ENERGY INTEGRATION OF HYDRODESULPHURIZATION PLANT USING PINCH ANALYSIS

(A CASE STUDY OF KADUNA REFINING AND PETROCHEMICAL COMPANY HDS PLANT)

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IN

JUNE, 2011

DECLARATION

I, **OLORUNFEMI DELE JOSEPH**, declare that this project is a product of my personal research work and has not been presented elsewhere for the award of any certificate.

Information derived from published and unpublished works have been dully acknowledged.

Afemiss
STUDENT SIGNATURE

14/10/2011
DATE

CERTIFICATION

This is to certify that this Project titled: (ENERGY INTEGRATION OF HYDRODESULPHURIZATION PLANT USING PINCH ANALYSIS (A CASE STUDY OF KADUNA REFINING AND PETROCHEMICAL COMPANY HDS PLANT) was carried out by OLORUNFEMI DELE JOSEPH (M.ENG/SEET/2007/2008/1590) meet the regulation governing the award of the Master of Engineering (Chemical Engineering) of Federal University of Technology, Minna and is approved for its scientific contribution to knowledge and literary presentation.

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DEDICATION

This work is dedicated to ***God Almighty The Preserver Of Life***

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ABSTRACT

Energy management is a critical activity in the developing as well as developed countries owing to constraint in the availability of primary energy resources and the increasing demand for energy from the industrial non industrial users. Energy consumption is a vital parameter that determines the economic growth of any country. Pinch technology is a vital tool to achieve this objective. Therefore energy Integration of Hydrodesulphorization Unit of Kaduna Refining and Petrochemicals Company was carried out using Pinch Technology. Optimum minimum approach temperature of 20^oC was obtained. The pinch point was found to be 523 K and the hot and cold pinch temperatures were found to be 247 and 231 ^oC respectively. The utilities targets for the minimum approach temperature were found to be 2,420.51 kW and 3366.86 kW for hot and cold utilities respectively. The utility and capital cost for optimum MTA of 10^oC are \$1.5 x 10^6 and \$0.22 x 10^6 respectively. The cold utility requirements of traditional energy approach and pinch analysis of HDS obtained are 15,144.62 kW and 3366.86 kW respectively while the hot utility requirements of traditional energy approach and pinch analysis are 15,332.667 kW and 2,420.51 kW respectively. Pinch analysis as an energy integration technique saves more energy and utilities cost than the traditional energy technique.

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A	Area	m^2
Q	Enthalpy	kW
H_{ST}	Hot Stream	
C_{ST}	Cold Stream	
T	Temperature	^oC
s	Source	
t	Target	
h	Film Transfer Coefficient	$kW/m^{2o}C$
ΔT_{MIN}	Minimum Temperature Approach	^oC
a	Area Efficiency	
ΔT	Temperature Difference	^oC
I	Interval	
T*	Shifted or Interval Temperate	^oC
ΔH	Interval Enthalpy Balance	kW
C_p	Specific Heat Capacity	$kJ/kg\,^oC$
C_P	Heat Capacity Flow rate	$kW/\,^oC$
M	Mass Flow rate	kg/s
ρ	Density	kg/m^3
VR	Volumetric Flow rate	m^3/s
ΔT_{LMTD}	Log Mean Temperature Difference	^oC
k	Watson Characterization Factor	

ΔT_{LMTD}	Log Mean Temperature Difference	^oC
k	Watson Characterization Factor	
G	Specific Gravity	
API	American Petroleum Institute	
CDU	Crude Distillation Unit	
KRPC	Kaduna Refinery and Petrochemicals Company	
HEN	Heat Exchangers Network	
ΔN	Number of heat exchangers	
ΔQ_H	Hot utility savings	kW
ΔQ_C	Cold utility savings	kW
ΔA	Additional area	m^2
MTA	Minimum Temperature Approach	

CHAPTER ONE

1.0. INTRODUCTION

Process Industries consume substantial quantum of energy in various forms like fuel, electricity, steam, thermal energy imported from other units, recycle heat from its own system design, waste heat etc. Energy consumption is specific and varies with the system design, technology selection, equipment efficiency, operational methods adopted etc. Every manager is interested in knowing the energy efficiency of each process so as to take appropriate and timely action to control the same. This information is vital to the executive as the operating costs are directly related to the energy efficiency of the system (Rajan, 2003). Then is therefore the need to incorporate energy management techniques such as pinch technology in process industries.

The use of Pinch Technology as Process Integration technique has been widely accepted in the process industries, especially in refineries and petrochemicals industries, to reduce energy cost. However in many situations, the implementation of energy cost reduction projects is faced with limited capital availability. Therefore, efforts to reduce energy costs have been limited to those that can be achieved with little or no capital investment (Linnhoff, 1994) .

Pinch technology is a complete methodology derived from simple scientific principles by which it is possible to design new plants with reduced energy and capital costs, as it can also be used where the existing processes require modification to improve performance as the case of Hydrodesulphurization unit (HDS) being considered in this work. An additional major advantage of the Pinch approach is that by simply analyzing the process data using its methodology. energy and other design targets can be predicted in a way that makes it possible to assess the consequences of a new design or a potential modification before embarking on actual implementation.

Pinch analysis originated in the petrochemical sector and is now being applied to solve a wide range of problems in mainstream chemical engineering. Wherever heating and cooling of process materials take place, there is a potential opportunity to adopt pinch technology as energy management tools. The technology, when applied with imagination, can affect reactor design, separator design and the overall process optimization in any plant. It has been applied to process problems that go far beyond energy conservation. Situation also resolved that pinch technology has been employed to solve problems as diverse as improving effluent quality, reducing emission, increasing product yield and debottlenecking, increasing throughput and improving the flexibility and safety of the process (Sahdev, 2002). Hence the proper utilization of pinch technology as a tool for energy management will reduce the problems of energy that are currently affecting the performance of process industries in Nigeria.

Since the ability of any nation to survive economically depends upon its ability to produce and manage sufficient supplies of low cost, safe energy and raw materials and the fact, the world consumption of limited fossil fuel resources currently increases annually by 3 percent. Projection in this trend shows that all known reserves will be exhausted in the next 50 years. Therefore, any sustained attempt to reduce rates of energy consumption even as little as 1 percent per annum ensures an effective eternal future supply as the world moves slowly toward renewable energy economies (Callagha, 1981).

As the demands on process industries to increase profitability and reduce emissions continue, many industries are focusing on improving energy efficiency to provide attractive solutions. It is ago when energy reduction projects required significantly better economics than yield projects, for example, to be considered for funding (Linnhoff et-al, 1982).

While oil prices continue to climb, energy conservation remains the prime concern for many process industries. The challenge every process engineer is faced with is to seek answers to questions related to their process energy patterns. Frequently asked questions are:

- Are the existing processes as energy efficient as they should be?
- How can projects be evaluated with respect to their energy requirements?

Determination of the energy requirement of an existing plant is very important as it helps to know whether that plant is saving or wasting energy. The answer to these questions is what this project intends to do by considering Process Integration of Hydrodesulphurization unit of Kaduna Refinery and Petrochemical using Pinch Technology.

Process integration using Pinch Technology offers a novel approach to generate targets for minimum energy consumption before heat recovery network design. The Pinch design can reveal opportunities to modify the core process to improve heat integration. Pinch Analysis is used to identify energy cost and heat exchanger network (HEN) capital cost targets for a process and recognize the pinch point. The procedure first predicts, ahead of design, the minimum requirements of external energy, network area, and the number of units for a given process at the pinch point. Next a heat exchanger network design that satisfies these targets is synthesized. Finally the network is optimized by comparing energy cost and the capital cost of the network so that the total annual cost is minimized. Thus, the prime objective of energy integration is to achieve financial savings by better process heat integration (maximizing process-to-process heat recovery and reducing the external utility loads). The traditional process design has quite often clear how improving energy efficiency can benefit in profitability by considering the cost of energy. Process industries energy cost vary greatly depending on the process industry configuration and location, but typically may range within hundreds of millions of Naira per annum (Callagha, 1981). Improving on the energy efficiency by only a few percent, has an obvious financial reward. Similarly as the majority of the cost of production is associated with energy, any percentage improvement in its efficiency automatically corresponds to an equivalent reduction in emission of CO_2, SO_2, NO_X caused by burning fuel(Brown, 1998). The emission of these compounds results to environmental pollution, which has to be combated.

Energy saving in the Nigerian industrial sector has several possibilities, due to the fact that, almost all the industrial equipment stock in Nigeria were imported during the era of cheap energy. Consequently, they are inherently energy inefficient. Furthermore, given the fact that energy prices had been kept at a low level up to 1985, energy cost has not been a significant fraction of total production cost even for energy intensive industry like refineries in Nigeria. The improvement of energy efficiency can provide substantial benefit in general to all the sectors of the economy (Dayo, 1994).

Despite this lack of capital, cost reduction still remains very important in the oil refining industry. Besides the cost of crude, energy is the largest cost which can be influenced by improved operation and capital investment, and has therefore become a primary focus. It has been shown from reported literature that due to the low technical risk associated with energy cost reduction projects, longer payback criteria are being accepted compared with projects which depend on outside market forces (Dayo, 1994). This is a significant shift from only a few years involved optimization of individual unit operations and tended to be largely dependent on the designer's intuition and experience as well. It is therefore not surprising that many processes hardly achieve , in practice the level of optimality envisage ed. However, with the increasing concern worldwide for the conservation of energy resources, preservation of the environment, coupled with the drive towards capital cost efficiency, it has much resulted in the need for all refineries around the world to improve their economic margins in order to increase their competitiveness. Using pinch technology to analyse important units of refinery could be an effective way to enhance its economic performance.

The Hydrodesulphurization unit of (Kaduna Refining and Petrochemical Company) is one of the most strategic unit for the production of Linear Alkylbenzene (LAB). Petroleum fractions contain various amounts of naturally occurring contaminants, the most important ones being organic sulphur, nitrogen, and metals compounds .These contaminants, if not removed, would increase the levels of air pollution and equipment corrosion, and in some cases would cause difficulties in the further processing of the material. The purpose of an Hydrodesulphurization Unit(HDS) process is to remove sulphur and nitrogen from the feed without greatly changing its boiling range ,therefore a catalytic hydrogenation method for upgrading the feed quality of the downstream in Molex Unit is employed. This upgrading is called hydro-treating. The process involves passing the feedstock over a fixed bed of Universal Oil Product (UOP) Hydrobon catalyst in the presence of high temperatures and pressures along with large amount of hydrogen in the two series arranged reactors. The high temperatures simply meant high energy cost for removing from raw kerosene the contaminants that are poisonous to the sieves in Molex Unit. The temperature and pressure required will depend on the nature of the feed as well as the amount of contaminant removal required. In view of the current high cost of energy, the production cost has moved up drastically which calls for careful study.

Over the years the trade-offs between energy and capital cost of process industries have changed drastically; it is, therefore, important to check the validity of this traditional configuration. Furthermore, the plant is now facing a lot of problems ranging from inadequate energy conservation and recovery, obsolete and corroded equipment, shortage of raw water, and low quality boiler feed water.

It is therefore crucial to carryout energy synthesis of the Hydrodesulphurization units so as to redesign the heat exchanger network of the units using Pinch technology design method. This will show areas in which the process can be improved and solve the above mentioned problems, together with reduction in the environmental pollution (flue gas emission and waste water discharge) and raw water consumption of the units (Akande, 2007).

1.1 Aim and Objectives

The aim of this research work is to develop an energy integration of Hydrodesulphurization plant of Kaduna Refinery and Petrochemical Company (Kaduna Refining and Petrochemical Company).

This aim will be actualized through the realization of the following objectives:

1. Collection of Design data, Operating Data and Piping and Instrumentation Diagram of Hydrodesulphurization (HDS) Unit of Kaduna Refining and Petrochemical Company (KRPC).
2. Identification of Hot, Cold, and Utility Streams in (HDS).
3. Thermal Data Extraction for HDS Process and Utility Streams

4. Selection of Initial ΔT_{min} Value
5. Construction of Composite Curves and Grand Composite Curve.
6. Estimation of Minimum Energy Cost Targets
7. Estimation of Heat Exchanger Network Capital Cost Targets
8. Estimation of Optimum ΔT_{min} Value
9. Estimation of Economic Trade-off between Operating Costs and Capital Cost
10. Optimization of energy by calculating the net present cost total for utilities and capital over a range of dT_{min} values and obtaining the optimum
11. Preparation of ΔT_{min} Optimization Plot, Capital Cost Plot and Utilities Cost Plot.

1.2 Scope of the Work

Pinch technology has now found application in different areas, such as water pinch, hydrogen pinch, and energy pinch. But its use for energy conservation purpose remains the most attractive, hence the technology has being applied to process unit (Akande, 2008). In this work, design of heat exchanger network for the Hydrodesulphurization Unit of Kaduna Refining and Petrochemicals Company will be analyzed. This is due to the energy intensive nature of the unit and the positive contribution of the product produced from the unit (LAB) to the development of the Nigerian economy.

1.3 Justification

Considering the negative effects of escalation in prices of oil, regular shut down of Hydrodesulphurization Unit (HDS) and utilities unit of our refineries and petrochemical Companies (due to poor management of material and energy resources), Federal Government has embarked on privatization of the downstream sector of its oil industries. Government policy has shifted from establishing Refineries and Petrochemical Companies for providing energy and petrochemicals at a subsidized rate for its populace to a full-fledged profit making companies. Hence energy saving is a pivot to achieving this goal. The prime objective of this project, which is to achieve financial savings by better process heat integration (maximizing process-to-process heat recovery and reducing the external utility loads) for our petrochemical company, is therefore justified.

The completion and implementation of the research will be of benefit to the Nigerian National Petroleum Corporation and other Nigerian process industries in the following ways:

a. It will provide an energy auditing methodology for the HDS plant, which can be applied in monitoring energy usage and management of the unit.
b. It will establish the principles for process modification i.e. retrofitting.
c. It will minimize the cost of energy supply in petroleum refining by efficient fuel consumption, hence energy savings and optimization.
d. It will reduce to a minimum the environmental pollution as the quantity of energy generating these pollutants from combustion product is reduced, if the utility usage (prime objective of this project) is conserved.

CHAPTER TWO

2.0 LITERATURE SURVEY

This chapter presents the fundamental concepts upon which pinch technology is based. These concepts evolved over the years as a result of research work carried out by Linnhoff and other researchers. (Bassey, 1995) It also presents a review of the research work done in the development of pinch technology with particular focus on heat exchanger network design and its retrofit application.

The methodology for accomplishing retrofit application of pinch technology is quite different from that used for grassroots design. This is because of the inherent constraints of an already existing process and the need not to dramatically alter the existing structure of the heat exchanger network.

This review is presented under two subsections. The first subsection is intended to review the general progress that has been made over the years. The second subsection presents the progress made in terms of heat exchanger network design, and retrofitting.

2.1 General Survey on Pinch Technology

The novel work of Linnhoff (1978) led to the evolution of pinch technology. Since then several attempts have been made by subsequent researchers to further understand and develop the concepts upon which pinch technology is presently based while the horizon gets further widened as it finds application in various process industries.

Sequel to his discovery of the existence of pinch point in heat exchanger network, Linnhoff et-al. (1979) proceeded to give further insight into the understanding of this pinch analysis. Linnhoff and Turner (1980) illustrated how simple concepts like problem table calculation and composite curve can give energy savings and elegant designs. This was further updated by the duo of Linnhoff and Turner in 1981. Townsend, et.al. (1982) attempted designing total energy systems by systematic methods using pinch. Linnhoff and Hindmarsh (1983) developed a systematic method of designing heat exchanger network using pinch. The first attempt to predict the surface area requirement of a heat exchanger network was made by Townsend and Linnhoff in 1984. Their methodology subsists till date despite its associated flaws. Kotjabasakis and Linnhoff (1987) showed how the cost of heat exchanger fouling can be reduced through better design using pinch technology method .Results of their findings made it possible that minimum number of heat exchangers required in a network can be determined prior to actual design by the postulation of Ahmed and Smith (1989). With this development, the full methodology had evolved for pinch technology with the philosophy of predicting energy and area requirements of a process prior to the actual design. Thus, Linnhoff and Vredeveld (1984) summarized this in their paper.

In a similar manner the first attempt to carry out retrofit was made by Linnhoff and Parker, (1984) when they studied process modifications with the heat exchanger network. In view of the difficulties encountered in this first attempt, Linnhoff and Tjoe (1986) evolved a detailed methodology for accomplishing process retrofit using pinch technology which takes cognizance of the specifics of an already existing process. Fraser and Gillespie (1992) applied Pinch technology to retrofit an entire oil refinery with the view to save energy. Shokoya (1992) also carried out retrofit of heat exchanger networks for debottlenecking and energy saving and found that it could be inferred from the available literature that the methodology devised from the preceding research can not handle heat exchanger Network problems below ambient temperatures because of their special feature, (Wu Shokoya, 1992). A methodology to determine the feasible region for shell and tube heat exchanger designs on a pressure drop diagram has been proposed by Shenoy et.al. (2000); their proposal does not address the gap between targeting and detailed design directly. It is also difficult to implement it on a network where multiple streams and multiple heat exchangers are involved. Liebmann and Dhole (1995) presented a systematic and integrated approach to the design of energy efficient crude distillation systems. Integrated implies simultaneous consideration of the options in the distillation system and the heat exchanger network. The proposed procedure is applicable for grass-roots as well as revamps situations. The approach is based on a combination of insights in distillation and pinch analysis. Zhu (1995) developed an algorithm for automated synthesis of heat exchanger network which is based on the block concept with the purpose of simplifying a design problem by decomposing it into a number of blocks, Usman(2002) reported an improved models for area prediction based on the allowable pressure drop fir the streams and a sensitivity analysis to take into account the flexibility, operability, and contrl bility requirement of the network in the retrofit procedures.

2.2 Pinch Technology

The ability of most processes to c me energy at one temperature level and reject it at another level are aided with available ut . Energy is provided to a process using such utilities as steam, hot water, flue gas etc. i ected to cooling water, air, refrigerant or in heat recovery steam rising. Heat recovery i o reduce the utility cost of a process. Evaluation of heat recovery involves a balancin ty against the capital cost of the heat recovery system. The utility cost not only depend amount of energy consumed and rejected but on the utility actually used .Cooling w per than refrigerant, low pressure steam is cheaper than high pressure steam (Adefi onsidering the importance of pinch technology on the evaluation of energy utilization by the process industries ,there is the need to report the working principle of the technology.

2.2.1 Principle of Pinch technology

Pinch analysis is a rigorous, structured approach that can be used on a wide range of process and site utility related problems. Such as lowering operating costs, de-bottlenecking processes, raising efficiency and reducing capital investment.

The majority of processes consist of streams that need to be heated up and streams that need to be cooled down. For each stream that requires heating or cooling, there are two basic choices. The heat can either be exchanged between two process streams or it can be exchanged between the process and the utility system. A fundamental strength of pinch analysis is that it determines the most appropriate set of heat exchange matches. In doing so, it reduces the cost of hot and cold utilities by minimizing the cascade of heat from the expensive, high temperature region down to ambient and also from ambient down to expensive, sub ambient temperatures.

The power of pinch technology lies in two factors:

1. Its ability to quickly evaluate the economics of heat recovery for a given process.
2. The guidance it provides regarding how a process can be modified in order to reduce associated utility needs and costs.

It is these two factors that attract the use of pinch technology to analyze and design the heat exchanger network of any system.

Here, only the source and target temperature, heat capacity and mass flow rates of the process streams are required to carry out the analysis and it works on certain established principles or concepts such as Problem Table Calculation, Composite Curve, Grand Composite Curve, Super Target, Grid representation etc. This chapter presents these underlying principles. However, in order to demonstrate the basic principles for easy understanding of the aforementioned concepts and to facilitate better appreciation of their significance we can consider a simple process that has two streams that require cooling and two that require heating. The streams are considered in terms of their start and target temperature along with an associated heating and cooling duty. The data set for this process is given in Table 1. Note that the last column (CP) is given as the product of mass flow rate time's heat capacity.

Table 2.1: **Streams thermal properties.**

Stream	Stream type	T_s °C	T_t °C	Duty kW	C_P kW°C^{-1}
1	Hot	180	80	2000	20
2	Hot	130	40	3600	40
3	Cold	60	100	3200	80
4	Cold	30	120	3240	36

2.2.1.1 Scenario I: Heat recovery

In this design all heat requirements are met by external source of energy (utilities). Hot utilities such as steam or flue gas is used to heat up cold streams 3 and 4 from source to their target temperature, while cold utilities such as cooling water is used to cool the hot streams 1and 2 from their source temperature to target temperature. This means that a total of 5600kW hot utility and a total of 6440kW cold utility are supplied. This leads to gross wastage and poor engineering design, since the energy released by hot streams could have been used to provide the need of the cold streams.

2.2.1.2 Scenario II: First law analysis

Consider using the hot streams to heat up the cold streams. Going by the first law implication this is a possibility. Thus if hot and cold streams were matched for energy recovery, only a net deficit of 800kW is to be met by hot utilities. However, this is not realistic due to energy flow restriction, a hot stream at a lower temperature (low quality energy) can not be used to heat up a cold stream at a higher temperature; this is a thermodynamic infeasible.

2.2.1.3 Scenario III: Second law analysis

The second law of thermodynamics places restriction on the direction of heat interchange. Energy can only be transferred from a region of higher temperature (high quality, source) to a region of lower temperature (lower quality, sink). Hence, match is only possible between a hot and cold stream if the latter is at a lower temperature; of course there can be several matches depending on the number of streams in the process. The design here depends on how to match the streams in order to attain maximum possible energy recovery and minimum use of external utilities. This leads to the concept of problem table.

2.2.2 The methodology for carrying out pinch

This takes into account the restriction placed by second law of thermodynamics in heat interchange. The second law dictates that there must exists a finite temperature difference between the hot and the cold streams for energy interchange to be feasible.

This consideration was incorporated into the energy integration analysis by Linnhoff and Flower (1978) following the pioneering work of Hofmann (1971). This entails establishing temperature intervals based on a pre specified minimum temperature approach (MTA). The minimum temperature approach use in the following example is $10^{o}C$

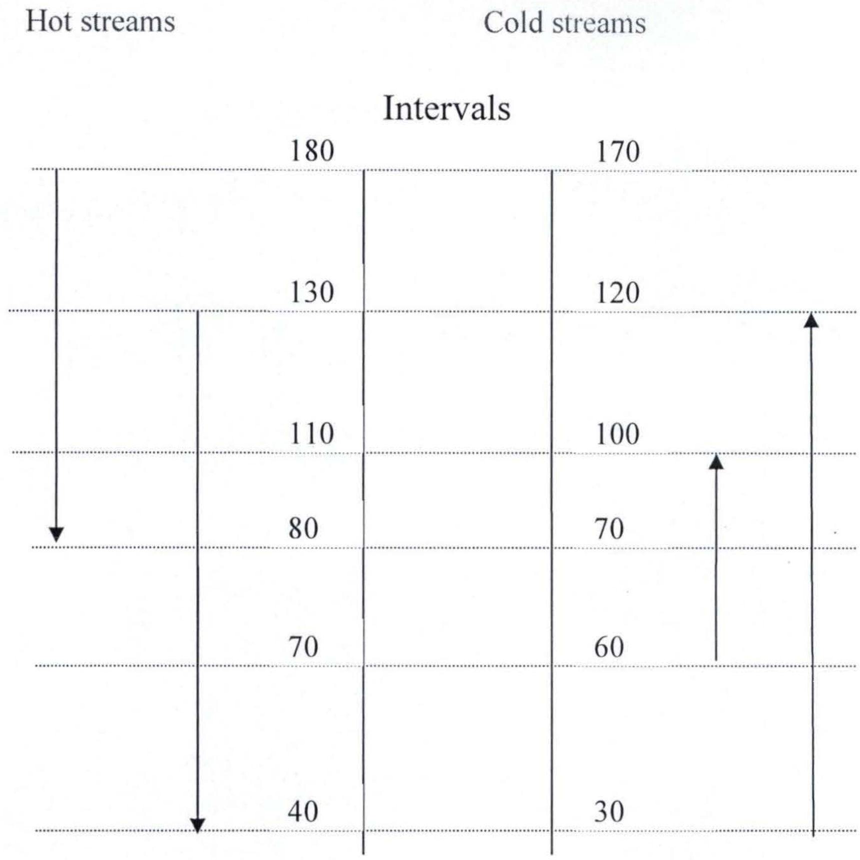


Figure 2.1: Temperature Intervals.

$$Q(I) = [\Sigma CP_{coldi} - \Sigma CP_{hoti}].\Delta T_i \quad \ldots\ldots 2.1$$

Thus, for the five intervals we get:

Q (1) = (0 – 20) (180 – 130) = – 1000kW (1 hot, 0 cold)

Q (2) = (36 – (20 + 40)) (130 – 110) = – 480kW (2 hot, 1 cold)

Q (3) = ((80 + 36) – (20 + 40)) (110 – 80) = 1680kW (2 hot, 2 cold)

Q (4) = ((80 + 36) – 40) (80 – 70) = 760kW (1 hot, 2 cold)

Q (5) = (36 – 40) (70 – 40) = – 120kW (1hot, 1 cold)

This result can be represented in the Figure 2.2

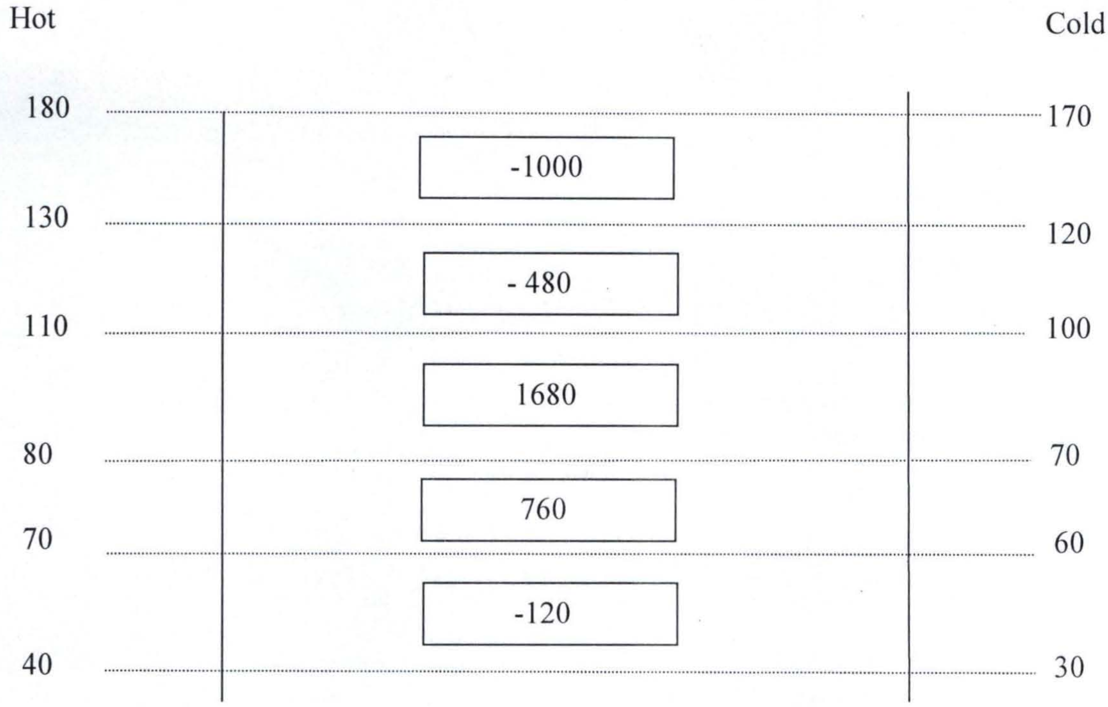


Figure 2.2: Surplus and Deficit Heat Energy in Intervals.

Heat can be transferred from any of the hot streams in the higher temperature intervals to any of the cold stream at the lower temperature intervals. Therefore, the surplus energy at higher intervals can be cascaded down, as shown in the Figures 2.3a and 2.3b. Figure 2.3a is cascaded with zero utility while Figure 2.3b is cascaded with the highest negative heat energy value in the intervals of Figure 2.3a.

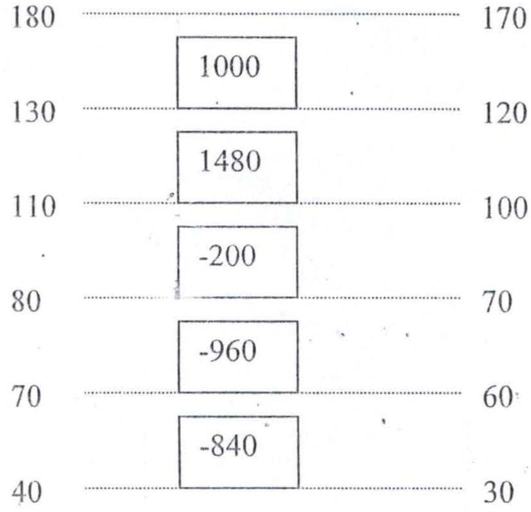


Figure 2.3 a: Cascade without utility.

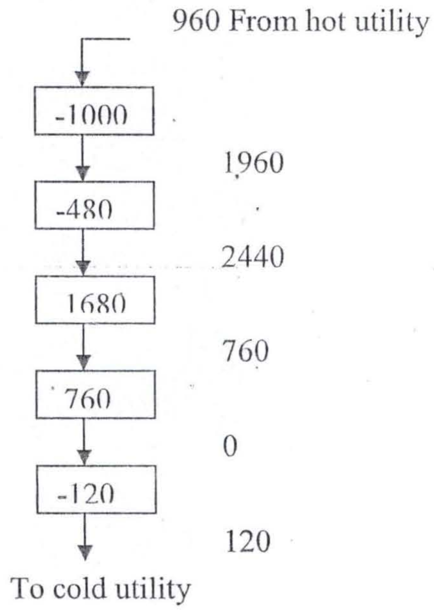


Figure 2.3 b: Cascade with utility.

The deficit of 960kW is still left in the fourth interval after this cascade. This is the maximum energy recovery possible for this system. This deficit would have to be supplied by hot utility as shown in Figure 2.3b. This leaves us with no deficit in any of the intervals. The cascade from the fourth to the fifth interval is zero. This represent the pinch and it implies that, the hot pinch temperature is 70^0C, while the cold pinch temperature is 60^0C. The surplus energy in the fifth interval 120kW would be rejected to a cold utility. Thus, the pinch temperature provides a decomposition of the design problem; therefore, above the pinch we only supply heat from hot utility while below the pinch we can only reject heat to a cold utility. The minimum heat requirement of the process is 960kW while the minimum cooling requirement is 120kW. This is the problem table calculation.

The problem table above can also be calculated using a single temperature scale incorporating both hot and cold temperature interval. The procedure is highlighted below.

1. Convert the actual streams temperature Tact into interval temperature Tint by subtracting half the minimum temperature approach (1/2 MTA) from the hot streams temperature and by adding half the minimum temperature approach (1/2 MTA) to the cold stream temperatures.

   Hot stream $T_{int} = T_{act} - 0.5\ \Delta T_{min}$

   Cold stream $T_{int} = T_{act} + 0.5\ \Delta T_{min}$

2. These interval temperature (Tint) are then rank in order of magnitude showing the duplicate temperatures only once in the order, as shown in Table 2.2 (columns 1and 2)

3. The other steps are simply repeated as already described above.

   Note that the minimum temperature approach (ΔT_{min}) is 10^oC

Table 2.2: Problem table calculation.

