OPTIMIZATION OF HEAT INTEGRATED DISTILLATION COLUMN

A DISSERTATION

Submitted in partial fulfillment of the
requirements for the award of the degree
of

MASTER OF TECHNOLOGY
in
CHEMICAL ENGINEERING
(with specialization in Computer Aided Process Plant Design)

By

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CANDIDATE'S DECLARATION

I hereby declare that the work, which is being presented in this dissertation entitled "**OPTIMIZATION OF HEAT INTEGRATED DISTILLATION COLUMN**" in the partial fulfillment of the requirements of the award of the degree of **Master of Technology** in **Chemical Engineering** with specialization in **Computer Aided Process Plant Design**, submitted in the Department of Chemical Engineering, Indian Institute of Technology Roorkee, Roorkee, is an authentic record of my own work carried out during the period from July 2005 to June 2007 under the kind supervision of Dr. RAVINDRA BHARGAVA, Assistant Professor, Department of Chemical Engineering, Indian Institute of Technology Roorkee.

The matter, embodied in this dissertation has not been submitted for the award of any other degree.

Date: 21/06/07

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Place: Roorkee

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CERTIFICATE

This is to certify that the above statement made by the candidate is correct to the best of my knowledge and belief.

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ABSTRACT

With the continuous rise in energy cost, depletion of conventional energy resources and growing environmental concern, there is need to reduce the energy consumption by optimization of existing processes. In refineries besides the cost of crude, energy is the largest cost which can be influenced by improved operation and/or capital investment, and has therefore become a primary focus.

Process-integration techniques based on pinch technology represent a new and powerful way to optimize process designs, yielding results superior to those achievable using conventional methods. These new techniques permit the design engineer to track the energy flows in a manufacturing process more clearly and to modify the process to reduce energy consumption. Pinch technology also enables the design of an optimum interface between the process and the utility systems.

Heat integration of a given process can give higher benefits to an existing industry. The use of heat integration for distillation columns holds a great promise of energy savings up to 45-50%. In addition to saving energy, heat integration reduces the environmental impact of a process, reduces site utility costs and can give possible reduction in capital costs.

There are several approaches for the analysis of heat integrated distillation systems, with their specific merits and demerits. These approaches are Pinch Technology, Evolutionary Algorithm Approach, Linear Programming (LP), Nonlinear Programming (NLP) and Mixed Integer Nonlinear programming (MINLP). Out of these tools Pinch Technology is a prominent tool because it allows the operator and designer to take part in each step of process integration.

The present work is related to heat integration of different processes available in Crude Distillation unit by carrying out Pinch study and identifying schemes that enable maximum energy recovery. Two possibilities are conceptualized to increase furnace inlet temperature of crude by heat integrating distillation column pump around and circulating reflux streams with preheat exchanger train in existing unit.

Target crude furnace inlet temperature of around 267 ^{0}C is achievable by introducing new heat exchanger matches i.e. Crude/ LVGO CR, Crude/ Column overhead vapors and Crude/ Sweet VGO.

By doing new heat exchanger matches reduction in furnace duty can be achieved corresponding to the heat recovered at lower temperature. However additional heat exchanger area needs to be provided. The economic analysis for both cases indicates the annual saving of Rupee 32.5 Crores and 61.8 Crores with payback period of 1.28 years and 0.88 years respectively.

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NOMENCLATURE

A	Area, m^2
CP	Product of steam flow rate and specific heat, kcal/ ^0C
DT_{min}	Δt_{min}, ^0C
Ft	LMTD Correction Factor
h	Heat transfer coefficient, kcal/ (m^2. ^0C)
HAGO	Heavy atmospheric gas oil
HK	Heavy kerosene
HVGO	Heavy vacuum gas oil
KBPSD	Kilo barrels per stream day
LK	Light kerosene
LVGO	Light vacuum gas oil
MCP	CP, kcal/ ^0C
N	Number of streams including utilities
Q	Difference in Enthalpy. kcal/ ^0C
RCO	Reduced crude oil
S	Heat Exchanger
U	Overall Heat transfer coefficient, kcal/ (m^2. ^0C)
u_{min}	Minimum number of units, including heaters and coolers
VR	Vacuum residue

Subscripts

c	Cold
h	Hot
i	Temperature interval number
in	Inlet temperature
min	Minimum
opt	Optimum
out	Outlet temperature

s	Supply temperature
t	Target temperature

CHAPTER-1
INTRODUCTION

Chemical industry employs a variety of separation processes like absorption, crystallization, distillation, evaporation, extraction etc in order to separate a binary or multi-component mixture into its components. Distillation is the most widely used separation process in chemical industry. This high usage rate is primarily due to distillation's flexibility, low capital investment relative to other separations technologies, and low operational risk. Unfortunately, the energy efficiency of a commercial distillation column is low with a thermodynamic efficiency of less than 10 percent being typical (Gadalla et al. [5]). On the surface, distillation would appear to be a favorable R&D target area with huge potential energy savings. In reality, it will be a formidable challenge to develop improvements or replacements for distillation to achieve significant energy savings, due to the large sunk capital investment in existing plants, the slow rate of plant replacements, and the diverse numbers of applications where distillation is utilized.

In the chemical industries, the task of separation is a very energy consuming process, where distillation is the process most widely used for fluid separations. Distillation columns are used for about 95% of liquid separations and the energy use from this process accounts for an estimated 3% of the world energy consumption (Bauer et al. [1]). With rising energy awareness and growing environmental concerns there is a need to reduce the energy use in industry. For the distillation process any energy savings should have an impact on the plant energy consumption.

The use of heat integration and more complex configurations for distillation columns holds a great promise of energy savings. In addition to saving energy, heat integration reduces the environmental impact of a process, reduces site utility costs and can give possible reduction in capital costs. However, there are a number of different methods or designs that can be applied to save energy in distillation, for example integration of distillation columns with the background process, heat pumps, multi-effect distillation and complex arrangements such as pre-fractionators or thermally coupled columns. Deciding which heat integrated arrangement to use is not a straightforward task as the best arrangement is very much dependent on the given separation task. When considering heat integration of distillation columns the distillation unit should also be looked at in terms of the whole process, using Pinch analysis.

There are several approaches for the analysis of heat integrated distillation systems, with their specific merits and demerits. These approaches are Pinch Technology, Evolutionary Algorithm Approach, Linear Programming (LP), Nonlinear Programming (NLP) and Mixed Integer Nonlinear programming (MINLP).

In the late 1970s, Pinch Analysis was first applied to the optimization of industrial energy systems, in response to high fuel prices during the 'energy crises'. 'Pinch technology', a term introduced by Linnhoff and Vredeveld in 1979, has its origin in the works of Hohmann. Pinch technology emerged as a tool for the design of Heat Exchanger Network (HENs). Its key contribution was to give the engineer simple concepts which were used interactively for the evolution of an optimal Heat Exchanger Network. In 1983, Linnhoff and Hindmarsh [12] first applied this technology to design Heat Exchanger Networks for maximum energy recovery. In the next major advance, Pinch Analysis was extended to the analysis of onsite utilities, such as boilers, turbines, heat pumps and refrigeration systems, and techniques were developed for optimum design of Combined Heat and Power (CHP) systems.

More recently, Pinch Analysis has been applied to the problems of minimizing environment pollution (NOx, SOx, VOC and Waste water emissions), and further reduction in energy costs through better integration of unit operations (e.g. distillation columns) within a process unit, as well as better thermal integration between various process units at a given site. Benefits of total site integration are quite significant, comparable in magnitude to the energy saving from intra-unit integration.

1.1 OBJECTIVE OF PRESENT STUDY:

Refineries are the largest energy consumer among the process industries. In a refinery, crude oil distillation systems are complex configurations that interact strongly with the associated heat recovery systems. In the Crude distillation unit, the crude oil is preheated in two stages before entering the distillation column. The first stage is Heat exchanger network, where oil is heated to an intermediate temperature by cooling distillation process streams and recovering the heat from the condensers. Afterwards, the crude oil enters furnace to reach the required processing temperature. The more fuel consumed in the furnace, the larger the operating cost. Any heat recovered from the distillation process streams reduces the hot utility consumption in the furnace. The energy efficiency of the distillation process can be improved by designing the distillation column to create opportunities for heat recovery and designing the heat exchanger network to exploit these opportunities.

In the light of above mentioned fact a theoretical study has been undertaken to achieve the following objectives:

1. To carry out Pinch study of existing Crude Distillation unit to identify the schemes which enable maximum energy recovery.
2. To maximize the crude preheat temperature of furnace by heat integrating different streams of the Crude Distillation Unit distillation column with existing preheat train, without increasing utility consumption.
3. To see the feasible possibility of maximizing the preheat temperature by heat integrating the Crude Distillation Unit processes with process streams of other unit, without increasing utility consumption
4. To do Pinch analysis to determine process Pinch point, target for utility consumption which involves identifying cold and hot streams, thermal data extraction, generation of Composite curves and Grand composite curves.
5. Area Targeting for Heat Exchanger network
6. Preliminary cost estimates of energy saving and payback period

CHAPTER-2

OVERVIEW OF HEAT INTEGRATED DISTILLATION SYSTEMS

When distillation is the choice of separation for a given mixture there are a number of ways that the separation by distillation can be carried out. There are many factors that are important in the design process, such as safety, operating costs, capital costs and operability. Deciding on the most appropriate distillation system can be difficult, as in most cases there are usually trade-offs between the various objectives.

In the chemical and petroleum industry distillation usually accounts for around 25-40 % of the energy usage, and it has been estimated that for a wide range of distillation applications 10-15 % of the energy use is in excess (Bauer et al. [1]). The potential for energy savings therefore exists and design and operation of energy efficient distillation systems will have a substantial effect on the overall plant energy consumption and operating costs.

For a given separation different distillation sequences have different energy consumption levels, there is a substantial economic incentive in selecting the most appropriate distillation sequence. Moreover, even higher energy savings can be achieved by incorporating Heat Integration techniques in distillation sequences. The use of integrated distillation systems can lead to a significant reduction of energy consumption in comparison with conventional distillation columns. With heat integrated distillation column systems energy savings up to 45% can be achieved (Gadalla [7]).

2.1 IDEAL HEAT INTEGRATED DISTILLATION COLUMN:

Ideal heat integrated distillation columns are created by chasing reduction of energy consumption in distillation columns. The way to pursue energy reduction is to adopt heat integration between the whole rectifying and the whole stripping sections and these results in a seemingly far different counterpart of conventional distillation column.

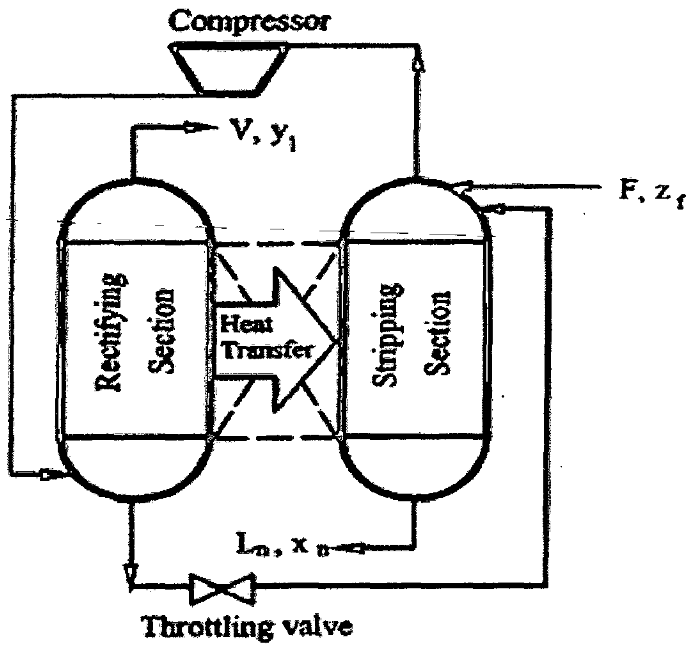


Fig. 2.1 Schematic representation of an ideal heat integrated distillation column

In ideal heat integrated distillation columns, its stripping section and rectifying section are separated into two columns, while heat is integrated through a number of internal heat exchangers. To accomplish internal heat transfer from the rectifying section to the stripping section, the rectifying section is operated at a higher pressure and a higher temperature than those of the stripping section. For adjusting the pressures a compressor and a throttling valve have to be installed between the two sections. Owing to the heat integration, a certain amount of heat is transferred from the rectifying section to the stripping section. It generates the reflux flow for the rectifying section and vapor flow for the stripping section. Thus the condenser and reboiler are, in principle, not needed for a single ideal heat integrated distillation column; as a result, both fixed and operating costs can be further reduced.

In the complex configuration of a heat and mass-integrated distillation system the condensing vapor from the top of the high pressure column (HPC) is used to heat the reboiler of the low pressure column (LPC). The double column system is superior to a single column in energy saving but disadvantageous in dynamic operability. Interactions and time delays leads to a more complicated controllability, so a double column system needs a higher effort in the design and the control systems than a single column.

2.1.1 ADVANTAGES OF HEAT INTEGRATED DISTILLATION COLUMN:

Advantages associated with heat integrated distillation column are following:

1. Each heat integrated (distillation) unit presents a saving which reduces both the hot and the cold utilities and hence decreases the total annual cost of process.
2. Additional inter-column heat exchanger area results in savings of both hot and cold utilities.

