98 - \frac{0.66 + 0.08\left(\frac{11.42 - 0.047\,\%O_2}{21 - \%O_2} + 0.047\right) \times (t_{flue\ gas} - 15.5)}{111.63 + \frac{11.42 - 0.047\,\%O_2}{21 - O_2}} \times 100 \quad (6.11)$$

Above equation (6.11) shows the relation between thermal efficiency of the furnace at different flue gas temperature and excess oxygen present in flue gas Variation of efficiency at different stack temperature and excess oxygen in flue gas is shown in Table 6 5

Table 6.5 Efficiency of furnace at different stack temperature

		Efficiency of furnace at different stack temperature °C						
% O_2 in dry flue gas	% Air in supply	471	450	460	470	480	490	500
		%	%	%	%	%	%	%
0.00	0.00	78.22	79.10	78.68	78.26	77.84	77 42	77.00
0.50	2.23	77.83	78.73	78.30	77.87	77.44	77.01	76.58
1 00	4.57	77.42	78.34	77.90	77.46	77.02	76.58	76.14
1.50	7.03	76.98	77.93	77.48	77.03	76.58	76.13	75 68
2.00	9.62	76 53	77 49	77 03	76.58	76.12	75.66	75.20
2.50	12.34	76.05	77.04	76 57	76 10	75 63	75.16	74.69
3.00	15.23	75.55	76.55	76.07	75.59	75.11	74 63	74.15
3.50	18.27	75.01	76.04	75.55	75.06	74.57	74 08	73.59
4.00	21.49	74.45	75.50	75.00	74.50	73.99	73.49	72.98
4.50	24 91	73.85	74.93	74.42	73.90	73.38	72.86	72.35
5.00	28.55	73.21	74.33	73.79	73.26	72.73	72.20	71.67
5.50	32.41	72.53	73.68	73.13	72.59	72 04	71.49	70.95
6.00	36.54	71.81	72.99	72.43	71.87	71.30	70.74	70.18
6.50	40.95	71.04	72.25	71.67	71.10	70.52	69.94	69.36
7 00	45.68	70.21	71.47	70.87	70.27	69.67	69.08	68.48
7.50	50.75	69.32	70.62	70.00	69.39	68 77	68 15	67.54
8.00	56.22	68.37	69 71	69.07	68.43	67.79	67.16	66.52
8.50	62.12	67.34	68.72	68 06	67 40	66.74	66.08	65.42
9.00	68.51	66 22	67.66	66.97	66.29	65.60	64.92	64 23
9.50	75.46	65.01	66.50	65.79	65.08	64.37	63.66	62.94
10.00	83 05	63.69	65.24	64.50	63.76	63.02	62.28	61.54

Analysis of Distillation furnace of ERL shows that, the efficiency of the furnace mainly depends on the air supply and temperature of the flue gas. Excess air supply can be determined by measuring the presence of oxygen in flue gas by Orsat apparatus.

CHAPTER 7

Energy Conservation

7.1 Introduction

Crude Distillation Unit of ERL was designed in the mid-60. Loss control at its early stage was not as stringent as it is now in a refinery of 2002. There are several approaches to effect energy conservation. Stress should be given to improve efficiency not to reduce fuel consumption. The main objective of energy conservation is to save energy during processing crude oil and to recover process heat by the proper arrangements

Energy conservation is broadly based on:

- Energy saving
- Energy recovery

7.2 Energy Saving

Energy saving is one of main objective of energy conservation. Energy can be saved at every step of energy use. In crude distillation energy can be saved in large scale by the following arrangements.

- Energy saving by flashing crude oil,
- Energy saving in distillation furnace.

7.2.1 Energy saving by flashing crude oil

Every kind of crude oil contains soluble or dissolve gas, which on increasing temperature or under low pressure evaporates. A reasonable amount of heat energy can be saved by flashing crude oil prior feeding to furnace Higher temperature in Desalter Unit/Flashing drum results more evaporation of soluble gas in crude oil that is directly feed to Distillation Column (C1101), helps in-lowering furnace heat duty, thereby saving fuel.

Flashing rate depends on:

- Type of crude oil processed
- Higher inlet temperature of flush drum
- Lower flash drum pressure

A study was made by ERL to increase the flashing rate of crude oil by improving desalter inlet temperature. Accordingly a modification has been made in 1995 by rearranging the existing heat exchanger in crude preheat train. ERL crude preheat train consists of 10 numbers of heat exchangers with six grades of hot product streams at different temperature The exchangers are marked as E1101, E1102, E1103, E1104, E1105, E1106, E1107, E1107, E1108, E1109AB (E1109A and E1109B). The Hot product streams are Light kerosene (K I), Heavy kerosene (K II), Light gas oil (LGO), Heavy gas oil (HGO), Circulating Reflux (Cir. Reflux) and Atmospheric residue.

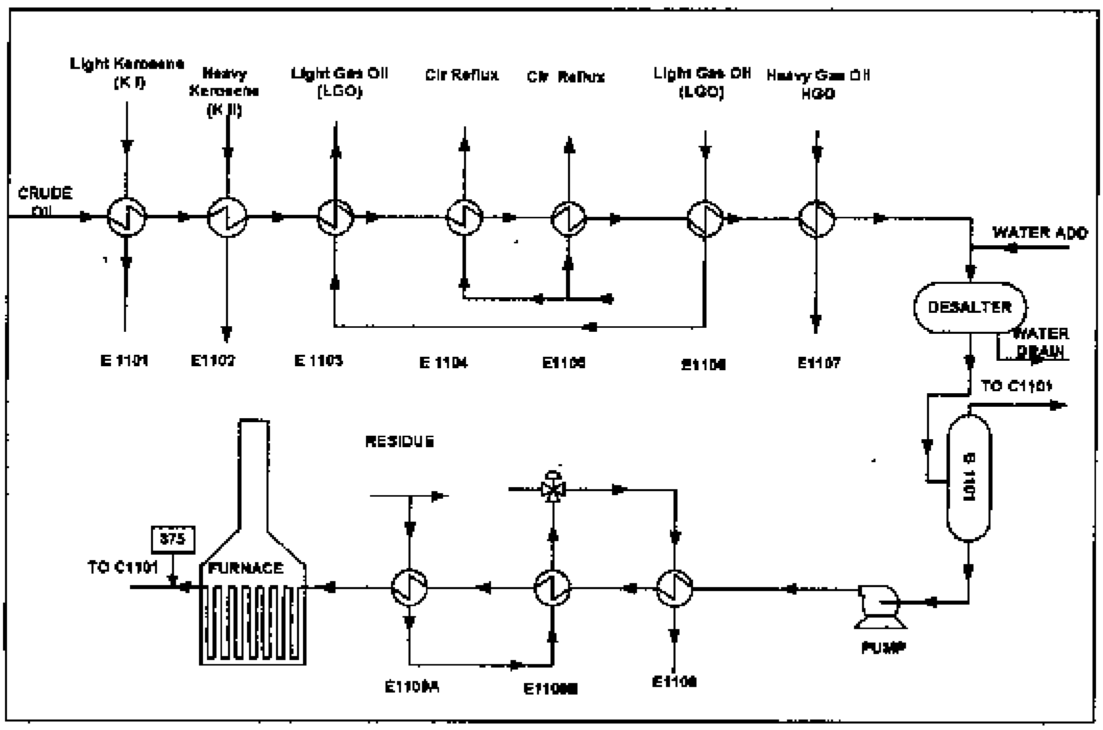

Figure 7.1 Crude preheat train (Before modification)

Before modification of preheat train the exchangers were placed in accordance with their number. The modification was made by placing Desalter after E1105 instead of

E1107 and crude from flashing drum feed to E1106 through E1108. Figure 7.1 and 7.2 shows the arrangement of exchangers in crude preheat train before and after modification respectively

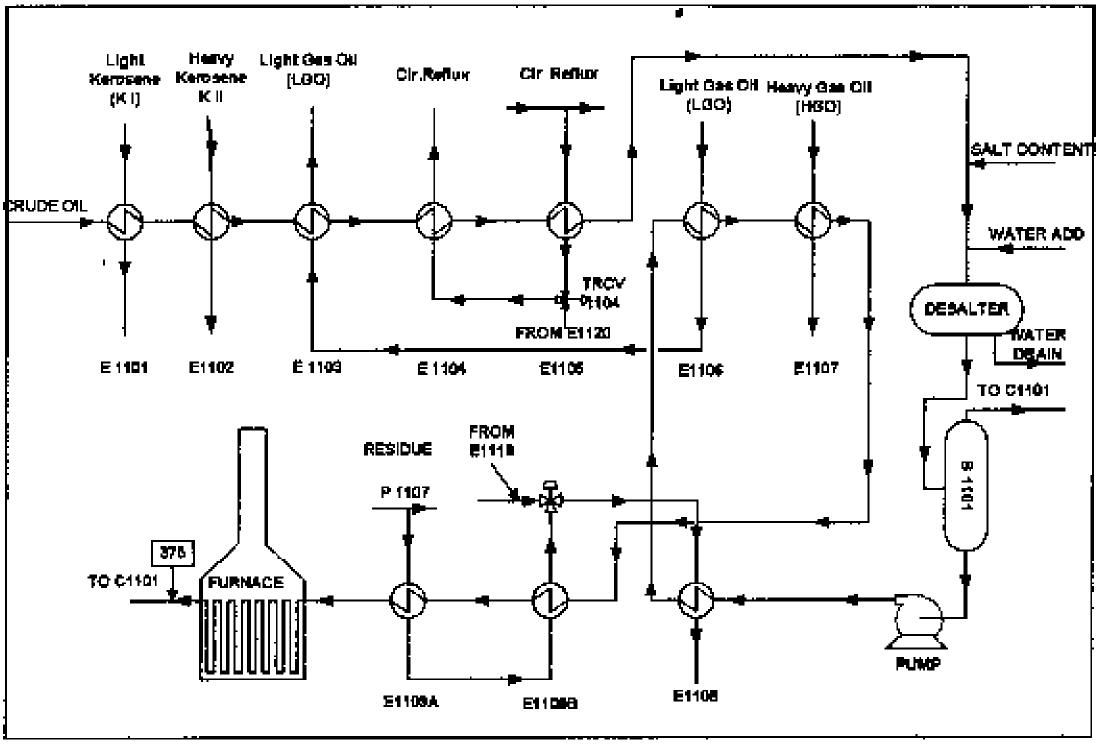

Figure7.2 Crude preheat train (After modification)

Present modification, made to ERL crude preheat train, shows marked improvement in desalting operation, more flashing from flush drum and rise in furnace inlet temperature.

Table 7.1 and Table 7.2 give the performance of crude preheat train before and after modification respectively (Study Report, ERL 1990).

Table 7.1 Performance before crude preheat train modification

Crude type	No. of records	Desalter inlet temperature	Furnace inlet temperature	Furnace outlet temperature	Del T °C
Murban	61	122	202	365	163
ALC	77	120	211	365	154

Table 7.2 Performance after crude preheat train modification

Crude type	No. of records	Desalter inlet temperature	Furnace inlet temperature	Furnace outlet temperature	Del T °C
Murban	64	154	203	371	168
ALC	58	148	215	369	154

7.2.1.1 Performance of crude oil flashing:

Sample calculation has been made to study the performance of crude oil flashing due to modification of crude preheat train. Modification of crude preheat train has increased the desalter/ flash drum inlet temperature remarkably and results more generation of crude oil vapor. Energy saving by feeding crude oil vapors directly to the distillation column, by-passing the furnace to avoid heating in the furnace for both ALC and Murban crude has been calculated below in details.

Flow rate of vapor from Flush drum (B1101) to Distillation column (C1101) through 4 inch diameter pipe with Butterfly control valve may be calculated by the follow empirical formula (Technip, 1967):

$$\text{Valve flow Coefficient, } C_V = \frac{W}{3.22\sqrt{(\Delta P(P_1 - P_2)G_f)}} \quad (7.1)$$

where,

P_1 = Pressure (higher side) in Psi

P_2 = Pressure (lower side) in Psi

W = Flow rate in lb/hr

G_f = constant (orifice)

For the following conditions (design basis)

C_v = 190

P_1 =Pressure in Flush drum =3.6 bar = 3.6 x 14.5 + 14.7 = 66.9Psia

P_2 = Pressure in Distillation column = 2.2 bar = 2.2x14 5+14.7 = 46.6Psia

G_f = 2.00

Putting the above values in equation (7.1)

Gas flow through the pipe, $W = 190 \times 3.22 \times \sqrt{(66.9 - 46.3)(66.9 - 46.3) \times 2}$

$$W = 41530.81 \text{ lb/hr}$$

$$= 18878 \text{ kg/hr}$$

For 100% valve opening vapor flow rate is 18878 kg/hr

Again, flow rate through the valve, can be obtained by the following relation (Technip, 1967):

$$Q = KY^2 \quad (7.2)$$

where, K = constant

Y = % valve opening .