I	$I_{TEMP.}$ ^0C	ΔT_C ^0C	$[\sum CP_{coldi} - \sum CP_{hoti}]$ $kW/^0C$	ΔH kW	Cascade I	Cascade II
	175				0	960
1	125	50	– 20	–1000	1000	1960
2	105	20	– 24	– 480	1480	2440
3	75	30	56	1680	–200	760
4	65	10	76	760	–960*	0
5	35	30	– 4	–120	–840	120

*Highest Negative

2.2.2.1 Composite curve

Composite curve is a plot of temperature against enthalpy for hot and cold streams. It represents the sum of the energy changes for a given temperature range.

The basis of all pinch analysis is the set of composite curves, which can be drawn for any process to represent all the heating and cooling duties in that process. The composite curve allow the designer to calculate hot and cold utility targets ahead of design, to understand driving forces for heat transfer and locate the heat recovery pinch. The degree to which the curves overlap is the measure of the potential for heat recovery.

The utility target depends on the value of ΔT_{min}. For instance a small ΔT_{min} bring the curves closer together, reducing hot and cold utility demands and given lower operating costs. This is at the expense of large heat exchanger area and hence greater capital cost. The optimum choice of ΔT_{min} depends on the trade-off between capital and energy. The optimum temperature difference must be calculated for every application or system, but usual values are 10 ^{o}C for liquids and 30^{o}C for gases.

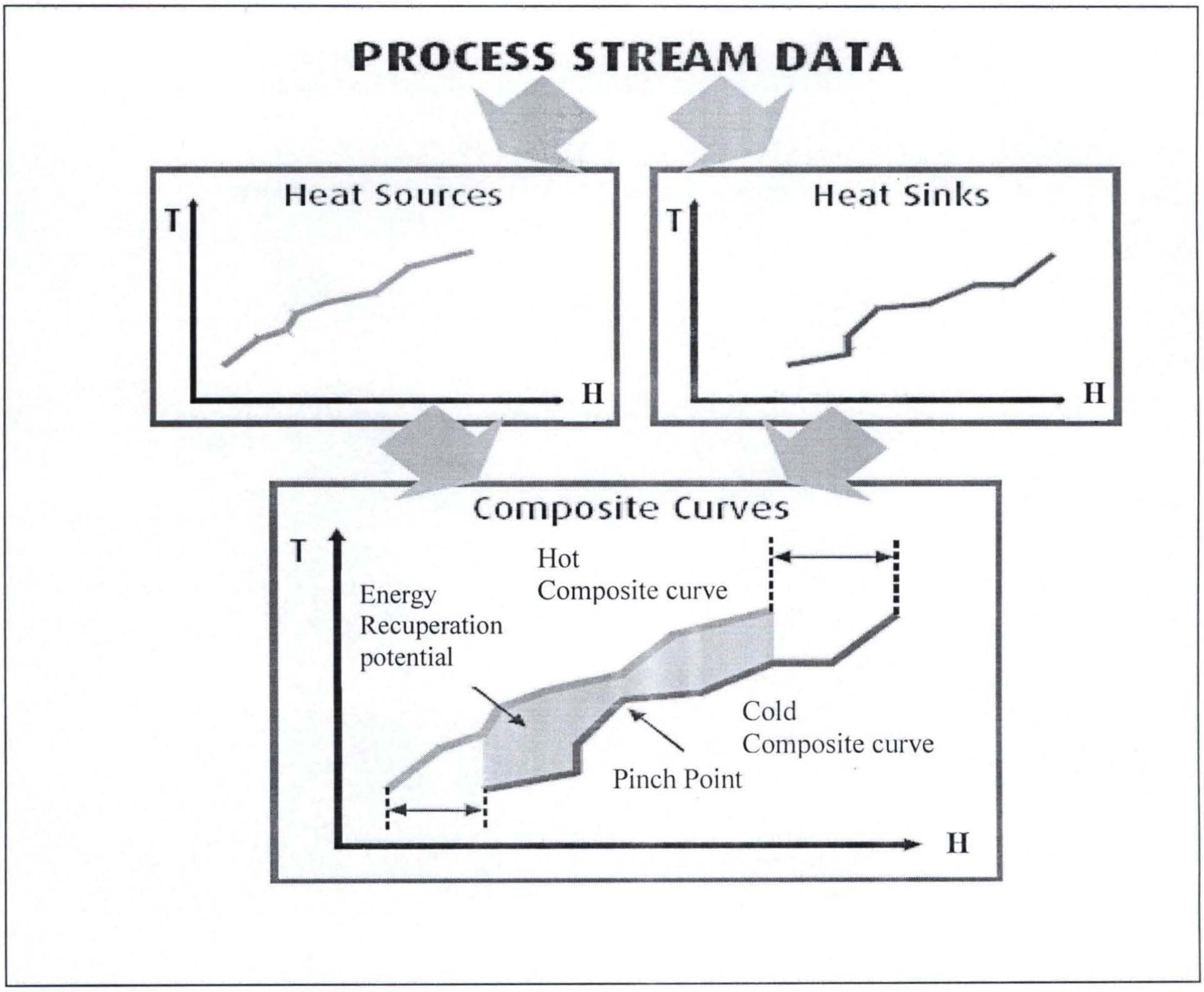


Figure 2.4: Composite Curves

To demonstrate the principle, which considers the process shown in figure 2.4 with two streams that require cooling and two that require heating, the streams are considered in terms of their start (source) and target temperatures along with an associated heating or cooling duty. The data set for this example process is given in Table 2.1 Note that the last column (CP) is defined as the product of mass flowrate time's heat capacity.

For example, stream 1 is cooled from 180^0C to 80^0C, releasing 2000 kW of heat, and so has a CP of 20 kW/^{0}C.

This information is now translated into the composite curve presentation. The hot composite curve in Figure 2.4 is constructed by adding the enthalpy changes of the hot streams in the respective temperature interval. In the temperature interval 180^0C to 130^0C only stream 1 is present. Therefore, the CP of the composite curve equals 20kW/^{0}C, the CP of stream 1, in the temperature interval 130^0C to 80^0C both stream 1 and 2 are present. The CP of the hot composite here is therefore 60kW/^{0}C, the sum of the CPs of the two streams. In the temperature interval 80^0C to 40^0C only stream 2 is present, so the CP for the composite is 40kW/^{0}C, and the slope of the lines is given as the reciprocal of their CP.

The construction of the cold composite curve is analogous to that of the hot composite curve, combining the T-H curves for the cold streams. To determine the minimum energy target for the process, the hot and cold composite curves are plotted on a single diagram as shown in Figure 2.4. The closest vertical separation between the two curves is defined as the minimum allowable temperature difference ΔT_{mi} In practical terms, ΔT_{min} represents the closest permitted temperature approach in a process heat exchanger. This value is normally chosen based on economic considerations and experience of the process involved. A value of 10^0C has been used for this example. The overlap between the composite curves shows the maximum process heat recovery possible as illustrated in figure

The point of closest approach of the composite curves, where ΔT_{min} is reached is known as the "Pinch". Recognizing the implication of the pinch being the point at which energy targets is realized in practice.

Composite curve has a role in pinch technology, because from it, it is possible to obtain abundant information such as;

- The minimum temperature difference that is normally observed at only one point between the hot and the cold composite curves, called the heat recovery pinch. This point has a special significance because it is the point at which there are more restrictions in the design of the heat exchanger network. In our example, the hot pinch temperature is 70°C, and the cold pinch temperature is 60°C obviously, the difference between these temperatures is the minimum temperature difference, 10°C.
- As occurred with single - stream composite curves, the horizontal overlapping of the curves is the maximum amount of heat that can be recovered. The enthalpy intervals not overlapped on the left and on the right are the enthalpy requirements that cannot be fulfilled with process streams, and thus they are the minimum requirements of cooling and heating utilities, these requirements are 120kW and 960kW respectively.
- The required utilities depend on the minimum temperature difference: if this difference is increased, the overlapping enthalpy interval shrink, and it will be necessary to spend more utilities. But as the temperature driving force increases, heat exchangers will be

smaller. It is necessary to calculate an optimum temperature difference that gives a trade off between energy consumption and capital cost.

- It must be emphasized that the minimum energy consumption calculated with composite curves is not an ideal minimum energy, but a minimum energy that actually brought into a design. To calculate this minimum energy, we have to established a minimum temperature difference bigger than zero, and thus we can build a network with real heat exchangers (that is, exchangers that do not have an infinite area of exchange), which reach these minimum energy requirements.
- An important characteristic of this method is that it enables the users to calculate the minimum energy requirements without a full design of the topology of the heat exchanger network. Later it will be seen how the minimum number of units and its area can also be calculated without a full design. This characteristic makes pinch technology a very useful tool for comparing several design alternatives without a big calculation effort.

The process can be considered as two separate systems. One above and the other below the pinch. The system above the pinch requires a heat input and is therefore a net heat sink. Below the pinch the, system rejects heat and so is a net heat source.

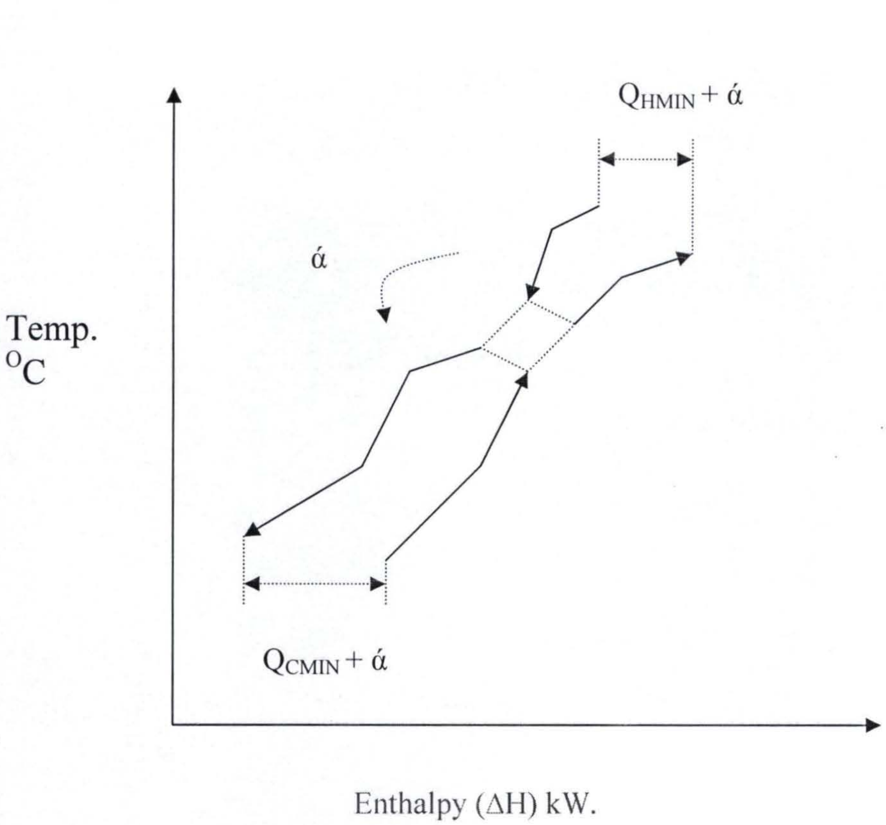


Figure 2.5: Intersystem Heat Transfer.

There are three golden rules listed below for achieving the minimum energy targets for a process, and violating any of these will result in an energy requirement that is greater than the target, as demonstrated in figure 2.5.

Rule 1: Heat must not be transferred across the pinch.

Rule 2: There must be no cold utility used above the pinch.

Rule 3: There must be no hot utility used below the pinch.

The story of energy pinch analysis continues through techniques use in optimization of multiple utilities, principles for process modifications, heat exchanger network design method and the development of total site analysis.

2.2.2.2 Grand composite curve

The Grand composite curve (GCC) shown in figure 2.6 which is derived from the same process data as the composite curve shows the net heat flow through the process. It highlights the process/utility interface and guides in the selection of different utilities sources and sink demonstrated earlier.

The key to process change is to split the GCC, taking a sub system such as unit operation, process or utility and plotting it on the same graph as the reaming 'background process' GCC. By applying appropriate placement concept it is possible to study how the sub- system relates to the rest of the process.

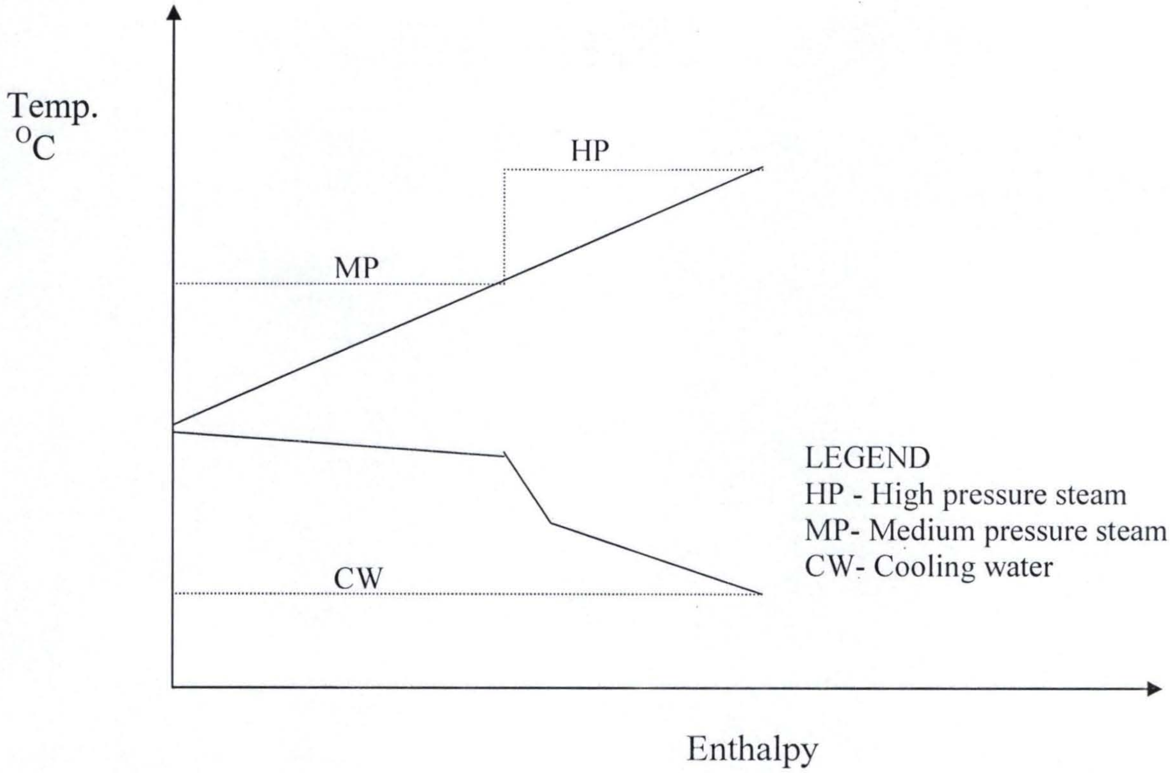


Figure 2.6: Grand Composite Curve

2.2.2.3 Surface area target

The heat recovery brought about by the technology can only be accomplished in heat exchanger network; therefore, a précised estimate of heat exchanger network area is necessary in order to appreciate the effect of minimum approach temperature (pinch) on it. However, to be in a position to accurately predict the optimum minimum approach temperature the procedure for predicting network area and the heuristic for determining the number of heat exchangers required in a network has to be adhesively followed. Since the surface area of the heat exchanger in a network depend on the specified minimum temperature difference ΔT_{min} just like energy. Thus it is possible for the network area to be determined with the relevant properties of the streams involved in the network prior to the actual detailed design of the heat exchangers.

To derive the equation for overall heat transfer coefficient, consider the composite curve in Figure 2.7 where the intervals are indicated by the change in slope of the curves. The duty for such exchanger and the values of the temperature driving forces at the end of each exchanger. Assuming the cooling and heating curves correspond to a single stream each, the overall heat transfer coefficient (u) can be estimated as

$$\frac{1}{U} = \frac{1}{hi} + \frac{1}{ho} \quad \text{..........} 2.2$$

Where; the individual film coefficient includes the fouling factor.

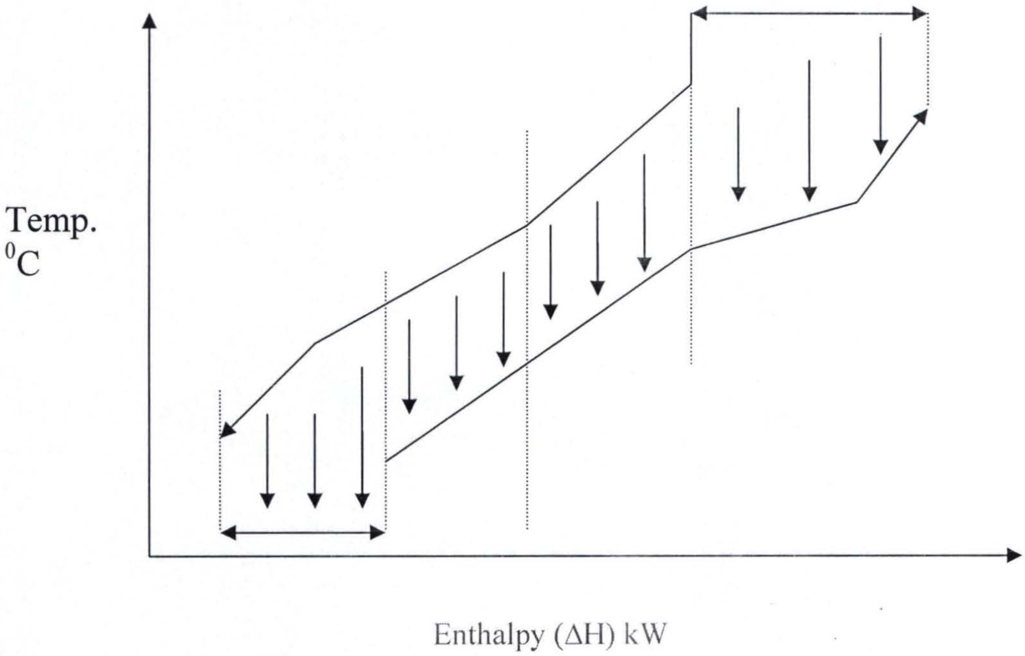


Figure 2.7: Vertical match up.

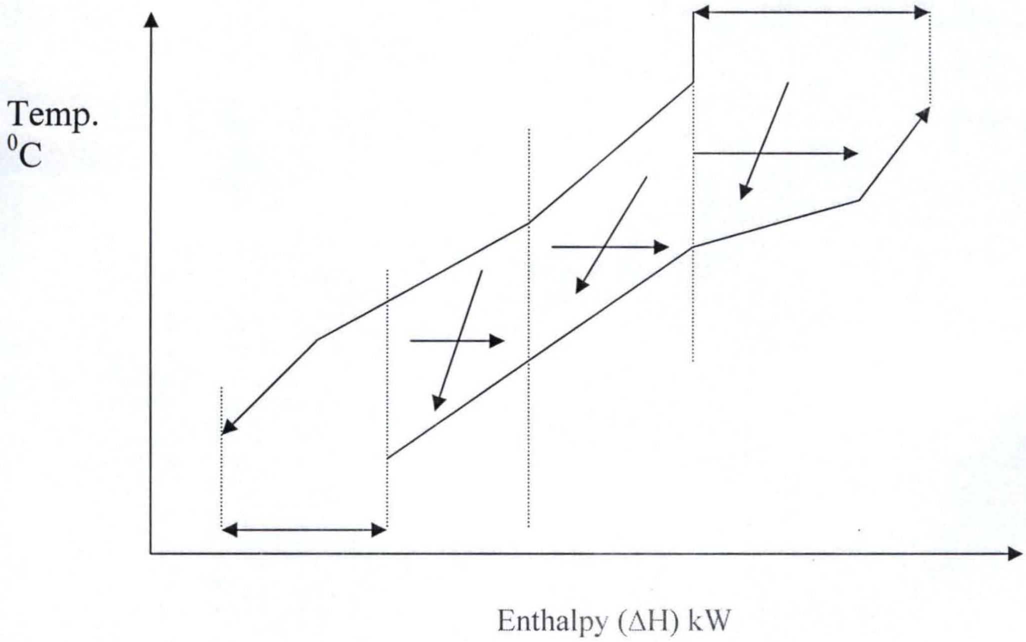


Figure 2.8: Criss-Cross Match Up.

The area of the heat exchanger is given by:

$$A = Q / U.\Delta T_{LM} \quad \text{......2.3}$$

Consider an interval where two hot streams (1,2) are matched against two cold streams (3,4), if stream 1 is matched with stream 3 and stream 2 with stream 4, we have two exchangers E1 and E2 respectively. The heat loads and the logarithmic mean temperature driving force for each of the heat exchanger will be the same. Their respective overall coefficients are:

Exchanger E1

$$1 / U_{E1} = 1 / h_1 + 1 / h_3 \quad \text{......2.4}$$

Exchanger E2

$$1 / U_{E2} = 1 / h_2 + 1 / h_4 \quad \text{......2.5}$$

The total area of the two exchangers is given as:

$$A_T = A_{E1} + A_{E2} = [Q / \Delta T_{LM} + Q / \Delta T_{LM}] \ldots\ldots 2.6$$

$$A_T = Q / \Delta T_{LM}.[1 / h_1 + 1 / h_2 + 1 / h_3 + 1 / h_4] \ldots\ldots 2.7$$

If on the other hand the stream matching has been on the other way round i.e. stream 1 with stream 4 and stream 2 with stream 3 an identical result is obtained.

Exchanger E3

$$1 / U_{E3} = 1 / h_2 + 1 / h_3 \ldots\ldots 2.8$$

Exchanger E4

$$1 / U_{E4} = 1 / h_1 + 1 / h_4 \ldots\ldots 2.9$$

$$A_T = A_{E3} + A_{E4} = Q / \Delta T_{LM}.[1/ U_{E3} + 1/ U_{E4}] \ldots\ldots 2.10$$

$$A_T = Q/\Delta T_{LM}.[1/ h_2 + 1/ h_3 + 1/ h_1 + 1/ h_4] \ldots\ldots 2.11$$

Hence this result can be generalized for any number of streams hot or cold in an interval.

The area in any interval J is given by

$$A_J = Q_J/\Delta T_{LM}.[\sum 1/h_{cold} + \sum 1/ h_{hot}] \ldots\ldots 2.12$$

Summing this expression over the entire intervals gives the area requirement for the network as follows:

$$A_{Target} = \sum A_J = \sum 1/\Delta T_{LM,J}.[\sum Q_J/h_{cold} + \sum Q_J/h_{hot}] \ldots 2.13$$

$$= \sum 1 / \Delta T_{LM, J}.[\sum Q_J / h_K] \ldots\ldots 2.14$$

Where:

K – The stream number, having values from 1 to stream K

Whether hot or cold.

Q_J – Heat load on stream K interval J (same for all streams in the interval)

ΔT_{LM} – Logarithmic temperature difference in interval J

The area target given by Equation 2.14 is the minimum possible for the network and it is true if the process streams in a network interchange heat in such a way that the streams matches are "Vertical" between the composite curves. This arrangement is equivalent for pure counter current flow in a single pass Heat Exchanger.

Any match away from the vertical as in Figure 2.8 will gain the local advantage of a large ΔT. This is referred to as Criss-Crossing and its net effect is an increase in the area requirement.

If Heat Exchangers with more than a shell pass or a tube pass are used, the minimum temperature difference should be corrected with effectiveness factor:

$$\Delta T_{LM} = \Delta T_{LM} . F_{TK} \quad \text{.......} 2.15$$

F_{TK} - range between (0.75 – 0.90)

This method will yield rather accurate estimations of area of the network, with maximum errors of about 10%. Usually, the real area will be greater than the area target, because in the final design, due to safety or operatability consideration, it will be necessary to install additional Heat Exchanger.

2.2.2.4 Cost target

The cost of a Heat Exchanger can be calculated as a function of its area of exchange by a cost correlation such as:

$$Cost = a + b.A^{C} \quad \text{.......} 2.16$$

This correlation usually yield good estimate of heat exchanger cost, because Heat Exchanger construction is much standardized.

The cost of Heat Exchanger Network can be calculated from the area target and the minimum number of Heat Exchanger assuming that every Heat Exchanger in the network has an area given as:

$$A_E = [A_{MIN} / N_{MIN}] \quad \text{.......................} 2.17$$

$$C_{MIN} = N_{MIN} [a + b [A_{MIN} / N_{MIN}]^c] \quad \text{.......................} 2.18$$

Despite the apparent simplicity of these calculations, it yields surprisingly accurate estimate of the Heat Exchanger Network cost.

Equation 2.18 assumes that every heat exchanger in the network follow the same cost law. When this is not the case, costs are calculated by assigning a fictitious area to these exchangers that follow a different cost law. This area can be calculated by multiplying the real area of exchanger by this correlation factor, which is a function of the cost correlation.

$$A_{MIN} = \sum 1 / \Delta T_{LM,K} \,.\, \sum q_i / h_i.F_i \quad \text{.......................} 2.19$$

Where:

$$F_{ESP} = [b_{REF} / b_{ESP}]^{1/C_{REF}}.[A_{MIN}/N_{MIN}]^{1-C_{ESP}/C_{REF}} \quad \text{.......................} 2.20$$

In Equation 2.20 the suffix REF refers to the cost parameter chosen as the references (which are the parameter that will be used for the calculation of C_{MIN} and suffix ESP refers to the cost parameter which differs from the reference parameter. The commonest type of Heat Exchanger should be chosen as the reference in order to get better accuracy.

2.2.2.5 Optimum temperature approach

The importance of the minimum temperature approach, ΔT_{min} was emphasized clearly as $\Delta T_{min} \rightarrow 0$ the true pinch is approached at which the area for heat transfer approaches infinity, while the minimum utility requirements are reduced to the absolute minimum. At the other extreme, as

$\Delta T_{min} \rightarrow \infty$, the heat transfer area approaches zero and the utility requirements are increased to the maximum, with no Heat Exchange between the hot and cold streams. The variation in heat transfer area and utility requirement with ΔT_{min} translate into variation in capital and operating cost as shown schematically in Figure 2.9 As ΔT_{min} increases, the capital cost is reduced toward zero as the heat transfer area diminished. Similarly, as ΔT_{min} decreased from large values, the cost of utilities decreases linearly until a threshold temperature difference ΔT_{thres} is reached, below which the cost of utilities is not reduced. Furthermore, when $\Delta T_{min} < \Delta T_{thres}$, there is no pinch and consequently, the trade offs between the capital and utility cost as ΔT_{min} varies are not applicable.

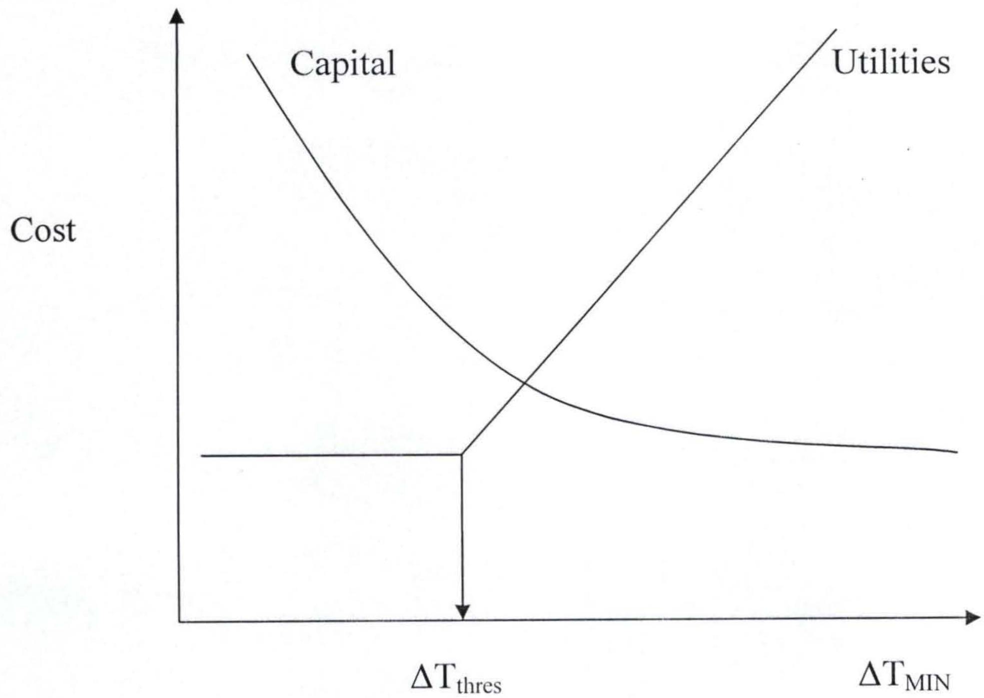


Figure 2.9: Trade-off between Capital and Utilities

Cost as a Function of ΔT_{MIN}

2.2.2.6 Minimum number of heat exchanger

Having designed a Heat Exchanger Network that operates with the minimum hot and cold utilities using either the method of Linnhoff and Hindmarsh (1983) or the MILP of Papoulis and Grossman (1983) it is common to reduce the number of that exchangers towards the minimum while permitting the consumption of utilities to rise, particularly when small heat exchangers can be eliminated in this way, lower annualized cost may be obtained, especially when the cost of fuel is low relative to the purchase cost of the heat exchangers. Before proceeding, it is important to recognize that, as pointed out by Hohmann, (1971) under most circumstances the minimum number of heat exchangers in a Heat Exchanger Network is given by Equation 2.21

In order to facilitate capital cost estimation prior to detailed design; the minimum number of heat exchangers required for a process must be known, in addition to the total surface area. This can be evaluated by applying the Euler terrain. It states that the minimum number of connections (Nmin) required in a network is one less than the number of streams, N, including the utilities (Spinoff, 1994). Thus:

(Number of Exchangers) = (Number of streams) + (Number of utilities) – 1

$$N_{min} = N_S + N_U - 1 \quad \text{.......... 2.21}$$

Where:-

N_S- is the number of streams

N_U- is the total number of hot and cold utilities.

In order to incorporate the second law of thermodynamics requirement i.e. pinch point into the calculation of the minimum number of Exchangers, we must use Equation 2.21 separately for the sub system above and below the pinch. The expression is however, not true for a system with loops. Each loop introduces an extra exchanger.

2.2.2.7 Pinch and thermodynamics

Thermodynamics is the study of the interconversion of heat, work and other forms of energy. The laws of thermodynamics provide the limits within which such energy transformation can take place. The existence of pinch in a Heat Exchanger Network (HEN) is in agreement with the finite temperature difference requirement of the second law. However, the most salient characteristic of pinch is that it constitutes heat recovery bottleneck. This can be explained using the thermodynamic concept of equitability of energy which is an offshoot of the second law.

Accompanying any heat change is a loss of energy ($T_0\Delta S$). What is actually available for an interchange energy (ΔH) is less than the supposed enthalpy change. The useful energy for a stream (stream availability) is defined by the following expression.

$$B = \Delta H - T_0\Delta S \qquad (2.22)$$

It is evident from above equation that there is a loss of availability following any heat interchange while the cold stream experiences availability increase. However, in a heat exchanger, the increase in availability of a cold stream is always less than the decrease in availability of a hot stream. Therefore there is a net loss of availability accompanying any operation of heat interchange. The pinch can thus be viewed as the point of zero availability (energy) in a system. Thus, any energy transfer across the pinch must come from external utilities and should equally be ejected to an external utility.

2.2.2.8 Grid representation

This is a pattern in which a heat exchanger network is represented when developing a design, whether new or retrofit. This entails drawing the process streams as horizontal lines, with the stream number shown in square boxes. Hot streams are drawn at the top of the grid, and flow from left to right, while cold streams are drawn at the bottom, and flow from right to left. The stream's heat capacity is shown in a column at the end of the stream line.