2.1.2 DRAWBACKS OF HEAT INTEGRATED DISTILLATION COLUMN:

There are some drawbacks associated with heat integrated distillation column which prevents the process from being commercialized. They are as following:

1. Operational difficulties due to the nonlinear, multivariable, and interacting nature of the process.
2. Energy integration makes the control loop coupling severe.
3. For optimal usages of utility, investment in equipment increases.
4. The relationships between the design variables (like column pressure, feed condition) and column parameters (like heat loads, temperatures, number of trays) are highly non-linear.
5. Efficient heat integration methods are not available to calculate minimum utility requirements when temperatures of process streams vary continuously. In particular, the condenser and reboiler temperatures vary with the operating

pressure of a column. Therefore, the feed and product stream temperatures of columns also vary with the operating pressure and feed condition.

6. Practical constraints such as forbidden matches, maximum number of exchangers for a reboiler or condenser, etc., need to be accounted for when optimizing a heat integrated distillation sequence for reasons of safety, start-up and control.

2.2 PARAMETERS EFFECTING EFFICIENCY OF HEAT INTEGRATED DISTILLATION COLUNM:

Efficiency of ideal heat integrated distillation depends on many variables like pressure difference between rectifying and stripping sections, thermal condition of feed, feed flow rate, feed composition, feed location, number of stages etc. The effect of these variables on the operation and efficiency of heat integrated distillation column are discussed below:

2.2.1 FUNCTION OF PRESSURE DIFFERENCE:

The pressure difference, p_r-p_s plays an important role in heat integration between the rectifying and the stripping sections. It is one of the two manipulated variables that are used for the operation of the ideal heat integration distillation column. After conceptual process design, higher pressure difference p_r-p_s is needed for high purity overhead and bottom product. It is extremely necessary to assess the flexibility of the ideal heat integration column with respect to the variations of the pressure difference, p_r- p_s. One character associated with the pressure difference, p_r-p_s is that it can only influence internal flows.

2.2.2 FUNCTIONS OF FEED THERMAL CONDITION:

Feed thermal condition q is another manipulated variable that must be used for the operation of the ideal heat integrated distillation column. It is readily to understand that the feed thermal condition q influences only the material balances of the process.. It must be within the range: $0<q\leq 1$, so as to provide a suitable basis for employing the heat pump principle in the ideal heat integrated distillation column. Effectively defining an

appropriate value for the feed thermal condition q is also beneficial to the process energy efficiency.

2.2.3 INFLUENCE OF FEED FLOW RATE:

Pressure difference in distillation column is related to feed flow rate. Higher pressure difference, p_r-p_s can be achieved by larger feed flow rate. As the pressure difference, p_r-p_s, is enhanced, the superiority of the ideal heat integrated distillation column will diminish gradually because electricity is usually several times more expensive than heating steam. It is anticipated that up to a certain flow rate the potential of energy saving will be totally lost for the ideal heat integrated distillation column. On the other hand, when feed flow rate becomes too small, the necessary pressure difference p_r-p_s will go down drastically, and so does the process energy efficiency. Much low pressure difference, p_r-p_s, can even cause strong interactions between the rectifying and the stripping sections and introduce extra difficulty in process operations. So higher energy efficiency can be achieved only within a certain range of the feed flow rate.

2.2.4 INFLUENCE OF FEED COMPOSITION:

As the variations of pressure difference, p_r-p_s, is quite limited in magnitude, it is reasonable to consider that feed composition does not influence the process energy efficiency very much, and thus it does not impose strict requirements to the process flexibility.

2.2.5 EFFECT OF FEED LOCATION:

Feeding location is an important design variable in the distillation process design. In an ideal heat integrated distillation column, potential of energy saving can be achieved by proper feed location arrangement. As the feed must be vapor / liquid mixture ($0 < q \leq 1$), it is better that the vapor and liquid portions should be introduced to the process at different locations. The liquid portion should be fed to a relative higher position than the middle stage, namely in the rectifying section, and the vapor portion should be fed to a relative lower position than the middle stage, namely in the stripping section. This arrangement reinforces internal flows and therefore benefits mass transfer. With the same operating conditions, proper feed location needs lower pressure difference, p_r-p_S, than the original one, hence the process efficiency can be improved.

2.2.6 EFFECT OF TOTAL STAGES:

As the number of stages increases, the required pressure difference, p_r-p_S, between rectifying and stripping section decreases. In other words, the operating cost is reduced with the expense of fixed investment. When the total number of stages reaches a certain high value, the direction of heat transfer from the rectifying to the stripping sections becomes difficult to be maintained, especially when external disturbances occur. Inverse heat transfer is detrimental to process energy efficiency and can add difficulties to process operation, thus it should be avoided. As the number of stages is further increased, the pressure difference, p_r-p_S, reaches a minimum value and in those circumstances, inverse heat transfer occurs. Therefore, there exists an appropriate range for the choice of number of stages. Smaller number of stages may result in lower fixed investment, but higher operating costs. On the other hand, large number of stages leads higher fixed investment, but lower operating cost. Therefore a careful compromise should be maintained among number of stages, internal heat transfer area and the pressure difference, p_r-p_S.

2.3 VARIOUS METHODS OF HEAT INTEGRATION:

Any separation requires the addition of work or energy. In a distillation column the work requirements are supplied indirectly through heating in the reboiler and cooling in the condenser at a lower temperature. Usually, if simple columns (no heat integration) are to be used, the first approach may be to use rules of thumb to find a column sequence that has low total energy consumption. A number of potential column sequences can be made based on 'heuristic rules' or 'short-cut methods' and then the most promising schemes can be tested using more rigorous simulation tools.

Various methods used for heat integration the distillation column are given below:

2.3.1 COLUMN INTEGRATION WITH PROCESS:

When the best column sequence has been identified further energy savings can be made by integrating the reboiler and condensers with other process streams, either by using the product/feed streams from the distillation sequences or integration with the rest of the process. The best combination for this heat exchanger network can be found using Pinch methods. The pinch method looks at the heating and cooling of the whole process and gives targets for the minimum heating and cooling requirements. A distillation column can also be integrated with the process where the heating or cooling duty is provided by the process streams. A grand composite curve can be used to see if a distillation column can be integrated with the plant. The rule for integration with the process is that a column should not be integrated across the pinch. Either the condenser should be integrated above the pinch (where heating is required) or the reboiler should be integrated below the pinch (where cooling is required).

2.3.2 INTERHEATING AND INTERCOOLING WITHIN COLUMN:

Another method for saving energy is the use of interheating or intercooling, where heat is added or removed at intermediate points in the column. There will be no reduction in the total energy requirement, however the temperature in the intermediate reboiler will be lower than in the bottom reboiler and the temperature in the intermediate condenser will be higher than the top condenser. This allows for the use of cheaper hot or cold utility, if different heat sources and heat sinks are available. Using intermediate reboilers and condensers may also improve opportunities for heat integration.

2.3.3 INTEGRATION WITH HEAT PUMP:

There are two different schemes for heat pumps. In the first scheme the overhead vapour is compressed and then condensed in the reboiler. This gives Latent heat of condensation to reboiler. Second, an external working fluid can be used for heat transfer. Heat pumping can save energy in the distillation of close boiling mixtures. The disadvantage is too high compression cost for multi-component mixtures or mixtures with large differences in boiling points.

2.3.4 DIRECT/ INDIRECT SPLIT COLUMN ARRANGEMENT:

For a multi-component mixture the distillation is usually carried out in a series of distillation columns that are operated in a direct split or indirect split fashion. In a direct split column the lightest component is split off first (at the top) and in an indirect split column the heaviest component is split off first (in the bottom). This is shown in Figure 2.2 and Figure 2.3for a ternary separation.

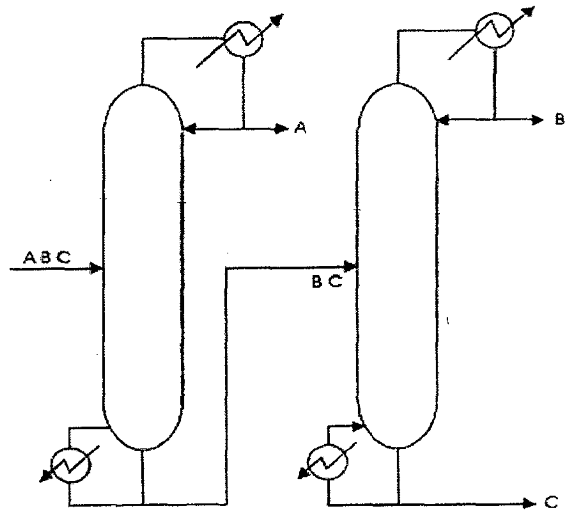


Fig 2.2 Separation of a ternary mixture by Direct split column arrangement

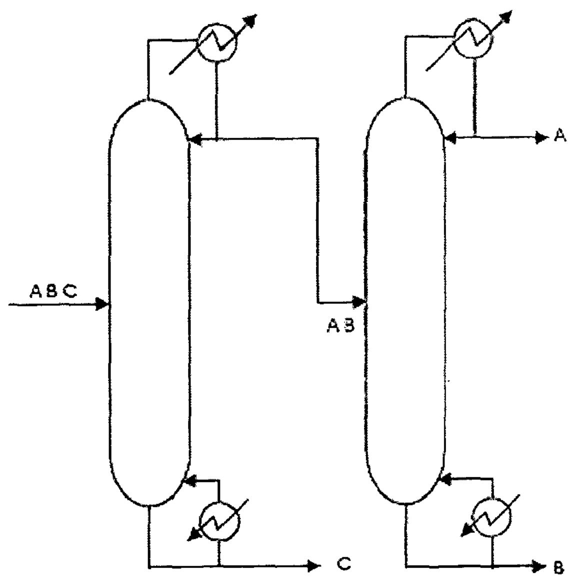


Fig 2.3 Separation of a ternary mixture by Indirect split column arrangement

2.3.5 INTEGRATING PREFRACTIONATOR WITH COLUMN:

Using a prefractionator is an effective way of reducing energy consumption. Prefractionators are thermodynamically more efficient than conventional two-column arrangements and require typically 20-30 % less energy. For a ternary separation (ABC) this arrangement consist of a prefractionator column which splits the lightest (A) and the heaviest (C) component and a main column with a side-stream which carries out the A/B split in the top part and the B/C split in the bottom part. The products are recovered in the main column with A in the distillate, B in the side-stream and C in the bottom stream.

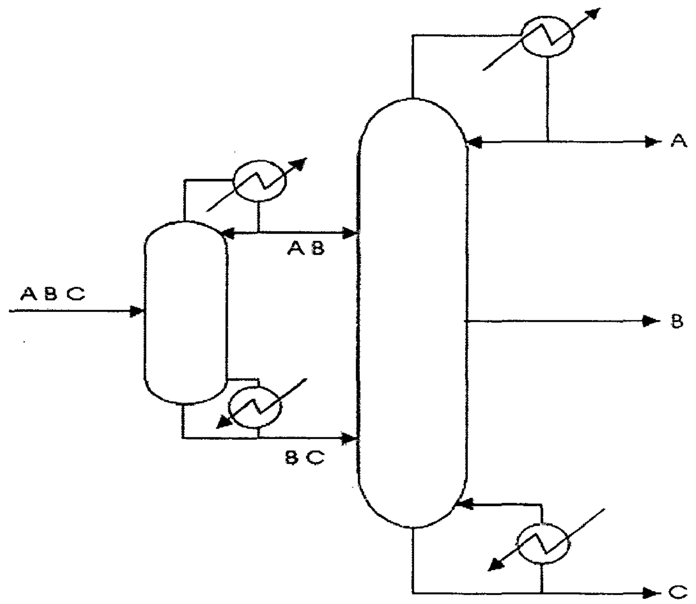


Fig 2.4 Prefractionator arrangement

2.3.6 THERMALLY COUPLED DISTILLATION COLUMN:

In the thermally coupled distillation column (Petlyuk arrangement) the liquid stream from the second column replaces the condenser of the first column and the vapour stream from the second column replaces the reboiler. The Petlyuk arrangement is equivalent to the prefractionator arrangement in terms of cooling and heating duties. Energy savings in these systems are achieved as the prefractionator prevents the re-mixing effect of the middle component, which otherwise occurs in conventional direct split or indirect split columns.

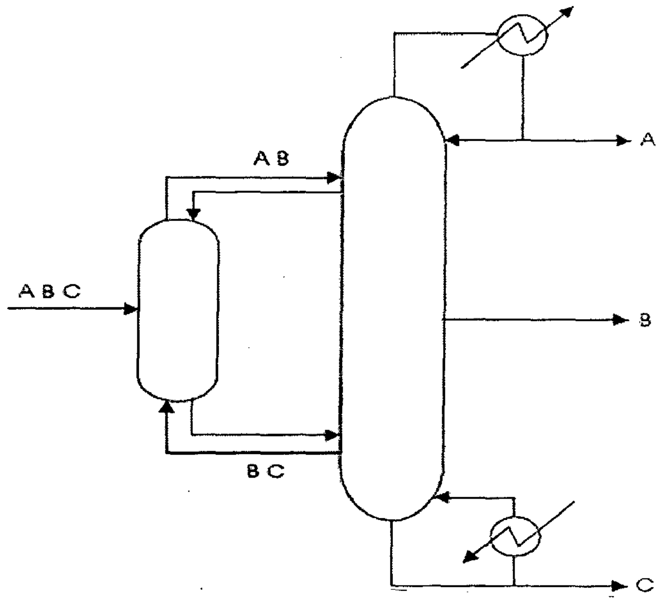


Fig. 2.5 Thermally coupled columns - Petlyuk arrangement

2.3.7 MULTIEFFECT DISTILLATION COLUMN:

Heat integration by using multi-effect columns involves using two (or more) columns with different pressures. The pressures are selected so that the condensation temperature of the overhead vapour is higher than the boiling temperature of the low-pressure column. The overhead vapour stream from the high-pressure (HP) column is then condensed by boiling the bottoms in the low-pressure (LP) column, so rather than having a separate condenser and reboiler these two are combined in one. For this type of arrangement there can be additional large savings in energy.