For Y_1% valve opening flow rate, $Q_1 = KY_1^2$ (7.3)

From. Equation (5 2) and (5.3),

$$Q_1/Q = (Y_1/Y)^2$$

$$Q_1 = Q \times (Y_1/Y)^2 \quad (7.4)$$

Here,

Q = 18878 kg/hr

Y = 100% valve opening

Putting different values of Y_1, in Equation (7.4), gives:

Valve opening (%)	Flow rate (kg/hr)
15	425
20	755
30	1699
50	4720

Flow rate of crude oil vapor from Flash drum (B1102) to Distillation column (C1101) is a complete balanced operation, flashing rate depends on evaporation rate of crude oil, higher inlet temperature and lower flash drum pressure will increase the evaporation rate. Excess flow of vapor to column may cool down the bottom temperature of distillation column, which may create an adverse effect on processing, besides this excess flow rate may entrain certain portion of liquid crude to column which is not desirable To obtain the maximum performance, it requires close control and proper monitoring of

- Vapor generation by crude oil in flash drum
- Flow rate of vapor from flash drum to distillation column by regulating valve
- Temperature of bottom level of distillation column
- Liquid level in flash drum.

A) Sample calculation: (Murban crude)

(Before modification of preheat train)

Specification of crude

Crude oil	: Murban crude
Sp. gravity	: 0.828
API gravity	: 39.4
Characterization factor (K)	: 11.85

Furnace inlet temperature	=202 °C
Heat content	= 215 Btu/lb
(After "K" factor correction)	=212 Btu/lb
Furnace outlet temperature	= 365 °C
Heat content	= 430 Btu/lb
(After "K" factor correction)	= 424
Heat duty of furnace	= (424-212) =212 Btu/lb
Desalter inlet temperature	= 122 °C
Heat content (liquid)	= 125 Btu/lb
(After "K" factor correction)	= 123 Btu/lb
Heat content (vapor)	= 250 Btu/lb
(After "K" factor correction)	= 235 Btu/lb
Heat carryover from B1101 to C1101	= Heat of vaporization
	= (235 - 123) =112 Btu/lb

Net heat saving for flashing in B1101 = Heat duty of furnace - Heat of vaporization

=(212 – 112) = 100 Btu/lb

For 1 MT/hr Crude oil Flashing:

Heat saving = 1000 kg/hr x 2.2 Ib/kg x 100 Btu/Ib x 0.252 kcal/Btu x 24 hr/day /0.8/10^6 = 1.66 MMkcal/day

Under balance condition of temperature, pressure and evaporation rate maximum valve opening for vapor flow from Flush drum (B1101) to Distillation column is about 21%,from equation (7.4).

Quantity flashed = 0.84 MT/hr

Therefore, saving = (0.84 x 1.66) = 1.39 MMkcal/day

B) Sample calculation (ALC)

(Before modification of preheat train)

Specification of crude oil:

Crude	: ALC
Sp. gravity	: 0.8524
API gravity	: 34.5
Characterization factor (K)	: 11.75

Furnace inlet Temperature	= 211 °C (413.6 °F)
Heat content (Liquid)	= 223 Btu/lb
After "K" factor correction	= 217.43 Btu/lb
Furnace outlet temperature	= 365 °C (689 °F)
Heat content	= 425 Btu/lb
After "K" factor correction	= 414.38 Btu/lb

Heat duty of furnace = (414.38 – 217.43) x 0.98 =197 Btu/lb

Desalter inlet temperature	= 120 °C (248 °F)
Heat content (liquid)	= 123 Btu/lb
After "K" factor correction	= 120 Btu/lb
Heat content (vapor)	= 245 Btu/lb
After "K" factor correction	= 228 Btu/lb
Heat carrying from B1101 to C1101	= Heat of vaporization
	= (228-120) = 108 Btu/lb

Heat saving for flashing in B1101= Heat duty of furnace - Heat of vaporization

= 197 - 108 = 89 Btu/lb

For 1 MT/hr Crude oil Flashing:

Net energy saving =1000 x 2.2 x 89 x 0.252 /0.8x24/10^6

= 1.48 MM kcal/day

Under above condition of temperature and pressure, the maximum allowable valve opening from B1101 is 15 %, than,

Quantity flashed = 0.424 MT/hr (Equation.5.4)

Therefore, total saving = 0.63 MMkcal/day

C) Sample calculation: (Murban Crud

(After Modification of preheat train)

Specification of crude oil:

Crude oil : Murban crude

Sp. gravity : 0.828

API gravity : 39.4

Characterization factor (K) : 11.85

Furnace inlet temperature =203 °C

Heat content = 225 Btu/lb

(After "K" factor correction) = 222 Btu/lb

Furnace out let temperature = 371 °C

Heat content = 435 Btu/lb

(After "K" factor correction) = 424 Btu/lb

Heat duty of furnace =202Btu/lb

Desalter inlet temperature = 154 °C

Heat content = 157 Btu/lb

(After "K" factor correction) = 155 Btu/lb

Heat content (vapor) = 278 Btu/lb

(After "K" factor correction) = 263 Btu/lb

Heat carrying from Flush drum (B1101) to Distillation column (C1101)

= Heat of vaporization

= (263 - 155) Btu/lb = 108 Btu/lb

Net heat saving for flashing in B1101 = Heat duty of furnace - Heat of vaporization

= (202 - 108)

= 94 Btu/lb

For 1 MT/hr Flashing:

Heat saving = $1000 \times 2.2 \times 94 \times 0.252 \times 24/0.8/10^6$

= 1.56 MMkcal/day

Maximum allowable valve opening from B1101 is 51 % and from equation (7.4)

Quantity flashed = 4.91 MT/hr (App)

Therefore, energy saving = (4.91 x 1.56) MMkcal/day

= 7.66 MMkcal/day

D) Sample calculation; (ALC)

(After modification of preheat train)

This modification causes higher furnace and desalter inlet temperature.

Crude	: ALC
Sp. gravity	. 0.8524
API gravity	: 34.5
Characterization factor (K)	: 11.75

Furnace inlet temperature	= 215 °C
Heat content	= 225 Btu/lb
(After "K" factor correction)	= 221 Btu/lb
Furnace outlet temperature	= 369 °C
Heat content	= 430 Btu/lb
After "K" factor correction	= 422 Btu/lb
Heat duty of furnace	= (422 - 221) Btu/lb
	= 201 Btu/lb
Desalter inlet temperature	= 148°c
Heat content (liquid)	= 151 Btu/lb

(After "K" factor correction) = 148 Btu/lb

Heat content (vapor) = 260 Btu/lb

After K" factor correction = 244 Btu/lb

Heat carryover from Flush Drum (B1101) to distillation column C1101

= Heat of vaporization

= (244 - 148) Btu/lb

= 96 Btu/lb

Net heat saving for flashing in B1101, = (Heat duty of Furnace - Heat of vaporization)

= (201 - 96) Btu/lb

=105 Btu/lb.

For 1 MT/hr Crude oil Flashing:

Heat saving = $1000 \times 2.2 \times 105 \times 24/.8 \times 0.252/10^6$

= 1.75 MMkcal/day

After modification of crude preheat train the inlet temperature of desalter increased and thereby increased evaporation rate, consequently the valve can be opened up to 28%.

From Equation (7 4), for 28 % valve opening,

Quantity flashed = 1.48 MT/hr (APP)

There fore, total saving = (1.48 x 1.75) MMkcal/day

= 2.59 MMkcal/day

Above sample calculation shows marked improvement in flashing operation after heat train modification. Overall desalting operation is quite satisfactory because of higher crude oil inlet temperature to the desalter unit A comparison statement has been made to observe the extent of energy saving due to modification of crude preheat train with different grades of crude oil (Murban and ALC) processed in ERL (Table 7.3)

Table 7.3 Comparison Statement:

Description	Before Modification of heat train		After Modification of heat train	
Crude type	ALC	Murban	ALC	Murban
Sp. Gravity	0.8524	0.824	0 8524	0.828
API Gravity	34.5	39.4	34.5	39.4
Furnace inlet temperature °C	211	202	215	203
Furnace outlet temperature °C	365	3365	369	371
Furnace Duty Btu / lb	197	202	201	202
Desalter inlet temperature °C	120	122	148	154
Heat saving per M.Ton MMkcal/day	1.48	1.66	1.75	1.56
Optimum valve opening %	15	21	28	51
Net heat energy saving per M.Ton MMkcal/ day	0.63	1.39	2 59	7.66

From the above statement, it is observed that increase in desalter inlet temperature improves the evaporation and thereby increases valve opening. To increase more evaporation rate it may require to increase the size of existing flash drum and to install a separating column to resist flow of liquid to the Distillation column (C1101). A proper design is essential for flashing more crude oil under balanced condition of temperature, pressure and flow rate.

7.2.2 Energy saving in distillation furnace

In refinery, maximum energy is consumed by furnace. Energy may be saved by reducing fuel supply or improving furnace efficiency by regulating the furnace in cost effective manners. Without reduction of fuel consumption, the efficiency of furnace can be improved greatly by –

- Reducing excess air supply to the furnace
- Lowering stack temperature
- Eliminating air leakage into the firebox

7.2.2.1 Energy saving by reducing excess air supply to the furnace

Burning or combustion is an exothermic reaction resulting from rapid combustion of oxygen with fuel. Most fuels contain hydrocarbons and some sulfur. Since perfect mixing of fuel and air is not possible, excess air is needed to ensure complete combustion of fuel Excess air is expressed as a percentage of theoretical quantity of air required for perfect combustion. For each part of oxygen, four parts of nitrogen enter the combustion process and leave without reacting. They absorb some of the heat generated and carry it to the stack. It is necessary to minimize excess air to avoid excessive heat loss. It is also undesirable to operate with less than stoichiometric combustion air, as it will lead to a smoking stack and incomplete combustion. The relations between excess oxygen in flue gas verses excess air supply can be represented by the following Equation (6.7).

$$\% \text{ excess air} = \frac{91.351 \ \% O_2}{21 - \% O_2}$$

Graphical representation of excess oxygen in dry flue gas verses excess air supply is shown in Figure 7.3

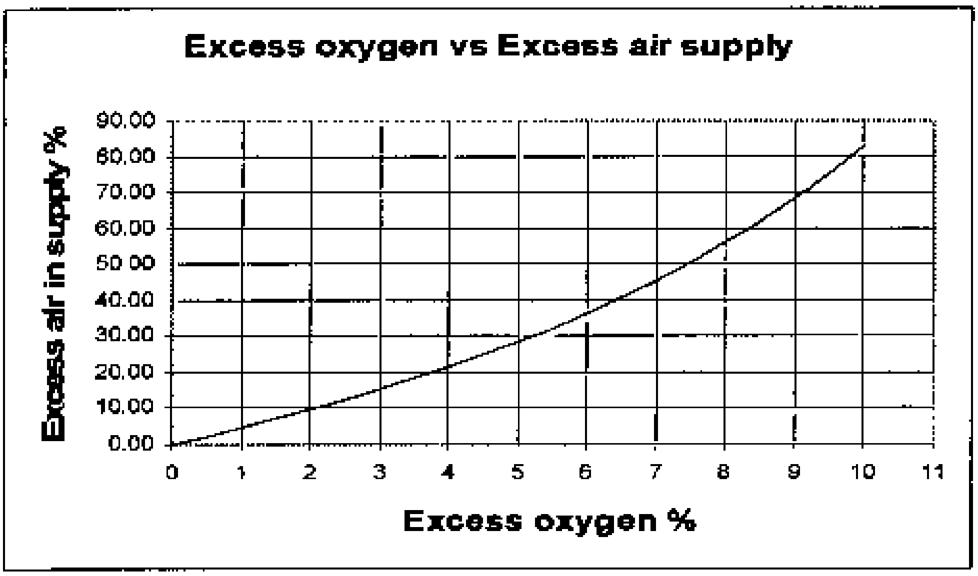

Figure 7.3 Relation between excess oxygen and excess air in supply

Table 7.4 gives the recommended excess air levels for natural draft and forced draft furnace for different grades of fuel:

Table 7.4 Recommended excess air levels (Garg, 1997)

Fuel	Natural draft	Forced draft
Fuel gas	15 – 20 %	10 – 15 %
Light fuel oil	20 – 25 %	15 – 20 %
Heavy fuel oil	25 – 30 %	20 – 25 %

7.2.2.2 Energy saving by lowering stack (Flue gas) temperature

The efficiency of furnace varies with flue gas temperature The Equation (6.11) (Chapter-6) reproduced below shows the relationship between excess oxygen in flue gas, flue gas temperature with thermal efficiency.