The pinch division is represented in the diagram by dividing the stream data at the appropriate temperatures, with a separation of ΔT_{min} between the hot and cold steams. Process heat exchangers, are represented by vertical lines and circles on the cold streams with H inserted and coolers by circles on the hot streams with C inserted. Stream temperatures are indicated on top of the stream lines while heat loads are placed underneath each exchanger. Figure 2.10 is an illustration of a grid representation.

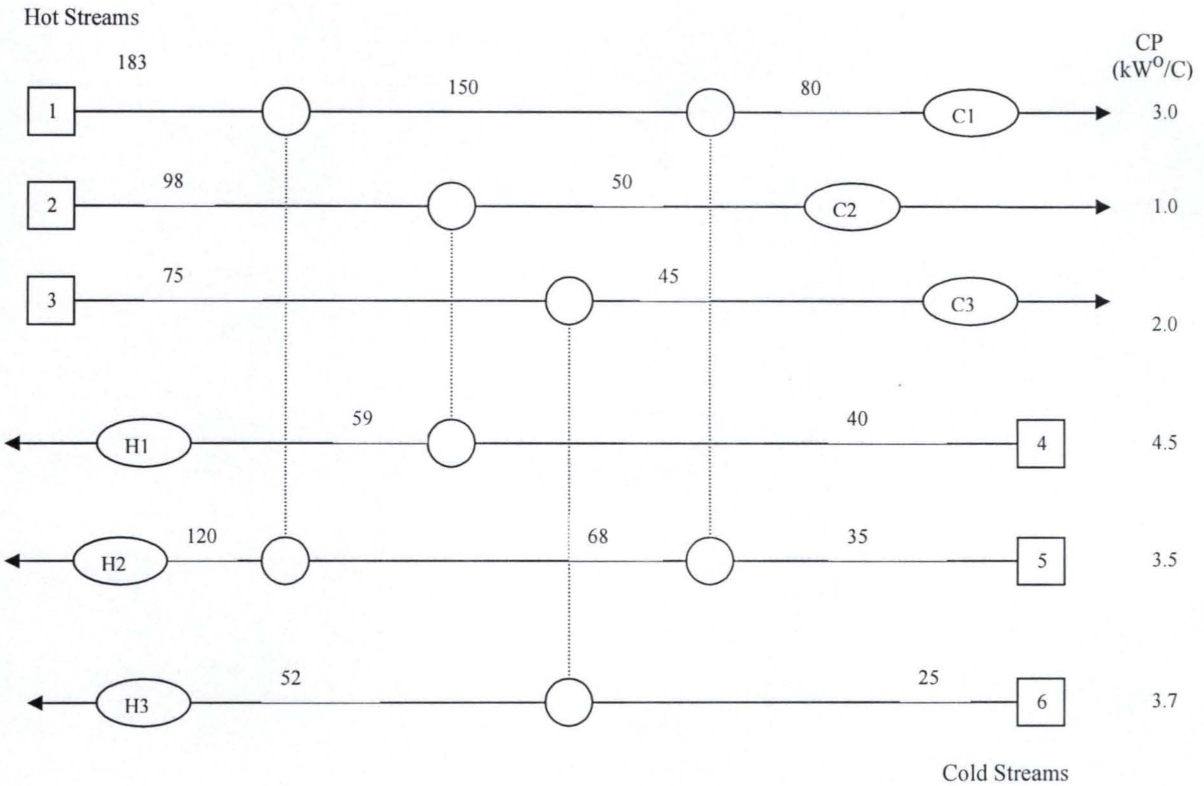


Figure 2.10: Grid Representation of Heat Exchanger Network

2.2.2.9 Trading off capital cost, energy costs and energy relaxation

In most processing systems the capital cost depends largely on the number of units involved. This is also true of heat exchange networks. The aim of the design is to reduce the capital cost (number of exchangers) and energy requirement both hot and cold utilities. The minimum number of exchanger unit as well as the minimum heating and cooling loads can be determined without even specifying the network.

In practice, designing the network involves designing a network for above the pinch and below the pinch. The total number of units of such a design is always greater than or equal to that predicted by the minimum energy requirement. This is because the combined network will have loop that cross the pinch.

By breaking the loops the number of units and hence capital costs for the network can be reduced. But breaking the loops, particularly, those that straddle the pinch (where the design is most constrained) inevitably leads to a violation of the minimum temperature approach ΔT_{min} and even in some cases to thermodynamically impossible temperature differences.

To re-establish the $\Delta T_{min,}$ it involved the use of paths through the network. This involves the addition and subtraction of heat loads in a network. The resulting heat load of both the hot and cold utilities are as a result increased beyond the minimum value established. The design is no longer based on the minimum heating and cooling restriction. The energy slack is what is termed energy relaxation. The restriction of the design based on the minimum heating and cooling requirement has to be relaxed in order to restore any impossible ΔT_{min} resulting from the breaking of the loop.

Although breaking of the loop and shifting of the heat load along a path has reduced the number of units (decreased capital cost) it has also increased the energy requirement. But a minimum of both the number of exchanger units and energy are the aims of the design. It thus appears that one of the aims can only be achieved at the expense of the other. Hence, a trade off between the number of units or capital for the network and the energy requirement for the network becomes necessary. This can be done by further shifting of the heat load along paths. This is because often the relaxed solution will result in new ΔT_{min} much greater than the minimum approach temperature ΔT_{min}. By the appropriate shifting of load along path the temperature difference between streams ΔT can be brought closer to ΔT_{min}. This will decrease the utilities requirement to some extent.

Thus the energy relaxation method for achieving a trade off between the capital cost and minimum energy requirement of the network can be summarized in the following steps.

1. Identify a loop (across the pinch)

2. Breaking the loop by the addition and the subtraction of heat load

3. Recalculate all the network temperature in order to identify any minimum approach temperature, ΔT_{min} violation

4. Find a relaxation path and restore the ΔT_{min} violation

5. Minimize the energy sacrifice by bringing the ΔT closer to ΔT_{min}.

2.2.3 Process modification

Experience has shown that industrial process can be significantly improved in a cost effective manner using pinch analysis. The method provides a guide as to the scope for improvement and how it should be carried out.

The analysis starts with the collection of data on temperature, mass flows, pressure, flow rates and the definition of the mass and heat balance. This is followed by the construction of the composite curves, similar to that in Figure 2.4 for the process and from where the minimum energy and other targets are determined.

The location of the pinch gives an important insight into how the process conditions may be changed to achieve lower utility targets. The relevant concept for carrying out the process change is the "plus/minus principle". Since the excess use of utilities in an existing plant is as the result of cross pinch heat transfer, this can be minimized by shifting cold streams from above to below the pinch and hot streams from below to above the pinch through the design of Heat Exchanger Network (HEN) the design starts from the pinch and work outwards.

It may however, be noted that not all processes involve Heat Exchanger Network (HEN) where the above approach may be most appropriate. There are processes with a few numbers of streams and a couple of unit operations where some direct heat exchange between energy source and energy sinks occurs. These processes require other method for process improvement.

The targets which illustrate cross pinch heat transfer apply to only a fixed set of process conditions. It is possible to change the key parameters and hence the process heat and material balance, to yield improved energy and capital cost. This may be achieved through the use of Grand Composite Curve (GCC) and appropriate placement concept.

2.2.3.1 Sensitivity considerations

One of the areas of intense criticism of pinch technology especially as it relates to its retrofit application is based on the operability, flexibility and controllability of the retrofitted process. Process integration critics reasoned that it would impact on the flexibility and operability of the process since the process necessarily becomes more complex. Also they posited that the controller scheme already in place might not fit into the new process. Work done in a bid to redress this problem is reviewed in this subsection.

Calandrains and Stephanopoulos (1986) developed procedure for examining the structural operability of heat exchanger networks. Linnhoff and Kotjabasakis (1986) also carried out a similar work towards developing operable process design. Kotjabasakis and Linnhoff (1986) again devised what they called sensitivity Tables for the design of flexible processes. With table, they could determine the contingency in Heat Exchanger Networks. Fludas and Grossmann (1987) devised a methodology for the synthesis of flexible Heat Exchanger Networks for multipurpose operations. Ahmad and Hui (1991) attempted to address the problem associated with network structure by restricting energy recovery between streams that fall within what they termed areas of integrity. These are identifiable regions having associated processing tasks. These regions are also associated with practical considerations such as flexibility, safety and plant layout. Amid pour and Pulley (1997) made further improvement in this direction by decomposing the overall problem into a number of sub-problems associated with specific parts of the flow sheet. However, this development has largely remained unpopular because it is cumbersome and does not directly address the controllability problem associated with process retrofit.

2.2.3.2 Existing retrofit methodology

Application of pinch technology to retrofits is not as explicit as in grassroots design owing to the idiosyncrasies of an existing process. Though the philosophy of setting target prior to design is common to both retrofit and grassroots applications, the procedure for establishing target differ significantly.

Unlike grassroots application where minimization of total cost (ΔT_{opt}) is the sole objective; three parameters need to be considered to arrive at a realistic target in retrofit. These are saving/annum, investment cost and payback period. One or two of these factors can form the retrofit objective upon which a target MTA is obtained. This section presents the retrofit target procedure currently in practice.

2.2.3.3 Setting retrofit target

Figure 2.11 shows an energy-area plot, which relates the energy requirement with the heat exchanger area used in a given process. Point A represents a case where the composite curves are close (low MTA), with corresponding high energy recovery but high investment in area. Point C relates to composite curves that are more widely spaced, yielding lower energy recovery but less investment. We have a continuous curve representing networks that are targets for both energy and area. Point B represents the optimum tradeoff with the lowest total cost. The area below the curve is marked "infeasible" because it is not possible for a design to be better than target. Point X shows the existing network. A design at point X does not take best advantage of its installed area when compared with point A or to put in another way, it does not recover as much energy as it should.

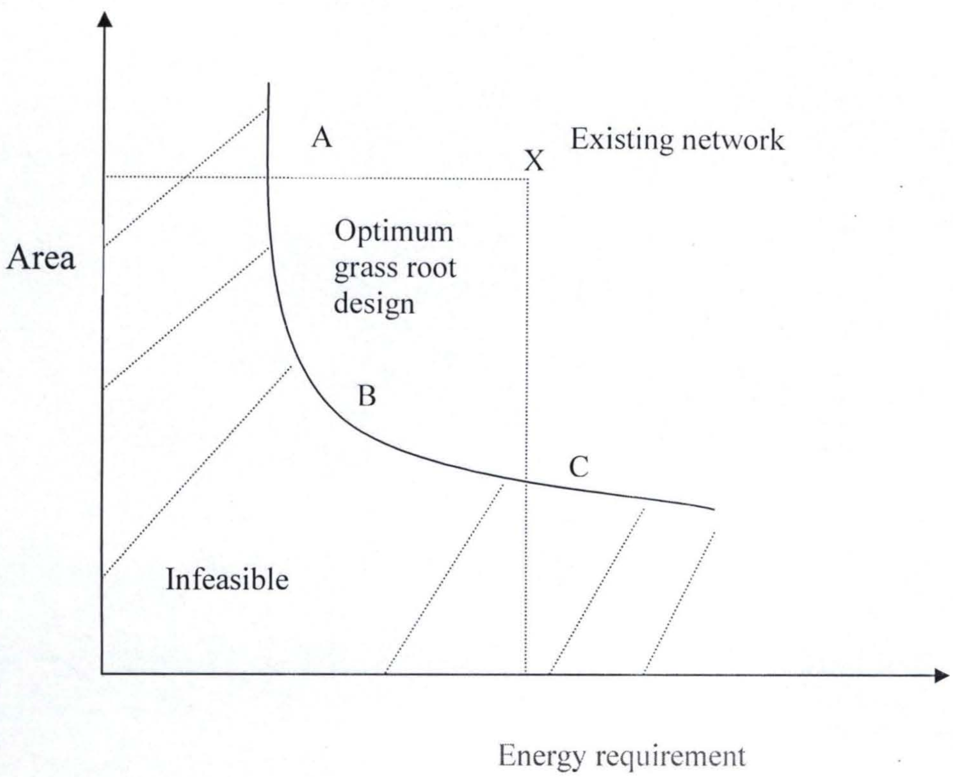


Figure 2.11: Energy Target Plotted against Area Target

2.2.3.4 Targeting philosophy

It is often assumed that retrofits should be conducted by aiming towards the optimum new design. This is usually not a viable option since we can not throw away area that has already been paid for, if an optimum new design calls for less area, this objective must be to use the existing area more effectively (Linnhoff and Tjoe, 1986).

In other word, we should try to improve on the ineffective use of area due to "criss-crossing" while shifting the composite curves closer to save energy. The ideal point to aim for from point *X* in Figure 11 would therefore be point A. Here we would save as much energy as possible using the existing area. However, in practice, we usually have to invest some capital to make changes to an existing network, thus increasing area.

2.2.3.5 Targeting procedure

In setting retrofit target, we assume that the network, after retrofit, will use area at least as effectively as before. If the project is good, then it is not likely to place new area in a manner that reduces the effectiveness of the area usage overall.

We can define area efficiency, ***a*** as the ratio of minimum area required (target) to that actually used for a specific energy recovery. The value of *a* can be expected to be less than unity in practical designs. A value of unity would indicate "no criss-crossing'. The lower the value of *a* the poorer the use of area, and the more severe the criss-crossing.

If we assume that, ***a*** is constant over the full energy span; we would obtain the curve shown in Figure 12. This curve forms a boundary for design. We can now distinguish four distinct regions in the Energy-Area plot: a region in which designs are infeasible (be they retrofit or new design); two regions in which economic retrofits are not expected; and a fourth region within which good retrofits should fall. From the constant *a* curve, we can determine what savings can be made for different levels of investment.

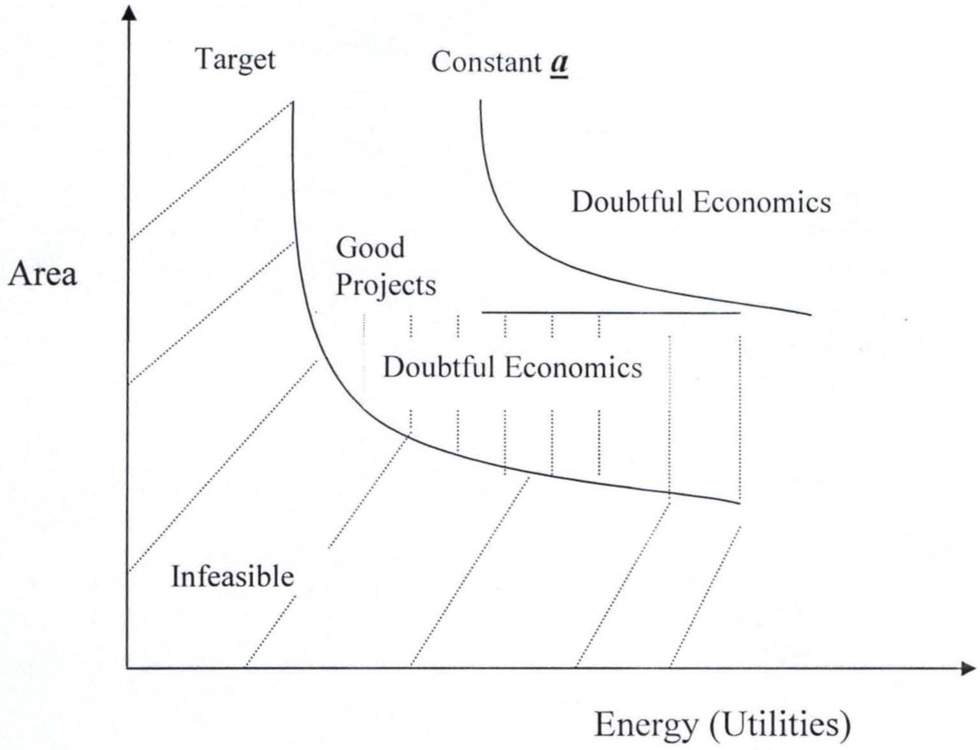


Figure 2.12: The Best Retrofits on the Area / Energy Plot

2.2.4 Retrofit methodology

The existing retrofit methodology entails the following steps (Linnhoff and Tjoe, 1986).

1. At various values of MTA, obtain the energy requirement of the network using the problem table calculation method and the target area using the existing model as described in Equation 2.13 of this Chapter.
2. Plot an area versus energy curve and locate the base case network on this curve.
3. Determine the area efficiency of the base case network and plot a constant ***a*** curve taking into consideration the infeasible and doubtful region for retrofits
4. Transform the constant ***a*** curve into savings/investment curve using appropriate cost index for the utilities and the heat exchangers.

2.2.4.5 Pinch design procedure

Without specifying the Heat Exchanger Network (HEN) we have been able to set targets for energy and area requirements of a network. We have to synthesize the network prior to its detailed design.

The procedure is as follows Linnhoff and Hindmarsh (1983):

1. Initiate the design by determining the optimum approach temperature.
2. Decompose the network at the pinch into two subsystems, above and below the pinch, and commence the synthesis at the pinch.
3. Just above the pinch the following conditions must be met to have a feasible match.

   $CP_{hot} \leq CP_{cold}$

   $N_{hot} \leq N_{cold}$

   On the other hand, just below the pinch

   $CP_{hot} \geq CP_{cold}$

   $N_{hot} \geq N_{cold}$

   If these conditions are not met at the pinch, the streams can be split. However, these conditions are only necessary for pinch matches, as we move away from the pinch we need not consider these constrains.

   1. Maximize the heat load on each of the matches so as to reduce the stream population (heuristic tick).
   2. The two separate subsystems design; above and below the pinch can now be brought together for a complete design.

5. Using the set retrofit criteria, either savings per annum or investment cost or payback period, establish the target MTA and the pinch temperatures.
6. Identify the cross pinch exchangers on the grid diagram of the existing network using the identified MTA in the targeting stage.
7. Eliminate cross pinch exchangers.
8. Complete the network by placing new exchangers, and where possible, reuse exchangers removed in the last step.
9. Evolve improvements by adequate manipulation of the loops and paths.

2.2.4.1 Drawbacks of the retrofit methodology

The state-of-the art methodology for accomplishing retrofit is bedevilled with two main drawbacks. The first drawback relates to the targeting procedure employed while the second one is the fact that operability, flexibility and controllability are taken for granted in the existing methodology.

2.2.4.2 Economic evaluation

Economic assessments must be made when considering power and heat exchange networks, and their design. The chemical process industries are very capital intensive. Therefore, costs must be fully accounted for as a measure of profitability established for any project of interest.

Initial capital costs can escalate at considerable rate, fixed capital costs including the cost of capital equipment, installation of process equipment, piping, insulation, instrumentation, engineering and constructing, consultant fees, commissioning cost, while the running cost, comprises cost of fuel, such items as cooling water, auxiliary services such as interest, insurance and management time, must be taken into account.

In the chemical process industries, potential projects are evaluated using some measure of profitability. Most popular profitability measure include: Return on investment, (ROI), Pay back period, (PBP) (in years), Net present worth or value (NPV), and the discounted cash flow rate of Return (DCFRR).

2.2.4.3 Super targeting

Since the minimum temperature approach between hot and cold streams occurs at the pinch, this is the most constrained region and where the designer must first concentrate his attention. Transfer of heat across the pinch is avoided through the simple strategy of first making stream matches at the pinch.

The choice of minimum temperature approach MTA specified in the design of Heat Exchanger Network (HEN) affects both energy consumption and the required surface area for the network. A high MTA will give low heat recovery, high energy target (high utility requirement) and thus high operating cost as evident in Figure 2.13. On the other hand, a low MTA gives high energy recovery, low energy target and low operating cost. However, a low MTA gives low temperature driving force in exchangers, thereby requiring a large surface area and therefore high capital cost.

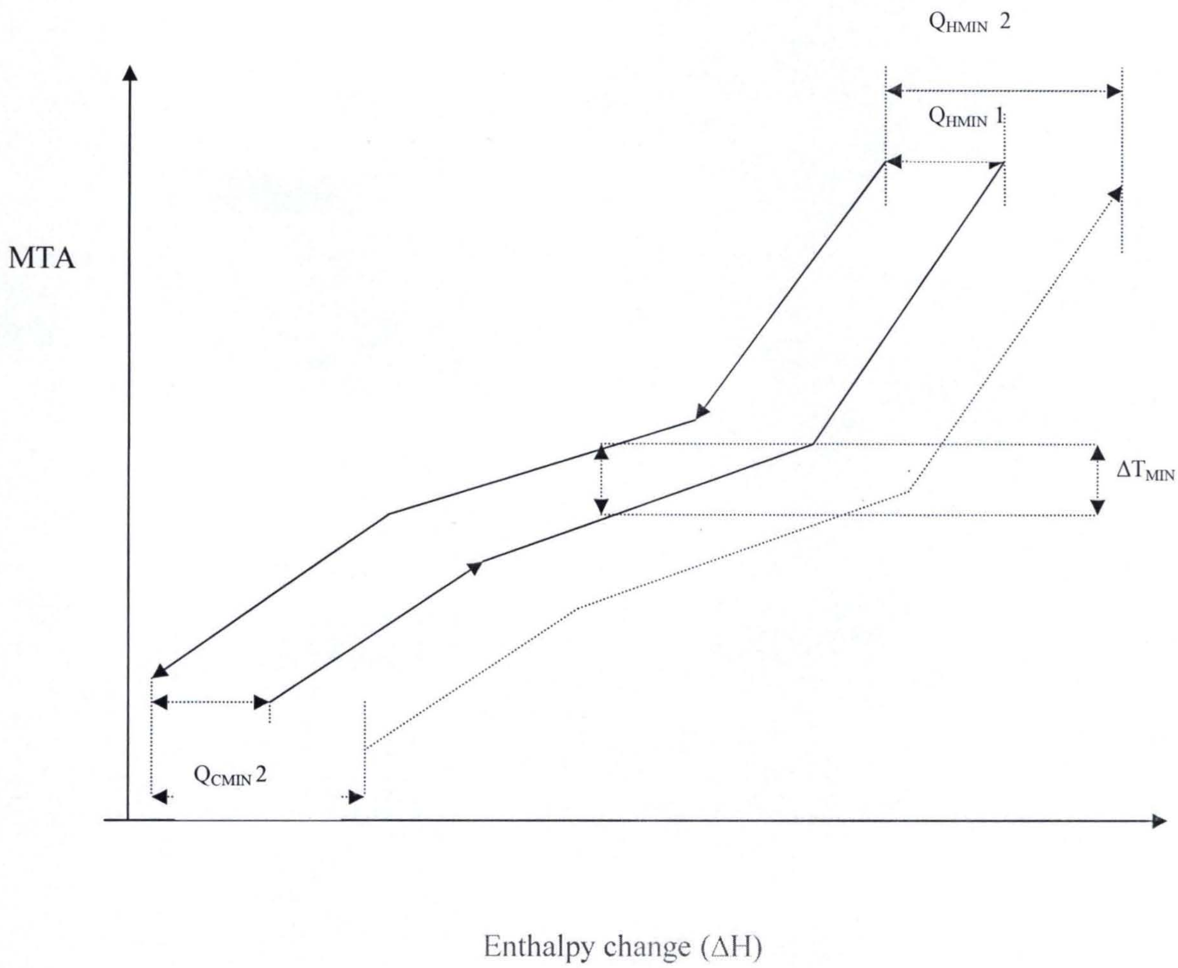


Figure 2.13: Energy Requirement against MTA

The operating and capital costs have opposing relation with MTA. Therefore, there is need for trade-off between the two in order to determine the optimum MTA. This is known as super targeting. It gives the optimum value of MTA that would yield the least total cost (operating and capital cost). This is shown in Figure 14.

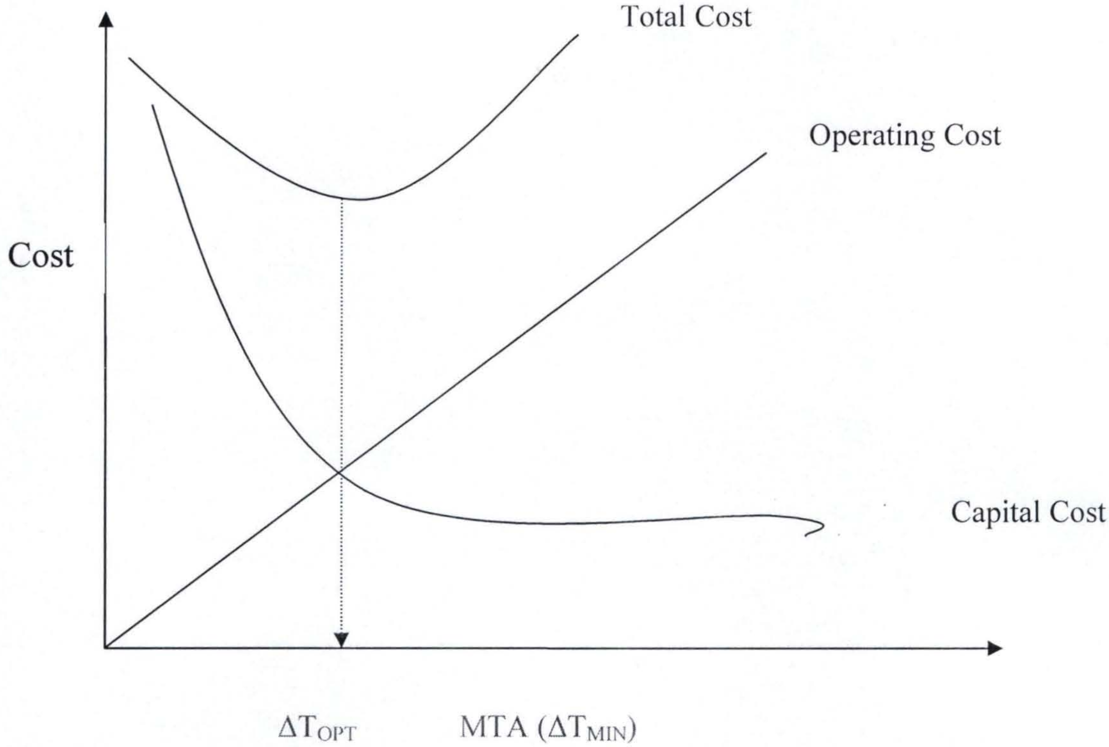


Figure 2.14: Surface Area and Energy Trade-off.

2.2.4.4 Stream matching at minimum utilities

Having determined the minimum utilities for heating and cooling using problem table or composite curve, it is common to design two Networks of Heat Exchanger, one above and one below the Pinch temperature as shown in Figure 15 and 16. Two methods are presented for this purpose. The first, introduced by Linnhoff and Hindmarsh (1983) places emphasis on positioning the Heat Exchangers by working out from the Pinch temperature. The second is an algorithmic strategy

There will always be two loops that cross the pinch, thus leading to two extra exchangers. This is evident from the application of Euler's equation to the total system. These loops can be broken resulting in reduced number of exchangers (lower capital cost), but at the expense of increased energy consumption (higher operating cost). A trade-off move has to be made to attain the optimum network. This design procedure is presented in Figures 15 and 16.

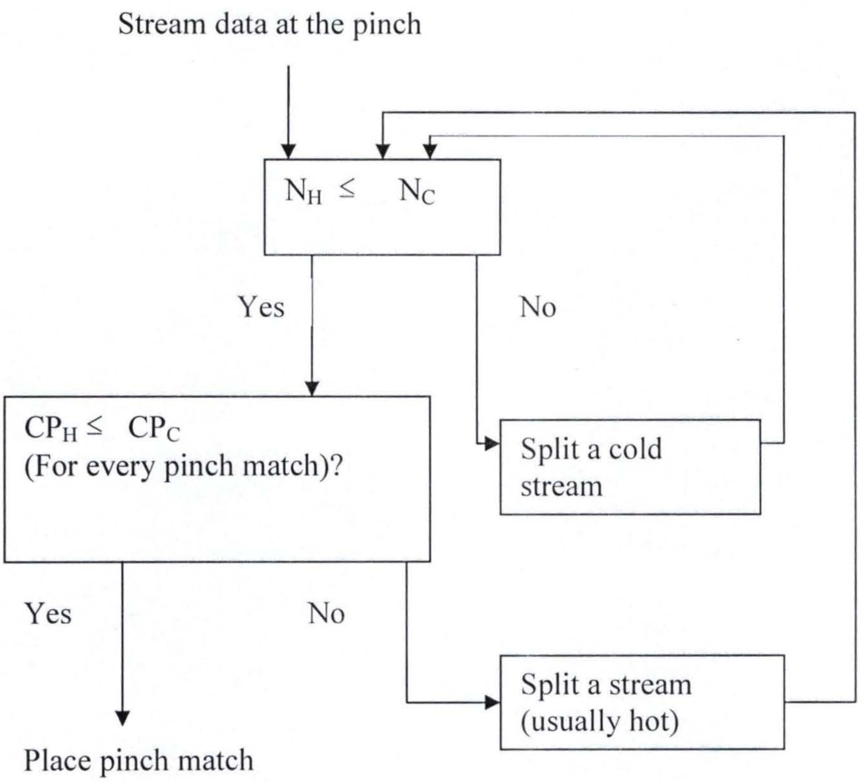


Figure 2.15: Design Procedure above the Pinch

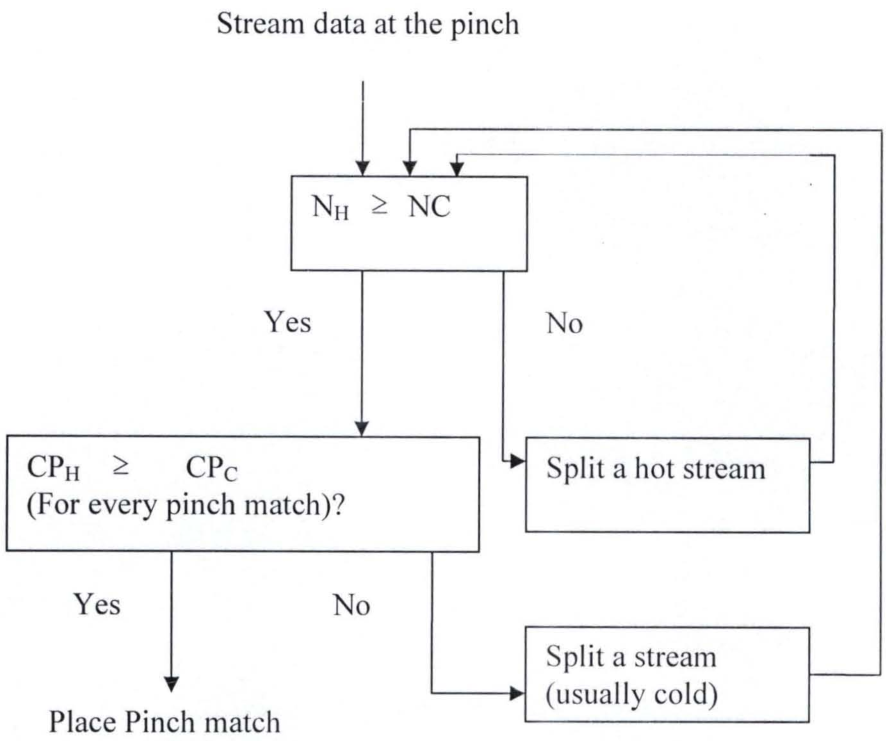


Figure 2.16: Design Procedure below the Pinch

2.3 Hydrodesulphurization Unit (HDS)

Petroleum fractions contain various amounts of naturally occurring contaminants, the most important ones being organic sulfur, nitrogen, and metals compounds. These contaminants, if not removed, would contribute to the increasing levels of air pollution and equipment corrosion, and in some cases would cause difficulties in the further processing of the material.

