INTEGRATION OF DISTILLATION COLUMN USING PINCH TECHNOLOGY

A DISSERTATION

Submitted in partial fulfilment 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, **"INTEGRATION OF DISTILLATION COLUMN USING PINCH TECHNOLOGY"** in the 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"**, 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 2007 to June 2008, under the esteemed guidance of **Dr. Bikash Mohanty**, Professor, Department of Chemical Engineering, Indian Institute of Technology Roorkee.

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

Date: 23rd June, 2008

Place: IIT Roorkee

(Gopalkrishna Pednekar)

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

In chemical industries, the task of separation such as distillation is an energy intensive process, and it is still the most widely used technique for fluid separations. Distillation columns are used for about 95% of liquid separations and the energy used for this process accounts for an estimated 3% of the world energy consumption. 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 hold great promise for energy savings.

The above fact makes it imperative to have a technique similar to pinch analysis for improving distillation column designs. Integration of the distillation column using pinch technology is one such technique based on thermodynamics for identifying appropriate design modifications with respect to energy conservation. These design modifications makes it possible to save substantial amount of energy resulting in optimum or near-optimum column design.

Integration of the column can be within the column or with the background process. A lot of work is carried out in energy savings in distillation column in which much of the work is related to integration of column with the background process and little work is done on internal integration of column.

The present study is related to internal integration of distillation column using pinch technology by considering the column modification using Column Grand Composite Curve (CGCC). CGCC is a profile of net heat surplus or deficit over various trays corresponding to minimum thermodynamic condition (MTC) of the column, on the temperature-enthalpy axis or stage-enthalpy axis. The MTC for a distillation column represent a reversible operation or zero thermodynamic loss in the column.

Aspen Plus software is used to simulate the column and to generate column targeting results, to plot required CGCC for the problem under study and to test different strategies such as feed location, reflux ratio modification, feed conditioning and side-reboiler/ side-condenser to decrease the heat load of the column so that savings in operating cost could be made.

By the addition of pre-heater the load on distillation column can be curtailed to some extent. An optimally designed pre-heater will obviously do the job in a better way. It is a well known fact that optimal design of heat exchanger is a tedious work and takes substantial time. In the present work a heat exchanger optimization technique is refined and tested using an equation based technique based on Bell's method in which constraints are plotted on pressure drop diagram to obtain minimum –area and – total annual cost.

A case study of the kerosene pre-fractionation unit of a refinery is considered and Column Targeting technique is applied. Two options are considered; Option A: Without capital investment and Option B: With capital investment. The economic analysis for Option A indicates the annual saving of Rs. 25.25 Crores whereas for Option B the annual saving is Rs. 48.14 Crores with an additional investment in preheater and side reboiler. The payback period of the additional equipment i.e. preheater and reboiler comes out to be 7.4 days. This payback period can be further reduced if preheater and reboiler are designed optimally. By adopting the optimization techniques developed in the present work the payback period of preheater can be reduced from 7.4 days to 3.8 days. Similarly the overall payback period for additional equipments can be reduced from 7.4 days to 6.5 days.

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NOMENCLATURE

A	Heat exchanger area
A_b	Clearance area between bundle to shell
A_w	Window area
A_{sb}	Shell to baffle clearance area per baffle
A_{tb}	Tube to baffle clearance area per baffle
B	Bottom product molar flow
$CGCC$	Column Grand Composite Curve
C_p	Specific Heat capacity of fluid
d_i	Diameter (inner) of tube
d_o	Diameter (outer) of tube
D	Distillate flow rate
D_b	Bundle diameter
D_s	Diameter of shell
F	Molar Feed flow
F	Correction factor for LMTD
F_b'	Bypass correction factor for shell side pressure drop
F_L'	Leakage correction factor for the pressure drop
G	Mass flow rate of fluid stream unit cross sectional area
h	Heat transfer coefficient for fluid stream
h_{oc}	Shell side heat transfer coefficient for ideal tube banks
H	Enthalpy
H_b	Height of the baffle chord to the top of tube bundle
k, k_s, k_t	Thermal conductivity of the tube wall, shell side fluid, and tube side fluid
L	Liquid molar flow
L	Length (effective) of tubes
L_{bc}	Baffle spacing
MTC	Minimum Thermodynamic Condition