Thermal efficiency,

$$\eta = 98 - \frac{0.66 + 0.08 \times \left(\frac{11.42 - 0.047\%O_2}{21 - \%O_2} + 0.047 \right) \times (t_{fluegas} - 15.5)}{111.63 + \frac{11.42 - 0.047\%O_2}{21 - \%O_2}}$$

Considering excess oxygen in flue gas 4 %, air temperature 30°C and furnace radiation loss 2%, the above equation gives a relationship between furnace efficiency and flue gas temperature, graphically represented in Figure 7.4.

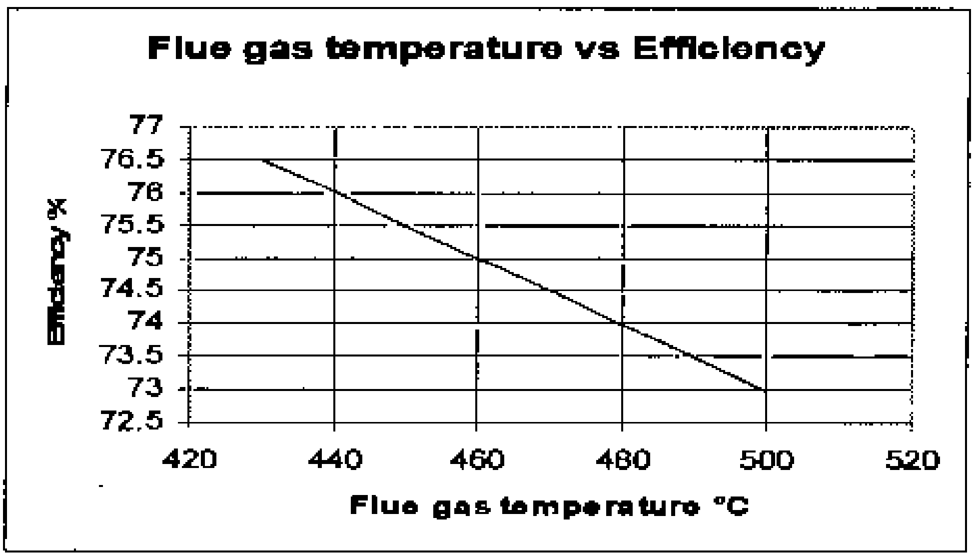

Figure 7.4 Relationship between flue gas temperature and furnace efficiency

The above graphical representation shows that, furnace efficiency increases with the decrease of flue gas temperature. It is observed for 30 ° C decrease in flue gas temperature, efficiency improvement is about 1.5 %.

Sample calculation:

Average value of oxygen in dry flue gas of Distillation furnace is about . 9.5 %,
Excess air supply to the furnace (equation 6.8) : 75.46 %
Design value of excess air supply for Distillation furnace . 40%
Excess air supply than the design value : 37 %
From Table 6.5 the efficiency of furnace at stack temperature 471 °C against excess air supply is given below.

	Excess O_2 %	Excess air %	Efficiency %
Designed value	6 4	40	71.20
Measured value	9.5	75.46	65.01

Heat absorbed by crude oil (useful heat from furnace) =18.88 MMkcal/hr
Cost of furnace oil = Tk. 7033.00/ M ton
Heating value of furnace oil = 10250 MMkcal/kg
Cost of heating value = Tk 686.15 /MMkcal
Heat fired:

For 71.20 % efficiency (18.88/0.7120) = 26.52 MM kcal/hr
For 65.01 % efficiency, (18.88/0.6501) = 29.04 MM kcal/hr

If it is possible to operate the furnace at efficiency 71.20% instead of 65.01 %, i.e for (71.20 - 65.01) 6.19 % efficiency improving:

Energy saving = 2.52 MM kcal/hr
Money saving = 2.52MMkcal/hr x 686.15 Tk. /MM kcal
= Tk.1729 01 /hr
= Tk.1, 38, 32,080.00 /year

For 1% efficiency improving

Energy saving = 0.407 MMkcal/hr = 3256 MMkcal/year
Money saving = Tk. 22, 34,104.00/year

For 1% efficiency improvement, energy saving is around 3256 MMkcal and money saving around Tk.22, 34,104.00 per year. Distillation furnace is running with excess air. ERL has scope to improve furnace efficiency significantly by controlling only the excess air supply. A properly designed modern burner and close control monitoring system for excess oxygen in flue gas is the prerequisite to achieve the target of energy saving.

7.2.2.3 Energy saving by eliminating air leakage into the firebox:

In most furnaces, the firebox pressure is kept close to atmospheric pressure. Figure 7.5 shows the places where air can leak into a heater. As flue gases progress through the unit, the pressure drops and may go down as low as –10 in. WC (Water Column) .Oxygen content in flue gas is determined by Orsat apparatus and it must maintain at minimum level as near as possible. Most of air leakage occurs in furnace chamber To minimize air leakage in to the heater, all peepholes must be kept closed

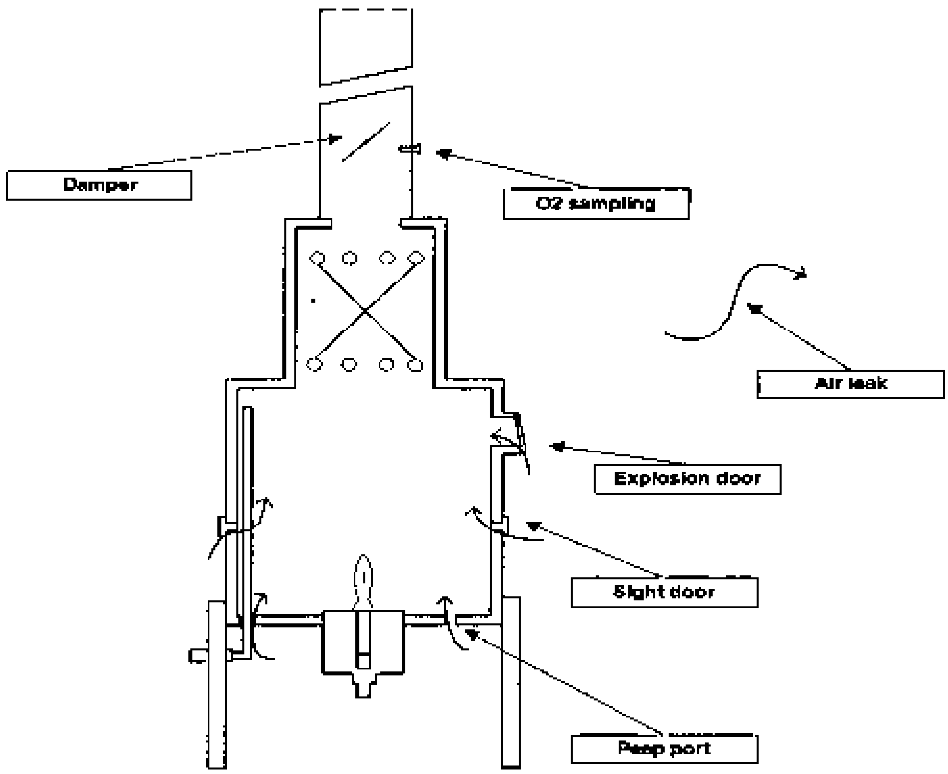

Figure 7.5 Air leakages in Furnace

The header box doors must be tightened to eliminate any air leakage. Explosion door should be kept closed. It should be ensured that there is minimal air leakage from the tube guide penetrations in the floor Optimizing fired heater performance is possible by making minor modifications and practicing good housekeeping.

Here is a summary of energy saving tips for fired heaters.

Furnace fuel can be saved in two major ways (Garg, 1997):

- Reduce excess air - every 15 % reduction in excess air saves about 1% in fuel
- Reduce flue gas temperatures - every 35 °F reduction in flue gas temperature saves about 1% in fuel

The following steps will help reduce excess air:

- Maintain a design draft (typical value is 0.1 in WC)
- Adjust burner registers and stack damper to control the draft
- Shut off air registers for the burners not online
- Close all peepholes and doors securely
- Close all header boxes in the convection section and radiant boxes
- Maintain clean combustion at all times.

The following steps will help reduce flue gas temperature:

- Reduce excess air to burners
- Clean convection section tubes with soot blowers or steam lances.

7.3 Energy Recovery

Tremendous amount of heat is utilized in heating crude oil in furnace for fractionation in distillation unit. Energy conservation is mainly based on recovery of excess heat from the system and from hot product streams Heat energy from furnace flue gas can be recovered by installing air preheater and process heat can be recovered by installing heat exchanger.

- Heat recovery from furnace stack gas
- Heat recovery from product streams

7.3.1 Heat recovery from furnace stack gas

Typically, the flue gas temperature leaving the furnace convection section ranges from 340 °C to 450 °C. To recover huge amount heat form flue gases it is wise to install air preheater in the system A typical installation of Air preheater is shown in Figure 7.6

The installation of air preheater recovers the heat from the stack and introduces it into the radiant section in the form of the preheated air. The preheated air combined with forced draught burners creates turbulence in the firebox. This raises the radiant heat flux and tube wall temperatures. Besides this, capacity of the heater increases. The air preheating system offers a huge energy saving potential. Every 35 °C drop in the exit flue gas temperature boosts thermal efficiency by 1%. Total fuel savings in this case would about 14% – 15 % (Garg, 1997).

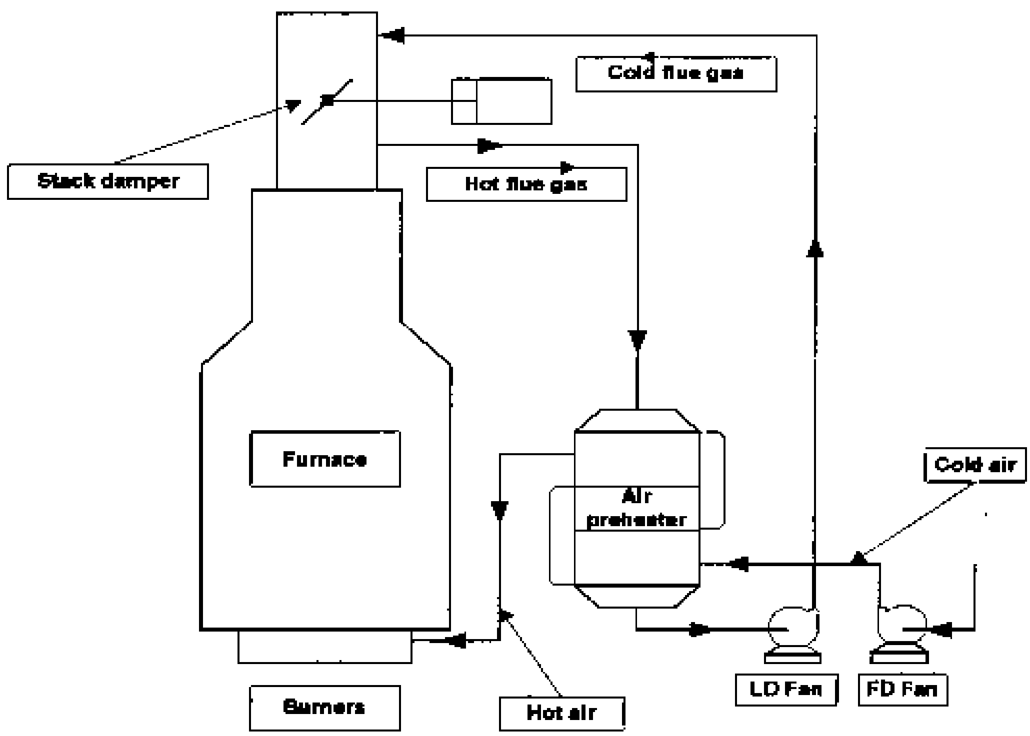

Figure 7.6 Air preheater system.

Installation an air-preheating system is a major revamp. It entails installing forced-draught burners, forced-draught and induced-draught fans, hot and cold air and flue gas ducts, and the air preheater Space must be available for the air preheater, fan, ducts and dampers.

Furnaces of ERL are not provided with air preheating system. Distillation furnace is a natural draft furnace equipped with duel fuel (heavy oil and Gaseous fuel) burning system. There is no provision to install air preheater in this furnace with the present facilities. Complete dismantling of existing furnace is required to install an air preheater. It is a major modification to the system. Due to financial constrains and considering return on investment, ERL has no any programme to install air heater in near future.