The purpose of Universal Oil Product(UOP) Hydrodesulphurization (HD) Unibon process unit is to remove sulphur and nitrogen from the feed without greatly changing its boiling range. If not removed, the sulphur and nitrogen would poison the sieve in the UOP Molex process unit. The HD Unibon process is a catalytic hydrogenation method for upgrading the feed quality. This upgrading, often called hydro treating is done by passing the feed stock over a fixed bed of UOP

HDS catalyst. But the catalyst alone is not enough to cause the decomposition of the sulphur and contaminants. It is also necessary to have high temperatures and pressure along with large amounts of hydrogen. The temperature and pressure required will depend on nature of the feed as well as the amount of contaminant removal required. For further processing in the Molex Unit, it is necessary to reduce the sulphur and nitrogen levels to 1 PPM or less. (UOP General Operating Manual 1983)

2.3.1 Reaction Chemistry

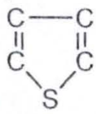
The following reactions represent in general what is taking place inside the reactor

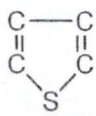
1. **Sulphur Removal**

MERCAPTAN $R-S-H+H_2 \rightarrow R-H+H_2S$

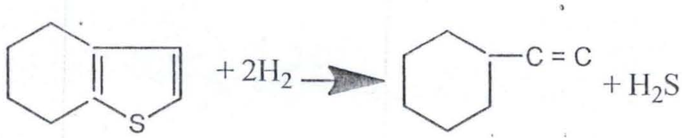
SULFIDE $\text{C-C-S-C-C} + 2H_2 \rightarrow 2\text{C-C} + H_2S$

DISULFIDE $\text{C-C-S-S-C-C} + 3H_2 \rightarrow 2\text{C-C} + 2H_2S$

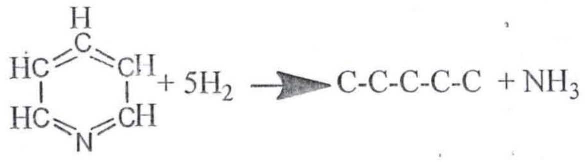
CYCLIC SULFIDE  $+ 2H_2 \rightarrow \text{C-C-C-C} + H_2S$

THIOPHENIC  $+ 4H_2 \rightarrow \text{C-C-C-C} + H_2S$

BENZOTHIOPHENES



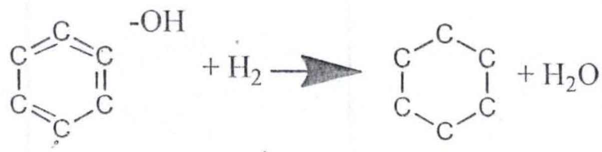
2. **Nitrogen Removal**

PYRIDINE 

3. **Metals Removal**

While mechanical of organo-metallic compound removal is not well understood, it is known that metals are retained on the catalyst. The limiting amount of metals the catalyst can remove is related to the amount of catalyst in the plant. Once this limit is exceeded, metals can be found in the reactor product stream. A process unit should be able to go through several operating cycles without exceeding the metals removing ability of the catalyst.

4. **Oxygen Removal**

PHENOL 

5. **Olefin Saturation**

LINEAR OLEFIN C-C=C-C-C $+ H_2 \longrightarrow$ C-C-C-C-C-C

6. Halides Removal

Organic halides are decomposed in the reactor to form inorganic salts. These salts will deposit downstream of the reactor if water is not injected to wash them away.

C-C-C-C-C-Cl $+ H_2 \longrightarrow$ C-C-C-C-C-C $+ HCl$

2.3.2 Other Reactions

Near the end of the run, temperature will be relatively high. This will compensate for the lower catalyst activity. With the higher temperature, there is an increased tendency to hydrocrack the feed. The increase in hydrocracking will be evidenced by higher hydrogen consumption and more net stripper overhead liquid production.

The products of the Unibon reactions are of a lower density than the feedstock. Therefore, the total liquid yield in most cases is more than 100 LV-, and may be as high 102 LV-. Total liquid yield will go up as hydrocracking increases, however, most of this increase will be in form of more net stripper overhead production accompanied by a lower stripper bottoms product make. A Unibon unit is designed for maximum bottoms production, thus, economic considerations will determine the amount of bottoms product that can be lost before the unit is shut down for regeneration or catalyst change.

2.3.3 Reaction Heats

All of the reactions we have just discussed give off heat. This is why there is a temperature rise across the reactor. Olefin saturation is the most exothermic reaction and it gives off about 856 cal per cubic meter of H_2 consumed. Desulfurization yields approximately 214 cal per cubic meter of H_2 consumed

2.3.4 Catalyst

The UOP Hydrobon catalyst consists of metal oxides impregnated on an alumnia base, and may be prepared either as a sphere or an extrudate. The particular catalyst selected will be based on type of feedstock, desired prod, and process design conditions. The most economical combination of these factors is considered in the basis for catalyst selections.

type of feedstock, desired prod, and process design conditions. The most economical combination of these factors is considered in the basis for catalyst selections.

2.3.5 Reactor Temperatures

The rector temperatures must at all time be the minimum necessary to achieve the desired desulfurization and nitrogen removal. Increasing reactor temperatures will accelerate the rate of coke formation, thereby reducing the time between regenerations. This same principle applies if the feed rate is changed, therefore it should always be common practice to lower the reactor temperature before reducing the feed rate, or to raise the feed rate before increasing the temperature.

Feed Boiling Range

As the feed end point rises, so will its sulfhur and nitrogen content. This requires higher temperatures for removal. Moreover, heavier feeds contain more coke precursors. These precursors deactivate the catalyst and shorten the operating cycle. The reactor pressure drop will also go up when the unit is operated on heavy feeds due in part to the accumulation of solids in the catalyst bed. Thus, the cycle length is reduced by the faster accumulation of coke and the higher reactor pressure drop. At the same time the ultimate catalyst life is reduced by the additional metals in the heavier feed.

2.3.6 Charge Rate

At greatly reduced charge rates, operation of the unit may become difficult due to hydraulic considerations, i.e. control values operating almost fully closed, e.t.c. also flow distribution in the reactor may become unequal and preferential flow paths establishes. This would result in less than full utilization of the catalyst, requiring higher reactor temperatures and increased coke formation. For these reasons, the unit should not be operated at feed rates below 70% of design for extended periods.

2.3.7 Hydrogen Purity-Hydrogen Partial Pressure

These two terms are interchangeable since at a given system pressure the purity of the recycle gas will determine the partial pressure of hydrogen in the reactor. At reduced hydrogen purities (or partial pressure) the reactions are not effectively completed and there is a greater tendency for coke to be formed. As the H_2 purity decreases, the catalyst appears less active. The operator may try to compensate by increasing reactor temperatures which will further aggravate the problem. If the H_2 purity can not be increased by either raising the makeup hydrogen purity or supplying more make up hydrogen. This action would include decreasing reactor temperatures and decreasing charge rate. Under no circumstances should the unit be operated at recycle gas hydrogen purities below 70 mol percent.

The hydrogen consumed by the HD Unibon reactions is supplied by the UOP Pacol Process Unit. Because is is supplied at a pressure lower than the system pressure, it must be boosted by reciprocating pressure.

Though the plant may be run at recycle H_2 purities less than 75 percent for a few hours, there will be adverse effects on the catalyst performance. Along with the adverse effects, there will be an increase in power required to pump the gas and a significant loss in cooling by the gas. Recycle hydrogen percentages less than 75 mol percent can only be tolerated on rare occasions for short periods of time.

Under normal operating conditions, the recycle H_2 purity will be measured once each day. If there are problems with the hydrogen sources, It may be necessary to sample more often. By maintaining the recycle gas hydrogen purity as high as possible, the catalyst deactivation rate can be kept to a minimum.

2.3.8 Recycle Gas Rate

The large quantity of gas from the high pressure separator to the reactor serves the following purposes.

a. Provides the excess hydrogen needed to assure that the reaction are carried to completion.
b. Absorbs some of the heat of reaction, thereby minimizing the catalyst bed temperatures.
c. Helps hold down charge heater and combined feed exchanger tube wall temperature by increasing the flow through the equipment, and the excess H_2 prevents the formation of coke as the charge is heated to reaction temperatures.

The charge heaters, combined feed exchangers, compressors, etc. have been designed to allow for the circulation of minimum ratio of hydrogen to hydrocarbon charge (H_2/CH). This ratio is expressed as standard cubic feet of H_2 per barrel of fresh feed (SCFB) or normal meters cubed of H_2 per meter cubed of fresh feed (Nm^2/m^2 H_2/HC). The hydrogen to hydrocarbon ratio is an economic optimum, balancing initial investment against catalyst life. For this implication, H_2/HC will normally be around. If the unit is operated at less than the design H_2/HC ratio, higher catalyst bed and charge heater tube wall temperatures will result, leading to accelerated coke formation and equipment wear. A complete loss of recycle gas can result in very serious damage to the catalyst and equipment; therefore, a safety device is incorporated into the unit to shutdown the charge heaters whenever the recycle gas flow drops below a predetermined minimum. Stop other steps to be taken during such as emergency are more completely described in the emergency procedures section of this manual.

2.3.9 Process Flow and Equipment

A11 HD Unibon Process units used in an LAB complex will consists of a high pressure reactor section and a lower pressure product stripper section. While the lines, vessels, pumps, compressors, et all. That make up these two sections vary in arrangement and number, they can still be described in general terms.

2.3.10 Reactor Section

Feed obtained from a crude distillation unit and/or storage and/or Prefractionation unit, is sent to the feed surge drum. In cases where the feed contain appreciable quantities of solid material, feed filters may be included before the surge drum. If filter are used, it will be necessary to provide a back flush system, or a means for periodic cleaning. The reactor charge pumps take suction from the feed surge drum and pump the raw kerosene to the combined feed exchangers. The charge pumps are high head machines capable of pumping large volumes of kerosene up to the pressure near 1000 psig (70kg/cm^2).

Before entering the combined feed exchangers, the oil is mixed with the recycle H_2 The make-up H_2 is usually obtained from the LAB pacol process unit and / or a hydrogen manufacturing plant ... pressure of 200 psig (14 kg/cm^2) or higher. Since the Unibon reactor section pressure may be ... psig (70kg/cm^2), the make-up gas often must be compressed before it can join the ...d. Reciprocating compressors are used to raise the pressure of the gas, with the ...mpression stages varying in accordance with the difference between the supply and ...ssures. If the make up gas available at 400psig (28kg /cm^2) or more, its pressure can be raised to 1000 psig (70kg/cm^2) with one stage of compression, however, if the supply pressure is 200 psig (14kg/cm^2) or less two stages of compression must be used. On multi-stage compressors the gas from the first stage must be cooled to about 120oF (50°C) before it can enter the second stage. The spillback gases, used to control the inter-stage suction drum pressures also need to be cooled before being returned to the suction drums.

The flow of make up H_2 may be controlled in a variety of ways with the objective in all cases being to achieve smooth high pressure separator pressure control while assuring that the compression ratio is always maintained within the allowable maximum for the compressor.

The most commonly used methods of control are as follows:

Fractionation Section

The hydrocarbon liquid collected in the high pressure separator is sent to the striper on level control. Before entering the column the separator liquid will be heated in the stripper feed/bottoms exchanger

The column strips out the water, light ends and H_2S in the separator liquid. The overhead vapors are condensed normally through an air cooled fin fan condenser and water cooled trim condenser, then enter the overhead receiver. The water phase is collected in a boot and sent to the sour water stripping unit. Condensed hydrocarbon is refluxed back to the column and a net overhead liquid draw is removed for further processing. The non condensable gases leave the overhead receiver on pressure control and are normally routed to the refinery fuel gas system.

The striper column is re-boiled typically with hot oil loop on flow control through a thermal siphon re-boiler. The H_2S free material from the column bottom is exchanged with the stripper feed and sent to the Molex unit on level control. A slipstream of the bottoms can be bypassed around the stripper feed/bottoms exchanger on TRC control to ensure a constant feed temperature to the Molex unit.

Before processing to the Molex feed surge drum, the stripper bottoms can alternately be routed to a Molex feed tank through an air cooled fin fan cooler and water cooled trim cooler in this manner feedstock can be accumulated to allow the Molex unit to continue running during short shutdowns of the Unibon unit. This tank is normally sized to hold a weeks supply feed back

In some designs the Molex feed surge drum may be located as a bottom extension of the stripper column. In this case the Molex feed pumps would also be located at the stripper column.

CHAPTER THREE

3.0 MATERIALS AND METHODS

This chapter presents all the steps involved in the analysis, designing and optimization of Heat Exchangers Network of Hydrodesulphurization Unit (HDS) of Kaduna Refining and Petrochemical Company. The procedures involved data extraction, process simulation and pinch analysis which are shown under Figure 3.1. Basically the existing Heat Exchangers Network of the Preheat train of the unit will be analyzed in order to extract all the necessary information required for the pinch analysis. As mentioned earlier, the use of pinch technology in the energy conservation area remains the focus of this work.

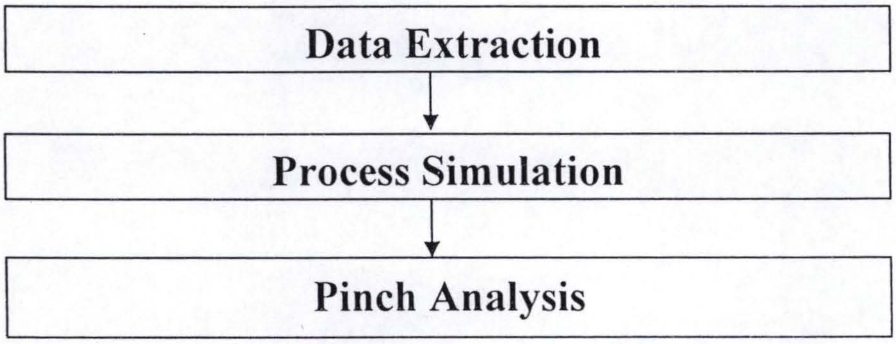


Figure 3.1: Steps involved in the energy integration of HDS unit of KRPC

3.1 Data Extraction

In the analysis of the existing network, a thorough study of the Process Flow Diagram (PFD) as shown in Figures 3.3 to 3.5, Piping and Instrumentation Diagram (P&ID) shown under Appendix A1 and Laboratory analysis of the raw kerosene shown under Figures 4.1 was carried out in order to extract all the necessary and available information require to carry out the process simulation pinch analysis of the HDS plant.

The feed and product composition of the laboratory analysis shown under Figures 4.1 was used in carrying out the process simulation. The stream temperatures, mass flow rates, pressures were also extracted from PFD and P&ID for carrying out the process simulation as shown in figure 3.1.

Feed and Product Compositions from the Laboratory Data

↓

Streams Temperatures, Pressures, Flow rates from PFD and P&ID

Figure 3.2: Data Extraction Steps

3.2 Process Simulation

3.2.1 Process Simulation using AspenONE

AspenONE is AspenTech's comprehensive set of software solutions and professional services designed to help process companies achieve their operational excellence objectives. It leverages the value of simulation models to help process companies increase operational efficiency and profitability across their global enterprise.

3.2.2 Introduction to Aspen Hysys

The simulations of the Hydrodesulphurization have been carried out using Aspen Hysys, which is chemical process simulation modeling software.

The flow sheet (PFD) includes a library of standard unit operation blocks and logical units (e.g. cooler, mixer, Heat-exchangers, separator, splitters, compressor, Recycle, spreadsheet, set, adjust), which represent processes taking place in an actual hydrodesulphurization plant. Aspen Hysys is a combination of tools that are used for estimating the physical properties and liquid-vapour phase Equilibrium of various inbuilt components. These components are the substances that are used within the plant for the feeds, within the reaction and separation sections. The program is such that it will converge energy and material balances and has standard unit operations typical of any processing plant. Aspen hysys updates the calculations as the user enters information. The successful completion of an operation is seen by the changes in colour on screen. Aspen hysys is not just a steady state program. A case can be transferred into a dynamic simulation where process controllers can be added, and hence, realistically evaluate a plant wide control philosophy

For the Hydrodesulphurization process to be modelled in Aspen hysys, there must be a foundation on which the components must be modeled. In this process, there is one component involved in the chemistry that is nitrogen. Nitrogen is selected as pure components within the simulation basis manager. The next task is to assign a fluids package, which is very critical to the successful calculation of the streams component as it is being used by the software (Aspen hysys). There are dangers of using an incorrect thermodynamics package because it is stated: "Everything from the energy balance to the volumetric flow rates to the separation in the equilibrium-stage units depends on accurate thermodynamic data." Peng-Robinson equation of state will be used in this research work.

3.2.2.1 Equation of state

In physics and thermodynamics, an equation of state is a relation between state variables. More specifically, an equation of state is a thermodynamic equation describing the state of matter under a given set of physical conditions. It is a constitutive equation which provides a mathematical relationship between two or more state functions associated with the matter, such as its temperature, pressure, volume, or internal energy. Equations of state are useful in describing the properties of fluids, mixtures of fluids and solids.

3.2.2.2 Peng-Robinson: equation of state.

Peng-Robinson is a Cubic equation of state that describe the relationship between thermodynamics properties such as Enthalpy and inlet conditions such as temperature and pressure to the system.

$$P = \frac{RT}{V_m - b} - \frac{a\alpha}{V_m^2 + 2bV_m - b^2} \tag{3.1}$$

$$a = \frac{0.45724 R^2 T_c^2}{p_c} \tag{3.2}$$

$$b = \frac{0.07780 R T_c}{p_c} \tag{3.3}$$

$$\alpha = (1+(0.37464+1.54226\omega - 0.26992\,\omega^2)\,(1-T_r^{0.5}))^2 \tag{3.4}$$

$$Tr = \frac{T}{T_c} \tag{3.5}$$

In polynomial form:

$$A = \frac{a\alpha p}{R^2 T^2} \tag{3.6}$$

$$B = \frac{bp}{RT} \quad 3.7$$

$$Z^3 - (1\text{-}B)\,Z^2 + (A\text{-}3B^2\text{-}2B)\,Z - (AB\text{-}B^2\text{-}B^3) = 0 \quad 3.8$$

where, ω is the acentric factor of the species and R is the universal gas constant and A, B and Z are constant.

The Peng-Robinson equation was developed in 1976 in order to satisfy the following goals:

1. The parameters should be expressible in terms of the critical properties and the acentric factor.

2. The model should provide reasonable accuracy near the critical point, particularly for calculations of the compressibility factor and liquid density.

3. The mixing rules should not employ more than a single binary interaction parameter, which should be independent of temperature pressure and composition.

4. The equation should be applicable to all calculations of all fluid properties in natural gas processes.

3.2.2.3 Simulation Environment

The Simulation environment contains the main flow sheet where you do the majority of your work (installing and defining streams, unit operations, columns and sub flow sheets). Before entering the Simulation environment, it is important to have a fluid package with selected components in the component list and a property package.

Fig: 3.3 simulation environment

The flow sheet in Aspen hysys shows the various components and the material streams needed to bring about the hydrodesulphurization of the raw kerosine. It consists of various apparatus(Object Palette) but few object which are in our use are as mixer, an isentropic compressor, a chiller, a LNG countercurrent heat exchanger, an isenthalpic J-T valve, a separator which performs flash separation operations and logical operation units Set, Spreadsheet and Recycle.

3.2.2.4 ASPEN HYSYS object

The description of the various components and the conditions at which they operate are described subsequently.

A. Mixer

The Mixer operation combines two or more inlet streams to produce a single outlet stream. A complete heat and material balance is performed with the Mixer. That is, the one unknown temperature among the inlet and outlet streams is always calculated rigorously. If the properties of all the inlet streams to the Mixer are known (temperature, pressure, and composition), the properties of the outlet stream is calculated automatically since the composition, pressure, and enthalpy is known for that stream.

B. Compressor

There are various type of compressor that are available in market but in Aspen Hysys option of isentropic centrifugal compressor is available. The centrifugal compressor operation is used to increase the pressure of an inlet gas stream with relative high capacities and low compression ratios. Depending on the information specified, the centrifugal compressor calculates either a stream property (pressure or temperature) or compression efficiency.

C. Cooler/Chiller

The Cooler operations are one-sided heat exchangers. The inlet stream is cooled (or heated) to the required outlet conditions, and the energy stream absorbs (or provides) the enthalpy difference between the two streams. These operations are useful when you are interested only in how much energy is required to cool or heat a process stream with a utility, but you are not interested in the conditions of the utility itself.

D. Heat Exchanger

The heat exchanger model solves heat and material balances for single-stream heat exchangers and heat exchanger networks. The solution method can handle a wide variety of specified and unknown variables. For the overall exchanger, you can specify various parameters, including heat leak/heat loss, UA or temperature approaches can be specify. Two solution approaches are employed; in the case of a single unknown, the solution is calculated directly from an energy balance. In the case of multiple unknowns, an iterative approach is used that attempts to determine the solution that satisfies not only the energy balance, but also any constraints, such as temperature approach or UA.

E. Separator

Multiple feeds, one vapour and one liquid product stream. In Steady State mode, the Separator divides the vessel contents into its constituent vapour and liquid phases

F. Air Cooler

The air cooler unit operation uses an ideal air mixture as a heat transfer medium to cool (or heat) an inlet process stream to a required exit stream condition. One or more fans circulate the air through bundles of tubes to cool process fluids. The air flow were specified or calculated from the fan rating information. The air cooler can be solve for many different sets of specifications including:

- The overall heat transfer coefficient, UA
- The total air
- The exit stream temperature

The air cooler uses the same basic equation as the heat exchanger unit operation however; the air cooler operation can calculate the flow of air based on the fan rating information.

The air cooler calculations are based on an energy balance between the air and process streams. For a cross-current air cooler, the energy balance is shown as follows:

$$M_{air}(H_{out} - H_{in})_{air} = M_{process}(H_{in} - H_{out})_{process} \qquad 3.9$$

Where: M_{air} = *Air stream mass flow rate* (kg/s)

$M_{process}$ = *Process stream mass flow rate* (kg/s)

H = *Enthalpy* (kJ/hr)

The air cooler duty, Q, is defined in terms of the overall heat transfer coefficient, the area available for heat exchange and the log mean temperature difference.

$$Q = UADT_{LM}F_t \qquad 3.10$$

Where U = overall heat transfer coefficient

A = surface area available for heat transfer (m^2)

ΔT_{LM} = Log mean temperature difference (LMTD)

F_t = correction factor

The LMTD correction factor, F_t, is calculated from the geometry and configuration of the air cooler.

G. Heat Exchanger

The hear exchanger performs two-sided energy and material balance calculations. The heat exchanger is very flexible and can solve for temperatures, pressures, heat flows (including heat loss and heat leak), material stream flows, or UA.

In aspen hysys, the Heat Exchanger Model was chosen analysis. The choices include an End Point analysis, an ideal (Ft=1) counter-current weighted model, a simple rating method for use with both steady state or dynamic simulations or third party heat exchanger design methods via OLE Extensibility.

The heat exchanger calculations were based on energy balances for the hot and cold fluids. In the following general relations, the hot fluid supplies the heat exchanger duty to the cold fluid:

$$M_{cold}(H_{out} - H_{in})_{cold} - Q_{leak} - (M_{hot}(H_{in} - H_{out} - Q_{loss}) = Balance\ Error \qquad 3.11$$

Where M = *Fluid mass flow rate* (kg/hr)

H = *Enthalpy* (kJ/hr)

Q_{leak} = *Heat Leak* (kJ/hr)

Q_{loss} = *Heat Loss* (kJ/hr)

The Balance Error is a heat exchanger Specification that will equal zero for most applications.

The subscripts are hot and cold designate the hot and cold fluids, while in and out refer to the inlet and outlet.

The Heat Exchanger duty were defined in terms of the overall heat transfer coefficient; the area available for heat exchange and the log mean temperature difference:

Because the reaction rate was considered spatially uniform in each subvolume, the third term reduces to $r_j V$. At steady state, the right side of this balance equal zero, and the equation reduces to:

$$F_j = F_{j0} + r_j V \qquad 3.14$$

3.2.2.5 Logical Units

A. SET

SET is used to set the value of a specific process variable (P V in the manuals) in relation to another PV. The relation must be of the form Y = mX + b and the process variables must be of the same type. For example, you could use the SET to set one material streams temperature to always be 20 degrees hotter than another material stream's temperature. SET may work both ways (i.e. if the target is known and not the source, the target will "set" the source).

B. SPREADSHEET

The Spreadsheet applies the functionality of Spreadsheet programs to flowsheet modeling. With essentially complete access to all process variables, the Spreadsheet is extremely powerful and has many applications in ASPEN HYSYS. The ASPEN HYSYS Spreadsheet has standard row/column functionality. You can import a variable, or enter a number or formula anywhere in the spreadsheet.

The Spreadsheet can be used to manipulate or perform custom calculations on flowsheet variables, because it is an operation calculations that are performed automatically; Spreadsheet cells are updated when flowsheet variables change.

$$Q = UA\Delta T_{LM}F_t \qquad 3.12$$

Where U = Overall heat transfer coefficient

A = Surface area available for heat transfer (m^2)

ΔT_{LM} = Log mean temperature difference (LMTD)

F_t = LMTD correction factor

The heat transfer coefficient (U) and the surface area(A) as shown in equation 3.12 are often combined for convenience into a single variable referred to as UA. The LMTD and its correction factor are defined in the performance section.

H. Distillation Column Theory

Multi-stage fractionation towers, such as crude and vacuum distillation units, reboiled demethanizers, and extractive distillation columns, are the most complex unit operations that ASPEN HYSYS simulates. Depending on the system being simulated, each of these towers consists of a series of equilibrium or non-equilibrium flash stages. The vapour leaving each stage flows to the stage above and the liquid from the stage above flows to the stage below. A stage may have one or more feed streams flowing onto it, liquid or vapour products withdrawn from it, and can be heated or cooled with a side exchanger. The following figure shows a typical stage j in a Column using the top-down stage numbering scheme. The stage above is j-1, while the stage below is j+1.

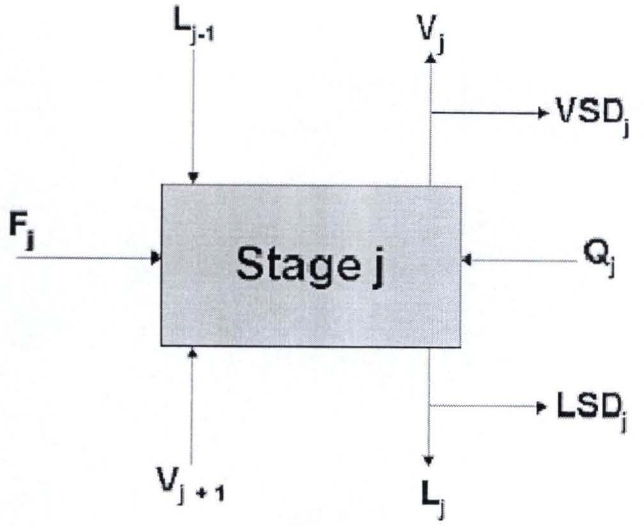


Figure 3.3: Equilibrium Stage Theory

Where:

F = Stage feed stream

L = Liquid stream travelling to stage below

V = Vapour stream travelling to stage above

LSD = Liquid side draw from stage

VSD = Vapour side draw from stage

Q = Energy stream entering stage

Column Initial Estimate

Initial estimates are optional values that are provided to help the ASPEN HYSYS algorithm converge to a solution. The better the estimates, the quicker ASPEN HYSYS will converge. It is important to remember that specifications become initial estimates, as one of the original default specifications (overhead vapour flow, side liquid draw or reflux ratio) was replaced with new active specifications, the new values become initial estimates. For this reason it is recommended that reasonable values were provided initially even if it will be replaced.

Initial estimates were provided via the Column Runner, either on the Monitor page, of the design tab, in the specification list or on the estimates page of the parameters tab. Although ASPEN HYSYS does not require any estimates to converge to a solution, reasonable estimates helped in the convergence process.

Temperature

Temperature estimates were given for any stage in the column, including the condenser and reboiler using the Estimates page in the Column Runner. Intermediate temperatures were estimated by linear interpolation. When large temperature changes occur across the condenser or bottom reboiler, it was helpful in providing an estimate for the top and bottom trays in the tray section.

I. Plug Flow Reactor (PFR) Property View

The PFR (Plug Flow Reactor, or Tubular Reactor) generally consists of a bank of cylindrical pipes or tubes. The flow field was modeled as plug flow, implying that the stream is radially isotropic (without mass or energy gradients). This also implies that axial mixing was negligible. As the reactants flow into the reactor, they were continually consumed, hence, there was an axial variation in concentration. Since reaction rate is a function of concentration, the reaction rate also varied axially (except for zero-order reactions). To obtain the solution for the PFR (axial profiles of compositions, temperature, etc.), the reactor is divided into several subvolumes. Within each subvolume, the reaction rate was considered to be spatially uniform. A mole balance as shown in equation 3.13 was done in each subvolume j:

$$F_{j0} - F_j + \int_v r_j dV = \frac{dN}{dt} \qquad 3.13$$

One application of the Spreadsheet is the calculation of pressure drop during dynamic operation of a Heat Exchanger. In the ASPEN HYSYS Heat Exchanger, the pressure drop remains constant on both sides regardless of flow. However, using the Spreadsheet, the actual pressure drop on one or both sides of the exchanger could be calculated as a function of flow. Complex mathematical formulas can be created, using syntax which is similar to conventional Spreadsheets. Arithmetic, logarithmic, and trigonometric functions are examples of the mathematical functionality available in the Spreadsheet. The Spreadsheet also provides logical programming in addition to its comprehensive mathematical capabilities

C. RECYCLE

This operation is used every time a stream is to be recycled. The logical block connects the two streams around the tear (remember that the tear does not have to be the official "recycle" stream itself, but instead should be the best place in the loop to make the break for convergence purposes). Before you can install the recycle, the flow sheet must have been completed. That means there is the need to be a value for both the assumed stream and the calculated stream. Once the Recycle is attached and running, HYS YS compares the two values, adjusts the assumed stream, and runs the flow sheet again. ASPEN HYSYS repeats this process until the two streams match within specified tolerances.

Those tolerances are set on the Parameters Page. There are tolerances for Vapour Fraction, Temperature, Pressure, Flow, Enthalpy, and Composition,it is important to ensures that the tolerances you enter are *not* absolute. They are actually multipliers for ASPEN HYSYS' internal convergence tolerances. For example, the internal value for Temperature is 0.01 degrees (note that is in Kelvin, because ASPEN HYSYS does all of its calculations in an internal unit set), so a multiplier often means the two streams must be within a tenth of a degree of each other.

On the Numerical Page, among other things, you may set the RECYC LE to either Nested (the operation is called whenever it is encountered in the flow sheet) or Simultaneous (all of the RECYCLEs are invoked).

3.3 Pinch Analysis.

This chapter presents all the steps involved in the analysis, designing and optimization of Heat Exchangers Network of Hydrodesulphurization Unit (HDS) of Kaduna Refining and Petrochemical Company and which was adopted by the Pinch analysis software (Maple). Maple was used for carrying out the heat integration of the HDS plant. The procedure involved analyzing the existing Heat Exchangers Network of the unit.

In the analysis of the existing network, a thorough study of the Process Flow Diagram of HDS was carried in order to extract all the necessary and available information required to carry out the Pinch analysis so as to come out with the minimum energy requirements or target. The parameters of interest here include, source and target temperatures of all the streams, mass flow rates of all the streams and specific heat capacities of each stream which was assumed to be constant.

The Process Flow Diagram was transformed into heat exchanger network representation, then into grid representation which is used in setting the problem table as detailed in Coulson and

Richardson (1999). Finally the design and optimization of the Heat Exchanger Network were carried as detailed in Smith (2005).

3.3.1 HEN Representation of Hydrodesulphurization Unit KRPC

The Process Flow Diagram of HDS I KRPC was carefully studied and all process streams were identified with their source and target temperatures. The stream heat capacities (which are assumed to be constant over their temperature range) were calculated from literature.

The network representation was drawn as shown in Appendix C according to procedure as detailed in Coulson and Richardson (1999).