The multi-effect columns use pressure to raise the boiling point of the overhead vapour from a HP column so that this can be condensed in the bottom of a LP column. The reboiler in the LP column and condenser in the HP column are thus combined and heat only has to be supplied at the bottom of the HP column. Heat also has to be removed in a condenser at the top of the LP column. This type of arrangement has potential for additional energy savings of up to 50-70 %, depending on the arrangement, feed composition and the difficulty of the separation (Engelien et al. [2]). The disadvantages are added capital costs, an increased temperature of the hot utility and potential control problems.

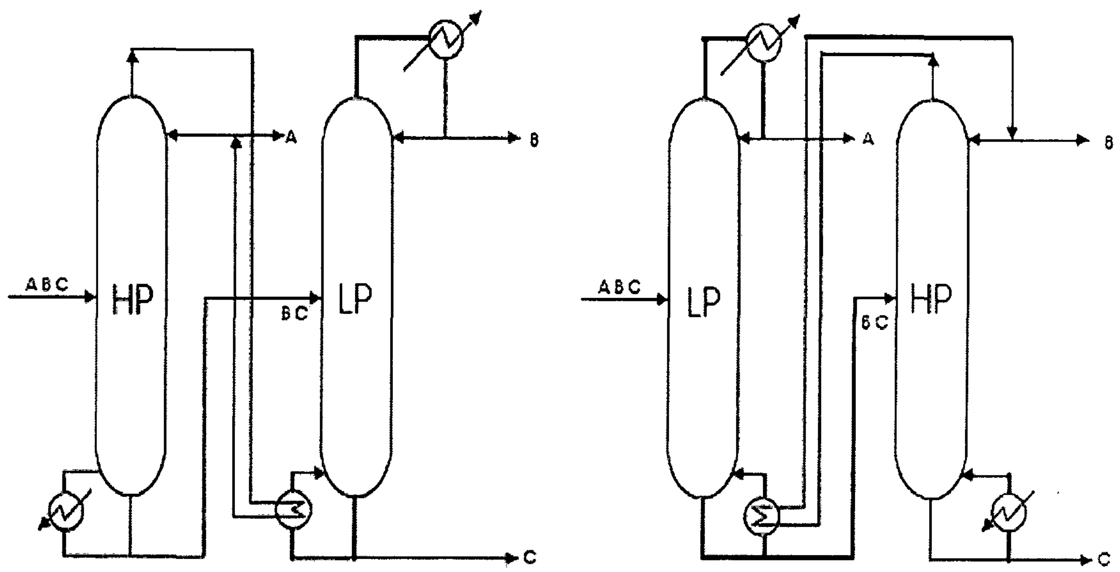


Fig 2.6 Multi-effect distillation for direct split

a) Direct split with forward integration   b) Direct split with backward integration

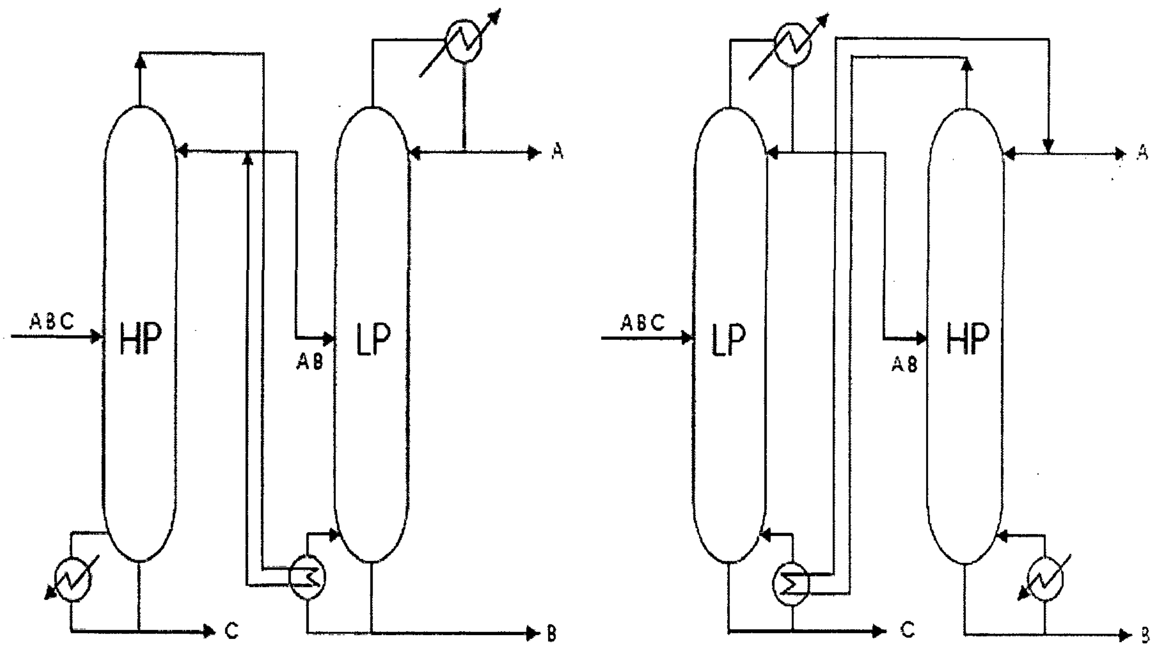


Fig 2.7 Multi-effect distillation for Indirect split

a) Indirect split with forward integration   b) Indirect split with backward integration

There are two types (or directions) of multi-effect heat integration: forward integration, where the integration of the energy is in the same direction as the flow and backward (reverse) integration where the energy integration is in the opposite direction of the flow. Distillation sequences are usually termed direct or indirect, where for the direct split the lightest component is split off as the distillate product from the first column and for an indirect split where the heaviest component is split off as the bottom product from the first column.

Figure 2.6 and Figure 2.7 shows, for a ternary separation, how direct and indirect sequence can be integrated in multi-effect arrangements. Figure 2.6 shows the direct split sequence with forward and backward integration and Figure 2.7 shows the indirect split sequence with forward and backward integration. The forward heat-integrated prefractionator arrangement is easy to control. The energy savings in backward integrated system assuming constant relative volatility will be identical to forward integrated system, but arrangement is difficult than forward integrated system.

It should be noted that when considering heat integration of distillation columns the distillation unit should also be looked at in terms of the whole process, using pinch analysis. Only then can it be decided whether or not a heat integrated system is the most effective.

CHAPTER-3

LITERATURE REVIEW

Separation processes, most commonly used in chemical industries are generally energy intensive. Therefore they have been the subject of investigation for the last few decades. In chemical industries, distillation columns are widely used to separate the desired components from a multi-component feed. Distillation being highly energy intensive, efforts was on for the past few decades to decrease the energy consumption. Past research has been concentrated mainly on distillation column as a single unit and very few researchers tried to probe deeper into heat integration issue i.e. energy saving by heat integrating the sequence of distillation columns or by heat integrating the distillation column with the process heat. For most separations heat integrated distillation columns are thermodynamically more efficient than conventional distillation column. The use of integrated distillation systems can lead to a significant reduction of energy consumption in comparison with conventional distillation columns. With heat integrated distillation column systems energy savings up to 45% can be achieved (Doukas et al. [1]). So work on this particular field has been progressive and modular with successive building up of the methods of exploiting the better heat integration aspects using various mathematical programming tools.

Energy optimization in distillation column can be done by two major techniques, one by depending on the Heuristic rules and other by using the Optimization techniques, which may be Linear Programming (LP), Mixed Integer Linear Programming (MILP) and Non Linear Programming etc. Heuristics methods are nothing 'but the rule of thumb' developed by experienced engineers and researchers.

In present chapter, a brief literature review of the methods used for optimization of energy in distillation column has been carried. As the present investigation deals with the energy optimization using Pinch technology, a detailed literature review on Pinch technology has been included.

Energy consumption in distillation column can be minimized by following number of methods.

3.1 INTERNAL HEAT INTEGRATION WITHIN SINGLE COLUMN:

A distillation column is generally divided into the rectifying and stripping sections. Thermal energy has to be supplied to the stripping section and removed from the rectifying section. The bottom reboiler and top condenser are the locations for the input and removal of thermal energy, respectively. If the energy removed from the rectifying section could be reused in the stripping section or waste heat was available, then energy savings would be achieved in a distillation column.

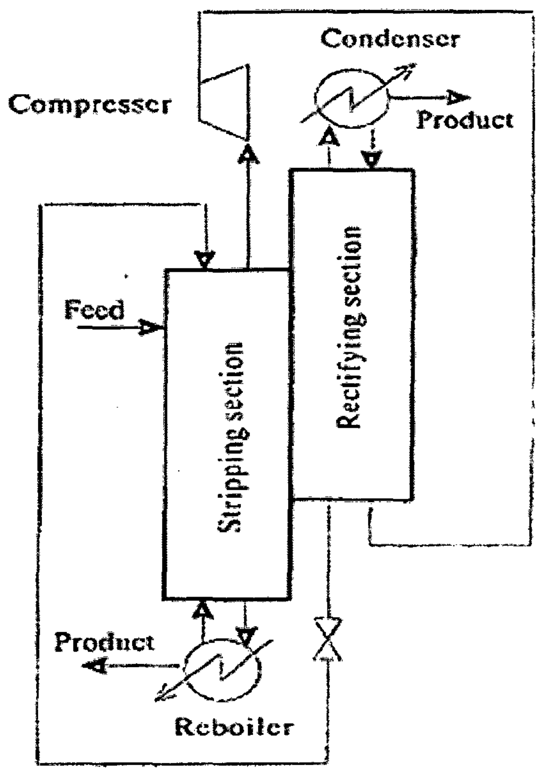


Fig.3.1 Internally heat integrated distillation column

An Internally heat integrated distillation column is constructed in such a manner that the rectifying and stripping sections are separated because a compressor and a throttling valve are installed between them as shown in Fig. 3.1. The manipulation is completed by exchanging heat between the two sections. To provide the temperatures necessary to serve as driving forces for heat transfers from the rectifying to the stripping section, the former must be operated at a higher pressure than the latter.

Since the 1960s, internal heat integration between the rectifying and the stripping sections of a distillation column has made significant advances in improving energy efficiency of distillation processes. Freshwater (1961) may have the first person to advocate this technique. Flower and Jackson (1964) further systematized the idea and classified the advantages of this approach through numerical simulations based on the second law of thermodynamics.

Nakaiwa et al. [14] continued the work on this subject both the theoretically and experimentally since 1985, and confirmed by large-scale experimental evaluations the advantages of these kinds of internally heat-integrated distillation columns. In 1995, they noticed that the degree of internal heat integration within a distillation column played a very important role in energy efficiency for a given separation. They proposed, therefore, to extend the internal heat to the whole rectifying and the whole stripping sections and this resulted in sharply different processes, configuration from conventional distillation columns.

3.2 HEAT INTEGRATION BETWEEN DIFFERENT COLUMNS:

Heat integration between different columns can be done by optimizing the column operating pressures. Although pressure is an important degree of freedom for heat integration, heat recovery can be further improved by optimizing feed condition, and by using multiple-effect columns, side exchangers, etc. Design variables for heat integration can be rigorously optimized by considering the heat integration opportunities between columns, between the other process and columns and opportunities for steam generation.

Gadalla et al. [6] presented a retrofit shortcut models for design of both reboiled and the steam-stripped distillation columns. The retrofit models were applicable for simple distillation columns, sequences of simple distillation column, and complex distillation configuration, including column with side-strippers and side rectifiers. The model fixed both the column configurations, and the operating conditions, including steam flow rates and calculated the product flow rates, temperatures and composition, and the various heat duties.

Engelien et al. [2] studied a multi-effect distillation where the condenser of a high pressure column was integrated with the reboiler of a low pressure column. The method of self-optimizing control has been used to provide a systematic procedure for the selection of controlled variables, based on steady-state economics. The heat integrated distillation system was optimized to find the nominal operating point. It was found that the constraint on the product composition, area in the combined reboiler/condenser, the purity constraint in low pressure column and the pressure in the low pressure column are active constraints. For the remaining unconstrained variable it was found that a temperature in the low-pressure column has good self-optimizing properties.

Wendt et al. [17] considered a pilot two-pressure column system for a theoretical and experimental study on startup strategies. A model concerning the different influential factors was used. A mathematical optimization was carried out to develop optimal operating policies for the column system. In the experimental study, different startup strategies were tested and the results verified the optimality of the optimal strategy.

From this work, the following simple operating policy can be suggested for starting up heat-integrated column systems in parallel operating arrangement. The feed splitting to the two columns can be set constant at the desired steady-state value. An enhanced value (i.e. approximately 1.2 times the steady state value) for the reflux flow of both columns and the reboiler duty of high pressure can be implemented in the first period to accelerate the startup. When the reboiler temperature of low pressure almost reaches its steady-state value, all the control variables should be switched to their steady-state value, allowing the column running to the desired steady state.

Doukus and Luyben [1] undertook a computational study of the alternative distillation configurations. They studied for different configurations the separation of a ternary mixture into their products. In addition to the classical direct and indirect sequences, they also included the following two configurations:-

1. a single column with side stream product.
2. a prefractionator column followed by a side stream column

Several values of the relative volatilities and the feed composition were explored. Their studies did not cover the wide domain of all possibilities, but they pointed to the following aspects:

1. The single side stream column is the most economical when the concentration of the most volatile component is very low (less than 10%). As its concentration increases above 10% it becomes rapidly uneconomical.
2. The effects of the changes in relative volatilities were very pronounced.
3. The heat and cooling costs apparently dominate the decision making.