M	Mass flow rate of fluid stream
N	Stage number
N_b	Number of baffles
N_c'	Number of rows crossed in series from end to end of the shell
N_{cv}	Number of constrictions, tube rows encountered in cross flow zone
N_t	Number of tube per shell
N_w	Number of tubes in window zone
N_{wv}	Number of restrictions for cross flow in window zone
NS	No of shell pass
NT	Number of tube pass
$PNMTC$	Practical Near Minimum Thermodynamic Condition
P_t	Tube pitch
q	Quantity (related to feed condition)
Q	Heat duty
R_{bs}	Ratio of baffle spacing to shell diameter
R	Reflux ratio (L/D)
R_d	Dirt factor
Re	Reynolds number for fluid stream
R_w	Ratio of number of tubes in window zones to the total number in the bundle
T	Temperature
TAC	Total annual cost
u	Velocity of fluid stream
u_w	Velocity of fluid stream in window zone
V	Vapor molar flow
x	Mole fraction in liquid phase
y	Mole fraction in vapor phase
z	Mole fraction in feed
Δ	Enthalpy difference
λ	Heat of vaporization
α	Relative volatality
μ	Viscosity of fluid stream

ρ	Density of fluid stream
ΔP	Pressure drop for fluid stream
ΔP_i	Ideal tube bank pressure drop

Subscript

B	Bottom product
C, c, cond	Condenser
CGCC	Column Grand Composite Curve
def	Deficit
D	Distillate
F, feed	Feed
H	Heavy key
L	Liquid, light key
max	Maximum
min	Minimum
N	Stage number
r, reb	Reboiler
R	Rectifying curve
s.reb	Side reboiler
s	Shell side
t	Tube side
V	vapor
w	Window zone

Superscript

*	Equilibrium condition

CHAPTER 1

INTRODUCTION

Distillation is highly energy intensive process comprising a sizeable portion of the energy consumption in any process 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 for 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.

On the other hand, pinch technology is a methodology for minimizing energy consumption of chemical processes by calculating thermodynamically feasible energy targets (or minimum energy consumption) and achieving them by optimizing heat recovery systems, energy supply methods and process operating conditions. The point of closest approach between the hot and cold composite curves is the pinch point and is where design is most constrained (Linnhoff et al., 1983). Similar case is observed in distillation column where a pinch point divides the column into two zones; one is a hot zone (reboiler side) and other is the cold zone (condenser side). The pinch point forms the basis and the starting point for column integration.

Integration of distillation column can be broadly categorized into two main categories.

- Integration of column with other columns or an overall process.
- Integration within the column

Over the years much literature has appeared on energy saving in distillation. Among these much of the work is on integration of the column with the background process (Naka et al, 1980; Linnhoff et al., 1983; Shenoy, 1995; Ficarella and Laforgia, 1999; Sunden; Hewitt et al., 1999) and very few studies are available on the integration of column internally (Dhole and Linnhoff, 1993; Bandyopadhyay et al., 1998).

The present study illustrates the techniques available for the integration of the column internally followed by a case study that eventually quantifies the work. When the feed point and the pinch point matches it is said that this is the point that will result in minimum reboiler and condenser duty. But pinch and feed point matches for a series of points. Therefore once this match is found one should search for other alternatives of column modifications such as Feed plate location, Reflux Ratio Modification, Feed Conditioning (Preheating/Precooling) and Side -reboiler/-condenser to further reduce energy consumption.

1.1 OBJECTIVE OF PRESENT STUDY

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

1. To carry out a detailed column analysis of the existing kerosene pre-fractionation unit of a refinery based on Column Grand Composite Curve (CGCC) and to identify options which enable maximum energy recovery.
2. To carry out necessary modifications to optimize column based on energy targets.
3. To carry out preliminary economical analysis to facilitate easy comparison between various options.
4. To apply equation based heat exchanger optimization technique based on Bell's Method to find minimum –area and –total annual cost and to show its impact on present study.

CHAPTER 2
LITERATURE REVIEW

Pinch technology has demonstrated that good process integration pays off through simplicity of plant design and good use of energy. The study of distillation columns using Pinch Analysis tools is the latest in the purview of pinch (Dhole and Linnhoff, 1993). It is called Column targeting. This new approach facilitates identification of improvements in column design as well as its synergetic integration with the background process.

2.1 INTEGRATION OF COLUMN WITH BACKGROUND PROCESS

The traditional heat integration of distillation column with a background process is based on the appropriate placement of the column in the temperature/pressure domain to make the best use of process cold stream in the reboiler and process hot stream in condenser (Smith and Linnhoff, 1988). Often, however the column box cannot be placed within the process composite curve. The result is that no decrease in overall heat load of the process is achieved. It then becomes clear that it is the position of the column relative to the pinch that is significant.