3.3.2 Grid Representation Methodology

As mentioned earlier in chapter two, it is convenient to represent a Heat Exchanger Network in a form of grid as shown in Appendix C. The process streams are drawn, as horizontal lines with the stream number shown in square boxes. Hot streams are drawn at the top of grid and flow from left to right and cold streams are drawn at the bottom, and flow from right to left.

The stream heat capacity CP is shown in a column at the end of the stream lines, while the heat exchanger is drawn as two circles connected by a vertical line. The circles connected the two streams between which heat is being exchanged, that is, the stream that would flow through the actual exchanger. Heaters and coolers are drawn as a single circle connected to the appropriate utility. The grid representation of HDS is shown in Appendix C.

3.3.3 Problem Table Methodology

This is the preferred method of analysis because of the need to draw the composite curves and manoeuvre the composite cooling curve in order to get minimum temperature difference on the curve. The problem table is the name given by Linnhoff and Flower (1978) to a numerical method for determining the pinch temperature and the minimum utility requirements.

The procedure is as follows:-

1. The actual stream temperatures T_{act} was converted into interval temperatures T_{int} by subtracting half the minimum temperature differences from the hot stream temperature and by adding half to the cold stream temperatures.

$$\text{Hot stream} \quad T_{int} = T_{act} - 0.5\Delta T_{min} \qquad 3.15$$

$$\text{Cold stream} \quad T_{int} = T_{act} + 0.5\Delta T_{min} \qquad 3.16$$

Here the use of the interval temperature rather that the actual temperature allows the minimum temperature difference to be taken in to account.

2. Duplicated interval temperatures with bracket was noted.

3. The interval temperatures were ranked in order of magnitude showing the duplicated temperatures only once in the order.

4. The heat balance for the streams falling within each temperature interval was carried out. For the nth interval.

$$\Delta H_n = (\sum CP_c - \sum CP_h)\Delta T_n \qquad 3.17$$

Where:-

ΔH_n = net heat required in the nth interval

$\sum CP_c$= sum of the heat capacities of all the cold streams

$\sum CP_h$= sum of the heat capacities of all the hot streams

ΔT_n = interval temperature difference (T_{n-1}-T_n)

5. The heat surplus was cascaded from one interval to the next down the column of interval temperature. Cascading the heat from one interval to the next implies that the temperature difference is such that the heat can be transferred between the hot and cold streams. The presence of a negative value in the column indicates that the

temperature gradient is in the wrong direction and that the exchange is not thermodynamically possible. This difficulty can be overcome if heat is introduced into the top of the cascade.

$$\Delta H_n^{prop} = \Delta H_{n-1}^{prop} - \Delta H_n \qquad 3.18$$

6. Introduce enough heat to the top of the cascade to eliminate all the negative values.

Comparing the heat surplus with the composite curve shows that the heat introduced to the cascade is the minimum hot utility required and the heat removed at the bottom is the minimum cold utility required. The pinch occurs where heat flow into the cascade is zero.

3.3.4 Heat Exchanger Network Design for Maximum Energy Recovery

From the problem table analysis, the minimum utilities target and pinch temperatures were determined. The grid representation of the streams was divided at the pinch temperatures, which represent two regions above and below the pinch. The design was carried out as detailed in pinch design procedure of chapter two.

3.3.5 Heat Exchanger Network Optimization

The network designed above to give the maximum heat recovery and minimum consumption and cost of the hot and cold utilities may not necessarily be the optimum design for the network. The optimum design will be that which gives the lowest total annual costs; taking into account the capital cost of the system, in addition to the utility and other operating costs. (The number of exchangers in the network and their sizes will determine the capital cost).

However, there is a scope for reducing the number of heat exchangers, but the heat loads of the cooler and heater were increased in order to bring some of the streams to their target temperature. Heat would cross the pinch and the consumption of the utilities will increase.

Whether the revised network would be better, more economical, would depend on the relative cost of capital and utilities.

For any network there will be an optimum design that gives the least annual cost, capital charges plus utility and other operating costs.

For optimum design it would be necessary to cost a number of alternative designs based on compromise between the capital costs determined by the number and size of the exchanger and the utility costs, determined by the heat recovery achieved.

The steps involved in the optimization of heat exchanger network are highlighted below as stated in Coulson and Richardson (1999).

1. Start with the design for maximum heat recovery. The number of exchangers needed will be equal to or less than the number for maximum energy recovery.
2. Identify loops that cross the pinch. The design for maximum heat recovery will usually contain loops.
3. Starting with the loop with the least heat load, break the loops by adding or subtracting heat.
4. Check that the specified minimum temperature difference ΔT_{min} has not been violated, and revise the design as necessary to restore the ΔT_{min}.
5. Estimate the capital and operating costs and the total annual cost.
6. Repeat the loop breaking network revision to find the lowest cost design.
7. Consider the safety, operability and maintenance aspects of the proposed design.

For any network, there will be a best value for the minimum temperature difference that will give the lower total annual costs. The effect of changes in the specified ΔT_{min} has to be investigated when optimising a heat recovery system.

3.2.6 Economic Analysis

The correct cost data is very important for a successful economical analysis of the project. The cost needs to be annualized to study the economics in terms of yearly savings and payback period. The basic economic data consist of yearly operating hours of 7920.

3.2.7 Operating cost

The utilities required for process operation formed the operating cost of the plant. The hot utility required is flue gas which will raise the temperature of the pre flashed liquid to the feed inlet temperature, while the cold utilities required to cool the products streams are cooling water and air.

The annualized utilities cost are estimated as follows, Coulson and Richardson (1999):

Cooling Water	21.04	USD/kW
Cooling Air	12.75	USD/kW
Flue Gas	71.23	USD/kW

The cooling water and air are supplied at 25 ^{o}C, the air is raise to 40 ^{o}C, while the cooling water is returned at 60 ^{o}C (Smith, 2005).

3.2.8 Investment cost

The existing total area of the heat exchangeris were calculated from the equipment data sheet of HDS of KRPC, this comprises of process to process heat exchangers area, cooling water and air heat exchangers area and heater area.

The capital cost of heat exchangers followed the relation, Coulson and Richardson (1999).

$$\text{HE cost} = A + B(\text{Area})^{c} \qquad 3.19$$

Where A = A fixed cost of installation independent of area, B = the exchanger cost per unit area, also account for material of construction in this case is carbon steel due to low temperature operation below 200°C

The investment cost considers only the cost of extra area required to achieve the energy recovery target.

The average size of the heat exchanger is calculated from the existing smallest process to process heat exchanger.

Investment cost is estimated using:

$$\text{Investment Cost} = \Delta N(A + B\,(\Delta A/\Delta N)^{c}) \qquad 3.20$$

ΔA = Additional area required to achieve the energy target.

ΔN = number of additional shell required

3.2.9 Total cost analysis

This comprises of both Operating Cost and Investment cost.

Energy saving is calculated by subtracting the target energy cost from the existing energy cost. The energy target for both cold and hot utilities were calculated from Problem table analysis.

For costing of energy consumption, it is assumed that the target design would use utilities in the same ratio as the existing, hot utility flue gas 100%, cold utilities, air cooling 80% and water cooling 20%.

The utilities cost which are the cost of flue gas and cold utility (Demineralized water) are shown below, Coulson and Richardson (1999):

Flue Gas	71.23 USD/kWy
Cold Utilities	14.41 USD/kWy

The Maple procedures for implementing the above algorithms for carrying out pinch analysis in this work are shown in Figure 3.4 and the calculation procedure shown in Appendix B.

3.2.10 Maple Program Procedure for Running Pinch Calculation

The Maple algorithm for carrying out pinch analysis is as shown below:

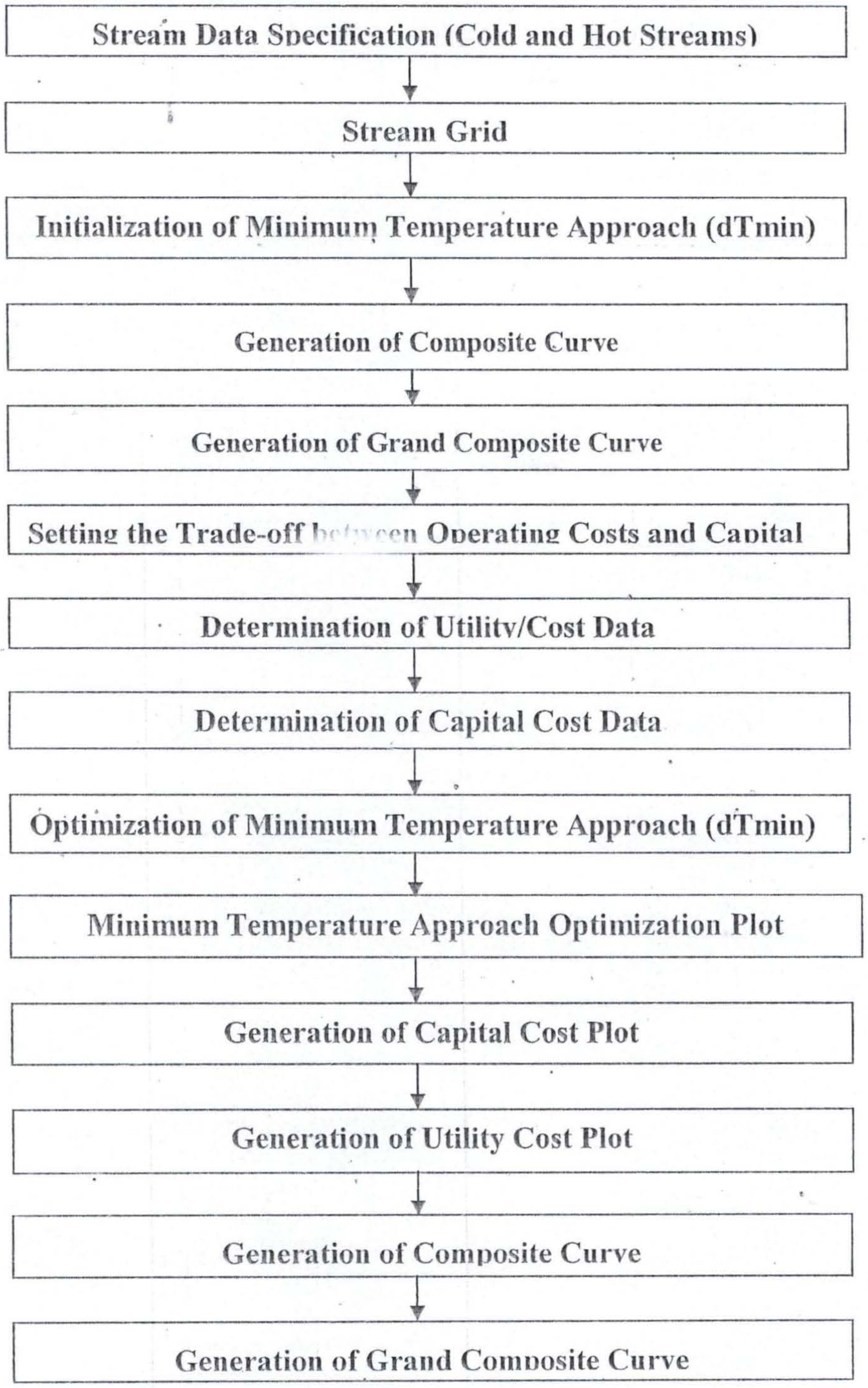


Figure 3.4: Pinch Analysis Procedural Steps using Maple

CHAPTER FOUR

4.0 RESULTS AND DISCUSSION OF RESULTS

4.1 Results

The results of the data extracted from HDS manual, laboratory data, PFD and process simulation results are shown below:

4.1.1 Data extraction and process simulation results

Table 4.1: Feed Conditions for Simulating the Process

S/N	Variables	Values
1	Temperature	120 °C
2	Pressure	3.871atm
3	Volume Flow	15 m^3/hr

Table 4.2: Feed Compositions of the Process for Simulation

Componet Volume	Volume Liquid Fraction
H2O	0.0024
H2S	0
Ammonia	0
123-MBenzene	0.0016
Benzene	0
Phenol	0.0002
Pyridine	0.0003
Quinoline	0.0002
Pyrrole	0.0004
Thiophene	0.0003
14-EBenzene	0.0001
n-Pentyl-BZ	0.0002
n-Hexyl-BZ	0.0001
n-Heptyl-BZ	0.0002
Naphthalene	0.0017
14CC6==	0.0006
1-Tridecene	0.0002
1-Tetradecen	0.0001
12MNaphthaln	0.0007
1PNaphthalen	0
cis-2-Decene	0.0002
1-Undecene	0.0001
1-C22=	0

Table 4.2: Continued

Cyclohexane	0.002
CC6one	0
33-Epentane	0.0014
33-Ehexane	0.0011
44-Mheptane	0.0003
n-Nonane	0.0001
n-Decane	0.0002
n-C11	0.0005
n-C12	0.0005
n-C13	0.0004
n-C14	0.0003
n-C15	0.0001
Nitrogen	0
S_Amorphous	0
nBMercaptan	0.0001
diE-Sulphide	0.0001
E-Mercaptan	0.0001
SULFOLANE	0.0001
NBP[0]350*	0.3359
NBP[0]319*	0.3296
NBP[0]287*	0.3172

Stream Name	Supply Temperature °C	Target Temperature °C	Heat Duty (kW)
35E01A	332	221	2803.000
35E01A	243	295	2803.000
35E01B	221	213	193.611
35E01B	203	243	193.611
35E01C	213	202	243.333
35E01C	190	203	243.333
35E01D	202	189	306.111
35E01D	173	190	306.111
35E01E	189	171	394.444
35E01E	150	173	394.444
35E01F	171	149	485.833
35E01F	120	150	485.833
35E01G	149	120	603.889
35E01G	80	120	603.889
35E01H	120	88	485.833
35E01H	34	80	485.833
35A01	87	45	756.500
35A02	172	112	1611.000
35H01	297	318	523.700
35E03A	46	68.7	2803.000
35E03A	101	65	2803.000
35E03B	68.7	97	558.6110
35E03B	140	101	558.6110
35E03C	97	120	453.3330
35E03C	170	140	453.3330
35E03D	120	145.6	313.6110
35E03D	190	170	313.6110
35E03E	145.6	182	322.5000
35E03D	210	190	322.5000
35E03E	182	217	758.3330
35E03E	255	210	758.3330
35E03F			

Table 4.3: Continued

	240	255	2577.0000
35E03F	318	298	2577.0000
35E04A	68	50	321.1110

35E04A	29	43	321.1110
35E04B	50	35	99.9170
35E04B	43	47	99.917
35E05A	240	255	897.45

4.1.2 Pinch analysis results

The results of the pinch analysis are shown below under energy target results. The result of the shifted composite curve for determining the hot and cold utilities of the plant is shown in Figure 4.1. Figure 4.2 shows the shifted composite curve for determining the pinch temperature while Figure 4.3 and Figure 4.4 shows the plot of net present cost against minimum temperature approach and target operating cost against minimum temperature approach. Figure 4.5 shows target capital cost against minimum temperature approach.

4.1.2.1 Energy target results

Table 4.4: Energy Target Values obtained from the PINCH Analysis Application

Minimum (target) hot utility requirement **(kW)**	Minimum (target) cold utility requirement (kW)	Temperature location where Minimum temperature approach (ΔT_{min}) occurs (°C)	Minimum number of heat exchangers to meet the design target
2420.51	3366.86	250	29

4.1.2.2 Optimization results

Table 4.5: Optimization Results from the PINCH Analysis Application

Optimum Minimum temperature approach(Optimum dtmin) (K)	**Net Present Cost at dTmin (million dollars)**
20	0.6234

4.1.2.3 Energy saving results

Table 4.6: Hot Minimum Utility Requirement for Traditional Energy Approach and Pinch Analysis

Hot Utility (kW) Traditional Energy Approach	Hot Utility (kW) Pinch Analysis
15,332.667	2,420.51

Table 4.7: Cold Minimum Utility Requirement for Traditional Energy Approach and Pinch Analysis

Cold Utility (kW) Traditional Energy Approach	Cold Utility (kW) Pinch Analysis
15,144.62	3366.86

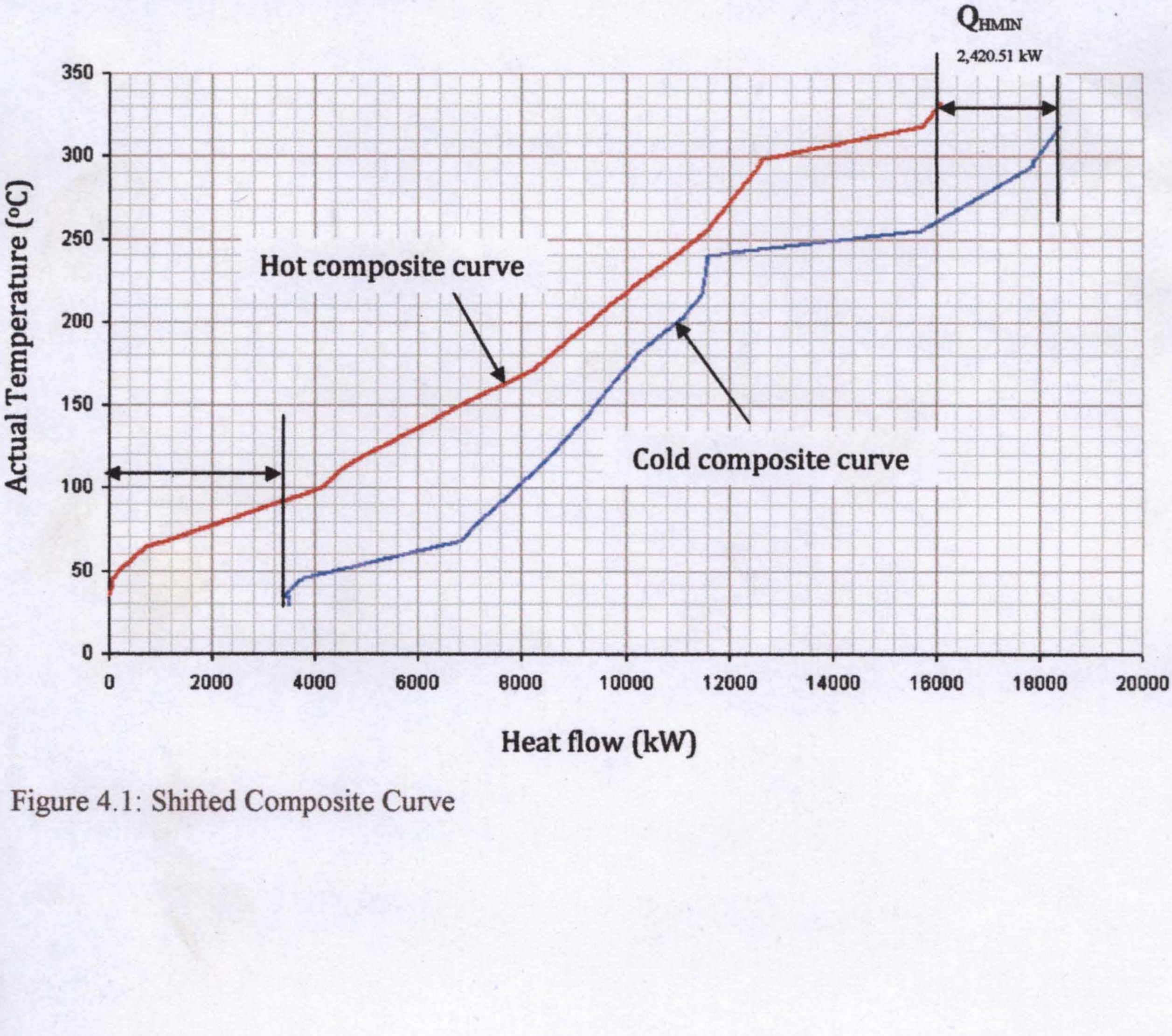


Figure 4.1: Shifted Composite Curve

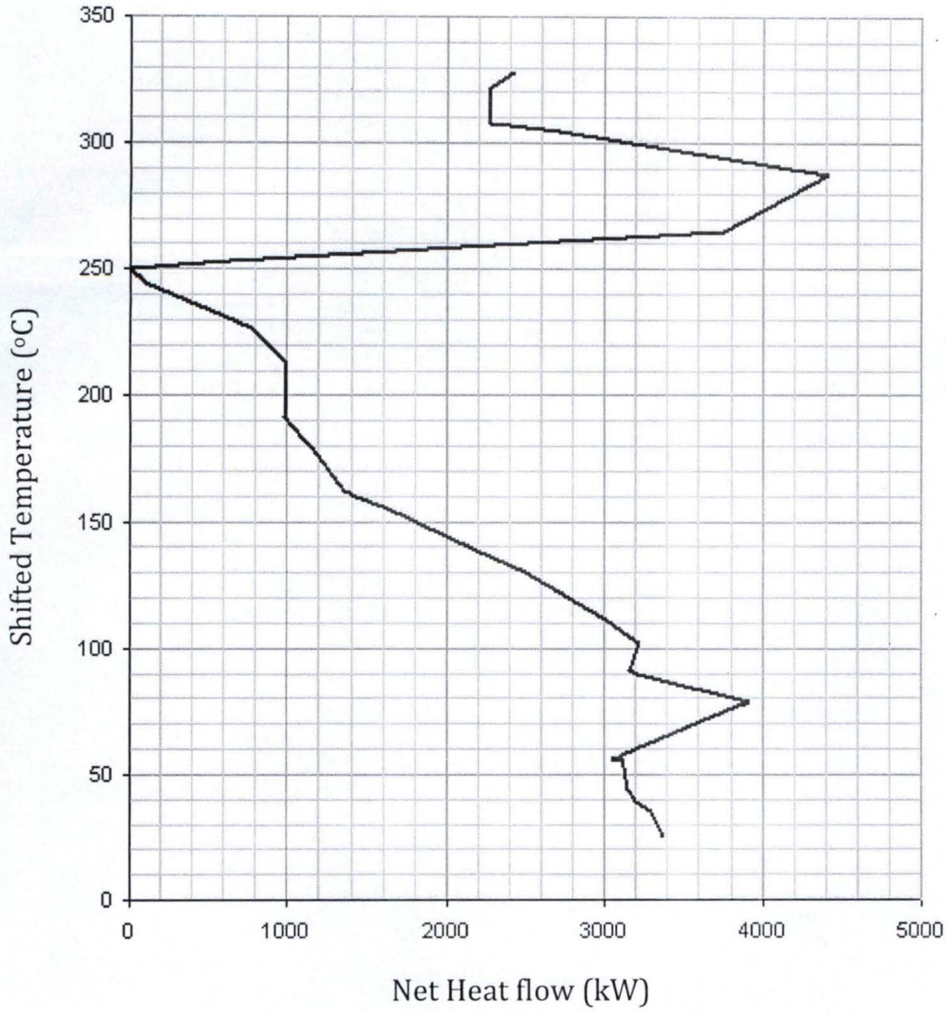


Figure 4.2: Shifted Grand Composite Curve

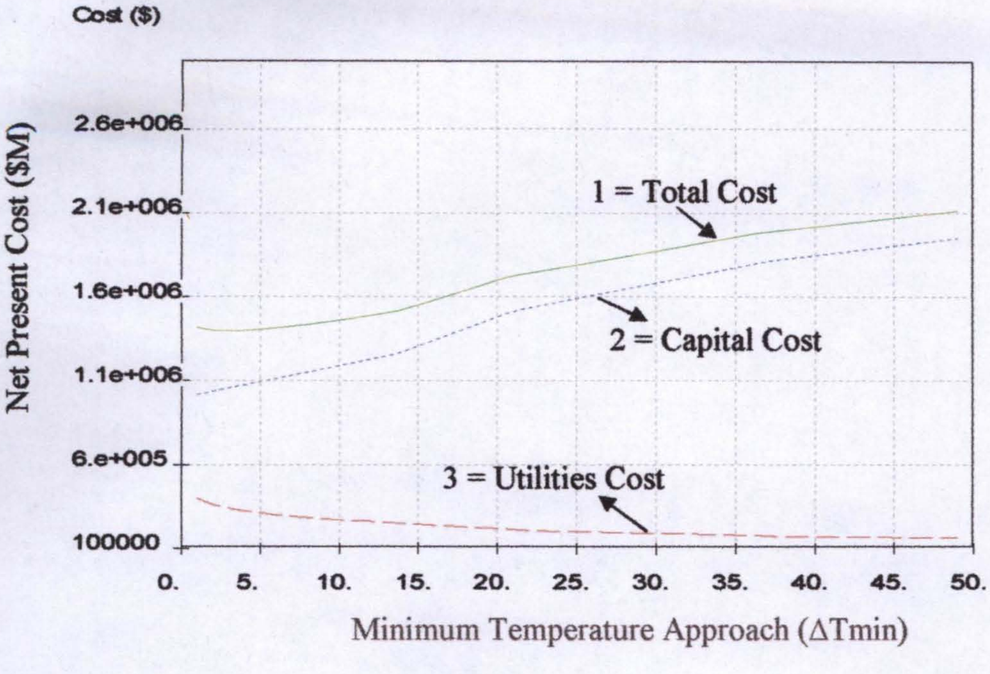


Figure 4.3: Net Present Cost against Minimum Temperature Approach (ΔTmin)

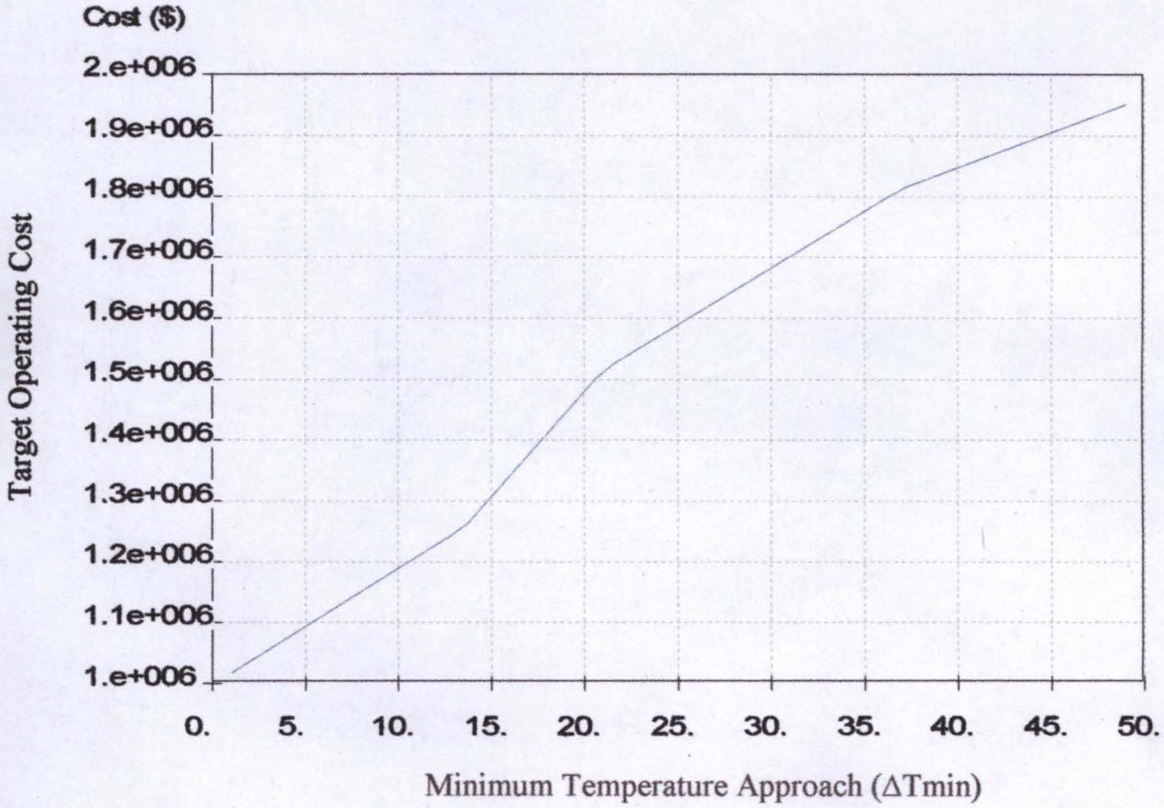


Figure 4.4: Target Operating Cost against Minimum Temperature Approach (ΔTmin)

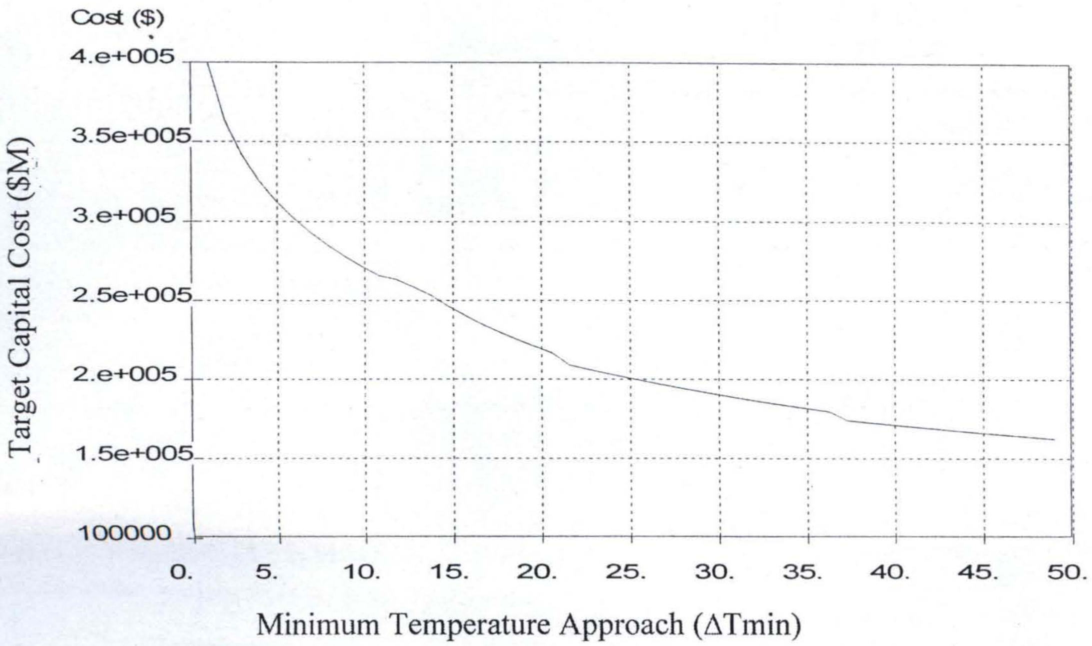


Figure 4.5: Target Capital Cost against Minimum Temperature Approach (ΔTmin)

CHAPTER FIVE

5.0 Discussion Conclusions and Recommendations — upper case

5.1 Discussion or result

5.1.1 Process simulation

The stream heat duties were computed in Aspen Aspen Hysys as shown in Table 4.4 and the values obtained were used for pinch analysis calculation to compute the cold and hot utilities and pinch point temperatures. The pinch analysis results are shown in Table 4.5 and Figure 4.2.

5.1.2 Pinch analysis

Figure 4.1 is the shifted composite curve (temperature-enthalpy) profile of heat availability in the process (the "hot composite curve") and heat demands in the process (the "cold composite curve"), plotted together. Table 4.5, Figure 4.1 and Figure 4.2 show that the heat available in the process is 2,420.51 kW while the heat demand in the process is 3366.86 kW. This is an indication that more heat is to be removed from the process than heat to be supplied to the system. Figure 4.2 (Grand composite Curve) show that the Pinch temperature of the process is 250 °C. The heat demand in the process is very high because high energy is generated in the removal of the sulphur group in the benzothiophene ring to produce Hydrogen sulphide (UOP, 1983).