3.3 EVOLUTIONARY ALGORITHM APPROACH:

Evolutionary Algorithms (EAs) are based on the model of evolution. Technical system to be optimized is represented in terms of EAs and called as *individual.* A population is a set of individuals competing with each other with respect to their target function values. In starting, population of random individuals is set and new populations are created by iteration and their target function values are calculated by evolution function. Only the best individuals are selected as the basis (*parents*) for the next population. The process of generating new individuals from two randomly chosen parents is modeled by the genetic operators *recombination* and *mutation.* The definition of the variations is strongly dependent on the representation of the individual, corresponding to different types of EAs.

The continuous iteration is done with recombination, mutation, evaluation and selection to improve the fitness of the average population and can be regarded as converged, if no progress is further made. EAs belong to the group of Probabilistic optimization techniques. Because of their robustness and generality they are especially suitable in cases where conventional strategies fail. They are suited to treat the complex optimization problem classes in chemical engineering, which result in mixed integer non linear programming (MINLP).

Evolutionary method can be used find the optimum sequences of heat integration between different distillation columns. Evolutionary method synthesizes new sequence by modification of previously generated sequence. Evolutionary synthesis includes the following three sub tasks:

(i) Generation of an initial separation sequence

(ii) Identification of the evolutionary rules

(iii) Determination of the evolutionary strategy

3.3.1 GENERATION OF AN INITIAL SEPARATION SEQUENCE:

For generating the initial sequence of columns and getting the better one in later stage, following six heuristics can be used:

(i) Identify the forbidden splits, which have relative volatility between the key components less than 1.05.

(ii) Easiest separations as characterized by their relative volatilities should be done first

(iii) When the mole percentage of feed components varies widely but the relative volatilities do not, sequence the splits to remove components in the order of decreasing molar percentage in the feed.

(iv) When neither relative volatility nor molar percentage in the feed varies widely, favor the direct sequence i.e. remove the components one by one as overhead product.

(v) When a mass separating agent is used remove it in the separator immediately following the one into which it is introduced.

(vi) When multi-component products are specified favor the sequence that produces the smallest product set.

3.3.2 IDENTIFICATION OF THE EVOLUTIONARY RULES:

Once an initial separation sequence has been generated, the evolution can begin. Evolutionary rules should have following properties:

(i) Efficiency, i.e. inventing separation sequences which are feasible.

(ii) Completeness, to guarantee by repeated application of the generation of all possible sequence.

(iii) Instructive reasonableness, to generate sequences which do not differ significantly from the current sequence undergoing evolution.

3.3.3 DETERMINATION OF EVOLUTIONARY STRATEGY:

Better and better designs can be evolved by following four evolutionary strategies:

(i) First generate all feasible separation sequences resulting from the current sequence through one modification. Size and cost all the alternatives rigorously and select cheapest one as the next current sequence.

(ii) A simple variation on the above strategy allows one to employ heuristics for the selection of the next flow sheet among all the alternatives generated from the evolutionary rules.

(iii) Apply the evolutionary rules selectively and repeatedly untill an apparent optimum has been encountered; then apply the other rules to break the impasse and generate improved sequences. This is known as "depth first" strategy.

3.4 MIXED INTEGER NONLINEAR PROGRAMMING (MINLP) APPROACH:

The optimal synthesis of distillation column sequences continues to be a major problem in the design of chemical processes due to the high investment and operating costs involved in these systems. The recent trends in this area have been to address models of increasing complexity through the use of mathematical programming. The high degree of nonlinearity and the difficulty of solving the corresponding optimization models, however, have prevented methods with rigorous models from becoming tools that can be readily used by industry. Problems involving discrete and continuous variables can be successfully represented by Mixed-Integer Nonlinear Programming (MINLP).

Floudas and Paules [3] deviated from the earlier work by formulating synthesis problem as MINLP model. Pressure was treated as continuous variable. Objective function comprising of total annual cost was then optimized subjected to heat duty constraints, Energy balance constraints, LMTD constraints, Material balance constraints, critical temperature bounds, Minimum temperature approach constraints, logical and integer constraints.

Kakhu and Flower [8] also formulated problem as MILP model but extended the idea by including thermally coupled or complex columns and multi effect columns. The temperature of the hottest hot utility and that of coldest cold utility were used as limits of problem with an optimal condition of lowest allowable pressure for each distillation task. Each column in superstructure is than solved for optimization phase with parameters comprising of number of trays, reflux ratio, cost of column as a linear function of feed flow rate. They observed that due to the large number of columns generated in the superstructure, especially for large problems (5 or 6 components), efficient shortcut method for solving distillation column are necessary.

Floudas and Aggarwal [4] proposed a systematic approach for the synthesis of distillation sequences involving non-sharp splits of components by considering the key component recovery explicitly as optimization variable. A superstructure is then formulated as a mixed integer nonlinear programming MINLP problem whose solution provides an optimal distillation sequences.

Later Floudas and Aggarwal extended the synthesis approach of non-sharp separations so as to allow for various possible heat integration alternatives. The pressure of each column and key component recoveries are treated explicitly as optimization variables and a two level decomposition report was proposed for the solution of the resulting MINLP model.

3.5 PINCH ANALYSIS METHOD:

3.5.1 INTRODUCTION:

In the late 1970s, Pinch Analysis was first applied to the optimization of industrial energy systems, in response to high fuel prices during the 'energy crises'. The term "Pinch Technology" was introduced by Linnhoff and Vredeveld in 1979 to represent a new set of thermodynamically based methods that guarantee minimum energy levels in design of heat exchanger networks. Over the last decades it has emerged as an unconventional development in Process design and Energy conservation. Pinch technology emerged as a tool for the design of Heat Exchanger Network (HENs). Its key contribution was to give the engineer simple concepts which were used interactively for the evolution of an optimal Heat Exchanger Network. In 1983, Linnhoff B. and Hindmarsh E. [12] first applied this technology to design Heat Exchanger Networks for maximum energy recovery. In the next major advance, Pinch Analysis was extended to the analysis of onsite utilities, such as boilers, turbines, heat pumps and refrigeration systems, and techniques were developed for optimum design of Combined Heat and Power (CHP) systems.

Dhole V. R. and Linnhoff B. [11] proposed model for distillation column targeting using Pinch technology. They done optimization of distillation column involving options such as different reflex ratios, pressures, side condensing / reboiling and feed reheating/cooling. Alongside heat load and temperature targets, the methodology clarifies the effect of design modifications on column capital cost, also ahead of design. Proposed methodology was applicable to non-ideal multi-component systems and complex distillation configurations.

Linnhoff B. et al. [9] presented the design of individual distillation columns into context with the heat integration with the overall process. They showed with case study that good integration between distillation and overall process can results in column operating at very less utility cost. They stated that optimizing the design of individual columns in isolation from rest of the process may sometimes have a counter-productive effect, spoil opportunities for the good heat integration and therefore adversely effects the performance of the overall process.

Pinch Technology provides a simple methodology for systematically analyzing chemical process and the surrounding utility systems with the help of First and Second Laws of Thermodynamics. The First Law of Thermodynamics provides the energy equation for calculating the enthalpy change in the streams passing through the heat exchangers. The Second Law of thermodynamics determines the direction of heat flow. That is, heat energy may only flow in the direction of hot to cold. This prohibits the '*temperature crossovers*' of the hot and cold stream profiles through the exchanger. In a heat exchanger unit neither a hot stream can be cooled below cold stream supply temperature nor can a cold stream be heated to a temperature more than the supply temperature of the hot stream. In practice hot stream can only be cooled to a temperature defined by the '*Temperature Approach*' of the heat exchanger. The temperature approach is the minimum allowable temperature difference (DT_{min}) in the stream temperature profiles, for the heat exchanger. The temperature level at which DT_{min} is observed in the process is referred as "*Pinch Point*" or "*Pinch condition*". The pinch defines the minimum driving force allowed in the heat exchanger.

The advantage of Pinch Technology arises from its ability to target minimum possible utility consumption of processes even prior to design of heat recovery systems. The Pinch Design Method then guides the process engineer to develop heat exchanger networks which would be able to achieve minimum energy targets.

Onion Diagram (Fig. 3.2) illustrates the role of Pinch Technology in the overall process design. The process design hierarchy can be represented by the "onion diagram" as shown below. The design of a process starts with the reactors. Once feeds, products, recycle concentrations and flow rates are known, the separators can be designed. The basic process heat and material balance is now in place, and the heat exchanger network can be designed. The remaining heating and cooling duties are handled by the utility system. The process utility system may be a part of a centralized site-wide utility system.

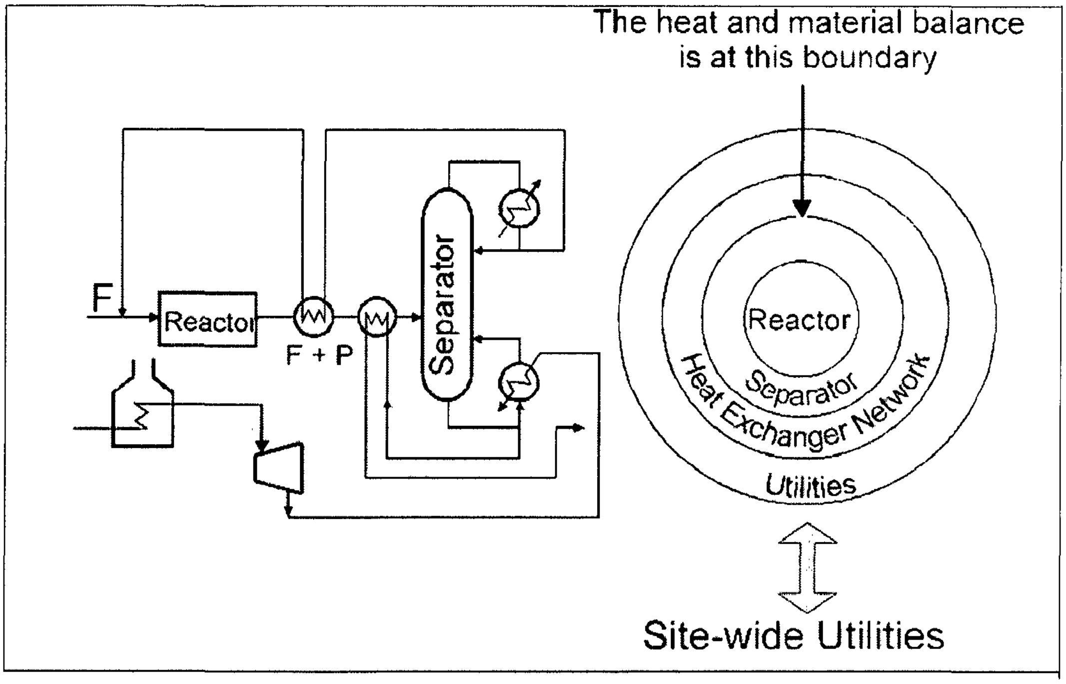


Fig. 3.2 Onion Diagram

A Pinch Analysis starts with the heat and material balance for the process. Using Pinch Technology, it is possible to identify appropriate changes in the core process conditions that can have an impact on energy savings (onion layers one and two). After the heat and material balance is established, targets for energy saving can be set prior to the design of the heat exchanger network. The Pinch design m0ethod ensures that these targets are achieved during the network design. Targets can also be set for the utility loads at various levels. The utility levels supplied to the process may be a part of a centralized site-wide utility system. Pinch Technology extends to the site level, wherein appropriate loads on the various steam mains can be identified in order to minimize the site wide energy consumption. Pinch Technology therefore provides a consistent methodology for energy saving, from the basic heat and material balance to the total site utility system.

3.5.2 OBJECTIVE OF PINCH ANALYSIS:

When the process involves single hot and cold streams, it is easy to design an optimum heat recovery exchanger network intuitively by heuristic methods. In any industrial set up the number of streams is so large that the traditional design approach has been found to be limiting in the design of a good network. With the development of pinch technology not only optimal network design was made possible, but also considerable process improvement could be discovered.

Most industrial processes involve transfer of heat either from one process stream to another process stream or from a utility system to a process stream. In the present energy crisis scenario all over the world, the target in any industrial process design is to maximize the process-to-process heat recovery and to minimize the utility (energy) requirements. To meet the goal of maximum energy recovery or minimum energy requirement an appropriate heat exchanger network (HEN) is required. The design of such a network is not an easy task considering the fact that most processes involve a large number of process and utility streams. With the advent of pinch analysis concepts, the network design has become very systematic and methodical.

Pinch analysis is used to identify energy cost and heat exchanger network (HEN) capital cost targets for a process and recognizing 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 pinch analysis is to achieve financial savings by better process heat integration i.e. maximizing process-to-process heat recovery and reducing the external utility load.

3.5.3 STEPS OF PINCH ANALYSIS:

In any Pinch analysis problem, whether a new project or a retrofit situation, a well defined stepwise procedure is followed. It should be noted that these steps are not necessarily performed on a once through basis, but additional activities such as re-simulation and data modification occur as the analysis proceeds and some iteration between the various steps is always required.

Following are the major steps performed for doing Pinch analysis:

Step 1: Hot, Cold and Utility streams identification in the process

Step 2: Thermal data extraction for Process and utility streams

Step 3: Selection of Initial DT_{min} value

Step 4: Construction of Composite curves and Grand composite curves

Step 5: Estimation of minimum energy cost targets

Step 6: Estimation of HEN capital cost targets

Step 7: Estimation of Optimum DT_{min} value

Step 8: Estimation of practical targets for HEN design

Step 9: Design of heat exchanger network (HEN)

3.5.3.1 HOT, COLD AND UTILITY STREAMS IDENTIFICATION:

- Hot streams are products streams which are required to be cooled to desired temperatures.
- Cold streams are feed streams, which are required to be heated to desired temperatures.
- Utility Streams are used to heat or cool process streams, when heat exchange between process streams is not practical or economic.