2.1.1 Placement of Distillation Column

2.1.1.1 Distillation column across the pinch

As shown in Fig. 2.1, heat Q_{reb} is required at a temperature higher than the pinch temperature and heat Q_{cond} is returned below the pinch temperature. In other words, heat is taken from the part of the process which is a sink and added to the part of the process which is a source. As a result extra Q_{reb} units of hot utility must be imported and an extra Q_{cond} units of heat is rejected. Heat must be transferred across the pinch through the column and we pay for this heat in increased utility usage, both hot and cold, over and above the minimum. Therefore, it is not advantageous to integrate the column across the pinch.

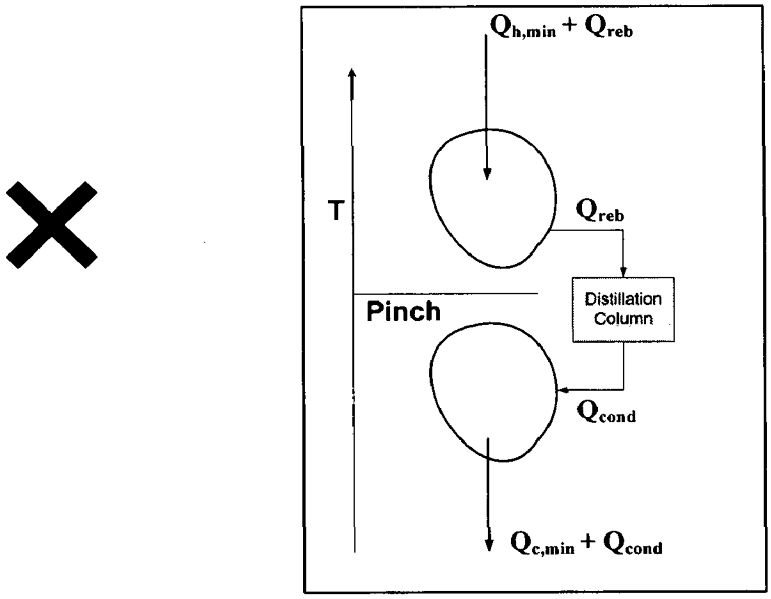
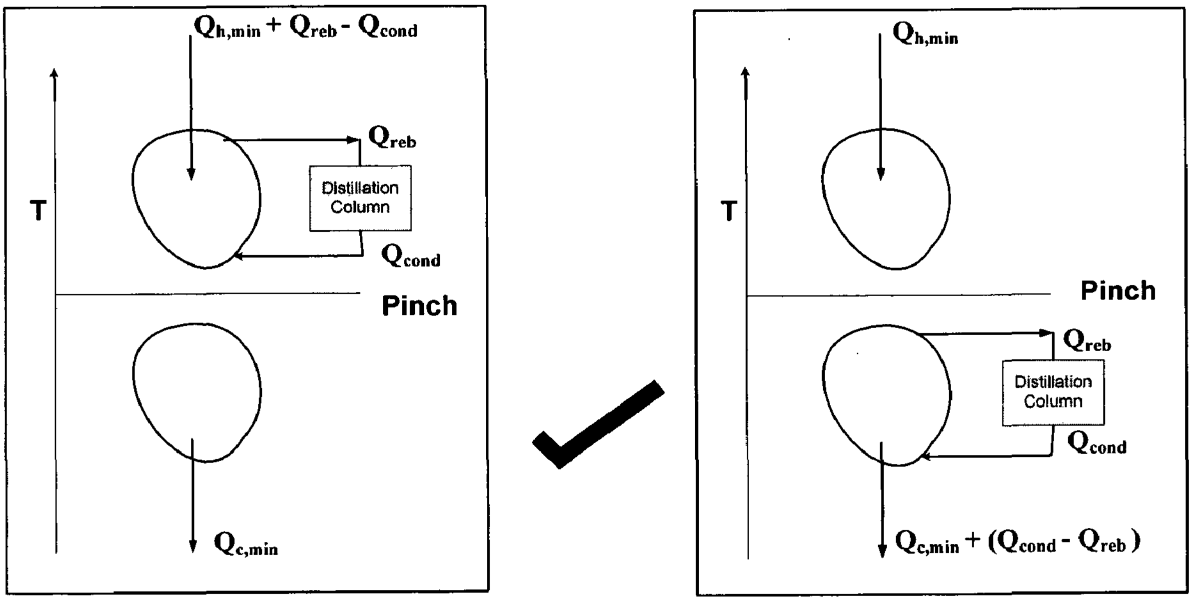


Figure 2.1. Column Across the pinch



(a) (b)