7.3.2 Heat recovery from product streams

The primary function of heat exchanger is to reuse process heat, to increase the furnace inlet temperature of feed stock by thermal exchange with hot product streams from the Distillation column, thus reducing the amount of fuel supplied to the furnace. ERL has installed ten numbers of cross flow, shell and tube type Heat exchanger to recover heat from six hot product streams from Distillation column at different temperature to preheat crude oil feed from storage. Design criteria of present exchangers are shown in Table 7.5.

Table 7.5 Existing ERL crude preheat train

Exchanger	Shell side fluid	Tube side fluid	Heat transfer Area (M^2)	Heat duty MM Kcal/hr
E 1101	K I	Crude oil	114.00	0.95
E 1102	K II	Crude oil	93.70	1.89
E 1103	LGO	Crude oil	93.70	1.54
E 1104	Cir.reflux	Crude oil	93.70	2.16
E 1105	Cir reflux	Crude oil	93.70	3.07
E 1106	LGO	Crude oil	93.70	3.00
E 1107	HGO	Crude oil	93 70	1.10
E 1108	Residue	Crude oil	237 00	0.42
E 1109AB	Residue	Crude oil	475.00	5.19

ERL should revamp the preheat train to increase the furnace inlet temperature, at the same time care should be taken, excessive heat may create gaseous phase in feed stock, which is undesirable due to low heat content.

7.3.2.1 Performance study of existing crude preheat train

A study was made with the existing heat exchangers used in crude heat train ERL processes mainly two types of crude oil named, Arabian light crude (ALC) and Murban crude. Performance of preheat train is analyzed based on type of crude oil processed. Net heat rejected by the hot product streams and net heat absorbed by the crude oil has been calculated for each type of crude oil. Following tables and figures show the details of the performance study on crude preheat train.

Table 7.6 Heat duty of ERL crude preheat train
Crude: ALC
Fluid: Shell side

Exchanger	Fluid	Sp. Gr. 15/15°C	Flow rate Kg/hr	Temperature °C				Cp kcal/kg°C	Heat duty MMkcal/hr
				Inlet(T_1)	Outlet(T_2)	ΔT	T avg		
E1101	KI	0.770	18820.00	154	54	100	104.00	0.632	1.0632
E1102	K II	0.803	22480.00	199	68	131	133.50	0.636	1.8716
E1103	LGO	0.857	31700.00	158	91	67	124.50	0.587	1.2470
E1104	Cir.reflux	0.787	58704.00	129	116	13	122.50	0.638	0.4867
E1105	Cir.reflux	0.787	58704.00	260	159	101	209.50	0.727	4.3122
E1108	Residue	0.968	79375.00	200	145	55	172.50	0.560	2.4444
E1106	LGO	0.857	31700.00	294	158	136	226.00	0.683	2.9449
E1107	HGO	0.888	11140.00	325	188	137	256.50	0.687	1.0488
E1109AB	Residue	0.968	79375.00	362	195	167	278.50	0.649	8.5977
Total heat rejected by hot fluids									24.0166

Specific heat of petroleum fraction lies in the range of 0.3 to 0.85 and depends upon temperature and gravity. Lighter fractions have higher values. With increasing density the specific heat decreases. In absence of relevant data, the following correlation can be used (Rao, 1997).

Bureau of Standards formula.

$$\text{Sp.heat (Cp)} = \frac{1}{\rho}(0.4024 + 0.00081 \times t) \qquad \text{Kcal/kg°C}$$

where, ρ = Sp.gravity at 15.6/15.6 °C and t = temperature °C

Heat duty, Q = Flow rate (MT/h) x Sp.ht (kcal/kg °C) x Temperature difference

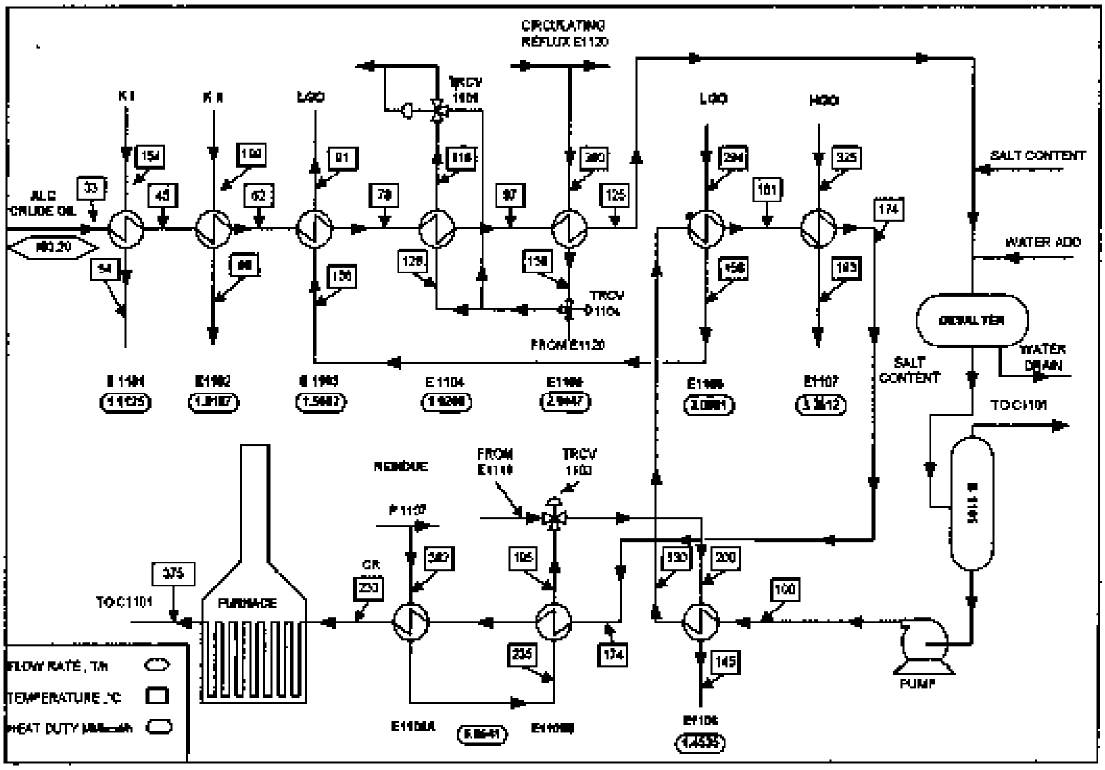

Figure 7.7 Flow diagram of Crude preheat train (ALC)

Table 7.7 Heat duty of ERL crude preheat train
Crude: ALC
Fluid: Tube side

Exchenger	Fluid	Sp. Gr.	Flow rate	Temperature °C				Sp. heat	Hear Duty
		15°/15°C	Kg/hr	Inlet(t_1)	Outlet(t_2)	Δ T	T avg	kcal/kg°C	MMkcal/hr
E1101	Crude oil	0.858	183200	33.00	45.00	12.00	39.00	0.506	1.1125
E1102	Crude oil	0.858	183200	45.00	62.00	17.00	53.50	0.520	1.6187
E1103	Crude oil	0.858	183200	62.00	78.00	16.00	70.00	0.535	1.5692
E1104	Crude oil	0.858	183200	78.00	97.00	19.00	87.50	0.552	1.9209
E1105	Crude oil	0.858	183200	97.00	125.00	28.00	111.00	0.574	2.9447
E1108	Crude oil	0.858	178202	100.00	130.00	30.00	115.00	0.578	3.0891
E1106	Crude oil	0.858	178202	130.00	161.00	31.00	145.50	0.607	3.3512
E1107	Crude oil	0.858	178202	161.00	174.00	13.00	167.50	0.627	1.4535
E1109 AB	Crude oil	0.858	178202	174.00	233.00	59.00	203.50	0.661	6.9541
Total heat gain by crude oil									24.0140

Table 7.8 Heat duty of Crude preheat train
Crude. Murban
Fluid· Shell side

Exchanger	Fluid	Sp. Gr. 15/15°C	Flow Rate Kg/hr	Temperature °C				Cp kcal/kg°C	Heat Duty MMkcal/hr
				Inlet(T_1)	Outlet(T_2)	*Del T	Tavg		
E1101	KI	0.7582	18179.00	158	55	103	106.50	0.645	1.2068
E1102	K II	0.8092	22087.50	204	69	135	136 50	0 634	1.8902
E1103	LGO	0.8460	32687.50	158	86	72	122 00	0 592	1.3944
E1104	Cir.reflux	0.7837	62518.05	140	110	30	125 00	0 643	1.2053
E1105	Cir.reflux	0 7837	62518.05	231	148	83	189 50	0.709	3.6807
E1108	Residue	0.9371	52625.00	194	129	65	161 50	0 569	1.9464
E1106	LGO	0.8460	32687.50	297	158	139	227 50	0 693	3 1508
E1107	HGO	0.8759	13362.50	329	188	141	258 50	0 698	1 3160
E1109AB	Residue	0 9371	52625.00	358	182	176	270 00	0 663	6.1388
Total heat rejected by fluids									21.9293

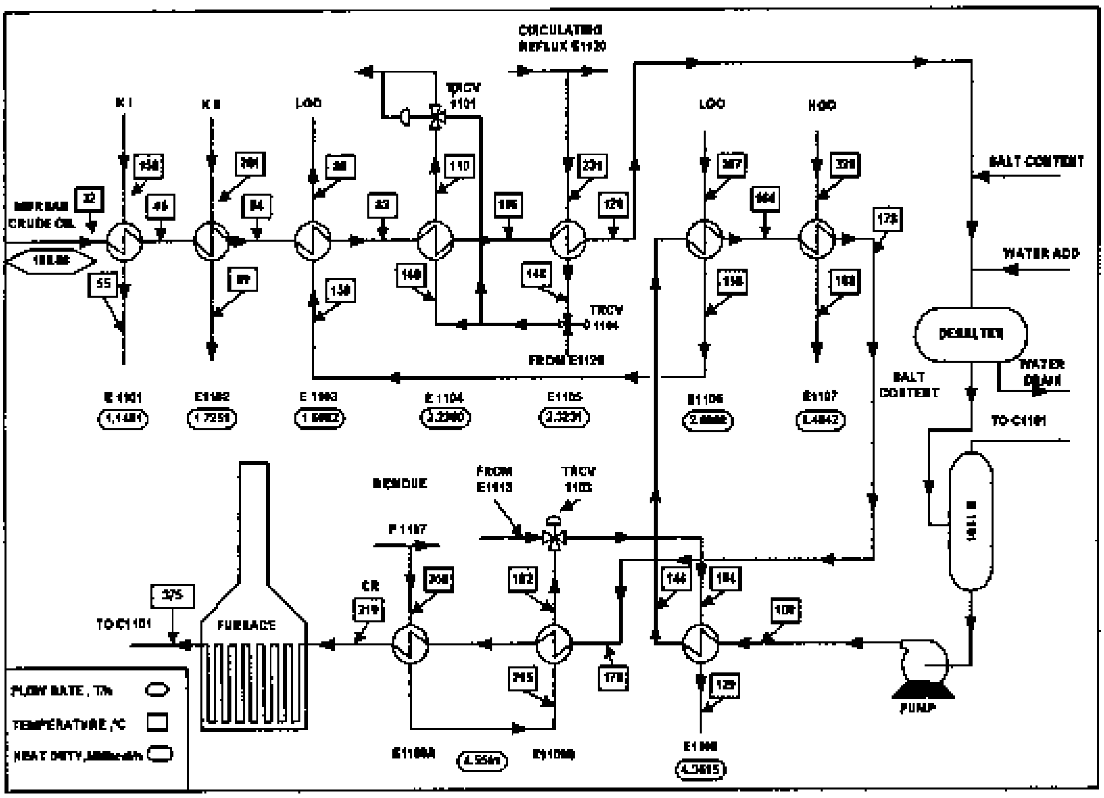

Figure 7.8 Flow diagram of Crude preheat train (Murban)

Table 7.9 Heat duty of Crude preheat train
Crude: Murban
Fluid: Tube side

Exchanger	Fluid	Sp Gr 15°/15°C	Flow Rate Kg/hr	Temperature °C				Sp heat kcal/kg°C	Hear Duty MMkcal/hr
				Inlet(t_1)	Outlet(t_2)	Del	Tavg		
E1101	Crude oil	0 8316	169083	32.00	45.00	13 00	38.50	0.521	1.1461
E1102	Crude oil	0.8316	169083	45 00	64.00	19.00	54 50	0.537	1.7251
E1103	Crude oil	0.8316	169083	64 00	82.00	18.00	73 00	0.555	1.6891
E1104	Crude oil	0 8316	169083	82 00	105 00	23 00	93.50	0.575	2 2360
E1105	Crude oil	0.8316	169083	105.00	128.00	23.00	116.50	0.597	2.3231
E1108	Crude oil	0.8316	164085	100.00	144.00	44.00	122.00	0.603	4.3515
E1106	Crude oil	0 8316	164085	144 00	164 00	20 00	154 00	0 634	2.0802
E1107	Crude oil	0 8316	164085	164.00	178.00	14.00	171.00	0.650	1.4942
E1109 AB	Crude oil	0 8316	164085	178.00	219 00	41.00	198.50	0 677	4 5561
Total heat gain by crude oil									21.6012

Table 7.10 Heat absorbed by crude oil

	unit	Design 190 T/H	ALC 190 T/H	Murban 173 T/H
Heat absorbed by crude oil pre-heating	MMkcal/h	19.8	24.01	21.60
Furnace inlet temperature °C	°C	209.00	233.00	219.00
Heat absorbed by crude oil in furnace	MMkcal/h	26.00	19.13	19 79

7.3.2.2 Performance study by rearranging exchanger

A study was made by rearranging existing exchangers in different positions. The main objective of this study is to increase the furnace inlet temperature. Calculations are made theoretically with the available data for hot product stream. The rise in temperature by crude oil through each exchanger is obtained by heat balance. Physical observation of heat train by rearranging the exchangers was not possible because, it needed a major modification in production pattern.