3.5.3.2 THERMAL DATA EXTRACTION FOR PROCESS AND UTILITY STREAMS:

For each hot, cold and utility stream identified, the following thermal data is extracted from the process material and heat balance flow sheet.

- Supply temperature (TS, ^{0}C)
- Target temperature (TT, ^{0}C)
- Heat capacity flow rate (CP, kcal/^{0}C), the product of flow rate and specific heat

  CP = m * Cp
- Enthalpy change (H, kcal) associated with the stream passing through the heat exchanger is given by 'First Law of Thermodynamics'

  Enthalpy Change = CP *(TS-TT)

3.5.3.3 SELECTION OF INITIAL DT_{min} VALUE:

The design of any heat transfer equipment must always adhere to the Second Law of Thermodynamics that prohibits any temperature cross over between any hot and cold stream i.e. a minimum heat transfer driving force must always be allowed for a feasible heat transfer design. Thus the temperature of any hot and cold streams at any point in the heat exchanger must always have a minimum temperature difference (DT_{min}). This DT_{min} value represents the bottleneck in the heat recovery. The value of DT_{min} is determined by the overall heat transfer coefficient and geometry of the exchanger.

For a given value of heat transfer load (Q), if a smaller value of DT_{min} is chosen, then area requirement in the heat exchanger rises and so capital cost increases. If a higher value of DT_{min} is selected then area requirement in the heat exchanger decreases, but heat recovery in the exchanger decreases and demand for external utility increases. Higher value of DT_{min} results in low capital cost, but high operating cost.

To begin the process an initial DT_{min} value is chosen and Pinch analysis is carried out. A few values based on Linnhoff March's application experience is tabulated below for the Shell and tube exchangers.

Table 3.1 DT_{min} values for different industrial sector

S. No.	Industrial Sector	Experience DT_{min} values
1	Oil Refinery	15-30^{0}C
2	Petrochemical	10-20^{0}C
3	Chemical	10-20^{0}C
4	Low temperature process	3-5^{0}C

3.5.3.4 CONSTRUCTION OF COMPOSITE AND GRAND COMPOSITE CURVES:

- **Composite Curves:**

The heat duty Vs temperatures data for all hot streams is combined to give hot composite curve. This represents the heat availability in the "combined hot stream" as a function of temperature. Similarly, the thermal data of all cold streams is combined to give the cold composite curve. The cold composite curve is shifted rightwards such that the minimum temperature difference between hot and cold composite curves is equal to a specified value (DT_{min}). The logic for shifting the cold composite curve is that heat can flow only from higher to lower temperature. Therefore, the hot composite curve should lie above the cold composite curve, at all temperature levels and minimum temperature difference should be equal to DT_{min}. The temperature at which the hot and cold composite curves come closest together is called "Pinch Point". The section of hot composite curve below which cold streams are not present has to be cooled by external utility (cooling water, refrigeration etc.) and heat duty of this section is the total cooling requirement of the process. Similarly heat duty of the section of cold composite curve above which no hot streams exist represents the total heating requirement of the process by external utilities (steam, furnace firing etc.).

The remaining part of hot and cold composite curves where both hot and cold streams are present represent the total heat interchange duty which must be recycled from hot to cold streams in order to achieve minimum utility requirements. By constructing the composite curves, the designer can establish a target for minimum utility requirements.

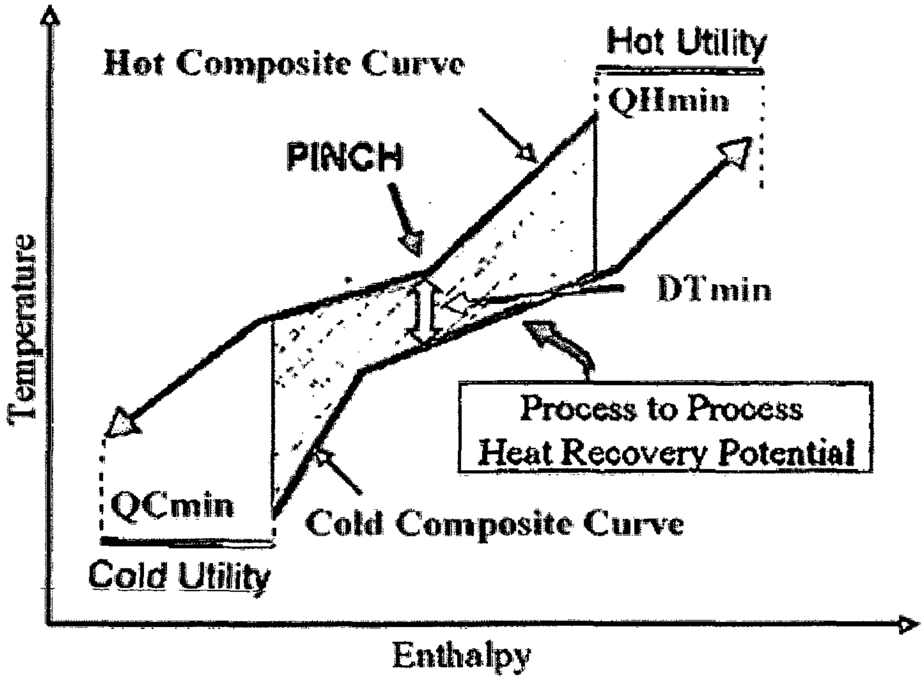


Fig. 3.3 General Combined Composite curve

- **Grand Composite Curve:**

The energy requirement for a process is supplied via several utility levels e.g. steam levels, refrigeration levels, hot oil circuit, furnace flue gas etc. The general objective is to maximize the use of the cheaper utility levels and minimize the use of the expensive utility levels. For example, it is preferable to use LP steam instead of HP steam, and cooling water instead of refrigeration. The composite curves provide overall energy targets but do not clearly indicate how much energy needs to be supplied by different utility levels.

The tool that is used for setting multiple utility targets is called the Grand Composite Curve, introduced in 1982 by Itoh, Shiroko and Umeda. The GCC shows the variation of heat supply and demand within the process. Using this diagram designer can find which utilities are to be used.

In summary GCC is one of the basic tools used in Pinch Analysis for the selection of appropriate utility levels and for targeting of a given set of multiple utility levels. The targeting involves setting appropriate loads for the various utility levels by maximizing the least expensive utility loads and minimizing the loads on the most expensive utilities.

3.5.3.5 ESTIMATION OF MINIMUM ENERGY COST TARGETS:

Once the DT_{min} is chosen, minimum hot and cold utility requirements can be evaluated from the composite curves, The GCC provides information regarding the utility level selected to meet minimum hot and cold utility requirements.

If the unit cost of each utility is known, the total energy cost can be calculated using the energy equation given below:

$$\text{Total Energy Cost} = \sum_{U=1}^{U} Q_U * C_U \qquad \text{.......................} (3)$$

Where Q_U = Duty of utility U, KW

C_U = Unit cost of utility U, Rs/KW

U = Total number of utilities used

3.5.3.6 ESTIMATION OF HEN CAPITAL COST TARGETS:

Capital cost of a heat exchanger network is dependent upon three factors:

1. the number of exchanger
2. the overall network area
3. the distribution of area between the exchangers

Pinch analysis enables targets for the overall heat transfer area and minimum number of units of a heat exchanger network (HEN) to be predicted prior to detailed design. It is assumed that the area is evenly distributed between the units. The area distribution cannot be predicted ahead of design.

AREA TARGETING: The calculation of surface area for a single counter-current heat exchanger requires the knowledge of the temperatures of streams in and out (Log Mean

Temperature Difference or LMTD), overall heat transfer coefficient (U-value), and total heat transferred (Q). The area is given by the relation

Area = Q / [U x LMTD] .... (4)

The composite curves can be divided into a set of adjoining enthalpy intervals such that within each interval, the hot and cold composite curves do not change slope. Here the heat exchange is assumed to be "vertical" (pure counter-current heat exchange). The hot streams in any enthalpy interval, at any point, exchanges heat with the cold streams at the temperature vertically below it.

The total area of the HEN is given by the equation 5.formula in Figure 8, where i denotes the *i*th enthalpy and interval j denotes the *j*th stream and LMTD denotes Log mean temperature difference in the *i*th interval.

HEN Minimum Area, $A_{min} = A_1 + A_2 + A_3 + \ldots\ldots\ldots A_i$

$$= \sum_i [(1/LMTD) * \sum_j q_j/h_j] \quad \ldots\ldots\ldots\ldots\ldots\ldots (5)$$

Where

i = *i*th enthalpy

j = *j*th stream

The actual HEN total area required is generally within 10% of the area target as calculated above. With inclusion of temperature correction factors area targeting can be extended to non counter-current heat exchange as well.

UNITS TARGETING: In designing for the minimum energy requirement (MER), no heat transfer is allowed across the pinch and so a realistic target for the minimum number of units is the sum of the targets evaluated both above and below the pinch separately.

Minimum number of heat exchanger units required for MER

$N_{min} = [N_h + N_c + N_u - 1]_{\text{Above Pinch}} + [N_h + N_c + N_u - 1]_{\text{Below Pinch}} \ldots\ldots (6)$

Where: N_h = Number of hot streams

N_c = Number of cold streams

N_u = Number of utility streams

HEN TOTAL CAPITAL COST TARGETING: The targets for the minimum surface area (A_{min}) and the number of units (N_{min}) can be combined together with the heat exchanger cost law to determine the targets for HEN capital cost (C_{HEN}). The capital cost is annualized using an annualization factor that takes into account interest payments on borrowed capital.

The equation used for calculating the total capital cost and exchanger cost law is given below.

$$C_{HEN} = [N_{min} \{a + b\,(A_{min} / N_{min})^{C}\}]_{\text{Above Pinch}} + [N_{min} \{a + b\,(A_{min} / N_{min})^{C}\}]_{\text{below Pinch}} \quad \ldots\ldots\ldots\ldots\ldots\ldots (7)$$

For the Exchanger Cost Equation shown above, typical values for a carbon steel shell and tube exchanger would be a = 16,000, b = 3,200, and c = 0.7.

3.5.3.7 ESTIMATION OF OPTIMUM DT_{min} VALUE:

To arrive at an optimum DT_{min} value, the total annual cost (the sum of total annual energy and capital cost) is plotted at varying DT_{min} values. Increase in DT_{min} values result in higher energy costs and lower capital costs and decrease in DT_{min} values result in lower energy costs and higher capital costs. So an optimum DT_{min} exists where the sum of total annual cost of energy and capital costs is lowest.

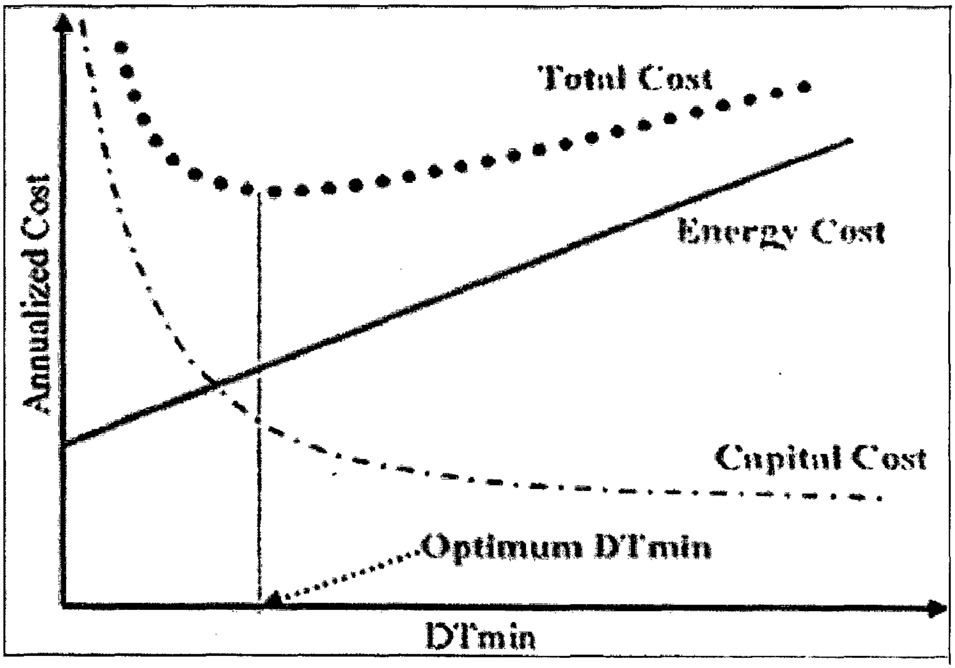


Fig. 3.4 Estimation of Optimum DT_{min} value

3.5.3.8 ESTIMATION OF PRACTICAL TARGETS FOR HEN:

The heat exchanger network designed on the basis of the estimated optimum DT_{min} value is not always the most appropriate design. A very small DT_{min} value, perhaps 8 ^0C, can lead to a very complicated network design with a large total area due to low driving forces. The designer, in practice, selects a higher value (15 ^0C) and calculates the marginal increases in utility duties and area requirements. If the marginal cost increase is small, the higher value of DT_{min} is selected as the practical pinch point for the HEN design.

Recognizing the significance of the pinch temperature allows energy targets to be realized by design of appropriate heat recovery network.