Figure 2.2 Column Above and below the pinch

2.1.1.2 Distillation above or below the pinch

Consider a column entirely above the pinch (Fig 2.2a), where only the process sink is affected. Heat Q_{reb} is taken from above the pinch and heat Q_{cond} returned at a temperature also above the pinch. The column borrows heat from the process and returns it while still usable. Here the change in the consumption of hot utility to keep the pinch flow at zero is only the difference between the two loads, i.e. an increase if $Q_{reb} > Q_{cond}$ or a decrease if $Q_{cond} > Q_{reb}$. However Q_{cond} is often similar to Q_{reb}, in which case there will be hardly any change in utility usage. Below the pinch we obtain the analogous result (Fig 2.2b). We need no extra cold utilities for $Q_{reb} = Q_{cond}$, a marginal increase for $Q_{reb} < Q_{cond}$ and a marginal decrease for $Q_{reb} > Q_{cond}$. In other words there will be no or marginal extra heat duty required if the column in placed above or below the pinch. Therefore, as far as possible column should be placed above or below the pinch.

It is not always necessary that the heat load Q_{reb} come from the process for the column totally above the pinch. It can be introduced directly from hot utility as shown in Fig. 2.3. In other words the reboiler need not be integrated with the rest of the process. However the condenser must be integrated since it is vital that it rejects heat into the process and not into cold utility. Below the pinch the logic is analogous (Fig. 2.3). The reboiler must be integrated but the condenser need not be. Thus only the condenser or the reboiler needs normally to be integrated with the process. This obviously simplifies operability problems associated with integrated distillation columns

There is a limit on the heat loads that can be borrowed from any process. Sufficient heat flow must remain in the process at all temperatures spanned by the column. In Fig. 2.4 the requirement is that Q_2 and Q_3 are greater than Q_{cond} prior to integration of the column. If the condenser only is to be integrated, as in Fig. 2.5 all heat flows above the condenser temperature must be greater than Q_{cond} to begin with. Analogous logic applies below the pinch.