Table 7.11 Placing E1108 parallel with E1101 (flow rate 50:50)
Crude ALC
Fluid: Shell side

Exchanger	Fluid	Sp. Gr. 15/15°C	Flow Rate Kg/hr	Temperature °C				Cp kcal/kg°C	Heat loss MMkcal/hr
				T_1	T_2	ΔT	Tavg		
E1101	KI	0.770	16820.00	154	54	100	104.00	0.632	1.0632
E1108	Residue	0.968	79375.00	200	145	55	172.50	0.560	2.4444
E1102	K II	0.803	22480.00	199	68	131	133.50	0.636	1.8716
E1103	LGO	0.857	31700.00	158	91	67	124.50	0.587	1.2470
E1104	Cir.reflux	0.787	58704.00	129	116	13	122.50	0.638	0.4867
E1105	Cir reflux	0.787	58704.00	260	159	101	209.50	0.727	4.3122
E1106	LGO	0.857	31700.00	294	158	136	226.00	0.683	2.9449
E1107	HGO	0.888	11140.00	325	188	137	256.50	0.687	1.0488
E1109AB	Residue	0.968	79375.00	362	195	167	278.50	0.649	8.5977
Total heat rejected by hot fluids									24.0166

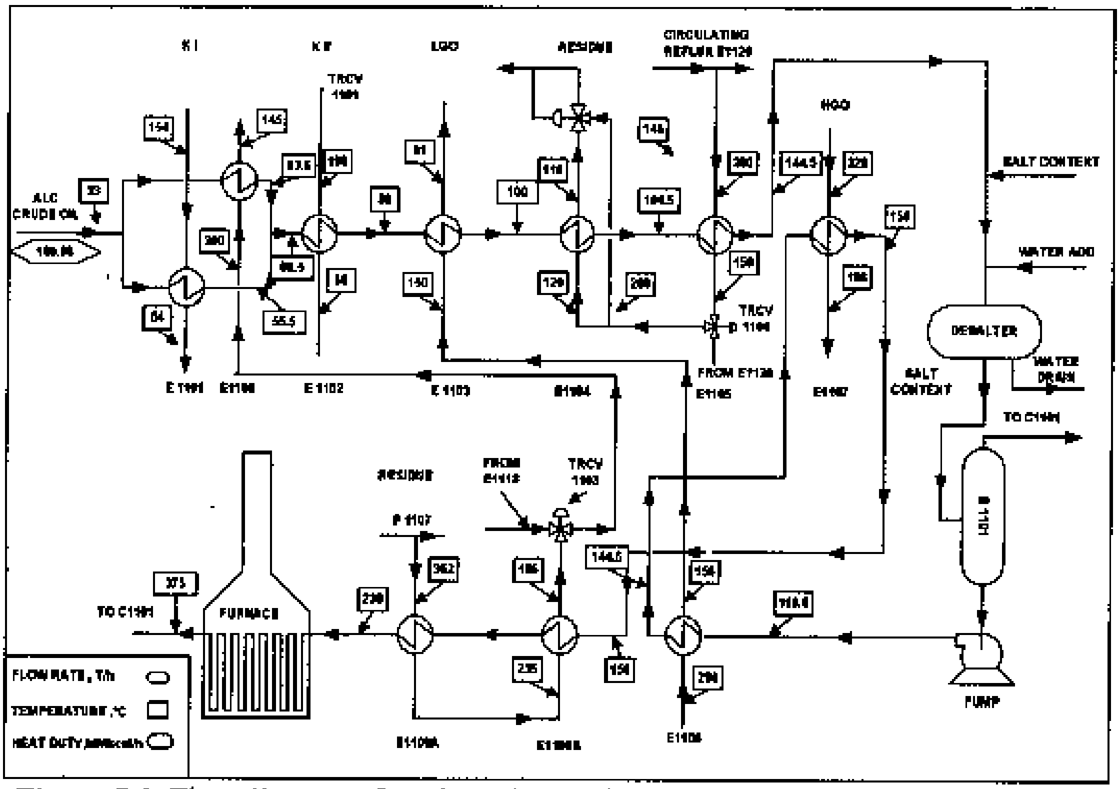

Figure 7.9 Flow diagram of crude preheat train, placing E1108 parallel with E1101.

Table 7.12 Placing E1108 parallel with E1101 (flow rate 50:50)
Crude: ALC
Fluid: Tube side

Exchanger	Fluid	Sp. Gr. 15°/15°C	FlowRate Kg/hr	Temperature °C				Sp heat kcal/kg°C	Hear gain MMkcal/hr
				t_1	t_2	Del T	Tavg		
E1101	Crude oil	0.858	91600	33.00	55.50	22.50	44.25	0.511	1.0532
E1108	Crude oil	0.858	91600	33.00	83.50	50.50	58.25	0.524	2.4250
E1102	Crude oil	0.858	183200	69.50	88.00	18.50	78.75	0.544	1.8424
E1103	Crude oil	0.858	183200	88.00	100.00	12.00	94.00	0.558	1.2267
E1104	Crude oil	0.858	183200	100.00	104.50	4.50	102.25	0.566	0.4664
E1105	Crude oil	0.858	183200	104.50	144.50	40.00	124.50	0.587	4.3001
E1106	Crude oil	0.858	183200	119.50	146.50	27.00	133.00	0.595	2.9423
E1107	Crude oil	0.858	178202	146.50	156.00	9.50	151.25	0.612	1.0362
E1109 AB	Crude oil	0.858	178202	156.00	230.00	74.00	193.00	0.652	8.5914
Total heat gain by crude oil									23.8836

From the above rearrangement it is observed that the furnace inlet temperature, that is, outlet temperature of exchanger E1109AB become 230 °C which is not satisfactory.

Table 7.13 ERL Crude preheat train, Placing E1108 parallel with E1103 (50/50), and placing E1104 after E1102
Crude: ALC
Fluid: Shell side

Exchanger	Fluid	Sp Gr 15/15°C	Flow Rate Kg/hr	Temperature °C				Cp kcal/kg°C	Heat loss MMkcal/hr
				T_1	T_2	Del T	Tavg		
E1101	K I	0.770	16820.00	154	54	100	104.00	0.632	1.0632
E1102	K II	0.803	22480.00	199	68	131	133.50	0.636	1.8716
E1104	Cir.reflux	0.787	58704.00	129	116	13	122.50	0.638	0.4867
E1103	LGO	0.857	31700.00	158	91	67	124.50	0.587	1.2470
E1108	Residue	0.968	79375.00	200	145	55	172.50	0.560	2.4444
E1105	Cir reflux	0.787	58704.00	260	159	101	209.50	0.727	4.3122
E1106	LGO	0.857	31700.00	294	158	136	226.00	0.683	2.9449
E1107	HGO	0.888	11140.00	325	188	137	256.50	0.687	1.0488
E1109AB	Residue	0.968	79375.00	362	195	167	278.50	0.649	8.5977
Total heat rejected by hot fluids									24.0166

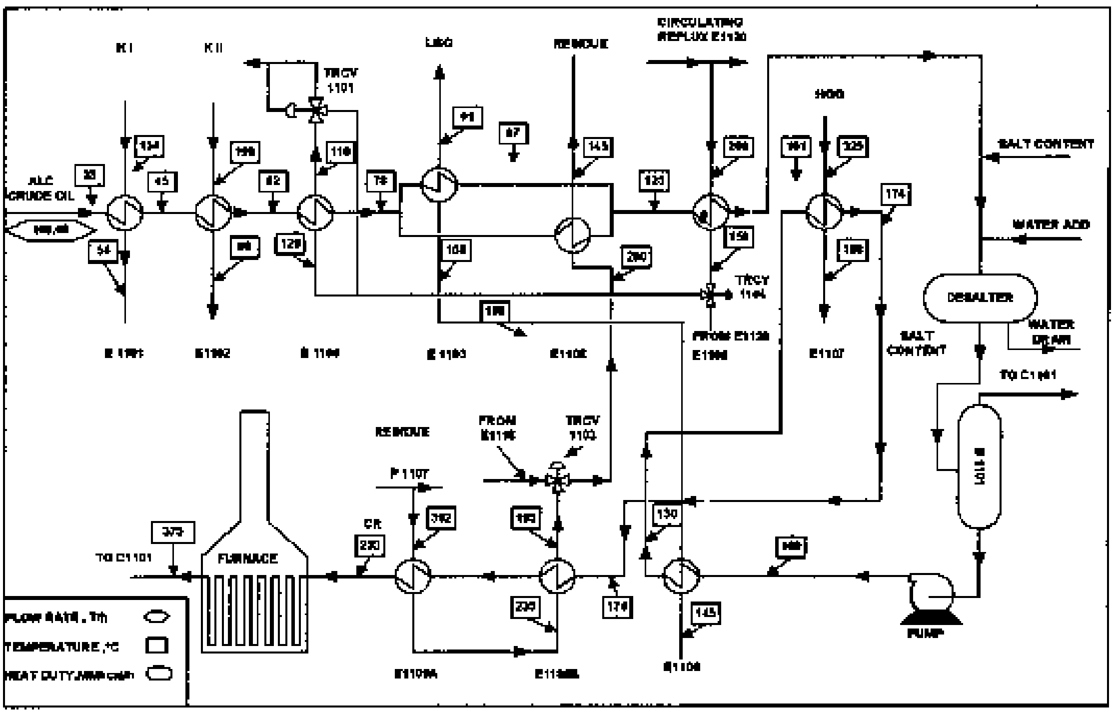

Figure 7.10 Flow diagrams of crude preheat train, Placing E1108 parallel with E1103 and placing E1104 afterE1102 (ALC)

Table 7.14 ERL Crude preheat train, Placing E1108 parallel with E1103 and placing E1104 after E1102
Crude: ALC
Fluid· Tube side

Exchanger	Fluid	Sp. Gr. 15°/15°C	Flow Rate Kg/hr	Temperature °C				Sp heat kcal/kg°C	Hear gain MMkcal/hr
				t_1	t_2	Del T	Tavg		
E1101	Crude oil	0 858	183200	33.00	45.00	12 00	39.00	0 506	1.1125
E1102	Crude oil	0 858	183200	45.00	62.00	17 00	53 50	0 520	1 6187
E1104	Crude oil	0.858	183200	62.00	67.00	5.00	64.50	0.530	0.4856
E1103	Crude oil	0.858	91600	67 00	92 00	25 00	79 50	0 544	1 2465
E1108	Crude oil	0.858	91600	67 00	115 00	48.00	91 00	0 555	2 4410
E1105	Crude oil	0 858	183200	103.50	143.50	40.00	123.50	0.586	4.2932
E1106	Crude oil	0.858	178202	118.50	147.50	29.00	133.00	0.595	3.0740
E1107	Crude oil	0 858	178202	147.50	157.00	9.50	152 25	0.613	1 0378
E1109AB	Crude oil	0.858	178202	157.00	231.00	74.00	194.00	0.652	8.6038
Total heat gain by crude oil									23.9130

Above arrangement of exchanger gives furnace inlet temperature 231°C which is nearer to the present set of arrangement.