PINCH TEMPERATURE: The pinch divides the process into two separate systems each of which is in enthalpy balance with the utility. The pinch point is unique for each process. Above the pinch, only the hot utility is required. Below the pinch, only the cold utility is required. Hence, for an optimum design, no heat should be transferred across the pinch. This is known as the key concept in Pinch Technology.

Pinch Technology gives three rules that form the basis for practical network design:

(A) DO NOT TRANSFER HEAT ACROSS THE PINCH:

The network should not feature any heat transfer between hot streams above the pinch point to cold stream below the pinch point. Such heat transfer would lead to an increase in hot and cold utility requirement by same amount. (Figure 3.5)

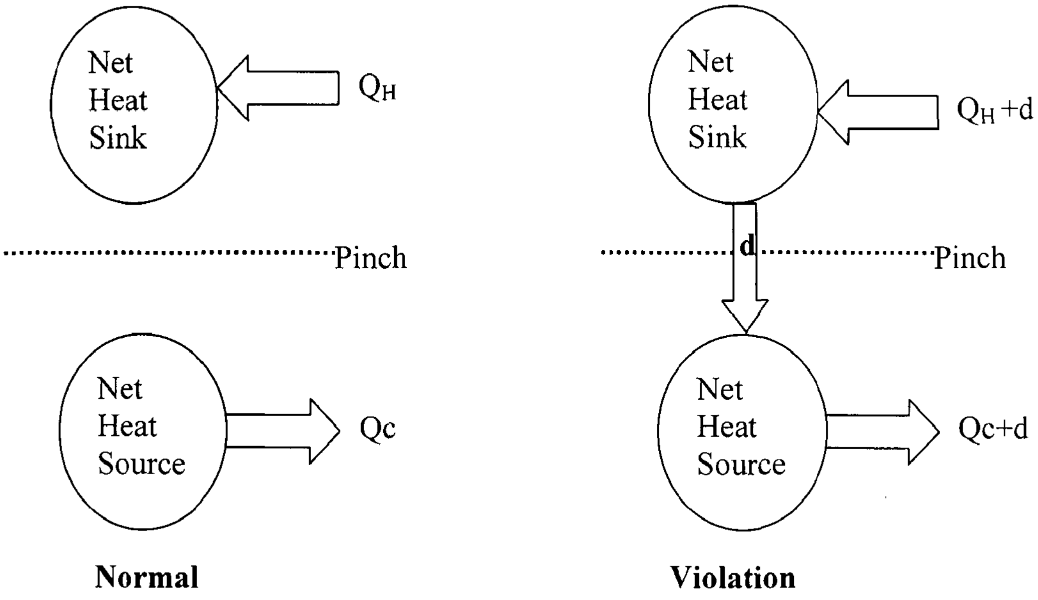
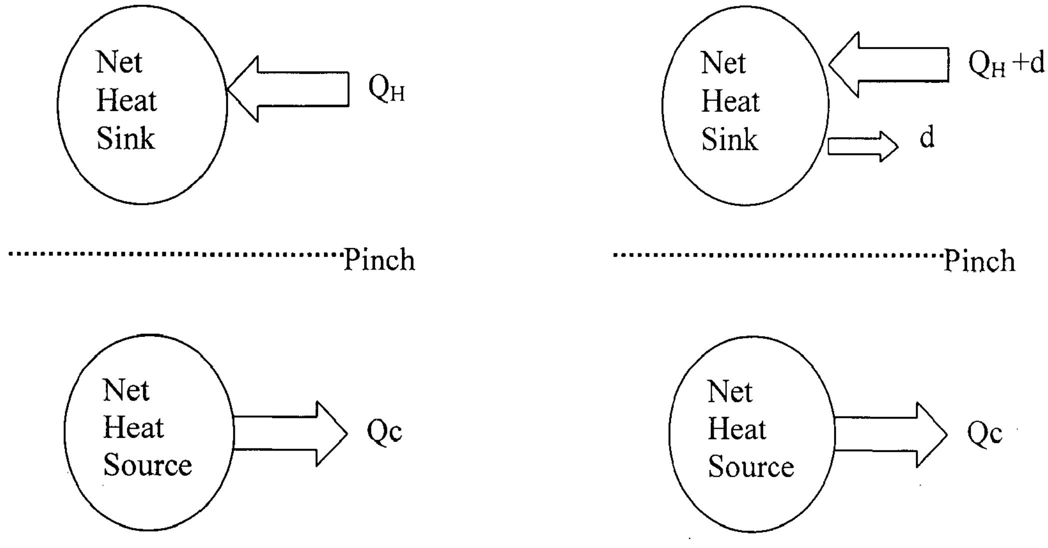


Fig. 3.5 Do not transfer heat across Pinch

(B) DO NOT COOL ABOVE PINCH POINT:

All the heat in hot process streams above the pinch point should be recycled to cold process streams and no heat should be rejected to cooling utility. Any cooling of hot streams above pinch point will result in equal increase in hot utility requirement (Figure 3.6). It may be noted that cooling would also include steam generation and although we rarely come across cases where heat above pinch point is being thrown in water or fin fan coolers, there are numerous incidents of steam being generated above pinch point resulting in equivalent increase in furnace duty.

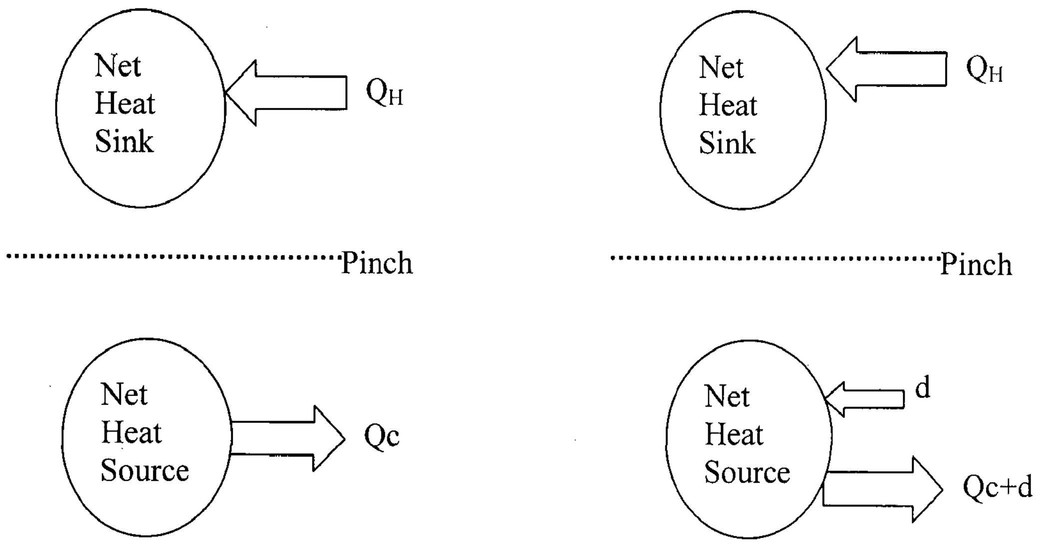


Normal **Violation**

Fig. 3.6 Do not cool above Pinch point

(C) DO NOT HEAT BELOW PINCH POINT:

None of the cold streams below pinch point should be heated by external utilities and the entire duty should be furnished through heat recycled from hot streams. Any heating of cold streams below pinch point will lead to equivalent cooling of hot streams (Fig 3.7).



Normal **Violation**

Fig. 3.7 Do not heat below Pinch point

In addition to the above pinch rules, a large number of factors must also be considered during the design of heat recovery networks. The most important are operating cost, capital cost, safety, operability, future requirements, and plant operating integrity. Operating costs are dependent on hot and cold utility requirements as well as pumping and compressor costs. The capital cost of a network is dependent on a number of factors including the number of heat exchangers, heat transfer areas, materials of construction, piping, and the cost of supporting foundations and structures.

The essence of the pinch approach is to explore the options of modifying the core process design, heat exchangers, and utility systems with the ultimate goal of reducing the energy and/or capital cost.

3.5.3.9 DESIGN OF HEAT EXCHANGER NETWORK (HEN):

The design of a new HEN is best executed using the "Pinch Design Method (PDM)". The systematic application of the PDM allows the design of a good network that achieves the energy targets within practical limits. The method incorporates two fundamentally important features:

1. Pinch design method recognizes that the pinch region is the most constrained part of the problem and consequently it starts the design at the pinch and develops by moving away
2. Pinch design method allows the designer to choose between match options.

In effect, the design of network examines which "hot" streams can be matched to "cold" streams via heat recovery. This can be achieved by employing "tick off" heuristics to identify the heat loads on the pinch exchanger. Every match brings one stream to it target temperature. As the pinch divides the heat exchange system into two thermally independent regions, HENs for both above and below pinch regions are designed separately. When the heat recovery is maximized the remaining thermal needs must be supplied by hot utility.

3.5.4 BENEFITS OF PINCH TECHNOLOGY:

1. Pinch technology tells the best that can be achieved in a given system.
2. Pinch provides lower energy consumption due to better process integration.
3. Pinch analysis provides lower energy cost due to lower consumption as well as shifting load from higher to lower cost utilities.
4. Pinch sets energy and capital cost target for an individual process or for an entire production site ahead of design.
5. Pinch gives practical targets by taking into account practical constraints and achieves better results than theoretical targets.
6. Pinch analysis on applying to debottlenecking studies leads to reduction in capital cost and decrease in specific energy demand.
7. Pinch shows, which waste heat streams can be recovered and lend insight into the most effective means of recovery.
8. Pinch gives a system-wide view of the problems.
9. Pinch helps in reducing combustion product by Emission targeting
10. Pinch Technology is in contrast to other design tools, which require detailed information about geometry, flow sheet structures etc.
11. Pinch shows the best type of Combined Heat and Power generation system that matches the inherent thermodynamic opportunities on the site.

3.5.5 APPLICATIONS OF PINCH TECHNOLOGY:

1. Heat integration – Heat exchange network
2. Mass integration- Mass exchange network
3. Distillation column targeting
4. Total site targeting
5. Water and waste water management
6. Hydrogen management in refineries
7. Emission targeting –Combustion products
8. Debottlenecking and retrofitting
9. Combined heat and power generation system

CHAPTER-4
PROBLEM STATEMENT

4.1 BACKGROUND:

For the present work a refinery situated in western India is selected. It consists of following units:

1. Crude Distillation units (CDU)
2. Diesel Hydrotreating units (DHT)
3. VGO Hydrotreating units (VGOHT)
4. Sat Gas Concentration units (SGCU)
5. Fluidized Catalytic Cracker unit (FCC)
6. Delayed Coker Unit (DCU)
7. Platformer
8. Light Naphtha Unionfining units (LNUU)
9. Heavy Naphtha Unionfining units (HNUU)
10. Aromatic unit
11. Propylene Recovery Unit (PRU)
12. Sulfur Recovery Unit
13. Captive power plant (CPP)

All these process plant utilize state of the art technologies with advanced process control schemes and have world scale capacities which makes this complex one of the most energy efficient refineries in the world. The complex has an integrated Cogeneration based Power Plant to generate power and process steam. Site steam system has steam headers at three pressure levels – HP (42 kg_f/cm^2-g), MP (17 kg_f/cm^2-g) and LP (4.0 kg_f/cm^2-g) which interconnect individual process plants with CPP.

The two CDU trains which were originally designed to process 225 KBPSD of middle east crude have since been de-bottlenecked to process 325 KBPSD of crude oil. The crude preheat trains were originally designed to deliver crude preheat temperature of 267 ^{0}C and revamped for a design preheat temperature of 259 ^{0}C for revamp case. However, the preheat trains have not been able to achieve the designed preheat temperatures. The current preheat temperatures are 240-245 ^{0}C. The crude heaters F-01 and F-51 are being operated at maximum efficiency and are not able to maintain design Heater coil outlet temperatures of 380 ^{0}C, because of which the crude throughput has been reduced to an average of 310-320 KBPD. In addition to reduced throughput, the product yields from atmospheric columns are also below design levels because of lower heater coil outlet temperature.

4.2 PROBLEM STATEMENT:

1. To carry out Pinch study of existing Crude Distillation unit to identify the schemes which enable maximum energy recovery with minimum additional heat exchanger surface area.
2. To maximize the preheat temperature of distillation column by heat integrating the Crude Distillation Unit distillation column with process streams of same unit , without increasing utility consumption
3. To see the feasible possibility of maximizing the preheat temperature of distillation column by heat integrating the Crude Distillation Unit processes with process streams of other unit , without increasing utility consumption
4. To do Pinch analysis to determine process Pinch point, target for utility consumption which involves identifying cold and hot streams, generation of Composite curves and Grand composite curves.
5. Area Targeting for Heat Exchanger network
6. Preliminary cost estimates of energy saving and payback period

4.3 PROCESS SCHEME:

Crude is pumped from Refinery Tank Farm and preheated in a series of heat exchangers (cold preheat train) to about 145^{0}C and sent to desalters V-02 and V-33 where salt is removed from the crude. Desalted crude is preheated in the second series of heat exchangers (warm preheat train) to about 180^{0}C and flashed in flash drum V-04 at about 4.5 kg_f/cm2.g pressure. The flashed vapors are piped to flash zone of atmospheric distillation column C-05. The flashed crude is further preheated in the third series of exchangers (hot preheat train), partially vaporized in crude heating furnaces F-01 and F-51, and sent to flash zone of crude column C-05. Crude is fractionated in C-05 to yield un-stabilized naphtha from column overhead and Reduced Crude Oil (RCO) from column bottoms. RCO is stripped by steam in bottom section. Four other side draws, Light Kerosene (LK), Heavy Kerosene (HK), Diesel (AGO) and Heavy Atmospheric Gas Oil (HAGO) are withdrawn from main column and stripped in strippers C-09, C-07, C-08 and C-06 respectively. Heat from vaporized crude is recovered through four circulating refluxes (Naphtha, HK, AGO, and HAGO CRs) and recycled back for crude preheating in exchangers S-01/S-04, S-11/S-02, S-08 and S-14/S-514/S-09 respectively.