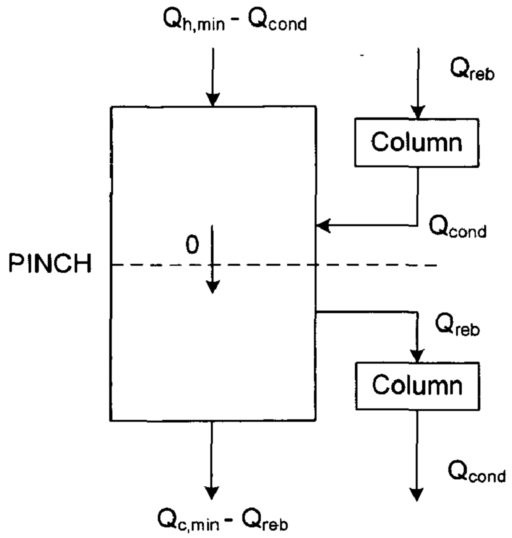


Figure 2.3. Other possible alternatives

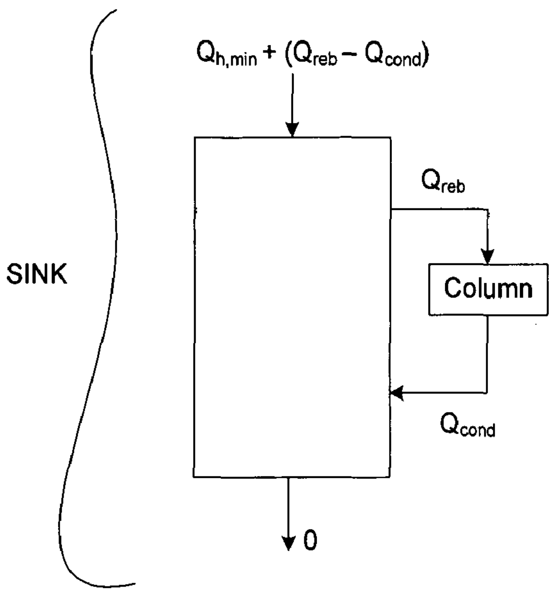


Figure 2.4 Heat load limit: General

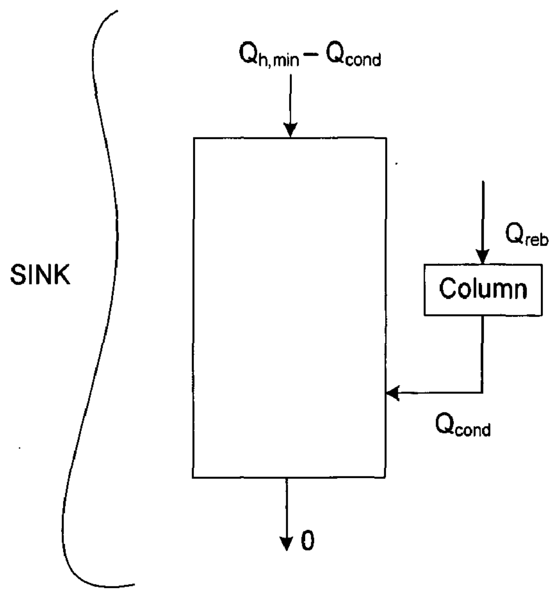


Figure 2.5. Heat load limit : condenser integration only

2.1.2 EFFECTS OF ALTERING COLUMN CONFIGURATIONS

It is now clear that provided a distillation column operates away from the pinch, and there is sufficient heat flow available, only marginal, or no, extra utilities are required for the distillation. If either of these conditions are not satisfied then we can alter column conditions to make integration possible away from the pinch.

2.1.2.1 Pressure changes

Many important design parameters, e.g. relative volatility, vapor density, shell thickness, etc. are influenced by pressure. However its most important influence in the present context is in determining the condenser and reboiler temperatures, and hence the levels of heating and cooling required. These temperatures are crucial as they determine the position of the column relative to the pinch. If the column is placed across the pinch we can in either increase or decrease the pressure, thus changing the column's position relative to the pinch.

Increasing the pressure. ~~Here we aim to integrate the column~~ condenser by lifting it above the pinch. The separation will generally become more difficult (the relative volatility decreases) requiring either more plates or a larger reflux ratio. However, the latent heat of vaporization decreases, compensating to some extent for the increased reflux ratio. The increase in the number of plates is offset by the reduction in column diameter because of increased vapor density. These conflicting trends usually result in there being little variation in column costs with increased pressure until some upper limit is reached. This limit will probably be defined by unacceptably high reboiler temperatures, either because of thermal decomposition of the bottom product or because of the lack of a sufficiently hot heating medium (process or utility).

Decreasing the pressure. By decreasing the pressure we hope to integrate the column reboiler. At lower pressures, in general, the separation is easier. Lower limits exist, however, and are usually fixed either by the desire to avoid refrigeration or by a reluctance to operate under vacuum.

2.1.2.2 Split column loads

It may be that even after all possible pressure changes have been explored there is no position which can totally accommodate the distillation heat loads. In such a situation one possibility is to split the column load into two or more smaller loads. This essentially means splitting the column feed and using two or more columns instead of one (Fig. 2.6). The pressures of each column must then be chosen such that no column operates across the pinch and all intermediate heat flows in the cascade are positive. Each column reduces the process heat flows by less than the original column would. Once again no extra utilities are needed. However two columns will be more expensive than one in terms of capital. The extra cost must be offset against the savings in energy. Usually, schemes like that in Fig. 2.6 would only be worth considering for large distillation loads.

2.1.2.3 Thermal coupling

An alternative solution when heat flows are limiting, integration possibilities is to reduce the heat load by thermal coupling. Thermal coupling is possible when multi-column arrangements produce a number of products from a multi-component mixture. A side stream rectifier is shown in Fig. 2.7. All of these arrangements consist of two columns coupled via liquid and vapor side-streams. This coupling eliminates at least one reboiler and/or condenser and reduces the total heat load to be handled as shown in Fig. 2.7. Thus, if the flows in the cascade are limiting integration opportunities thermal coupling is worth considering. It may be possible to accommodate the smaller loads required by the thermally coupled arrangement where larger loads associated with the conventional arrangements will not fit in.

2.1.2.4 Intermediate reboilers and condensers

In a conventional distillation column, all heat is added and removed at the extremities of the column, and hence at the most extreme temperature levels. It is possible, however, to add or remove heat at any plate within the column. In traditional design practice, this is only worthwhile if it allows cheaper heat sinks or sources to be used, e.g. lower pressure steam or less severe levels of refrigeration. Thus when considering a column in isolation, intermediate reboiling and condensing are only likely