Table 4.5, Figure 4.1 and Figure 4.2 indicate that the minimum utilities required for the minimum temperature approach of 20 °C are 2,420.51 kW and 3,366.86 kW of hot and cold utilities respectively, and that the pinch point occurred at 250 °C. Table 4.5 shows that the minimum number of heat exchanger required to meet the target is 29. The results show that the utility heating of the plant is far less than the utility cooling of the plant. Therefore any utility heating supplied to the process below the pinch temperature cannot be absorbed and will be rejected by the process to the cooling utility, increasing the amount of cooling utility required, hence waste of energy (cold utilities) by the HDS process.

5.1.3 Optimization of the trade-off between energy costs and capital Cost

Table 4.6 show that a minimum temperature approach (ΔT_{min}) required for the HDS plant is 20 K. This is the closest approach temperature that is allowable between two streams exchanging heat. Typically, in a petrochemical plant such as HDS plant, a value of 10 to 20 K is reasonable (Linhoff, 1983). However, the optimum value of 20 K of this parameter is significantly affected by the relative costs of energy and heat exchange area, and this is the primary parameter that was optimized in the *pinch* design program. The minimum approach temperature obtained affected both the capital costs and the operating costs. A low value of minimum approach temperature of 20 K means that hot streams approached the temperature of the cold streams more closely. The cold stream thus absorbs more heat from the hot stream. This reduces the utility heating required for the cold stream and also the utility cooling required for the hot stream, as the hot stream exits at a lower temperature after heat exchange with the cold stream. This also reduces the operating costs by lowering the utility costs, but it also increases the capital costs, since the lower approach temperature between the hot and cold streams reduces the Log Mean Temperature Difference (LMTD) in the heat exchanger. The lowered driving force and higher duty of individual heat exchangers resulted in larger heat exchanger areas being required, which increases capital costs. Similarly, a large value of the minimum approach temperature resulted in lower capital costs and higher utility (operating) costs.

5.1.4 Utility cost plot

Figure 4.4 shows that as ΔTmin increases the utilities operating cost also increased. This indicates that energy recovery cost decreases as ΔTmin increases, hence more hot utility cost is required to attain its target temperature. Likewise as ΔTmin increases, the energy removed from the hot streams decrease and this implies that more cold utilities are required to cool the streams to their target temperatures. The cost of utility for optimum ΔTmin of 20 K is \$1.5 x 10^6. This shows that the optimum cost of utilities required for the HDS plant is \$1.5 x 10^6.

5.1.5 Capital cost plot

Figure 4.5 shows that as ΔTmin increases the capital consumption cost also decreases. There was a sharp decrease in the capital cost from \$0.4 x 10^6 to \$0.2 x 10^6 as the ΔTmin increased from 1 to 25 K. The capital cost then decreased gradually from to \$0.2 x 10^6 to \$0.16 x 10^6 as the ΔTmin increased from 25 to 50 K. The cost of utility for optimum ΔTmin of 20 °K is \$0.22 x 10^6. This shows that the optimum cost of heat exchangers required for the HDS plant is \$0.22 x 10^6.

5.1.6 Energy saving between the traditional energy approach and Pinch Technology

The cold utility requirements of traditional energy approach and pinch analysis shown in Table 4.7 are 15,144.62 kW and 3366.86 kW respectively while the hot utility requirements of traditional energy approach and pinch analysis shown in Table 4.9 are 15,332.667 kW and 2,420.51 kW respectively. This shows that pinch analysis energy integration saves more energy and utilities cost than the traditional energy approach. This statement is in agreement with literature (Smith, 2005) which states that pinch analysis as an energy integration technique saves more energy than the traditional energy technique.

5.2 Conclusions

The following conclusions were drawn from the result of the analysis carried out on HDS unit of KRPC.

1. Within the range of minimum approach temperature 10 – 50 °C analysed the best minimum approach temperature was found to be 20 °C.
2. The utilities targets for the minimum approach temperature were found to be 2,420.51 kW and 3366.86 kW for hot and cold utilities respectively, whereas in traditional energy approach they were 3366.86 kW and 15,144.62 respectively.
3. The utility and capital cost for optimum ΔT_{min} of 20 °C are \$1.5 x 10^6 and \$0.22 x 10^6 respectively.
4. Pinch analysis as an energy integration technique saves more energy and utilities cost than the traditional energy technique

5.3 Recommendations

Based on the analysis carried out the followings are recommended:

1. The Nigerian National Petroleum Corporation should carry out retrofit of HDS of Kaduna Refining and Petrochemicals Company in order to increase its profitability.
2. The analysis should be carried out in other units (PACOL, HF and MOLEX units) of the petrochemical in order to check the validity of the design, so as to save energy cost
3. This analysis should be carried out in other Nigerian refineries in order to determine their level of energy efficiency and cost effectiveness.

4. The utilities unit of the refinery should also be integrated, so as to optimize heat recovery and generation; this will save a lot of operational cost and environmental pollution.

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APPENDIX A

A.1 Process Simulation Procedure.

The procedure for the simulation using Aspen Hysys is shown below.

1. The start button was clicked using the mouse
2. The "all programs" button was clicked which displayed a pop-out menu.
3. From the pop-out menu displayed, Aspen Hysys was clicked to launch the Aspen Hysys program displaying the Aspen Hysys windows.
4. The "tools" menu was clicked to drop down the menu list and "preference" was clicked to give the "preference" dialogue box.
5. On the preference dialogue box, the variable tab was clicked and the units selected to display the system of units available in Aspen Hysys.

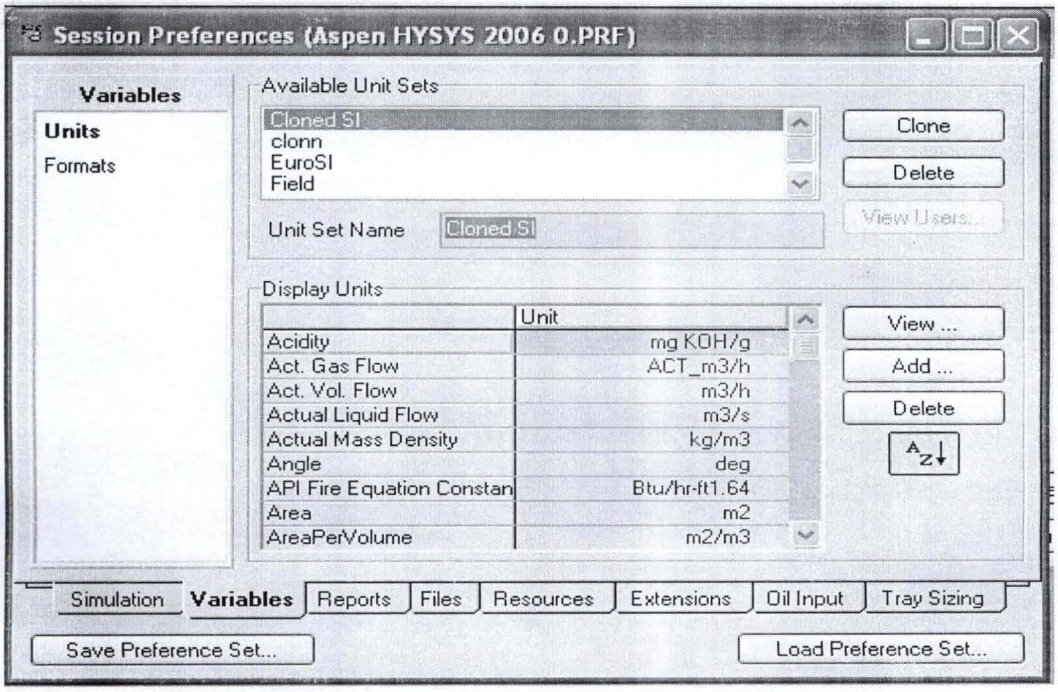


Figure A.1: A Preference Dialogue Box

A.2 Addition of a Property Package

1. A **New** Case icon was selected by selecting the new case icon under file menu.

2. The fluid package was created by clicking the Add button on the Fluid Package tab. On the displayed fluid package the default fluid package Basis 1 was changed to PRSV. This is shown in Figure A.2.

3. The **View** button in the Component List Selection section of the **Set Up** tab was clicked.

4. The **PRSV** Equation of State model was selected.

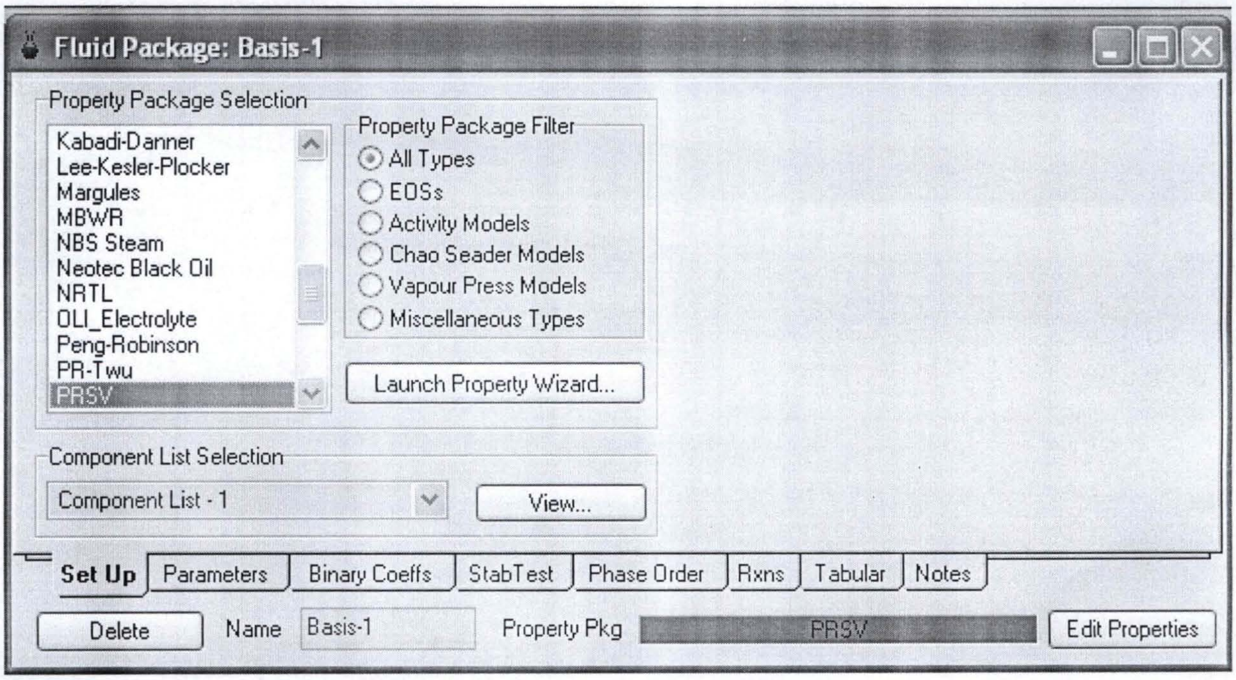


Figure A.3 Fluid Package Dialogue Box

3.3 Addition/Selection of Components

Components for the simulation were selected using Match Cell method. To do this:

1. Full Name/Synonym was selected.

2. The Match cell and the name of the component were entered.

3. Once the desired component was highlighted,

• The enter key was clicked

•The Add Pure button was also clicked.

• The components were added to the simulation case by double clicking on the component.

4. The Library components selected are shown in Figure A.4.

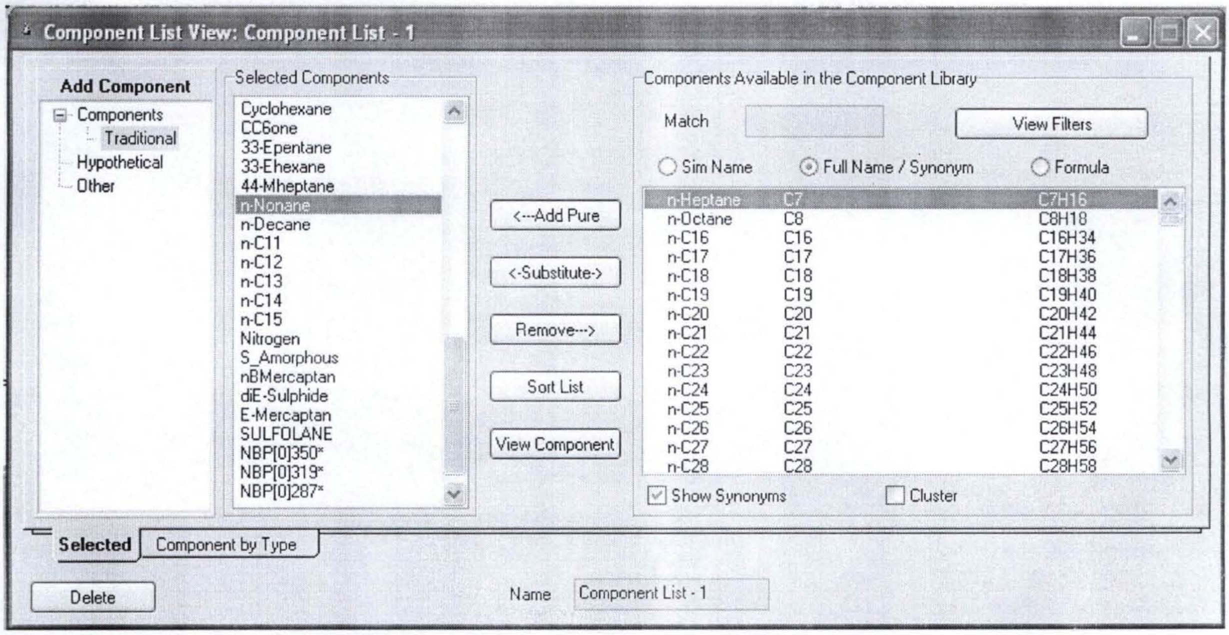


Figure A.4: Add Components Dialogue

A.4 Creation of Non Library Components

The non-Library components for the simulation are created by using Hypothetical Tab. To do this, three base properties are needed, they are Molecular Weight, Boiling point and Ideal liquid density.

1. Molecular weight of 71.06 kg/kgmol.
2. Boiling point of 40°C and,
3. Ideal liquid Density of 635.4kg/m^3 was used.

Then, Estimate Unknown Properties were clicked and other properties like Critical temperature, Critical Pressure, Critical volume, and acentricity were calculated. This is shown in Figure A.5.

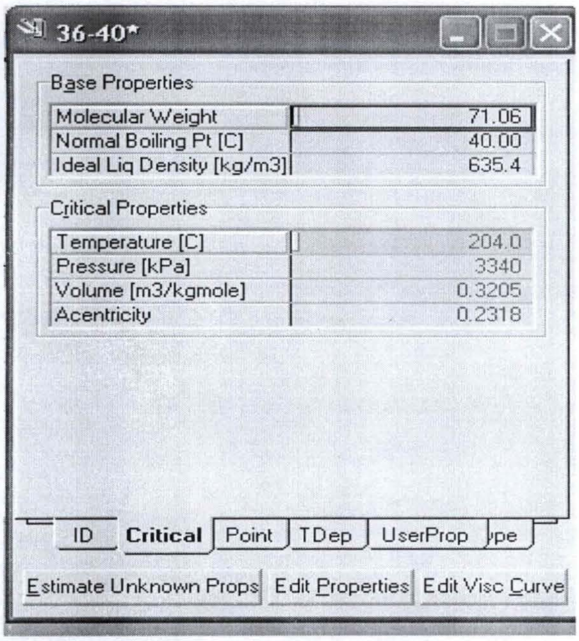


Figure A.5: Hypo component Dialogue

The same procedures were used to create other non library components. These are shown in Figure A.5 and A.6.

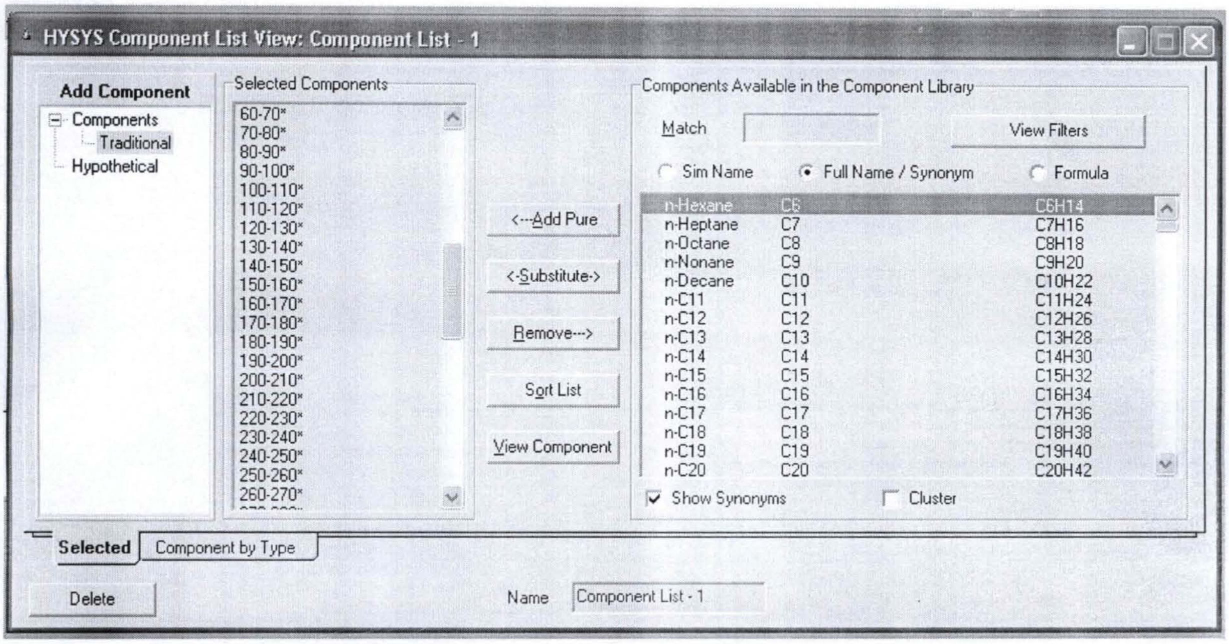


Figure A.6: Group Hypo components Dialogue

4. The Simulation Basis Manager was changed to Simulation Environment by clicking on Enter Simulation Environment tab

A.5. Defining Reactions

The reaction is:

Pyridine + $5H_2$ = n Butane + NH_3

Selecting the Reaction Components

From **Rxn Components**
The **Add component** was pressed.
Add This Group of Components was pressed.
The **Simulation Basis Manager** was returned to
A.6 Creating the Reaction
In the **Reaction, Add Rxn** was pressed
Conversion Rxn type was selected.
Add Reaction was pressed.
The **Conversion Reaction Rxn-1** page was displayed
In the **Component** column, the **Add** component was clicked on.
Select **Pyridine** was selected from the dropped **Drop-Down** arrow.
This was repeated for **Hydrogen, n-Butane** and Ammonia.
In the Stoichiometry **Coefficient** Corresponding to Pyridine

-1 was typed for **Pyridine**.

-5 was typed for **Hydrogen**
1 was typed for **n-Butane**
1 was typed for **Ammonia**

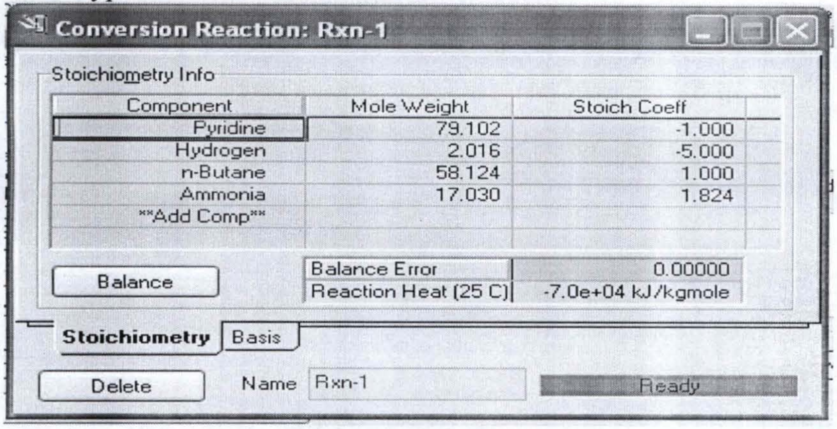


Figure A.7: The conversion reaction dialog box
It is noted that the **Balance Error** is 0.0 indicating that the reaction was mass balanced.
The next step was to move to the **Basis** tab.
The **Basis**
Rxn phase = Vapour phase
For the **Base Component, Pyridine** was chosen.
% Conversion Co=15 %
The status indicator at the bottom of the **Conversion Reaction** property view changes from **Not**

Ready to **Ready**, indicating that the Reaction was completely defined.

The **Conversion Reaction** property view was closed.
The **Reactions** view was closed.
The above step was repeated for other reactions
Creating a Reaction Set
The **Add Set** button in the **Reaction** Sets group was clicked on.
In the **Active List** for the cell called <**empty**>from the **Drop-Down** arrow, Rxn-1 was selected.
The Rxn-1 was renamed to Hydrodesulphurization reaction
The **Close** button was pressed.
Making the Reaction Set Available to the Fluid Package
Set-1 in the **Reaction Sets** on the **Reactions** tab was clicked on.
The **Add to FP** button was pressed; the **Add "Set-1"** view appeared.
The **Add Set to** Fluid Package button is pressed on.
. At this point, the work was saved as a **New Case** and a name Design Data was given to it as a choice.

A.8 Addition of the steam stream

A stream with the following data was added.

Table A.1: Feed Steam Stream Specification

In this cell	Enter
Name	Raw Kerosene
Temperature	36^0C
Pressure	15 kPa
Ideal Volumetric Flow	25 m^3/hr

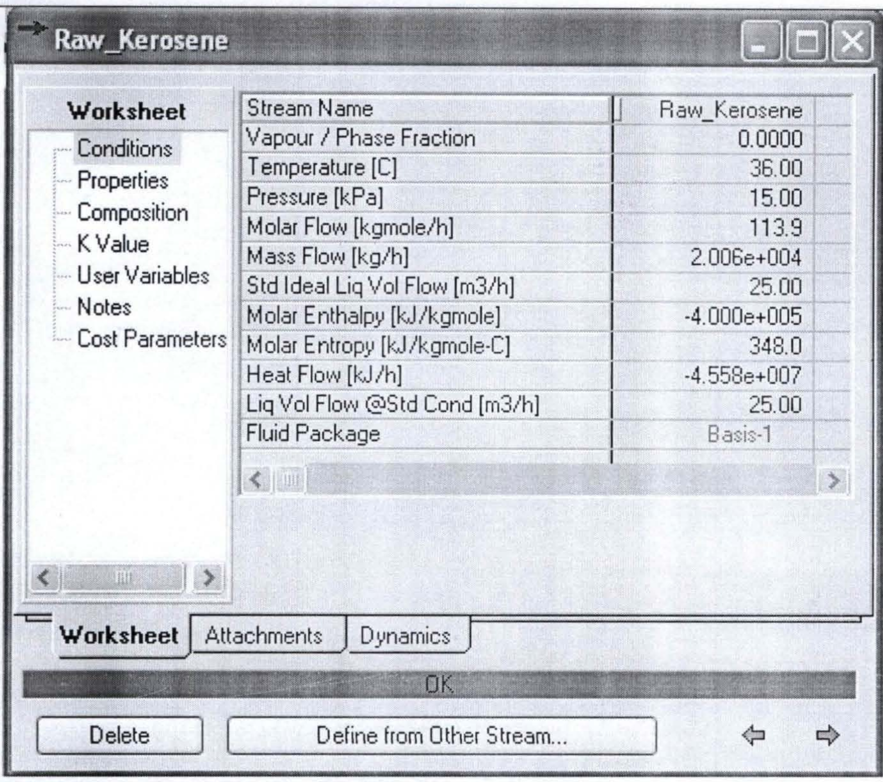


Figure A.8: Feed Stream Dialogue

Table A.2: Feed Steam Stream Specification

In this cell	Enter
Name	From Wash Water
Temperature	20^0C
Pressure	5410kPa
Ideal Volumetric Flow	2 m^3/hr

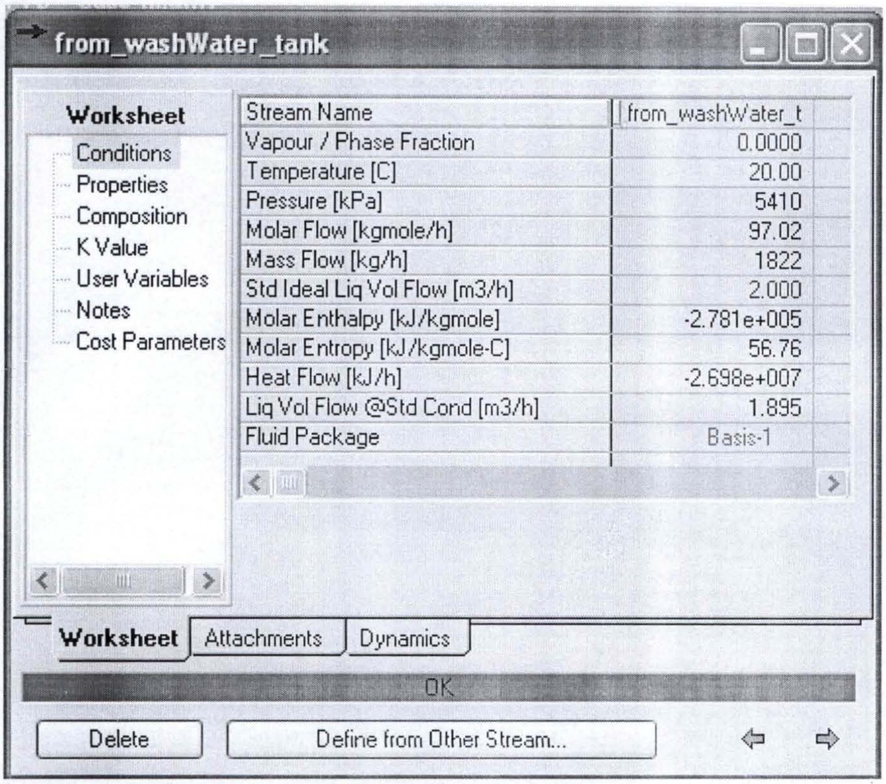


Figure A.9: Feed Steam Stream Specification

A.9 Modeling of the Hydrosulphurization Unit.

A.9.1 Three Phase Separator

3 Phase Separator was added from the unit operation palette and the following information were entered:

1. The design tab was selected and the connection tab was also clicked on to show the connections page.
2. The Inlet and outlet connections of the **3 Phase Separator** were entered.
3. The inlet stream Raw Kerosene was selected.
4. The outlet streams were named SepVap, Kero and SepLiq to represent Vapor, Light Liquid and Heavy Liquid respectively.
5. On the connections page, the pressure drop was specified as 0kpa.

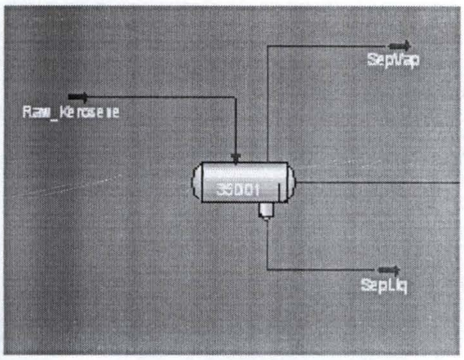


Figure A.10 Three Phase Separator Unit window

A.9.2 Pump

Pump was added from the Unit Operation palette and the following information were entered:

1. The design tab was selected and the connection tab was also clicked on to show the connections page.
2. The Inlet and outlet connections of the pump were entered.
3. The inlet stream From Wash water Tank was selected.
4. The outlet stream was named wash water.
5. On the worksheet the discharge pressure was specified as 6000kpa.

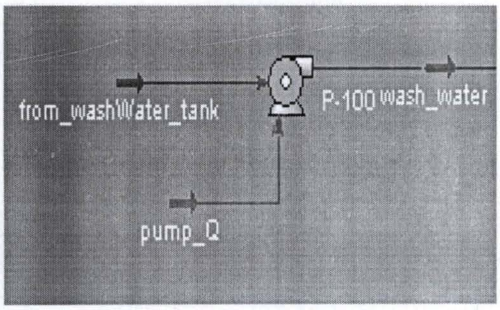


Figure A.11 Pump Unit window

A.9.3 Heat Exchanger

Heat Exchanger was added from the Unit Operation Palette and the following information were entered:

1. The design tab was selected and the connection tab was also clicked on to show the connections page.
2. The Inlets and outlets connections of the **heat exchanger** were entered.
3. The hot fluid inlet stream was named mixed product.
4. The Cold fluid inlet stream was named mixed kero.
5. The hot fluid and cold fluid outlet streams were named preheated kero and cool mixed product.
6. On the connections page, the pressure drop was specified as 10kpa for the Tube side and Shell Side.

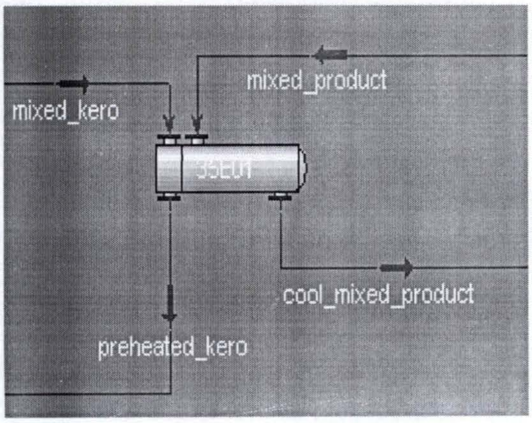


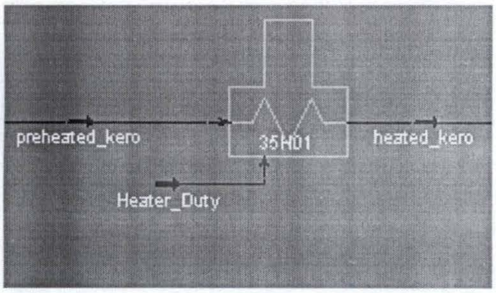
Figure A.12 Heat Exchanger Unit window

A.9.4 Furnace

Furnace was added from the Unit Operation Palette and the following information were entered:

1. The design tab was selected and the connection tab was also clicked on to show the connections page.
2. The Inlet and outlet connections of the furnace were entered.

3. The inlet stream was named Preheated Kero.
4. The outlet stream was named heated Kero.
5. On the worksheet the temperature of the furnace was specified as 320°C.



. **Figure A.13 Furnace Unit window**

A.9.5 Conversion Reaction

Conversion reactor was added from the Unit Operation Palette, and the following information were entered:

1. The design tab was selected and the connection tab was also clicked on to show the connections page.
2. The Inlet and outlet connections of the **conversion reactor** were entered.
3. The inlet stream was named heated Kero.
4. The outlet streams were named Rxt Vap and Rxt Liq to represent Vapor and Liquid products respectively.
5. On the connections page, the pressure drop was specified as 8.5kpa.