The atmospheric column overhead is equipped with air fin fan coolers A-05 to partially condense overhead vapors which are sent to vessel V-12. Vapors from V-12 are compressed, combined with liquid from V- 12, condensed, cooled in fin fan coolers A-06 and sent to overhead drum V-13 from which the liquid is withdrawn as unstabilised naphtha stream and sent to saturated gas concentration unit (SGCU) for LPG recovery. The net gas compressor makes it possible to operate Crude column C-05 at low pressures (~1.05 kg_f/cm2-g) which permit high product yield and good separation at reasonable Furnace Coil Outlet Temperature (COT).

The LK and HK strippers are equipped with reboilers which utilize heat from HAGO circulating reflux. The Diesel and HAGO strippers utilize steam for stripping. The stripped LK Product preheats crude in exchanger S-05 and is finally cooled in fin fan cooler A-01 and water cooler S-06 before being sent to storage. Diesel product is sent hot to Diesel Hydrotreater Unit. HK product is cooled in fin fan cooler A-03 and sent to diesel tanks.

The RCO from crude unit is mixed with turbulizing steam, heated in vacuum furnace F-02 and flashed in vacuum column C-16. The vacuum column consists of four packed beds. Three side draws are withdrawn – Light Vacuum Gas Oil (LVGO), Heavy Vacuum Gas Oil (HVGO) and Slop Wax. Vacuum residue (VR) is stripped in bottom section trays and withdrawn from the bottom. A quench VR stream is cooled in preheat exchangers and recycled back to column to keep the VR temp around 350^0C in order to prevent cracking. Slop wax stream is recycled back to RCO heater F-02 with an aim to reduce metal content of VGO below acceptable levels. The vacuum column is operated under a very high vacuum (top pressure ~ 20 mm Hg) with the help of a bank of MP steam driven ejectors in order to maximize VGO yield from the column. Sufficient liquid flows must be maintained in various packed beds to prevent cracking of petroleum fractions.

Part of LVGO and HVGO draws are recycled back as circulating refluxes after cooling in exchangers A-08/S-20 and S-13/S-513/S-12 respectively. LVGO and HVGO products are combined with HAGO product and sent to VGOHT unit.

4.4 DESCRIPTION OF CRUDE PREHEAT TRAIN:

The Crude preheat train is divided in three sections:

(i) COLD PREHEAT TRAIN:

Simplified process scheme of cold preheat train is shown in Figure A.1 of Appendix-A. Crude is split equally into two streams. First split is preheated in S-01 A/B by Naphtha circulating reflux(CR) and S-02 by HK circulating reflux. This stream is split again- first split is preheated in S-25 A/B by VR and the other is preheated in S-03 by MP steam.

The other main crude split is preheated in S-04 A/B by Naphtha CR, S-05 A/B by LK product and S-08 by Diesel CR.

Naphtha CR is split into three streams, first going to S-01 A/B, second to S-04 A/B and the third bypasses both these exchangers mixing with the outlet streams from these two. The bypass flow is manipulated to control Naphtha CR return temperatures to the Crude column. HK-CR and Diesel-CR are also equipped with bypass arrangement around exchangers S-02 and S-08 respectively to control reflux return temperatures to crude column.

Crude streams leaving S-25, S-03 and S-08 are combined and sent to desalters at around 145^0C. Fresh wash water preheated by outgoing brine in S-07 A/B and S-27 A/B is also fed into desalters. Temperature of crude reduces by 4-5^0C in the desalters due to addition of wash water.

(ii) WARM PREHEAT TRAIN:

Simplified process scheme of warm preheat train is shown in Figure A.2 of Appendix-A. Desalted crude is again split into two streams. The first crude split stream is preheated in S-08 A by Diesel CR, S-09 by HAGO-CR and S-10 by VR. The second split stream is preheated in S-11 by HK-CR and S-12 by HVGO-CR. The two splits streams are separately fed into a flash drum where light hydrocarbons and dissolved water in crude are flashed. Flashed vapors are sent to flash zone of crude column. Flash drum temperature and pressure are maintained around 180^0C and 4.4 kg_f/cm2-g respectively. Diesel CR provides heating for Saturated gas concentration unit stabilizer reboiler before preheating crude in S-08 A/B.

(iii) HOT PREHEAT TRAIN:

Simplified process scheme of hot preheat train is shown in Figure A.3 of Appendix-A. Flashed crude is again split into two streams. The first is preheated in S-13 by HVGO-CR, S-14 by HAGO-CR, S-28 by slop wax and S-15 by VR before being fed into crude heater F-01. The second hot train is identical to the first, and preflashed crude is preheated in S-513 by HVGO-CR, S-514 by HAGO-CR and S-513 by VR before being fed into crude heater F-51. HVGO-CR, HAGO-CR and VR streams are also split equally to be cooled in S-13 / S-513, S-14 / S-514 and S-15/S-515 respectively. HAGO-CR from crude column is sent to LK stripper reboiler S-18 and HK stripper reboiler S-16, before it is cooled in S-14 / S-514. The final crude preheat temperature from S-15 / S-515 is around 240-247°C.

CHAPTER-5

RESULTS AND DISCUSSION

This chapter discusses the salient results obtained by Pinch Analysis. To see the feasible possibility of maximizing the preheat temperature of distillation column by heat integrating the Crude Distillation Unit processes with process streams of same unit and of different units are analyzed, without much increase in both heat exchanger area as well as utility requirement.

5.1 ANALYSIS OF PROCESS:

The first step in Pinch Analysis study is to obtain the process stream data. The stream set data comprises information on both hot and cold streams. In this context, a hot stream is defined as a stream which requires to be cooled and a cold stream is which requires to be heated. Thus the streams such as crude, desalted crude, flashed crude, furnace inlet crude etc are cold streams while pumparounds and product streams are hot streams. For each hot and cold stream in the system, the following data is stabilized.

- The start or supply temperature of each stream
- The final target temperature for each stream
- The heat capacity flowrate CP (m.Cp) for each stream, which is the amount of heat required to raise stream temperature by one degree.
- The representative film heat transfer coefficient.

5.2 DIFFERENT SCHEMES FOR ENERGY SAVING:

For maximized preheat temperature by various possibilities two different cases Case-B and Case-C are conceptualized and then it is compared with the existing case Case-A.

5.2.1 CASE- A (EXISTING CASE):

Case A is existing case of Crude Distillation Unit having less preheat temperature of 247^0C. In this case, Preheat train consists of four cold streams and ten hot streams. Four cold streams Crude A, B, C, D are raw crude, desalted crude, flashed crude and heater inlet crude respectively. In this case, MP steam is used as external hot utility. Heat exchanger network for all cold, warm and hot preheat train for Case-A are shown in figure A-1, A-2 and A-3 respectively of Appendix A.

Stream details for Case-A are given in Table A-1 in Appendix A.

From data given in Table A-1 and Table A-3 Problem Table Algorithm for Case-A is generated below for a DT_{min} of 15^0C.

Table 5.1 Problem Table Algorithm for Case-A

Temperature	$\sum(CP)_c - \sum(CP)_h$	Q_{int}	Q_{cas}	R_{cas}
387.5	0.0	0.0	0.0	196.6
342.5	1.5	67.5	-67.5	129.1
331.5	1.1	12.4	-79.9	116.7
301.1	1.1	33.6	-113.5	83.1
280.5	0.8	17.2	-130.7	65.9
254.5	0.3	7.6	-138.3	58.3
252.5	-0.1	-0.2	-138.1	58.5
245.5	0.2	1.1	-139.3	57.4
242.8	0.2	0.5	-139.8	56.9
227.5	-0.2	-3.7	-136.1	60.5
214.5	0.0	0.4	-136.5	60.1
211.5	-0.2	-0.5	-136.0	60.6
207.5	-0.3	-1.1	-134.9	61.7
206.5	-17.3	-17.3	-117.6	79.0
194.0	-0.3	-3.4	-114.2	82.4
193.5	0.3	0.1	-114.3	82.3
187.5	0.3	1.6	-116.0	80.7
163.0	0.4	9.0	-125.0	71.6
152.5	0.8	8.4	-133.4	63.3
137.8	0.6	8.9	-142.3	54.3
125.5	0.8	9.9	-152.2	44.4
81.5	0.1	3.4	-155.7	41.0
45.9	0.8	28.8	-184.4	12.2
32.5	0.9	12.2	-196.6	0.0

It is clear by Problem Table Algorithm that present energy targeting problem is a threshold kind of Pinch problem, as no cold utility is required for this case. Pinch point occurs at a temperature of 32.5 ^{0}C. MP Steam at mass flow rate of 38 tonnes /h is used as hot stream in cold preheat train. In addition to MP steam used as a hot stream, furnace duty requirement is 196.6 Mkcal/h. This hot utility demand is fulfilled by furnace by combustion of fuel oil.

5.2.1.1 COMBINED COMPOSITE CURVE:

Hot and Cold composite curves are generated from data given in Table A-4. Hot and Cold Composite curves again verify that cold utility and hot utility requirement are 0 and 196.6 Mkcal/h respectively. Process to Process heat exchange is equivalent to 224.2 Mkcal/h.

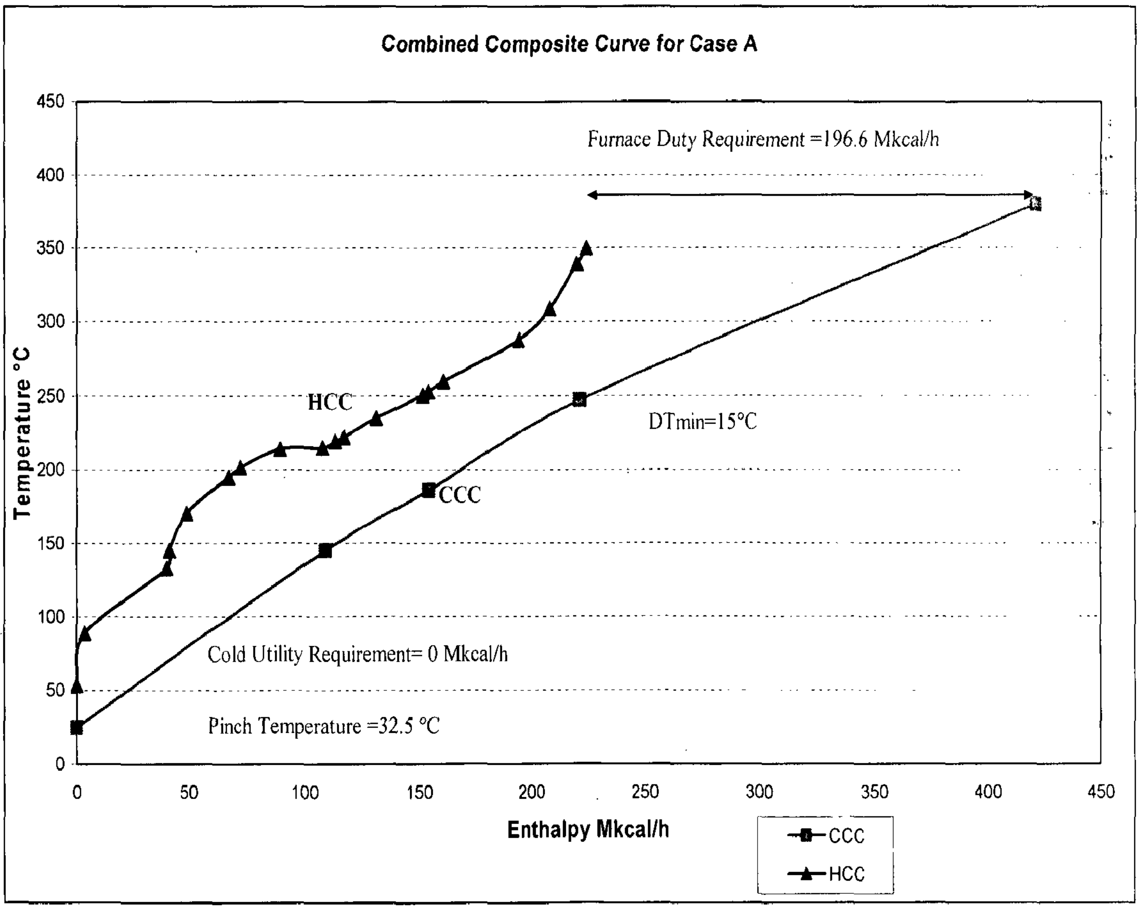


Fig 5.1 Combined Composite Curves for Case-A

5.2.1.2 GRAND COMPOSITE CURVE:

Grand Composite Curve is drawn with the help of Problem table algorithm table for a DT_{min} of 15^0C. Grand Composite Curve shows the heat available at various temperature intervals.