Table 7.15 Crude preheat train Placing E1108 after E1104
Crude: ALC
Fluid: Shell side

Exchanger	Fluid	Sp. Gr 15/15°C	Flow Rate Kg/hr	Temperature °C				Cp kcal/kg°C	Heat loss MMkcal/hr
				T_1	T_2	Del T	Tavg		
E1101	KI	0.770	16820.00	154	54	100	104.00	0.632	1.0632
E1102	K II	0.803	22460.00	199	68	131	133.50	0.636	1.8716
E1103	LGO	0.857	31700.00	158	91	67	124.50	0.587	1.2470
E1104	Cir.reflux	0.787	58704.00	129	116	13	122.50	0.638	0.4867
E1108	Residue	0.968	79375.00	200	145	55	172.50	0.560	2.4444
E1105	Cir.reflux	0.787	58704.00	260	159	101	209.50	0.727	4.3122
E1106	LGO	0.857	31700.00	294	150	144	222.00	0.679	3.1008
E1107	HGO	0.888	11140.00	325	188	137	256.50	0.687	1.0486
E1109AB	Residue	0.968	79375.00	362	195	167	278.50	0.649	8.5977
Total heat rejected by hot fluids									24.1725

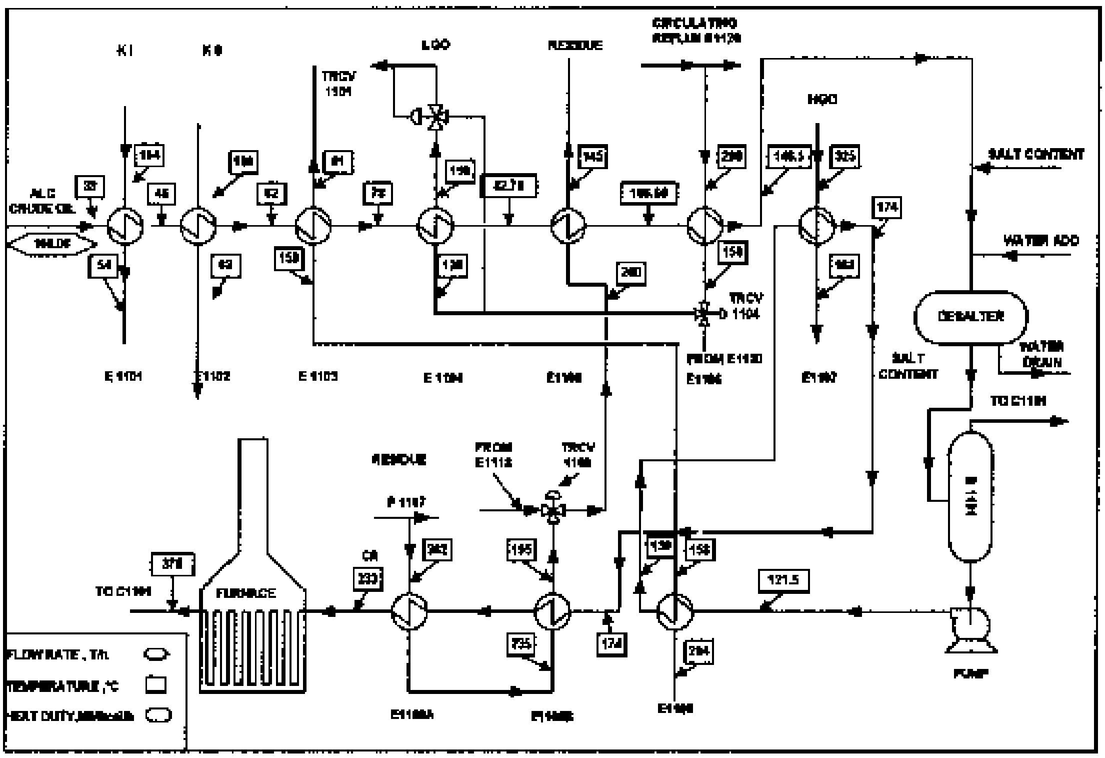

Figure 7.11 Flow diagram of crude preheat train, Placing E1108 after E1104 (ALC)

Table 7.16 Crude preheat train placing E1108 after E1104
Crude: ALC
Fluid: Tube side

Exchanger	Fluid	Sp Gr 15°/15°C	Flow Rate Kg/hr	Temperature °C t_1	t_2	Del T	Tavg	Sp heat kcal/kg°C	Heat gain MMkcal/hr
E1101	Crude oil	0.858	183200	33.00	44 00	11.00	38.50	0.506	1.0188
E1102	Crude oil	0 858	183200	44 00	63.50	19.50	53 75	0.520	1 8576
E1103	Crude oil	0.858	183200	63.50	76 00	12 50	69.75	0 535	1.2254
E1104	Crude oil	0 858	183200	76.00	80.50	4.50	78.25	0.543	0.4478
E1108	Crude oil	0 858	183200	80 50	104.00	23 50	92 25	0.556	2 3952
E1105	Crude oil	0.858	183200	104.00	144 00	40 00	124.00	0 588	4.2967
E1106	Crude oil	0.858	178202	119 00	148.00	29.00	133.50	0 595	3.0765
E1107	Crude oil	0.858	178202	148.00	157.50	9.50	152.75	0.613	1 0386
E1109 AB	Crude oil	0 858	178202	157 50	231.00	73.50	194 25	0 653	8.5488
Total heat gain by crude oil									23.9052

Above arrangement gives furnace inlet temperature 231 °C which is less than that of the existing arrangement. A comparison of performance of crude preheats train for different arrangement of exchanger has stated in Table 7.17.

Table 7.17 Comparison statement of crude preheat train:

Condition of preheat train (Crude oil : ALC)	Heat loss by hot product MMkcal/hr	Heat gain by crude oil MMkcal/hr	Furnace inlet temperature °C
Existing preheat train	24 0166	24.0140	233
Placing E1108 parallel with E1101 (flow rate 50:50)	24.0166	23.8836	230
Placing E1108 parallel with E1103 (flow rate 50:50) and placing E1104 after E1102	24.0166	23 9130	231
Placing E1108 after E1104	24.1725	23.9052	231

From above statement it can be concluded that, present arrangement of heat exchangers in crude preheat train is satisfactory.

7.4 Analysis of Crude Preheat Train by "Pinch Technology"

The design and lay out of present heat exchangers of crude preheat train is studied on basis of 'Pinch Technology'. In ERL crude preheat train there are 10 numbers exchangers with six hot streams at different temperature and one cold stream.

Construction of "Composite curve"

The "hot" streams (i.e streams that have to be cooled) and "cold" streams (i.e. streams that have to be heated) in a process can be represented on a temperature-heat content (enthalpy) graph once their input and output temperatures (or "supply" and "target" temperatures) and their flow rates and physical properties are known. Differential heat flow dq, when added to a process stream, will increase its enthalpy by Cp dT (ICE, 1983).

where,

Cp = "heat capacity flow rate" (kcal/°C)

= mass flow (kg/hr) x specific heat (kcal/kg°C)

dT = differential temperature change

Hence, with Cp assumed constant, for a stream requiring heating ("cold" stream) from a "supply temperature" (T_s) to a "target temperature" (T_T), the total heat added will be equal to the stream enthalpy change, i.e.

$$Q = \int_{T_S}^{T_T} Cp\,dT = Cp(T_T - T_S) = \Delta H$$

The T/H diagram can be used to represent heat exchange. Starting from the individual streams it is possible to construct one "composite curve" of all hot streams in the process and another of cold streams, by simple addition of heat contents over the temperature range in the problem. Construction of composite curve is illustrated in Figure 7.12 for a number of hot streams with heat capacities A, B and C being cooled through the temperature level indicated

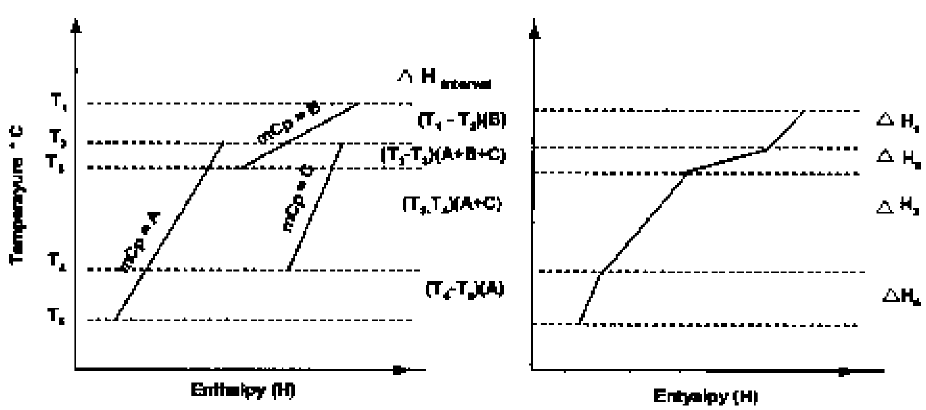

Figure 7.12 Construction of "composite Curves"

The inlet and outlet temperature of hot streams in each exchanger of present crude preheat train with corresponding enthalpy is shown in Table 7.18

Table 7.18 Crude preheat train (hot fluid stream)

Exchanger	Temperature, °C		Enthalpy MMkcal/h	
	inlet	outlet	inlet H1	outlet H2
E1101	154	54	2.34	1 46
E1102	199	68	3.75	2.06
E1103	158	91	4.58	3.35
E1104	129	116	7.34	6.76
E1105	260	159	12.24	8.32
E1108	200	145	13.46	11.03
E1106	294	150	7.44	4.46
E1107	325	188	2.85	1.81
E1109 AB	362	195	22.06	13.23

Graphical representation of hot product stream in temperature - enthalpy diagram is shown in Figure 7 13

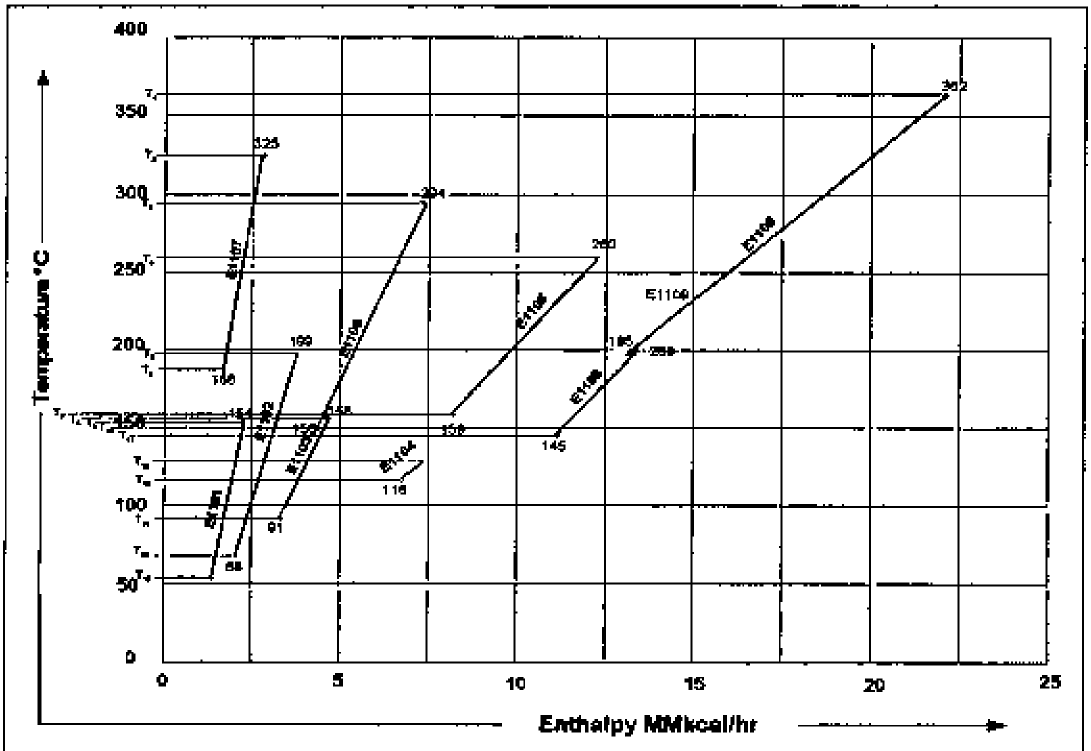

Figure-7.13 Representation of process stream ('hot stream').

The inlet and outlet temperature of cold stream in each exchanger with corresponding enthalpy is shown in Table 7.19 and graphical representation of cold product stream in temperature - enthalpy diagram is shown in Figure 7.14.