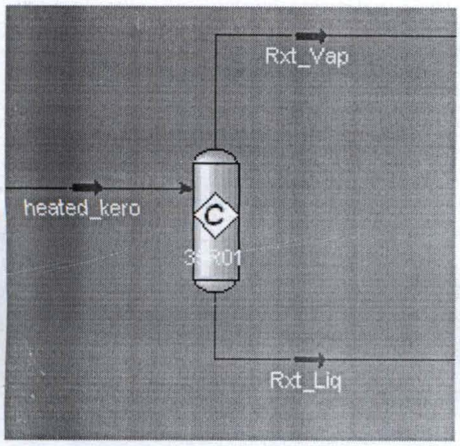


Figure A.14 Conversion reactor Unit window

A.9.6 Modeling of the Shortcut Distillation Column

A.9.7 Addition of Shortcut Distillation Column

Shortcut Distillation Column was added from the unit operation palette and the following information were entered:

Table A.3 Shortcut Distillation Specification

In this cell	Enter
Name	Shortcut Distillation Column
Condenser Duty	S-CondDuty
Distillate	S-Distillate

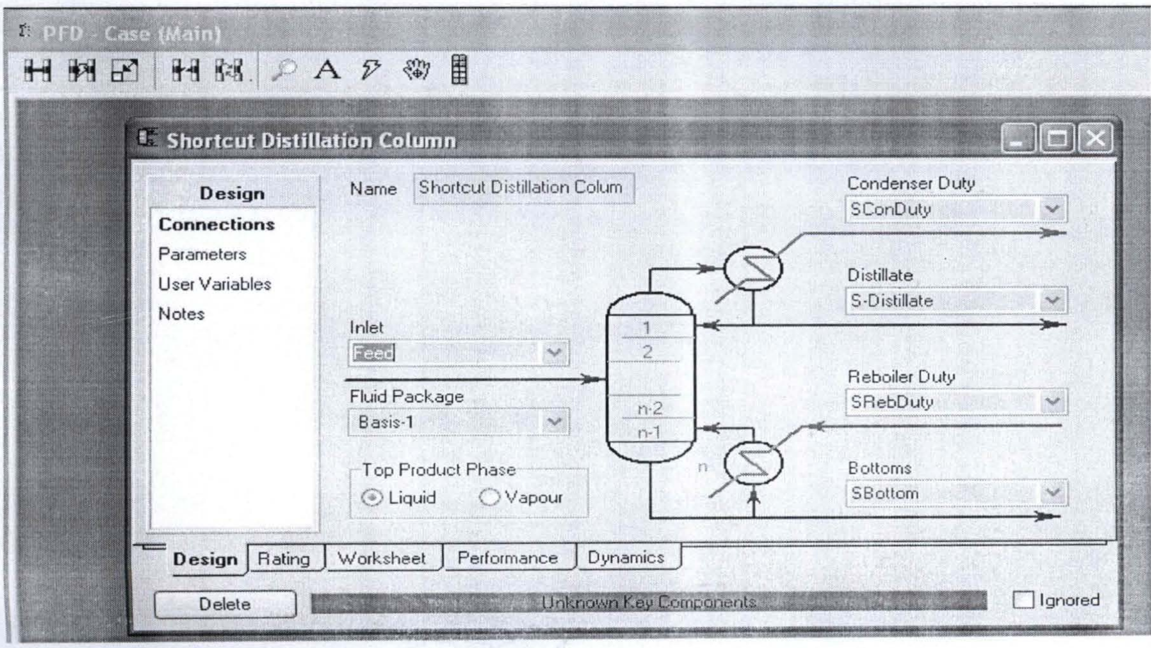


Figure A.15 Shortcut Distillation Column window

1. The design tab was selected and the connection tab was also clicked on to show the shortcut column connections.
2. The Inlet and outlet connections of the column were entered.
3. The parameter tab was selected and the key components, condenser pressure, reboiler pressure and external reflux ratio were specified as shown below

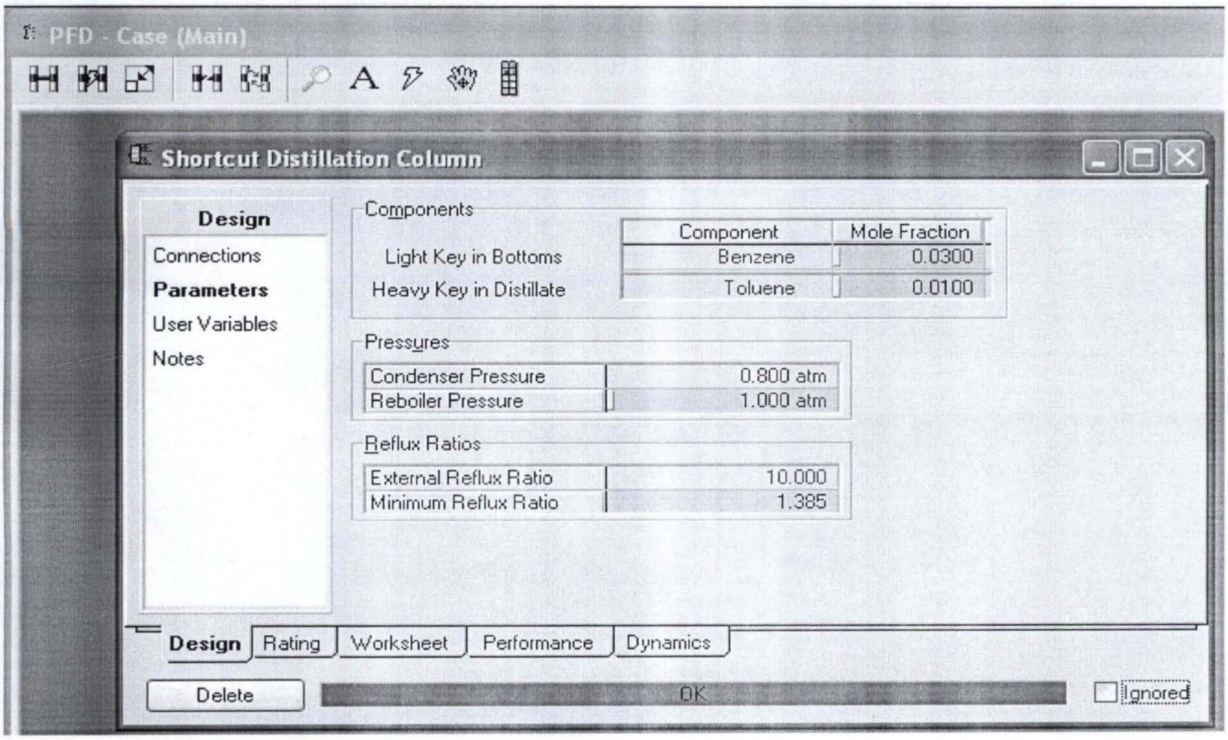


Figure A.16 Shortcut Distillation Column parameters

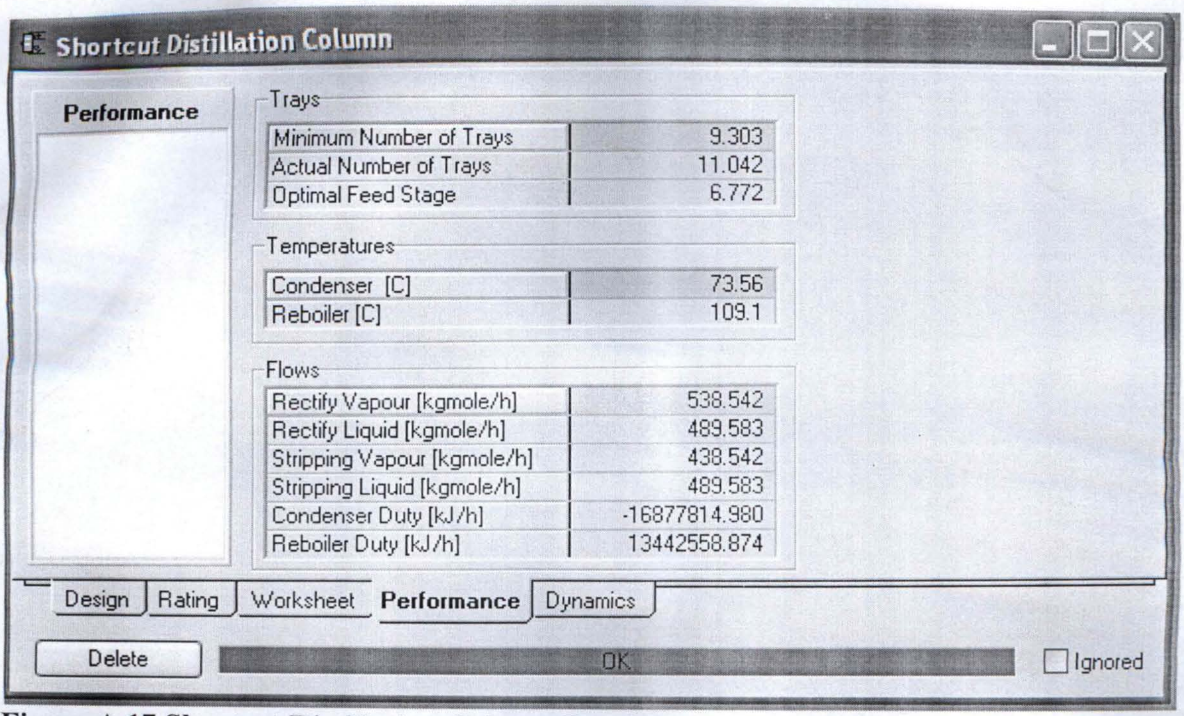


Figure A.17 Shortcut Distillation Column Results

A.9.8 Modeling of the Rigorous Distillation Column

A.9.9 Addition of Rigorous Distillation Column

Rigorous Distillation Column was added from the unit operation palette and the following information was entered:

Table A.4 Rigorous Distillation Specification

In this cell	Enter
Name	Rigorous Distillation Column
Inlet Stream	Feed
Stages	11
Condenser Energy Stream	R_{Con}Duty
Condenser	Total
Reboiler Energy Stream	R_{Reb}Duty

Ovhd Outlet	Column Ovhd
Bottoms Liquid Outlet	Column Bottom

1. The connections to the distillation column was specified as shown below:

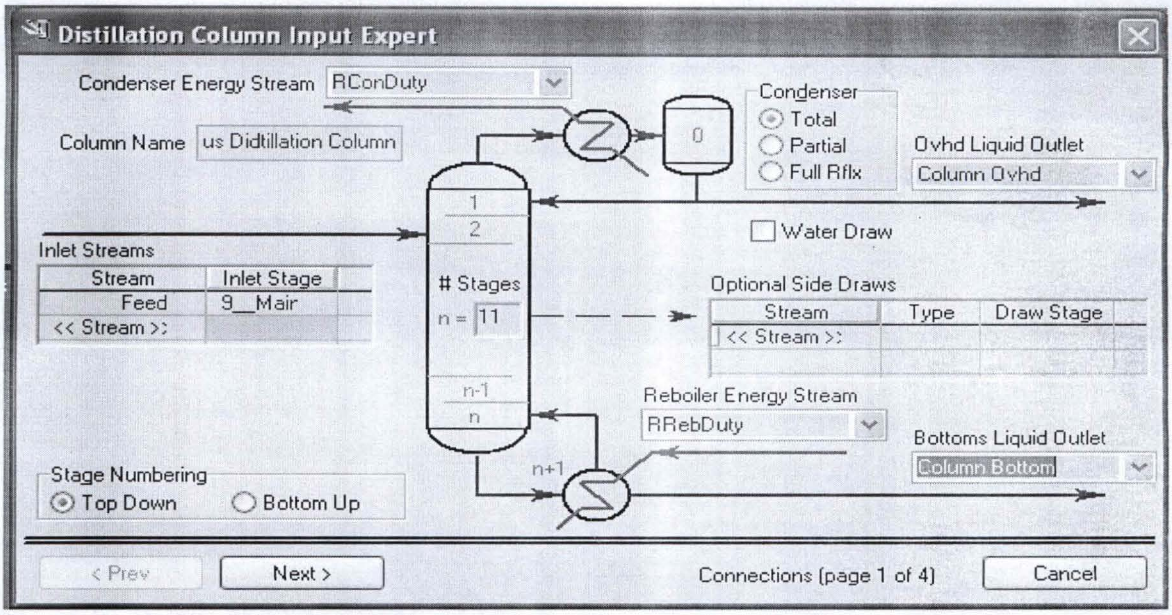


Figure A.18: The Distillation Column Input Expert showing the connections.

2. The Condenser and reboiler pressure were specified as shown below:

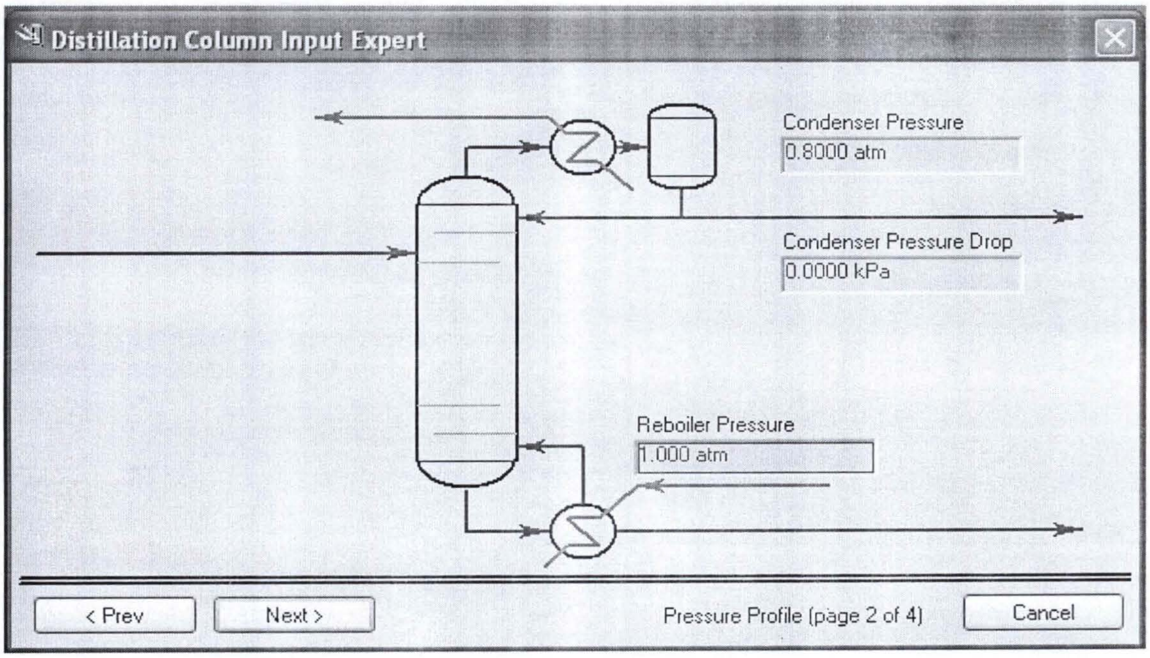


Figure A.19: The Distillation Column Input Expert showing the condenser and reboiler pressure.

3. The Condenser and reboiler temperature were specified as shown below:

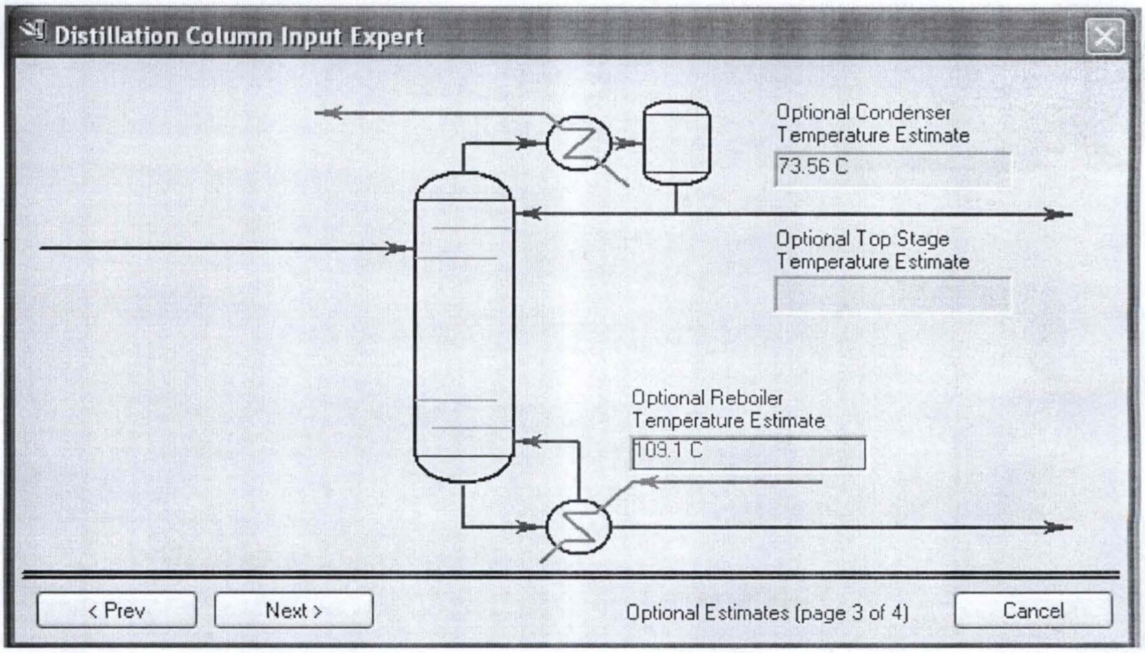


Figure A.20: The Distillation Column Input Expert showing the condenser and reboiler temperature.

4. The distillate liquid feed rate and reflux ratio were specified as shown below:

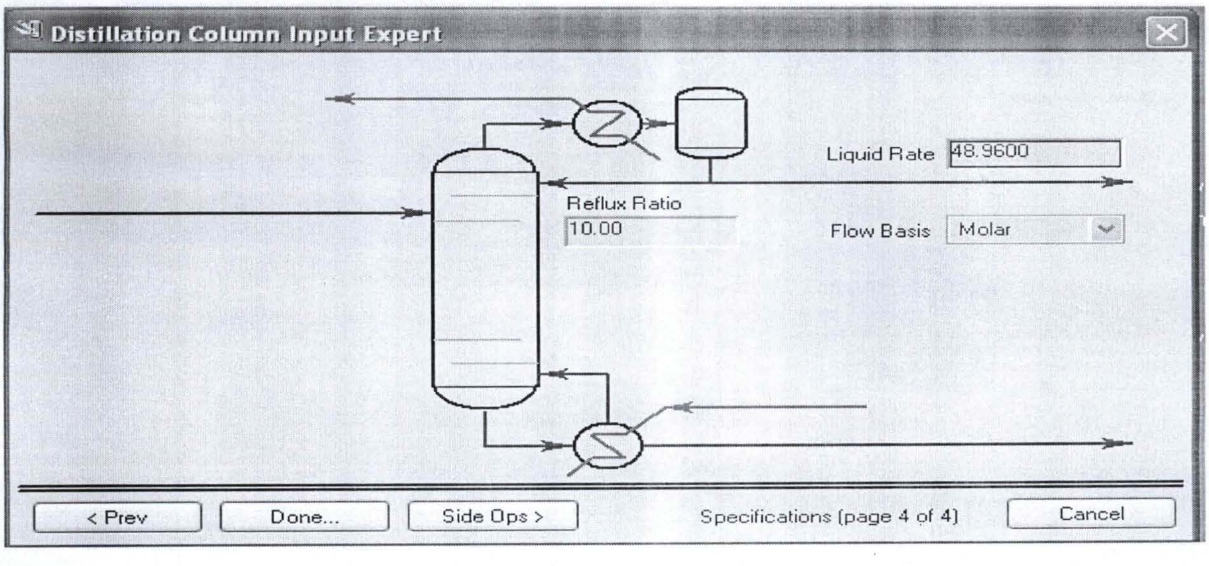


Figure A.21: The Distillation Column Input Expert showing the Liquid Rate and Reflux Ratio.

5. The Done button was clicked on and the following interface was displayed.

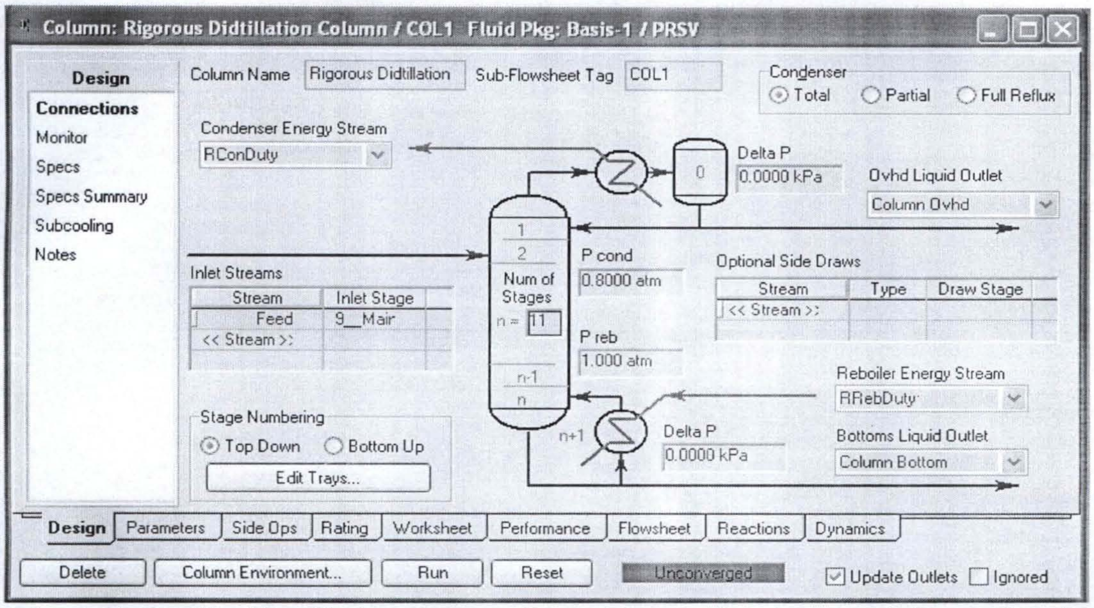


Figure A.22: The Distillation Column Input Expert showing the Column Conditions.

6. The Run button was clicked on and the converged simulation screen was displayed.

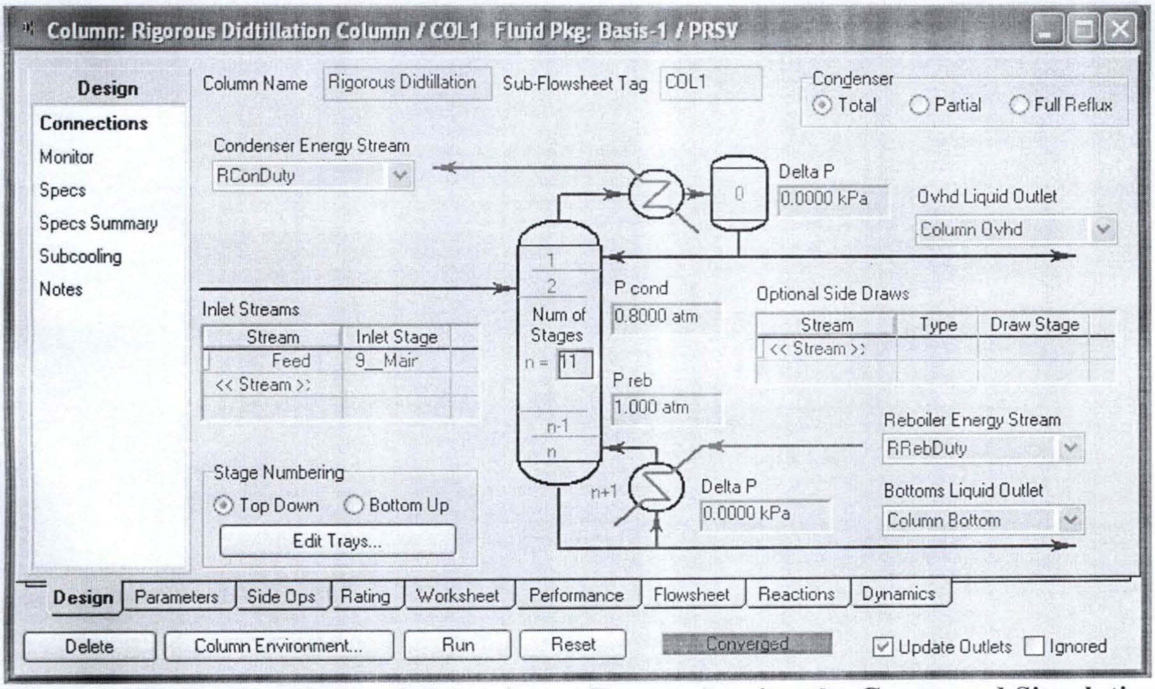
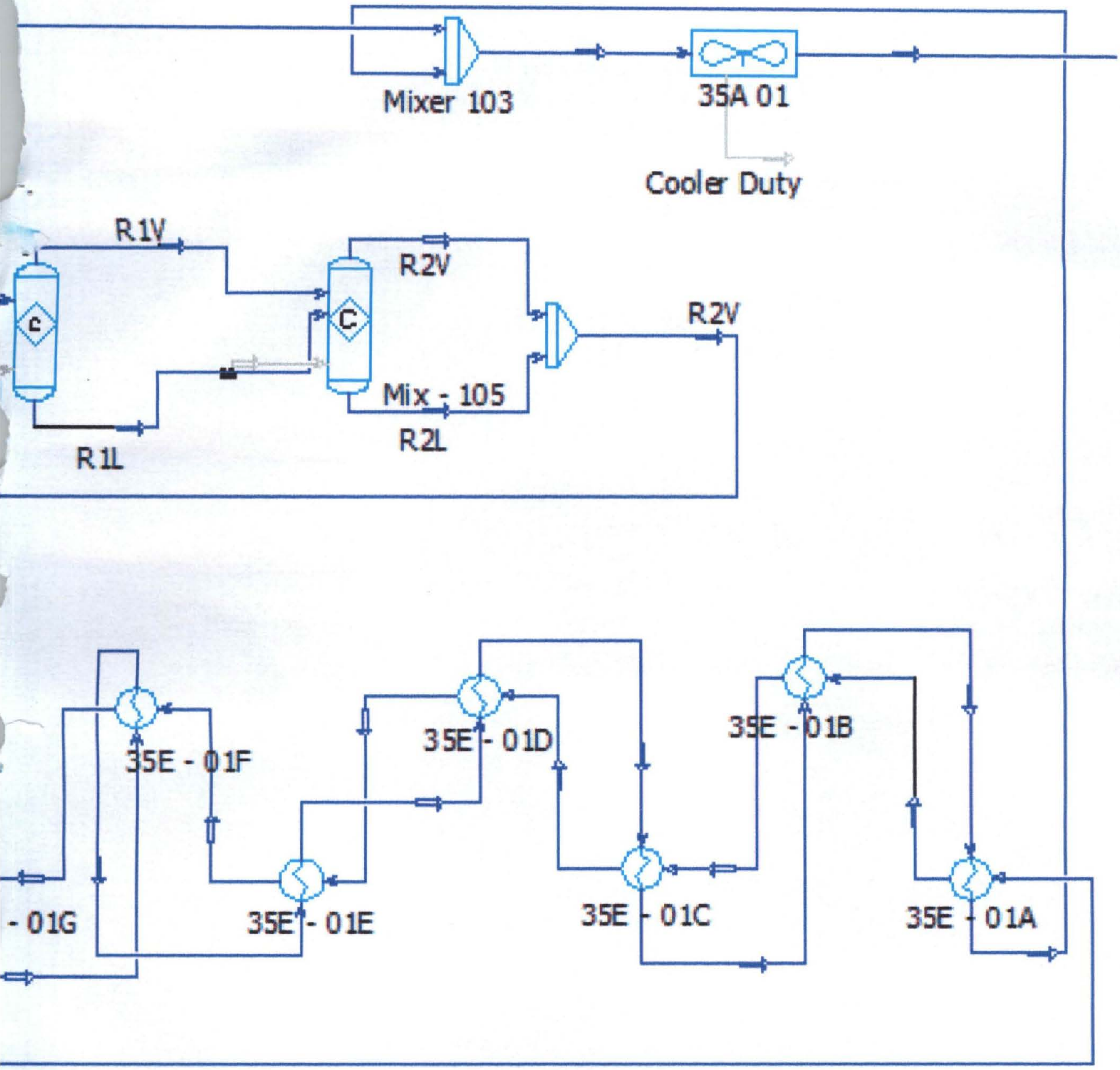
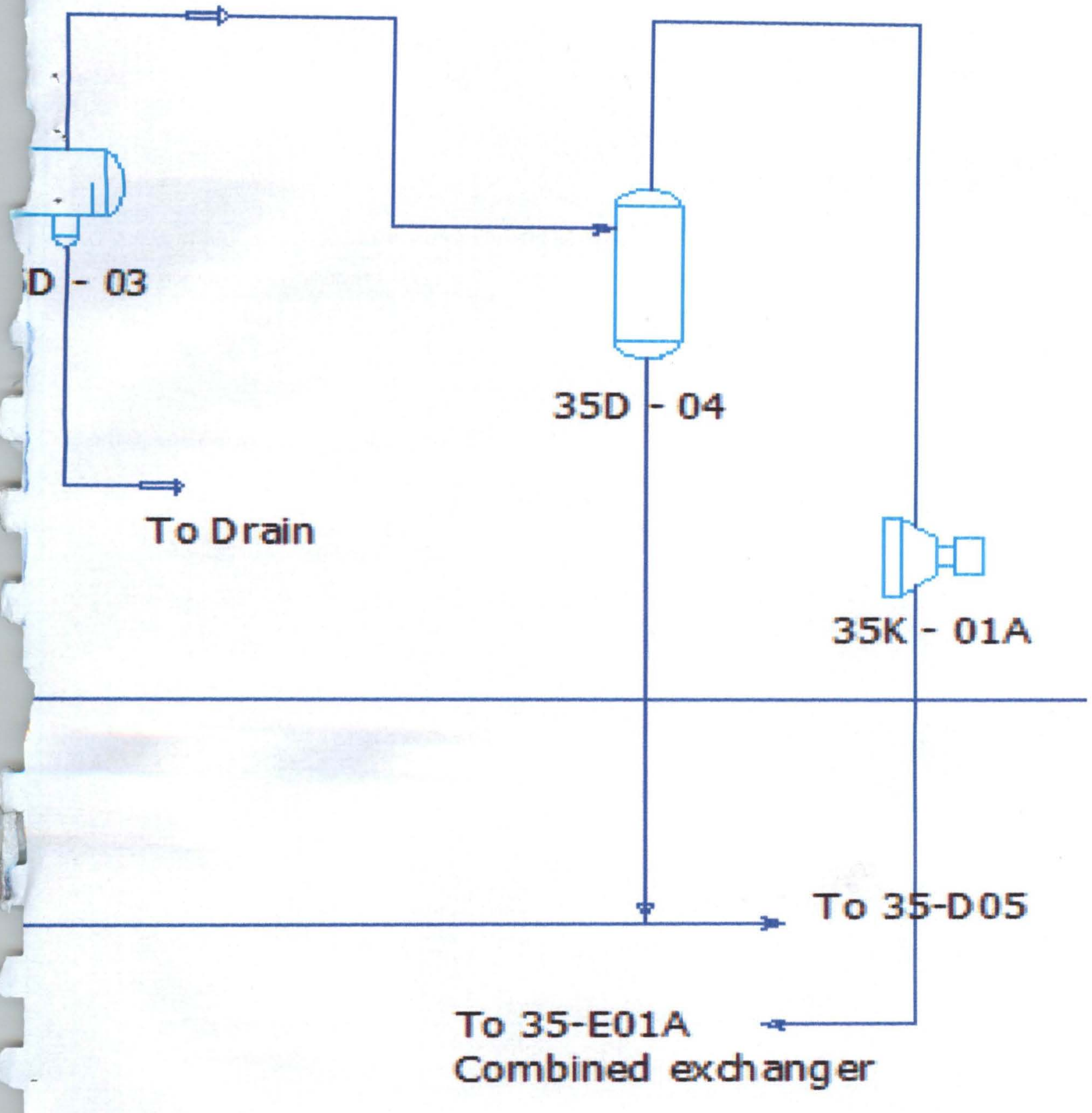


Figure A.23: The Distillation Column Input Expert showing the Converged Simulation.