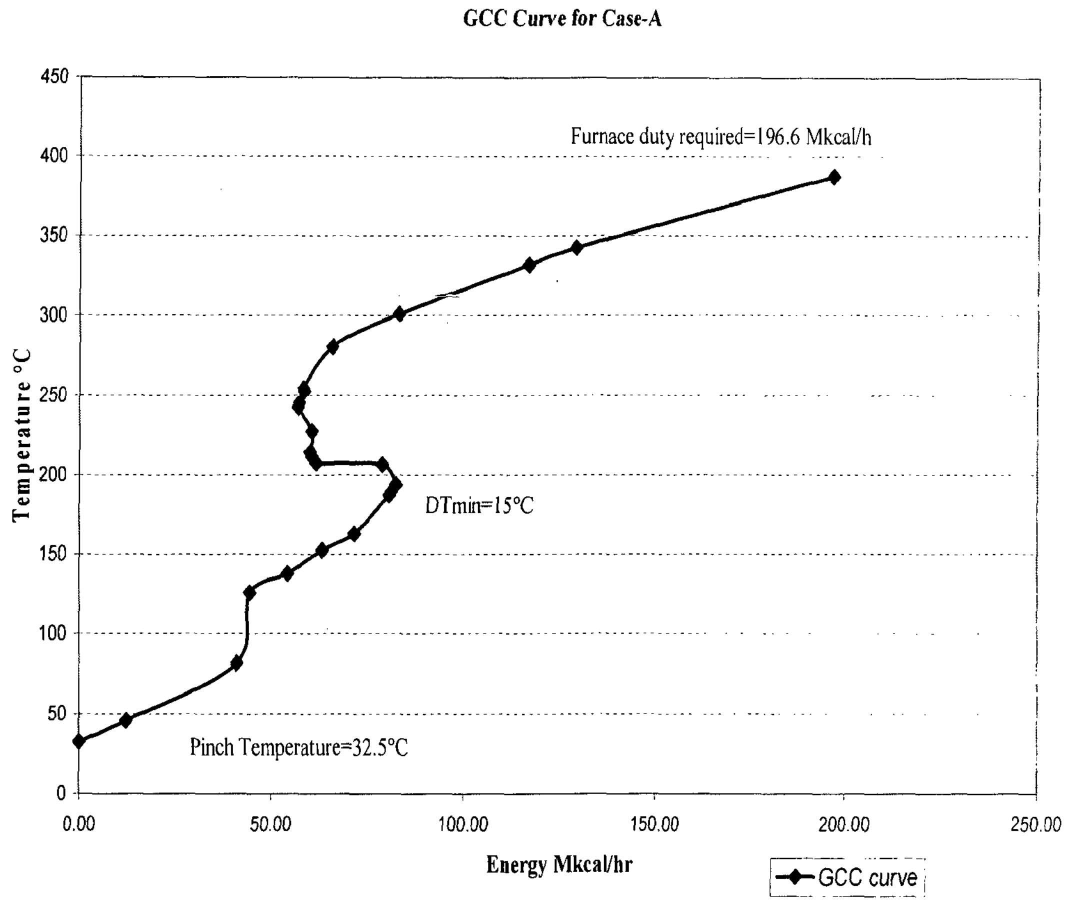


Fig 5.2 Grand Composite Curve for Case-A

Total Area available in Heat exchangers preheat train including stand by exchangers for Case-A is 27095 m^2.

5.2.2 CASE B:

In Case-B, Preheat exchanger train consists of, four cold streams and thirteen hot streams. Four cold streams Crude A, B, C, D are raw crude, desalted crude, flashed crude and heater inlet crude respectively. In this case, LVGO circulating reflux of Vacuum column and Sweet VGO from Vacuum Gas Oil Hydrotreater (VGOHT) are also considered as hot stream for achieving Furnace Crude inlet temperature of 267-270 ^{0}C. Sweet VGO of 310^{0}C is considered for heating flashed crude in hot preheat train. Sweet VGO from VGOHT unit is finally sent to FCC unit after cooling in steam generators. There does not appear to be any operation problem in pumping this hot stream to CDU unit for crude preheating instead of steam generation in VGOHT unit, before sending to FCC unit. In addition to above streams, LP and MP steam are considered as external hot utility in preheat train.

Heat exchanger network for all cold, warm and hot preheat train for Case-B are shown in figure A-5, A-5 and A-6 of Appendix A.
Stream details for Case-B are given in Table A-5 in Appendix A.

Problem Table Algorithm for Case-B is generated below for a DT_{min} of 15^0C.

Table 5.2 Problem Table Algorithm for Case-B

Temperature	$\sum(CP)_c-\sum(CP)_h$	Q_{int}	Q_{cas}	R_{cas}
387.5	0.0	0.0	0.0	133.1
342.5	1.5	67.5	-67.5	65.6
331.5	1.1	12.4	-79.9	53.2
302.5	1.1	32.0	-111.9	21.2
301.1	0.9	1.2	-113.1	20.0
280.5	0.6	12.0	-125.2	7.9
274.5	0.0	0.3	-125.4	7.7
252.5	-0.4	-7.8	-117.6	15.5
245.5	-0.1	-0.6	-117.0	16.1
242.8	-0.1	-0.2	-116.8	16.3
227.5	-0.5	-7.5	-109.3	23.8
214.5	-0.2	-2.9	-106.5	26.6
211.5	-0.4	-1.3	-105.2	27.9
207.5	-0.5	-2.1	-103.1	30.0
206.5	-17.5	-17.5	-85.6	47.5
204.5	-0.5	-1.0	-84.6	48.6
194.0	-0.3	-2.9	-81.7	51.4
193.5	0.3	0.1	-81.8	51.3
187.5	0.3	1.6	-83.5	49.7
163.0	0.4	9.0	-92.5	40.6
152.5	0.8	8.4	-100.9	32.3
142.5	0.6	6.1	-106.9	26.2
141.5	-25.4	-25.4	-81.5	51.6
137.8	0.8	3.0	-84.5	48.6
125.5	0.8	9.9	-94.4	38.7
99.5	0.1	2.0	-96.5	36.7
81.5	-0.1	-1.3	-95.2	38.0
61.5	0.7	13.2	-108.3	24.8
45.9	0.8	12.6	-120.9	12.2
32.5	0.9	12.2	-133.1	0.0

This energy targeting problem is also a threshold kind of Pinch problem, as no cold utility is required for this case. Pinch point occurs at a temperature of 32.5 ^0C. 38 tonnes per hour of MP Steam and 53 tonnes per hour of LP Steam are used as hot utility in cold preheat train. In addition to LP and MP steam as hot utility, furnace duty requirement is 133 Mkcal/h.

5.2.2.1 COMBINED COMPOSITE CURVE:

Hot and Cold composite curves are generated from data given in Table A-6. Cold utility and hot utility requirement for Case-B is 0 and 133 Mkcal/h respectively. Process to Process heat exchange is equivalent to 280 Mkcal/h.

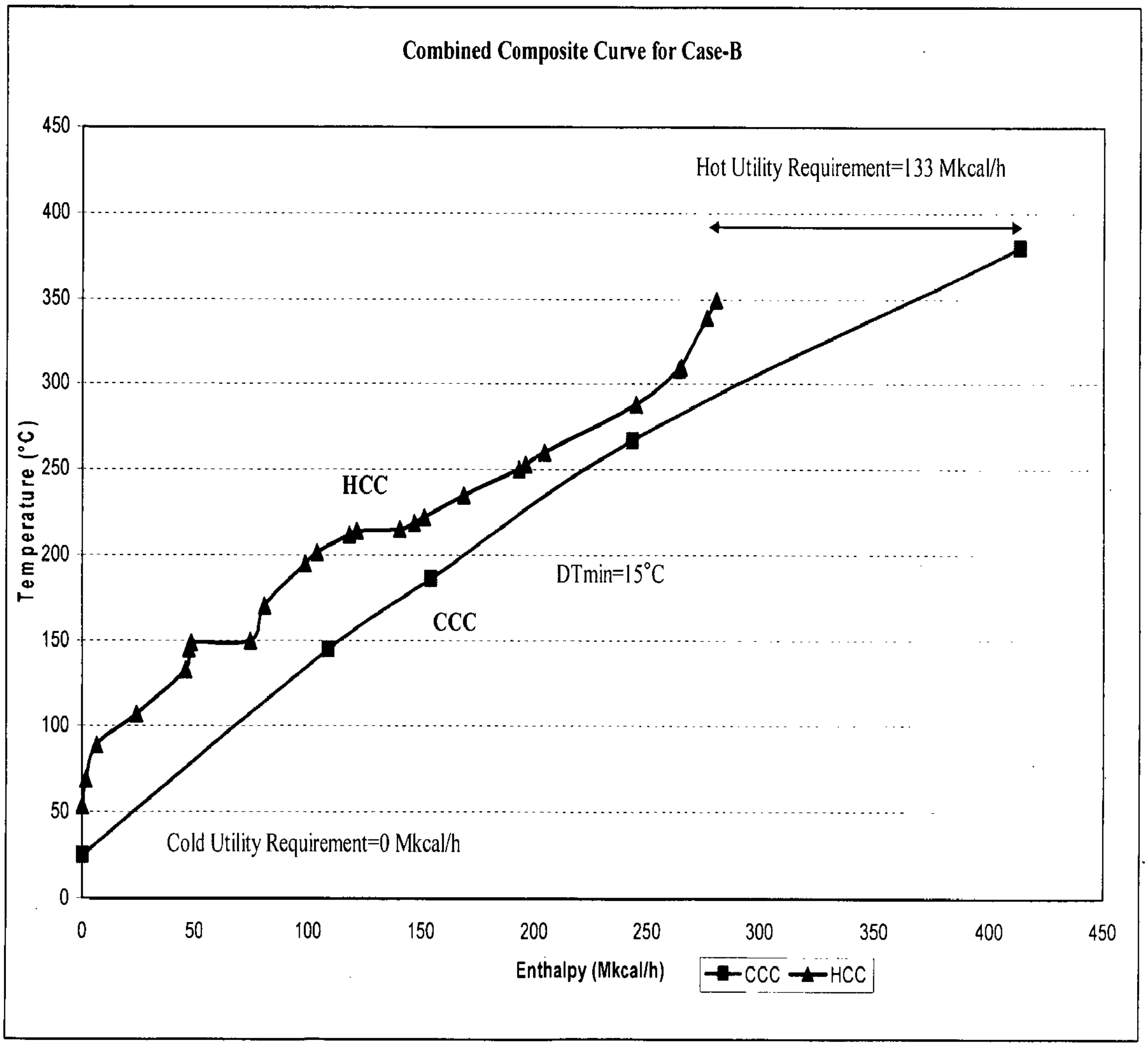


Figure 5.3 Combined Composite Curves for Case-B

5.2.2.2 GRAND COMPOSITE CURVE:

Grand Composite Curve is drawn with the help of Problem table algorithm table for a DT_{min} of 15^0C. Grand Composite Curve shows the heat available at various temperature intervals.

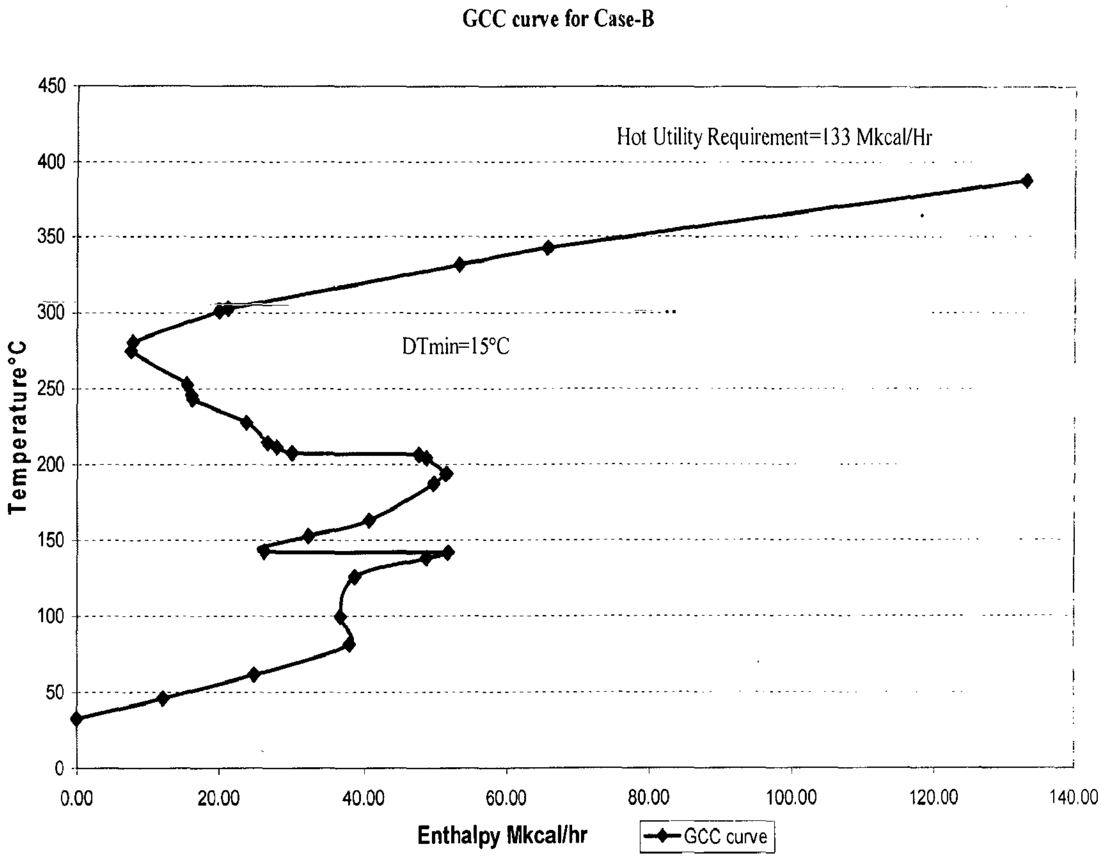


Figure 5.4 Grand Composite Curve for Case-B

For estimating the amount of heat exchanger surface required, Area targeting for Case-B is done and shown in Table A-7 of Appendix A. Total Area required in Heat exchangers preheat train for Case-B comes out to be 41785 m^2.