Table 7.19 Crude preheat train (Cold fluid stream)

Exchanger	Temperature		Enthalpy MMkcal/h	
	Inlet	Outlet	Inlet H_1	Outlet H_2
E1101	33	45	14 26	15 28
E1102	45	62	15.28	17.01
E1103	62	78	17.01	18 33
E1104	78	97	18.33	20.17
E1105	97	125	20.17	23.02
E1108	100	130	19.82	22.79
E1106	130	161	22.79	25.96
E1107	161	174	25.86	27.25
E1109 AB	174	233	27.25	34.18

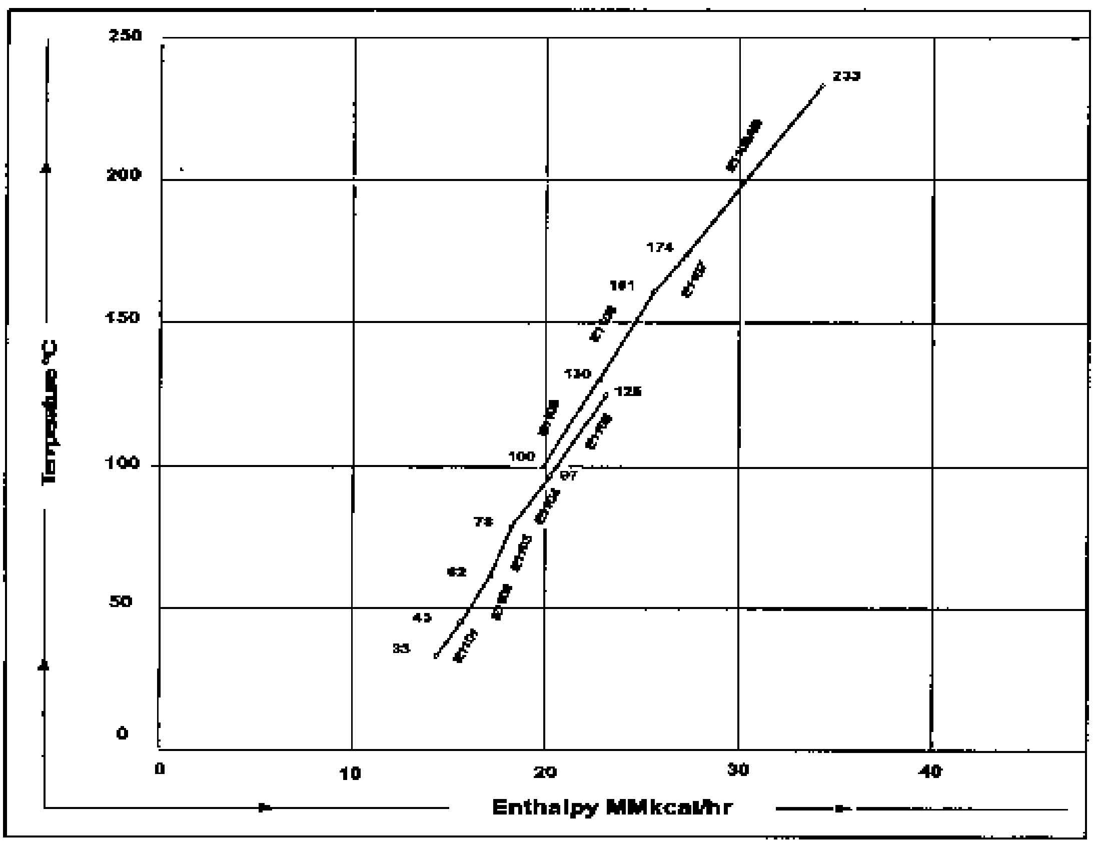

Figure 7.14 Representation of process stream (Cold stream)

Construction of composite curve of hot and cold stream (Pinch Analysis)

Heat gain / loss, Q = Mass flow rate x Sp.heat x Temperature difference

$= m \times Cp \times \Delta T$

$= E \times \Delta T$

$= \Delta H$

Table 7.20 Construction of Composite curve of hot stream

Exchanger	Product Shell side	Sp.gr. 15°C/15°C	Flow Rate Kg/h	Temperature		Cp	E(i)
				Inlet	Outlet		M x Cp
E1101	KI	0.770	16820.00	154	54	0.6320	0.0106
E1102	K II	0 803	22480.00	199	68	0.6358	0.0143
E1103	LGO	0 857	31700.00	158	91	0.5872	0.0186
E1104	Cir.reflux	0.787	58704.00	129	116	0.6374	0.0374
E1105	Cir.reflux	0.787	58704.00	260	159	0.7269	0.0427
E1108	Residue	0.968	79375.00	200	145	0.5600	0.0445
E1106	LGO	0.857	31700 00	294	150	0.6794	0.0215
E1107	HGO	0.888	11140.00	325	188	0.6871	0 0077
E1109 AB	Residue	0.968	79375.00	362	195	0.6487	0.0515

Table-7.21 Construction of composite curve

Shell side (Hot stream)					
Del H i = E i x (Inlet, temp. – Outlet,temp.)		E(mxCp)	Temperature °C		Del Hi
			Inlet	Outlet	
Del H1 =	E9 x(T1-T2)	0.0515	362	325	1.91
Del H2 =	(E9+E7) x (t2-T3)	0.0591	325	294	1.83
Del H3 =	(E9+E7+E6) x (t3-T4)	0.0807	294	260	2.74
Del H4 =	(E9+E7+E6+E5)X(T4-T5)	0.1234	260	200	7.40
Del H5 =	(E9+E7+E6+E5+E8)X(T5-T6)	0.1678	200	199	0.17
Del H6 =	(E9+E7+E6+E5+E8+E2)X(T6-T7)	0.1821	199	195	0.73
Del H7 =	(E7+E6+E5+E8+E2)X(T7-T8)	0.1306	195	188	0 91
Del H8 =	(E6+E5+E8+E2)X(T8-T9)	0.1091	188	159	3.16
Del H9 =	(E6+E8+E2)X(T9-T10)	0.0803	159	158	0.08
Del H10 =	(E6+E8+E2+E3)X(T10-T11)	0.0989	158	154	0.40
Del H11 =	(E6+E8+E2+E3+E1)X(T11-T12)	0.1095	154	150	0.44
Del H12 =	(E8+E2+E3+E1)X(T12-T13)	0.0880	150	145	0.44
Del H13 =	(E2+E3+E1)X(T13-T14)	0.0435	145	129	0.70
Del H14 =	(E2+E3+E1+E4)X(T14-T15)	0.0810	129	116	1.05
Del H15 =	(E2+E3+E1)X(T15-T16)	0.0435	116	91	1.09
Del H16 =	(E2+E1)X(T16-T17)	0.0249	91	68	0.57
Del H17 =	(E1X(T17-T18)	0.0106	68	54	0.15

Table 7.22 Construction of hot composite curve (Enthalpy against temperature):

Temperature °C	H i	Del Hi	Enthalpy H Hi+ Del Hi
54	1.46	-	1.46
68	1.46	0.15	1.61
91	1.61	0.57	2.18
116	2.18	1.09	3.27
129	3.27	1.05	4.32
145	4 32	0.7	5.02
150	5.02	0.44	5.46
154	5.46	0.44	5.90
158	5.90	0.4	6.29
159	6.29	0.08	6.37
188	6.37	3.16	9.54
195	9.54	0 91	10.45
199	10.45	0.73	11.18
200	11.18	0.17	11.35
260	11.35	7.4	18.75
294	18.75	2.74	21.49
325	21.49	1.83	23.33
362	23.33	1.91	25 23

Table 7.23 Construction of composite curve (cold stream)

Exchanger	Product	Flow rate kg/hr	Sp.gr. 15°C/15°C	Temperature °C		Cp kcal/kg °C	Ei= mCp kcal/hr°C
				inlet	outlet		
E1101	Crude oil	183200	0.8582	33	45	0.51	0.0926
E1102	Crude oil	183200	0.8582	45	62	0.52	0.0952
E1103	Crude oil	183200	0.8582	62	78	0.53	0.0980
E1104	Crude oil	183200	0.8582	78	97	0.55	0.1010
E1105	Crude oil	183200	0.8582	97	125	0.57	0.1051
E1108	Crude oil	178202	0.8582	100	130	0.58	0.1029
E1106	Crude oil	178202	0.8582	130	161	0.61	0.1080
E1107	Crude oil	178202	0.8582	161	174	0.63	0 1117
E1109 AB	Crude oil	178202	0.8582	174	233	0.66	0.1178

Table 7.24 Construction of cold composite curve

Del H i = E i x (Inlet,temp. – outlet,temp.)		E	Temperature °C		Del Hi
			Inlet	Outlet	
Del H1 =	E9x(233-174)	0.1178	233	174	6.95
Del H2 =	(E7) x (174-161)	0.1117	174	161	1.45
Del H3 =	(E6) x (161-130)	0 1080	161	130	3.35
Del H4 =	(E8)X(130-125)	0.1029	130	125	0.51
Del H5 =	(E8+E5)x(125-100)	0.2080	125	100	5.20
Del H6 =	(E5x(100-97)	0.1051	100	97	0.32
Del H7 =	E4x(97-78)	0.1010	97	78	1.92
Del H8 =	E3x(78-62)	0.0980	78	62	1.57
Del H9 =	E2x(62-45)	0.0952	62	45	1 62
Del H10 =	E1x(45-33)	0.0926	45.00	33	1.11

Table 7.25 Construction of cold Composite curve

(Enthalpy against temperature):

Temperature. °C	Hi	Del Hi	Enthalpy H= Hi + Del Hi
33	14.26	-	14.26
45	14.26	1.11	15.37
62	15.37	1.62	16.99
78	16.99	1.57	18.56
97	18.56	1.92	20.48
100	20.48	0.32	20.80
125	20.80	5 20	26.00
130	26.00	0.51	26.51
161	26.51	3.35	29.86
174	29.86	1.45	31.31
233	31.31	6.95	38.26

A combined composite curve of hot and cold product streams has been constructed in Temperature - Enthalpy diagram as shown in Figure 7.15 with the available data from Table 7 22 and Table 7.25 for hot and cold product respectively.

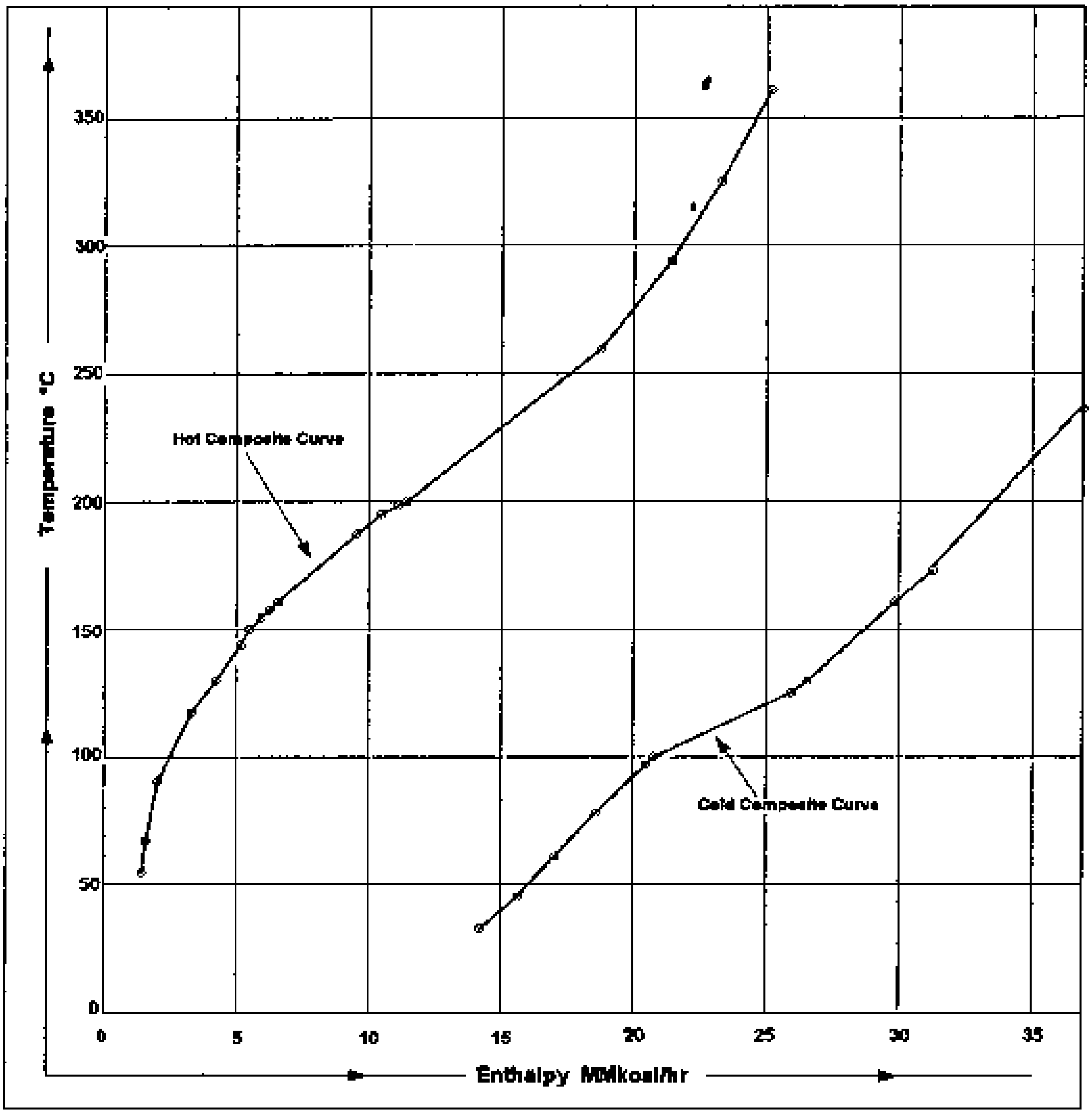

Figure 7.15 Composite Curve of Hot and Cold Stream.