REACTION SECTION



ection of Hydrodesulphurization Unit



ion Section of Hydrodesulphurization Unit

APPENDIX B

B.1 Stream Data Specification (Cold and Hot Streams)

The heat exchanger network has the task of heating and/or cooling a collection of process streams. For example, a hot water stream may require cooling before it is sent to a waste treatment facility. The engineer must first define all of the stream requirements. The heating/cooling requirements are met by interchanging the heat between streams or by using utilities.

Cold streams

The cold streams are the streams that were heated to a higher temperature. Each cold stream was assigned a stream ID as a sequential integer starting at 1. Four more values were specified for each cold stream:

C_P (kw/K), stream heat capacity x stream flow rate

T_{supply} (K), stream supply (initial) temperature

T_{target} (K), stream target (required) temperature which is higher than Tsupply

h (kw/sqm-K), estimated heat transfer coefficient for the stream within a heat exchanger

The cold stream data arrays were defined as follows:

Stream ID, C_P (kw/K), T_{supply} (K), Ttarget (K), h (kw/m^2-K)

Hot streams

The hot streams are the streams that were cooled to a lower temperature. Each hot stream was assigned a stream ID as a sequential integer following the assignment of cold stream numbers. Four more values were specified for each hot stream:

C_P (kw/K), stream heat capacity * stream flow rate

T_{supply} (K), stream supply (initial) temperature

T_{target} (K), stream target (required) temperature which is lower than Tsupply

h (kw/sqm-K), estimated heat transfer for the stream within a heat exchanger

The hot stream data array were defined as follows:

Stream ID, C_P (kw/K), T_{supply} (K), T_{target} (K), h (kw/m^2-K)

B.2 Stream grid

The stream grid gives a visual representation of the heating and cooling requirements. It contains the information from the stream tables along with the calculation of the heat requirement for each stream as H (kw) = CP * (Ttarget-Tsupply). The stream grid also provides a good tool for designing the heat exchanger network. The following command was typed under stream grid in maple code to display the Stream Grid:

```
> StreamGrid();
```

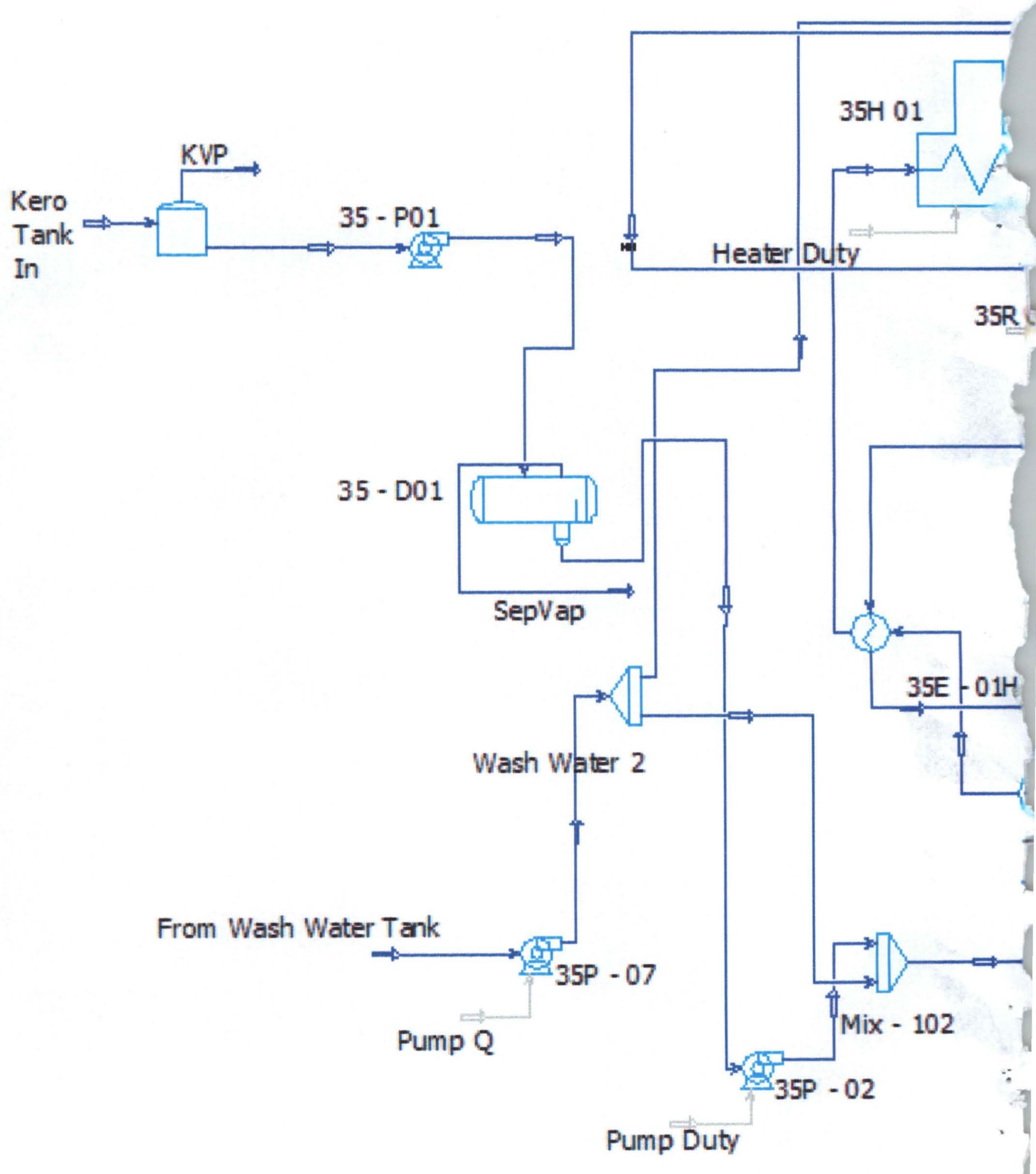


Figure 3.3: Process Flow Diagram of Reactio

B.3 Problem Definition Table

Pinch analysis defines the minimum temperature approach that occurs within the network as the Delta T min. Its value sets the heating and cooling targets for the network.

The minimum utility duties were calculated via a problem definition table. The hot stream temperatures were shifted by -1/2 dTmin and the cold stream temperatures were shifted by 1/2 dTmin. The hot streams were combined into the hot composite curve and the cold streams were combined into the cold composite curve. The problem data table then calculated the pinch temperature where dTmin was achieved along with the utility targets.

dTmin (K), delta T minimum of pinch analysis

QH (kw), minimum (target) hot utility requirement

QC (kW), minimum (target) cold utility requirement

Pinch T (K), temperature location where dTmin occurs

Umin, minimum number of heat exchangers to meet the design target

`The following command is typed in the maple program to calculate` dTmin (K), QH (kw), QC (kw), Pinch T (K) and Umin:

```
> dTmin:=10;  # 10 K is a common starting point for the
design
ProblemDataTable(dTmin);
```

B.4 Composite curves

The results from the problem definition table were commonly plotted as composite curves in shifted temperatures as shown in figure A.1. The hot composite curve shown in red and the cold composite curve shown in blue on the plot. The two curves touched at the pinch temperature. QC is the horizontal distance between the starting points of the composite curves and QH is the horizontal distance between the end points of the two composite curves. QC and QH were provided by cold and hot utilities. Heat was interchanged between streams where the curves overlap.

The following command was typed in the maple program to display the composite curves:

```
> p1:=plots[pointplot](HCdata,connect=true,color=red):
p2:=plots[pointplot](CCdata,connect=true,color=blue):
plots[display](p1,p2,title=CCtitle,labels=[`H(kw)`,`Tsh
ifted(K)`]
> );
```

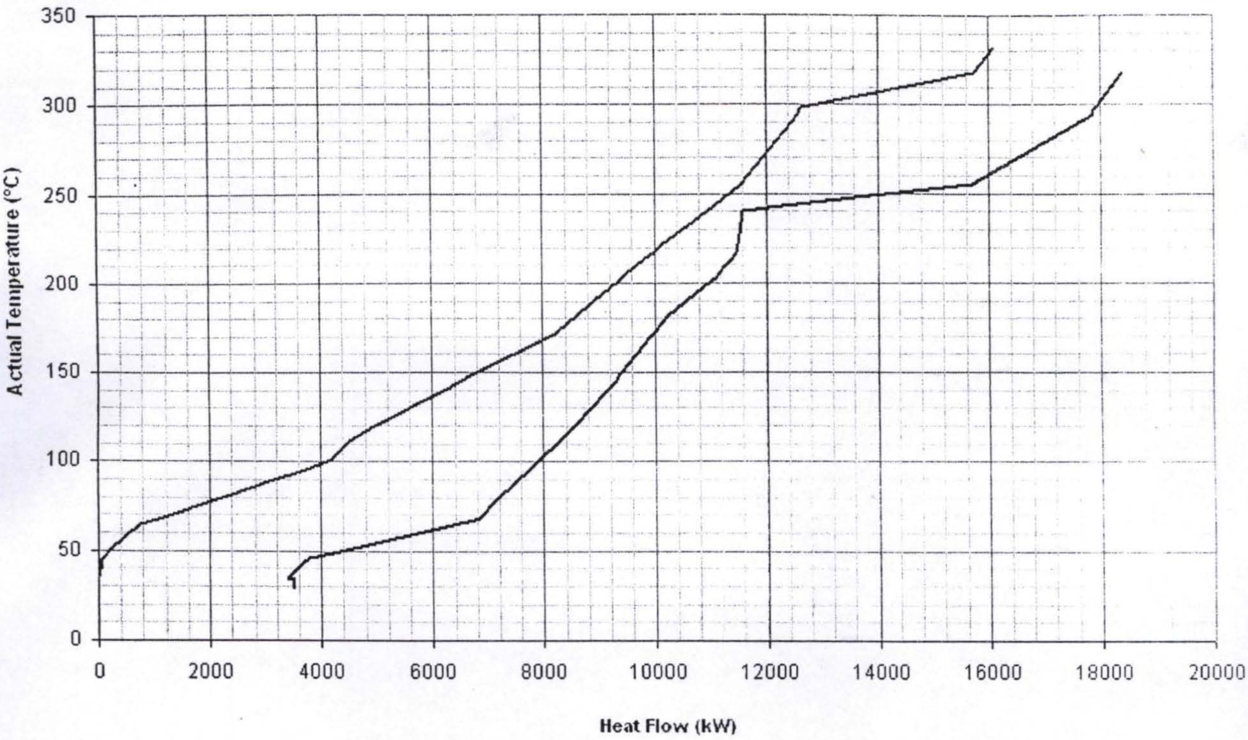


Figure B.1: Composite Curve

B.5 Grand composite curve

The Grand Composite Curve is a plot of the difference in enthalpies between cold and hot composite curves. It is commonly used in determining the temperature levels required for the hot and cold utilities. It was plotted as the shifted temperatures versus the enthalpy difference. The grand composite curve is shown in Figure A.2

The following command was typed in the maple program to display the grand composite curves:

```
>
plots[pointplot](GCCdata,title=GCCtitle,connect=true,
                 labels=[`H(kw)`,`Tshifted(K)`]);
```

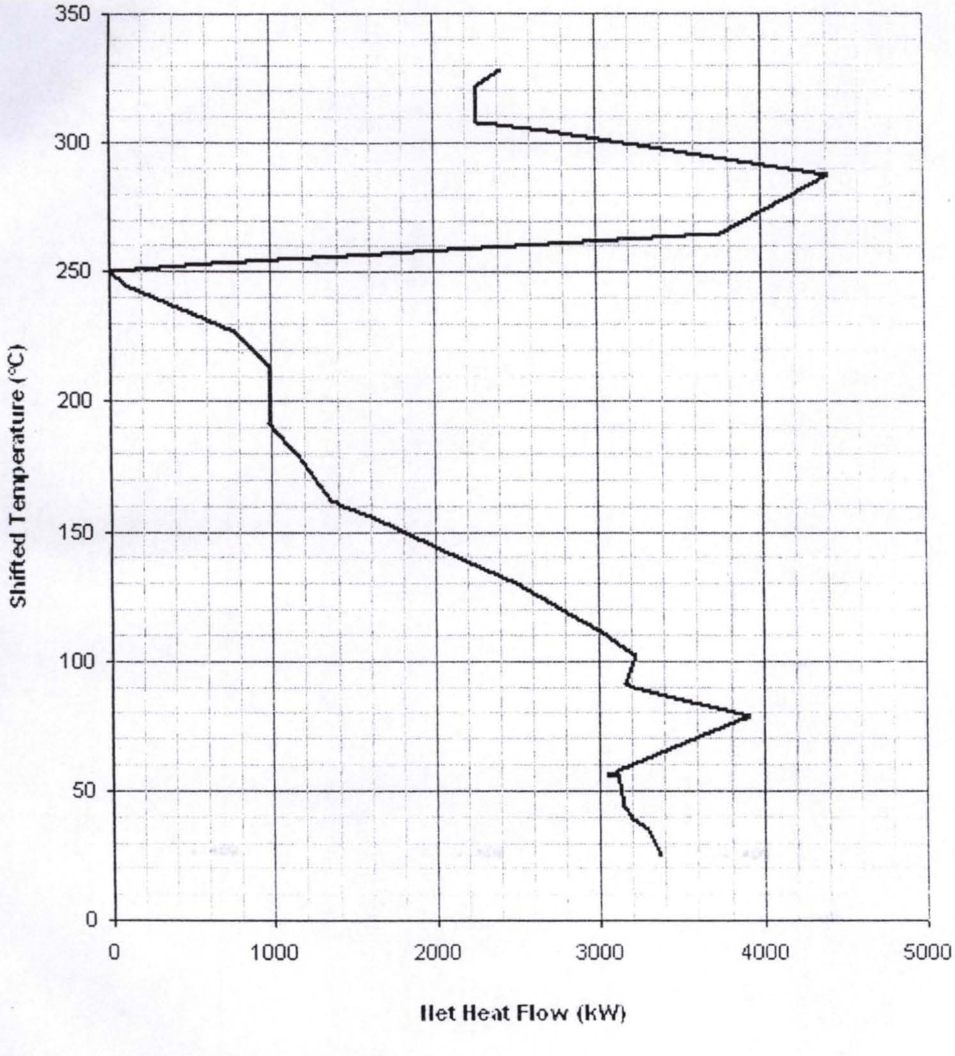


Figure B.2: Grand Composite Curve

B.6 Economic trade-off between operating costs and capital cost

The optimum selection of the dTmin value was made by optimizing the trade-off between operating costs and the installation cost of the network. The installation of larger heat exchangers can reduce operating costs for utilities but at the expense of a higher capital cost for the network. The trade-off was best made by minimizing the net present cost of capital and operating costs.

B.7 Utility data

The calculations need information about the available utilities.

Cold and hot utilities constants were entered in Maple code as:

```
> ColdUtilityNPCfactor:=70:
```

Cold utility T (K)

```
> TColdUtility:=290:
```

Cold utility heat transfer coefficient (kw/sqm-K)

```
> hColdUtility:=.4:
```

Hot utility usage net present cost factor ($NPC/kw)

```
> HotUtilityNPCfactor:=500:
```

Hot utility T (K)

```
> THotUtility:=500:
```

Hot utility heat transfer coefficient (kw/sqm-K)

```
> hHotUtility:=2.:
```

B.8 Capital cost data

The heat exchanger cost formula ($) used is:

$\ln(\text{cost}) = \text{capA} + \text{capB} \times \ln(\text{area,m}^2) + \mathbf{capC \times \ln(area,m^2)^2}$

The capital costs were entered into the Male code as follows:

```
> capA:=7.5:
capB:=0.24:
capC:=0.06:
```

The heat exchanger capital costs were adjusted to the purchase year by a factor.

```
> CostAdjust:=1.2:
```

The sum of heat exchanger costs were scaled to the installed network cost by a Lang factor.

```
> LangFactor:=5.:
```

The capital NPC factor was used to convert the capital cost to net present cost terms ($NPC/$capital)

```
> CapitalNPCfactor:=.90:
```

B.9 Performing the optimization

The optimization was performed by calculating the net present cost total for utilities and capital over a range of dTmin values. The following optimization starting value, range and increment were entered in Maple code as

```
> dTstart:=0:

Range:=20:

Increment:=2:

ans:=OptimizeData(dTstart,Range,Increment);
```

B.10 Minimum temperature approach (dTmin) optimization plot

The optimization plot was generated in Maple code by entering the following command:

```
plots[pointplot](NPCvalues,title=`NPC      vs.      dTmin;
`||ans,connect=true,labels=[`dTmin(K)`,`NPC($M)`]);
```

B.11 Capital cost plot

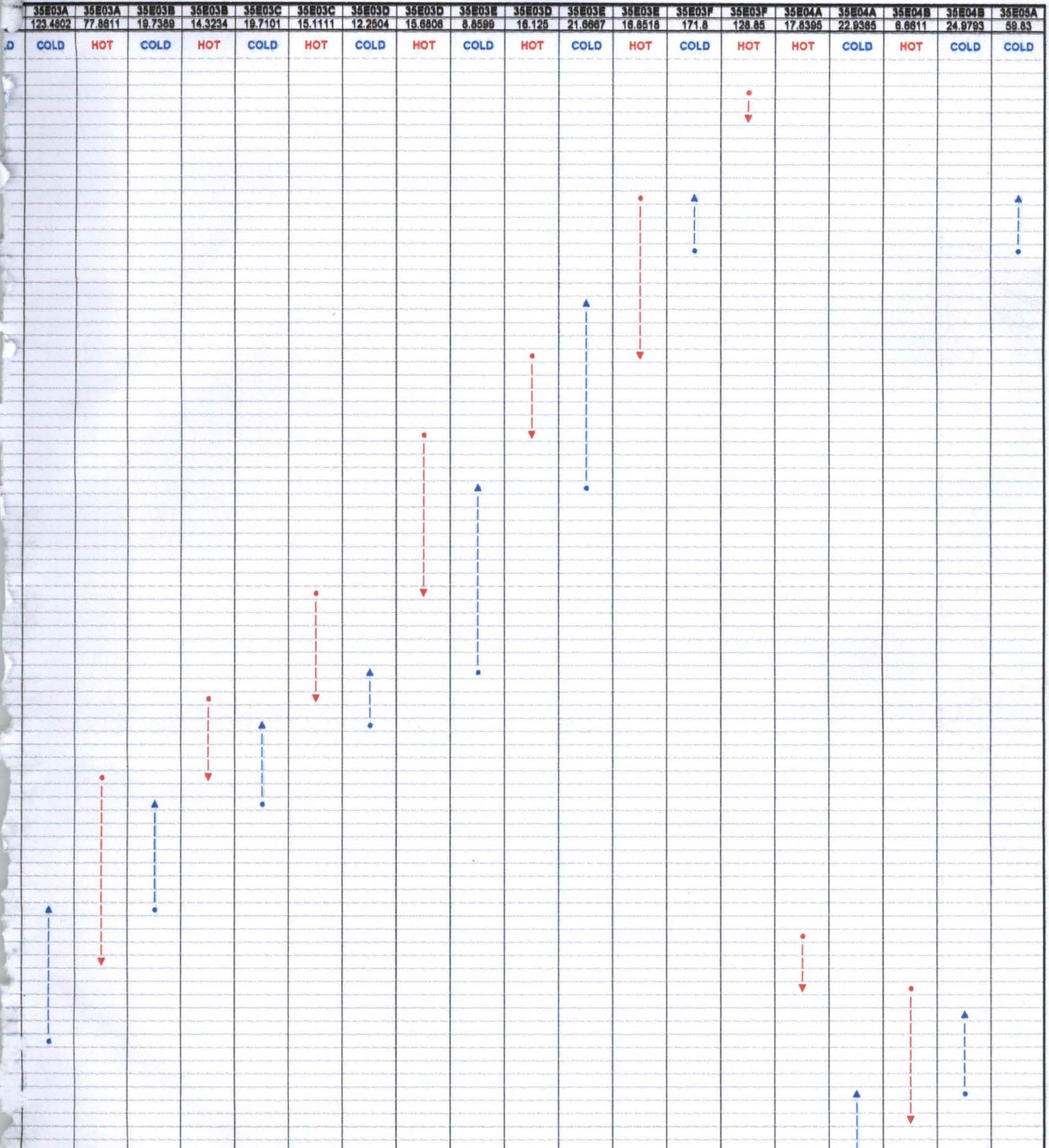
The capital cost plot was generated in Maple code by entering the following command:

```
> plots[pointplot](NPCvalues,title=`NPC vs. dTmin;
 ||ans,connect=true,labels=[`dTmin(K)`,`NPC($M)`]);
```

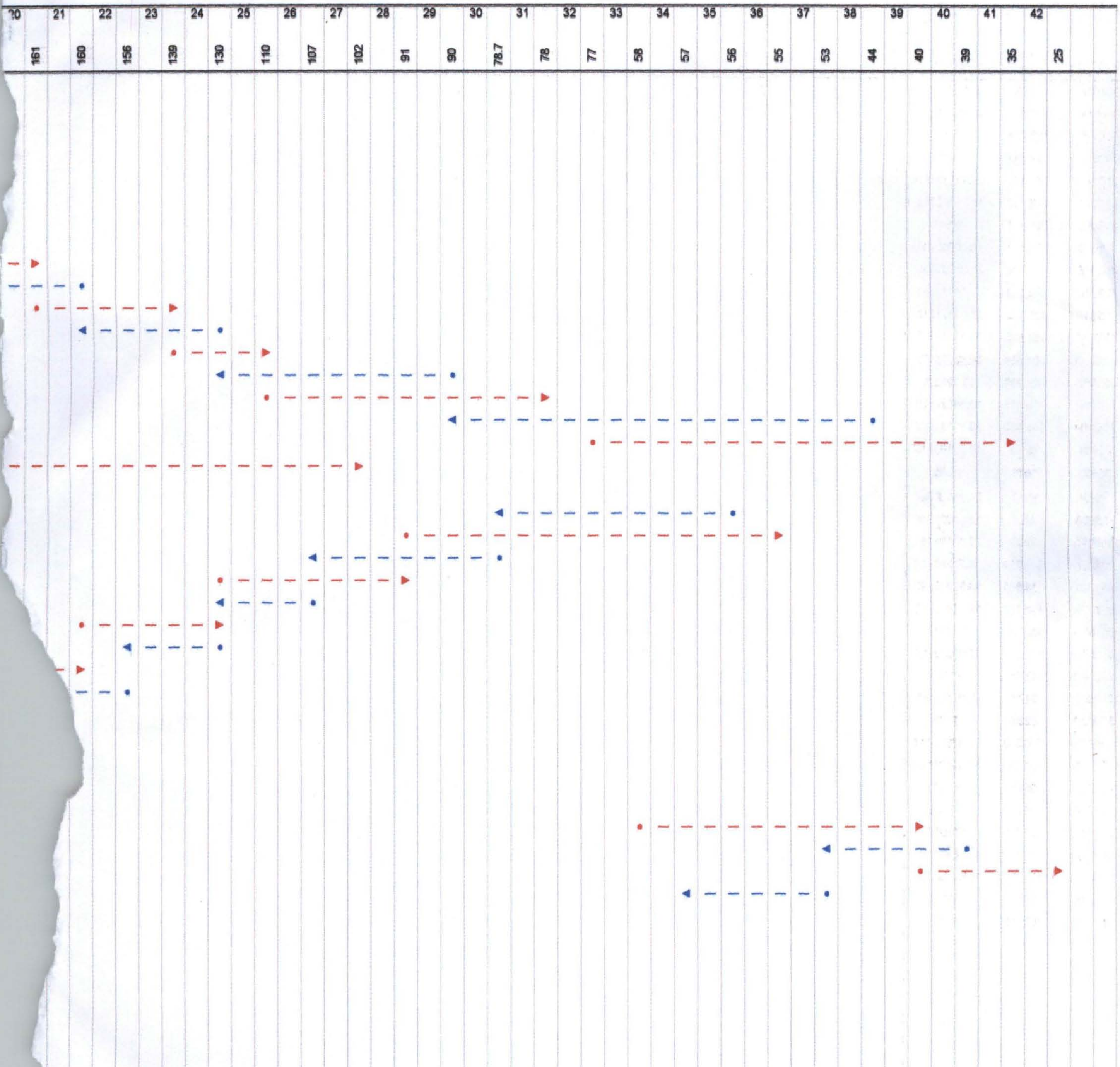
B.12 Utility cost plot

The utility cost plot was generated in Maple code by entering the following command:

```
> plots[pointplot](UtilityNPCvalues,title=`Target
Utility Costs ($M) vs.
dTmin`,connect=true,labels=[`dTmin(K)`,`NPC($M)`]);
```



35E03A
123.4802
COLD
35E03A
77.8611
HOT
35E03B
19.7389
COLD
35E03B
14.3234
HOT
35E03C
19.7101
COLD
35E03C
15.1111
HOT
35E03D
12.2504
COLD
35E03D
15.6806
HOT
35E03E
8.8599
COLD
35E03D
16.125
HOT
35E03E
21.6667
COLD
35E03E
16.8518
HOT
35E03F
171.8
COLD
35E03F
128.85
HOT
35E04A
17.8395
HOT
35E04A
22.9365
COLD
35E04B
6.6611
HOT
35E04B
24.9793
COLD
35E05A
59.83
COLD



20
21
22
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24
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26
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39
40
41
42
161
160
156
139
130
110
107
102
91
90
78.7
78
77
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44
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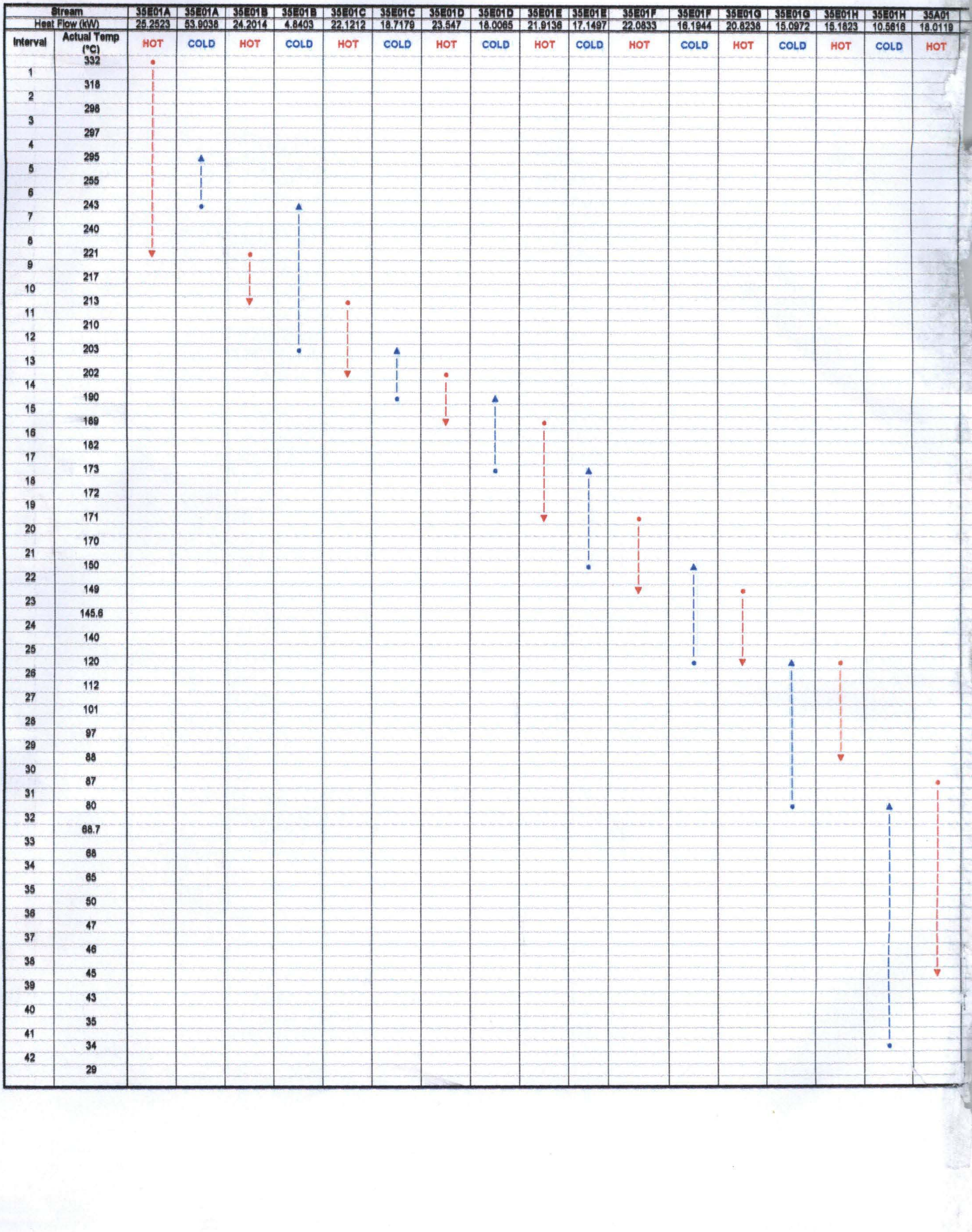
35
25

APPENDIX C1



Stream
Heat Flow (kW)
35E01A 25.2523 HOT
35E01A 53.9038 COLD
35E01B 24.2014 HOT
35E01B 4.8403 COLD
35E01C 22.1212 HOT
35E01C 18.7179 COLD
35E01D 23.547 HOT
35E01D 18.0065 COLD
35E01E 21.9136 HOT
35E01E 17.1497 COLD
35E01F 22.0833 HOT
35E01F 16.1944 COLD
35E01G 20.8238 HOT
35E01G 15.0972 COLD
35E01H 15.1823 HOT
35E01H 10.5616 COLD
35A01 18.0119 HOT
Interval
Actual Temp (°C)
332
1
318
2
298
3
297
4
295
5
255
6
243
7
240
8
221
9
217
10
213
11
210
12
203
13
202
14
190
15
189
16
182
17
173
18
172
19
171
20
170
21
150
22
149
23
145.6
24
140
25
120
26
112
27
101
28
97
29
88
30
87
31
80
32
68.7
33
68
34
65
35
50
36
47
37
46
38
45
39
43
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35
41
34
42
29

APPENDIX C.2

Grid Diagram

Stream Name	Heat Flow (kW)	mCp (kW/K)	
35E01A	2803	25.25225225	HOT
35E01A	2803	53.90384615	COLD
35E01B	193.611	24.201375	HOT
35E01B	193.611	4.840275	COLD
35E01C	243.333	22.12118182	HOT
35E01C	243.333	18.71792308	COLD
35E01D	306.111	23.547	HOT
35E01D	306.111	18.00652941	COLD
35E01E	394.444	21.91355556	HOT
35E01E	394.444	17.14973913	COLD
35E01F	485.833	22.08331818	HOT
35E01F	485.833	16.19443333	COLD
35E01G	603.889	20.82375862	HOT
35E01G	603.889	15.097225	COLD
35E01H	485.833	15.18228125	HOT
35E01H	485.833	10.56158696	COLD
35A01	756.5	18.01190476	HOT
35A02	1611	26.85	HOT
35H01	523.7	24.93809524	COLD
35E03A	2803	123.4801762	COLD
35E03A	2803	77.86111111	HOT
35E03B	558.611	19.73890459	COLD
35E03B	558.611	14.32335897	HOT
35E03C	453.333	19.71013043	COLD
35E03C	453.333	15.1111	HOT
35E03D	313.611	12.25042969	COLD
35E03D	313.611	15.68055	HOT
35E03E	322.5	8.85989011	COLD
35E03D	322.5	16.125	HOT
35E03E	758.333	21.66665714	COLD
35E03E	758.333	16.85184444	HOT
35E03F	2577	171.8	COLD
35E03F	2577	128.85	HOT
35E04A	321.111	17.8395	HOT
35E04A	321.111	22.9365	COLD
35E04B	99.917	6.661133333	HOT
35E04B	99.917	24.97925	COLD
35E05A	897.45	59.83	COLD