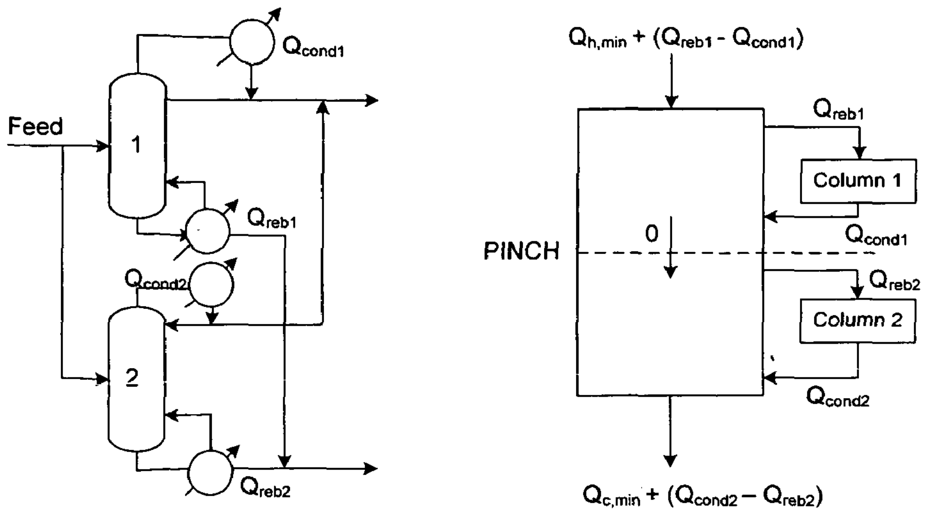


Figure 2.6. Splitting the column load

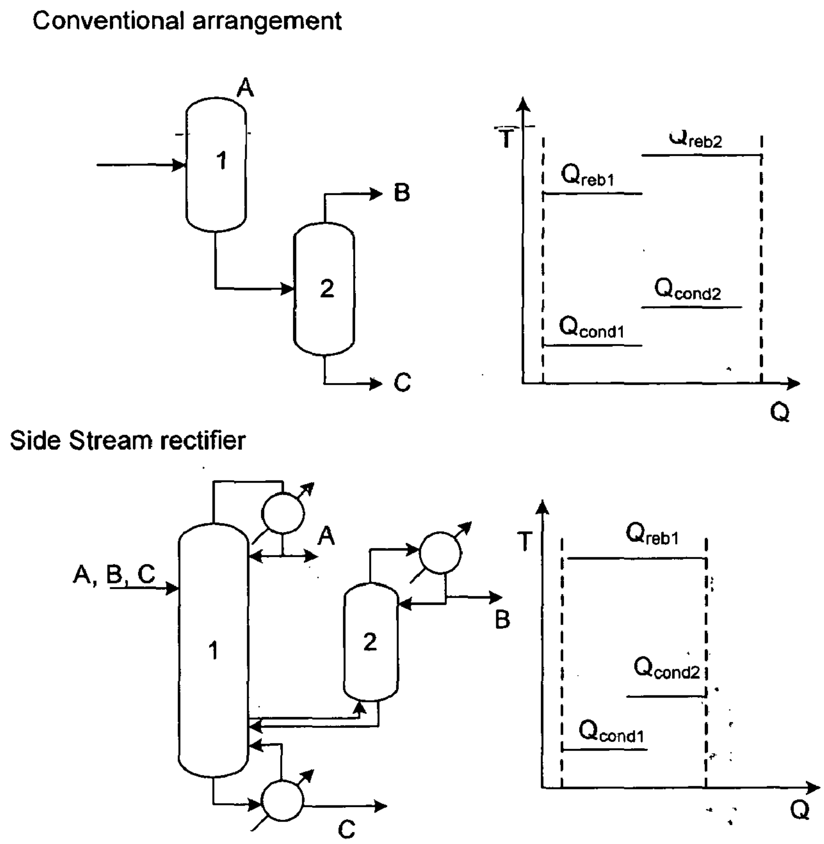
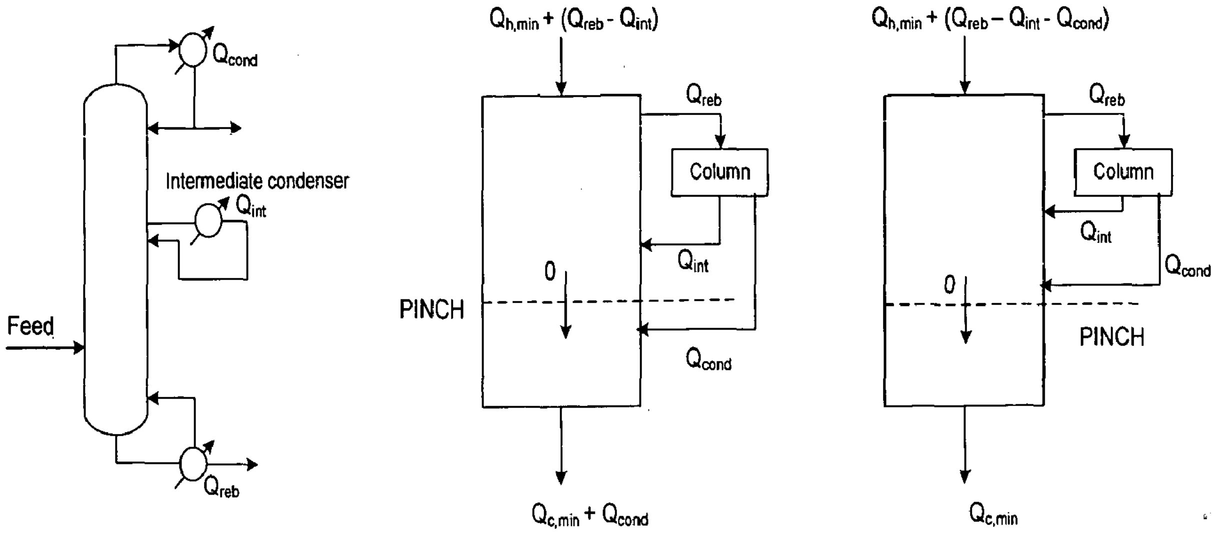