The above study on present crude preheats system of ERL based on Pinch Technology shows that, there is no remarkable "pinch" in the composite curve (Figure 7.15). It may be mentioned that, there is no bottlenecking in the present crude preheat heat train and the complete design and layout of the heat exchanging system is satisfactory.

7.5 Heat Rejected to Cooling Water

After recovery of heat from the product streams by the present Crude preheat train a reasonable amount of heat is rejected to cooling water during cooling hot products before storage in tanks. Table 7.26 shows the amount of rejected to cooling water by each product stream.

Table 7.26 Heat rejected to cooling water

Exchanger	Fluid	Sp.gr. 15°C	Flow rate kg/hr	Temperature °C Inlet(T_1)	Outlet(T_2)	Del T	T avg	Sp.ht Cp	Q(S) MMkcal/hr
E1110	TG	0.5805	300	70	45	25	57.5	0.77	0.0058
E1111	KI	0.7699	16817	58	45	13	51.5	0.58	0.1261
E1112	K II	0.8033	22483	68	45	23	56.5	0.56	0.2885
E1113	LGO	0.8571	31700	92	63	29	77.5	0.54	0.4989
E1114	HGO	0.8879	11140	165	66	99	115.5	0.56	0.6160
E1115	Residue	0.9682	79380	165	96	69	130.5	0.52	2.8744
E1117	TG	0.5805	300	75	75	45	75	0.80	0.0108
E1121	HG	0.7338	7154	130	39	91	84.5	0.64	0.4177
E1122	LG	0.6803	12771	55	37	18	46	0.65	0.1486
Total heat rejected to cooling water									4.9868

It is observed that total heat rejected to cooling water is 4.9868 MMkcal/hr of which heat rejected through water cooler E1114 and E1115 is about 3.49 MMkcal/hr. (about 70%).Extra heat exchanger can be placed before cooling through Exchanger E1114 in HGO product stream line and Exchanger E1115.in RCO product stream line to save the maximum amount of heat. This amount of heat can be used for preheating crude oil, boiler feed water, desalter flash water, or as convenient to the operations. Quality of the material of heat exchanger is to be selected in accordance with the type of liquid is selected to exchange heat with the hot product stream.

CHAPTER 8

Discussions

ERL is the only petroleum refinery plant of its kind in Bangladesh. It is a process plant of complex nature. The main function of this plant is to refine crude petroleum to furnish various fuel and non fuel petroleum products to cater the need of end users. ERL designed in the mid 60's went on production in 1968. Its installed capacity of crude oil processing is 1.5 million tons per year. It produces 16 grades of petroleum fuel products including 4 grades of non-fuel products. It has five processing units. Study on energy use in ERL and conservation possibilities was kept limited to Crude Distillation unit, the main processing unit of refinery. Crude distillation is referred to as complete operation, in which heating of crude oil, vaporization, fractionation, condensation and cooling are performed.

Energy cost is the single largest component of variable operating expenses and is key to refinery profitability. At early stage of ERL, due to low cost of fuel, the energy conservation was not so important as it is in a refinery of today. Energy conservation has emerged as a popular topic of discussion over the past 30 years due to OPEC regulation of crude oil price since Arab oil embargo in 1973. To keep pace with the Global market, ERL should take necessary steps to minimize production cost by reducing energy waste, improving equipment efficiency and ensuring highest degree of conservation of energy.

Heating of crude oil is done in furnace, at the cost of burning hydrocarbon fuel. Furnace is the main consumer of energy. A sample calculation shows that energy consumption of distillation furnace of ERL is about 28.67 MMkcal/hr, which are about 47.27% of the total energy consumption of ERL. It is estimated that total energy consumption in ERL is about 60.54 MMkcal/hr. It is found that, the furnace is running with thermal efficiency of about 65% and is operating with about 77% excess air (9.5% excess Oxygen in fuel gas) which is much higher than the designed value (40%). 1-2 % improvement in thermal efficiency is possible by reducing only excess air supply to the furnace and through proper monitoring. Calculation shows that 1 % improvement in furnace efficiency results in energy saving around 3256 MMkcal/year and in terms of money saving around TK 22,34,104.00 per year. A relationship has been developed between thermal efficiency versus temperature of flue gas and presence of excess oxygen in flue gas. Thermal efficiency of a furnace increases with the decrease of flue gas temperature (stack temperature) and excess oxygen present in flue gas. In practice there are many factors related in obtaining the maximum efficiency. Proper air-fuel mixture, type of burner, placement of burner (floor or wall), placement of feed tube (horizontal or vertical), size of convection and radiant section, nature of refractories etc. play vital role to optimize the furnace efficiency.

Steam is another from of energy; a significant amount of steam is used in distillation unit as stripping steam to reduce the partial pressure of hydrocarbon in distillation column. Reduction of partial pressure causes fractionation of crude oil at low temperature. It is found that, consumption of steam in ERL distillation unit is about 14.53 kg per cubic meters of crude oil processing.

Electrical power is the other form of energy that plays vital role in any refinery. Entire operation of crude oil processing depends on electrical power. Uninterrupted power supply in a refinery is very essential; any break down, even for a few seconds, plays adverse effect in processing. ERL has its own power generating units with capacity of 8,500 kW. It is observed that, in case operation of all electrical equipments of distillation unit simultaneously through out the year, the maximum power consumption in distillation unit is near about 62,79,280 kwh , in terms of money it amounts to TK. 2,07,21,624 00 per year. It is estimated that, in processing crude oil in ERL, energy consumption is about 0.05 MMkcal per barrel. Overall processing expense of crude oil is increasing. Since 1990-91, this expense has risen from Tk. 32.88 per barrel to Tk.58 70 per barrel in 1999-2000

The main objectives of conservation of energy are to save energy during processing by proper operations and to recover heat from the process through proper arrangements. A reasonable amount of heat energy can be saved by flashing crude oil prior feeding to furnace, the crude vapor thus generated in flash drum is fed to distillation column by passing furnace, thereby saving fuel by eliminating the unnecessary overheating of vapors. The rate of evaporation depends on type of crude oil used, temperature and pressure of the flash drum. It is observed that evaporation rate of Murban crude is higher than ALC crude, mainly used in ERL. Excess flow rate of crude vapor from flash drum to distillation column may cool down the bottom temperature of the column that may create an adverse effect on processing. Excess flow rate may carry certain portion of liquid to the column which is not desirable A proper design is essential to save more energy by increasing flow rate of flashed vapor under balanced condition. It is estimated that, in ERL a saving of 1 MMkcal/hr is worth TK.686.15 per hour.

Recovery of process heat is another objective of energy conservation. Excess heat energy of flue gas can be recovered partly by installing air preheater. It is observed that 33% of the total generated heat in furnace is rejected to air through flue gas. Heat of flue gas can be use to preheat air required for combustion in furnace. Every 35 °C drop in exit flue gas temperature boots up thermal efficiency by 1 % and total fuel saving is around 14-15 %. Installation of an air-preheater is a major revamp. Space must be available for the air-preheater, fan, ducts and dampers. Furnaces of ERL are not equipped with air-preheater and present design will not permit to install any air-preheating system. After fractionation of crude oil in distillation unit, the main product streams leaving the column carries huge amount of heat. Part of this heat is recovered by installing heat exchanger in each product stream to preheat crude oil prior to feed the furnace through thermal exchange with hot product. Crude Distillation Unit of ERL has 10 numbers of heat exchangers to recover heat from the hot product streams as well as to preheat the feed stock. It is estimated that net heat recovered through the heat exchangers in preheating crude (ALC) oil is about 24.00 MMkcal. Preheating of crude oil by present set of exchanger raises the furnace inlet temperature to about 233 °C. For 1°C increase in furnace feed temperature would result in energy saving about 0.12 MMkcal/hr. Necessary steps are to be taken for further increment of furnace inlet temperature through heat exchangers.

After recovery of heat from the product stream by the present crude preheat train a reasonable amount of heat is rejected to cooling water during cooling hot products before storage in tanks It is observed that total heat rejected by hot products to cooling water is about 4.9868 MMkcal/hr. Heat rejected by the products HGO and Residue is about 3.49 MMkcal/hr (about 70%) Extra heat exchanger can be placed to recover further amount of heat. This amount of heat can be used for preheating crude oil, boiler feed water, desalter flush water, or as convenient to the operations.

"Pinch Analysis" is a modern concept introduced in heat exchange net work design. A study was made on present set of crude preheat train on the basis of pinch technology. It is observed that composite curve of combined hot and cold product streams shows no 'Pinch' that indicates that the present design and layout of exchangers in crude preheat train is satisfactory.

CHAPTER 9

Conclusion

9.1 Conclusions

Based on study described in the preceding chapter, the following conclusions can be drawn:

1. A comprehensive study was made on crude distillation furnace. It is estimated that the thermal efficiency of furnace is about 65 %, excess air supply to the furnace is about 77 % which is about 37% higher than the design value. 1-2 % improvement in thermal efficiency is possible by reducing only excess air to the furnace. 1 % percent improvement in furnace efficiency results energy saving around 3256 MMkcal/year. It is observed that, 33 % of the total heat generated in furnace is rejected to air through flue gas. Excess heat of flue gas can be recovered partly by installing air preheater.

2. Study of crude preheats train shows that recovery of heat during preheating the Murban crude is about 21.60 MMkcal/hr and ALC about 24.01 MMkcal/hr. The temperature of crude oil (ALC) rises from 23 °C to around 233 °C at inlet to the furnace. Performance of the preheat train was observed by rearranging the present position of the exchangers for further improvement in furnace inlet temperature. The design and lay out of the existing exchanger was checked by 'Pinch Technology' and found satisfactory.

3. A reasonable amount of heat energy can be saved by flashing crude oil prior feeding to furnace. A modification has been made in ERL crude preheat train to improve flashing of crude oil. It is observed that the modification gives energy saving per day in case of Murban crude about 7.66 MMkcal/M.Ton and in case of Arabian Light Crude about 2.59 MMkcal/M.Ton. Further improvement in flashing can be achieved by providing extra arrangement like large flashing area, separator to ensure vapor flow to the column, etc.

4. It is observed that about 4.9868 MMkcal/hr. of heat is rejected to the cooling water during cooling hot products before storage in tanks. Heat rejected by HGO and Residue product stream to cooling water is about 3 49 MMkcal/hr, (about 70%). Extra heat exchanger can be placed to recover further amount of heat. This amount of heat can be used for preheating crude oil, boiler feed water, desalter flush water, or as convenient to the operations.

9.2 Recommendations

1 Necessary arrangement is to be made to increase evaporation rate by flashing crude oil in flushing drum and flow of vapor is to be made at a higher level of the Distillation column instead of bottom level

2 Excess air supply to the furnace should be maintained at minimum level

3. A continuous monitoring system for oxygen in flue gas should be provided to monitor and control the excess air supply to the furnace.

4. Air preheater is to be installed to improve furnace efficiency.

5 Heat exchangers may be placed in between Exchanger E1107 and Water cooler E1114 in HGO product stream to recover heat about 0 16 MMkcal/hr,

6. Another heat exchanger can be placed in between Exchanger E1108 and Water cooler E1115 in residue product stream to recover heat about 2.87 MMkcal/hr.

7. Comprehensive study on other units of Eastern Refinery Ltd. should also be conducted.

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