Figure 2.7. Side- stream stripper reduces heat load requirements

to be worthwhile when there is a large temperature difference across the column, i.e. the feed is a wide boiling mixture.

There are two situations in particular where intermediate reboiling and/or condensing should be considered.

1. Earlier it was suggested that if a distillation column is situated across the pinch then, if possible, the pressure should be changed to move the column away from the pinch. However, this may not be possible. In this case, intermediate reboilers or condensers can be used to get at least some of the savings resulting from good integration. Consider the column shown in Figs. 2.8a and 2.8b. It is operating across the pinch and the heat added to the reboiler is Q_{reb}. Q_{cond} is removed from the overhead condenser below the pinch but Q_{int} is removed above the pinch. Thus the hot utility requirements of the process must increase by only (Q_{reb}-Q_{int}). Extra utility is needed to run the column but not as much as the total load. Thus with a column forced to operate across the pinch it is still possible to "rescue" some heat and reduce the utility requirements of the overall process at least partly by good integration.
2. If a distillation column is not operating across the pinch but there is insufficient heat flow at some temperature levels in the cascade to integrate the total loads, then again intermediate reboiling and condensing may provide a remedy. Consider the situation illustrated in Fig. 2.8c. Here a column has to operate close to the pinch, where it is usual for the heat flows in the cascade to be low. In this case the flows Q_4 and Q_5 are too low for the load Q_{reb} to be accommodated. The situation can be remedied by an intermediate condenser and we again require no extra utilities at all. Below the pinch the logic is analogous. We can use intermediate reboilers to ensure that the cascade heat flows remain positive. Use of intermediate reboilers and condensers obviously introduces extra heat transfer equipment and hence increased capital cost. It also increases the number of plates required in the column. Again, as with splitting the column feed, the reduction in utilities must be traded off against the higher capital cost.



(a) (b) (c)

Figure 2.8. Appropriate placement of an intermediate condenser

2.2 INTEGRATION OF THE COLUMN INTERNALLY

The concept of the GCC has been extended to distillation (Dhole and Linnhoff, 1993; Bandyopadhyay et al., 1998) and is referred to as the column grand composite curve (CGCC). The CGCC forms the basis for the internal integration of distillation column. The CGCC depends not only on the operating reflux, but also on the feed location in the column.

2.2.1 Column Grand Composite Curve (CGCC)

The thermal analysis capability is useful in identifying design targets for improvements in energy consumption and efficiency. This capability is based on the concept of minimum thermodynamic condition (MTC) for a distillation column. The minimum thermodynamic condition pertains to thermodynamically reversible column operation or zero thermodynamic loss in the column. In this condition, a distillation column would operate at minimum reflux, with an infinite number of stages, and with heaters and coolers placed at each stage with appropriate heat loads for the operating and equilibrium lines to coincide (Dhole and Linnhoff, 1993; Sivakumar et. al., 1996). In other words, the reboiling and condensing loads are distributed over the temperature range of operation of the column. The stage-enthalpy (Stage-H) or temperature-enthalpy (T-H) profiles for such a column therefore represent the theoretical minimum heating and cooling requirements in the temperature range of separation. These profiles are called the Column Grand Composite Curves (CGCCs).

Several authors have reported various ways of plotting the CGCC both for the binary distillation as well as multi-component distillation.

Binary distillation: Several publications have been reported on the MTC for binary distillation and for generating the corresponding CGCC (Naka et al., 1980; Fitzmorris and Mah, 1980; Ishida and Ohno, 1983). The column will require infinite stages and infinitely many side exchangers. The operating line for each stage will be coincident with the equilibrium curve and thus the operating and the equilibrium curves will overlap on all points (King, 1980). Hence, the general approach employed for evaluating the CGCC involves simultaneously solving the operating line equations and the equilibrium line equations which is a relatively simple task for binary systems due to the small dimensionality of the problem.

Multicomponent distillation. The MTC for multicomponent separation has been reported by Fonyo, 1974 and Franklin and Wilkinson, 1982. A very important fact regarding MTC for multicomponent distillation is that the purity of separation is greatly limited. Within a single reversible column only the heaviest component can be removed completely from the overhead product and the lightest from the bottoms product. In other words, each column section can separate only one component, either the lightest or the heaviest. Certain sharp separations can be obtained reversibly by linking several multicomponent reversible columns. Fonyo, 1974 identifies one such a scheme for a four-component separation. However, for many practical multicomponent separation tasks it is impossible to devise a scheme for completely reversible separation. An example is a four-component mixture to be separated into two products each with two components (Franklin and Wilkinson, 1982).

multicomponent systems, there are several previous publications. Franklin and Wilkinson, 1982, propose an N-component model. The CGCC is obtained for each of the columns in a reversible scheme. These CGCCs are then added together to obtain an overall CGCC for the reversible scheme. Terranova and Westerberg, 1989, also propose an N-component model and adopt a similar approach to Franklin and Wilkinson, 1982. They use equations based on flash calculations to simultaneously solve the equilibrium and the component mass balance equations for all components to obtain the CGCC. However, they claim that such a CGCC can be obtained for any multicomponent separation. This does not seem to agree with the known limitations on sharpness for a reversible multicomponent separation. Fonyo, 1974 suggested a slightly simpler approach involving light and heavy key models. The light and heavy keys are the lightest and the heaviest components, respectively. They obtain the CGCC for individual columns in the reversible scheme using this model.

Most procedures for multicomponent mixtures (Franklin and Wilkinson, 1982; Fonyo, 1974) rely on a reversible scheme for a given separation. However, for many industrial applications involving sharp separations such a scheme is impossible to obtain as mentioned earlier. In these situations, the procedures do not provide practical guidelines. Even when a reversible scheme can be constructed, the procedures require iterations for relative volatilities or compositions which can be tedious. The CGCC developed by Terranova and Westerberg, 1989, evaluates the effect of different feed vapor fractions. However it requires the iterative procedure for the CGCC to be repeated for different feed conditions which can be cumbersome.

2.2.2 Practical Near Minimum Thermodynamic Condition

The practical near-minimum thermodynamic condition (PNMTC) relates to a minimum loss condition after accepting the inevitable losses due to the practicalities of column design and modification i.e. it represents the actual column. These losses include inevitable feed losses, losses due to sharp separation, chosen distillation configuration (e.g. multiple products, single column, side stripper etc.), pressure drop losses, etc. The column at PNMTC will still require infinite stages and infinite side exchangers. For generating CGCC, Dhole and Linnhoff, 1993, proposed a new approach that utilizes results from an already converged column simulation. The CGCC is generated using stage wise information on compositions and enthalpies from the output of a converged simulation of a distillation column. The CGCC depends not only on the operating reflux, but also on the feed location in the column. The procedure is based on PNMTC and is described in detail in Section 4.1.

2.2.3 Use of the CGCC

The CGCC is readily used for targeting for different possible column modifications. Fig. 2.9 describes the targeting procedure. A horizontal distance between the CGCC pinch point and the vertical axis represents the scope for reduction in reflux ratio (Fig. 2.9a). The CGCC pinch point indicates the minimum reflux condition for the column. As we reduce the reflux ratio, the CGCC will move towards the vertical axis, thus reducing the reboiler as well as the condenser load. The next modification to consider is feed conditioning. Inappropriate feed condition usually causes a sharp enthalpy change in the profile near the feed location. For example, a feed which is excessively subcooled, causes sudden "quenching". This will result in a sharp enthalpy change on the reboiler side. Such a sharp enthalpy change is particularly easy to see in the stage-enthalpy representation. As shown in Fig. 2.9b, the extent of sharp enthalpy change on the reboiler side determines the approximate heat load for feed preheating. Analogous arguments apply for feed cooling. Successful feed preheating and cooling will reduce the reboiler and condenser loads, respectively. After feed conditioning, we consider side condensing/ reboiling. Fig. 2.9c describes CGCCs which show potential for side condensing and reboiling. Feed conditioning always offers a more moderate temperature level than side condensing/reboiling. Also, feed conditioning is external to the column unlike side condensing/reboiling. Thus the sequence for considering different column modifications is recommended as follows:

1. Reflux and pressure modifications.
2. Feed preheating/cooling.
3. Side condensing/reboiling.

While setting the targets for the above-mentioned modifications it is assumed that the feed stage location for the column has been appropriately chosen beforehand. Inappropriate feed positioning could cause sharp enthalpy changes in the CGCC similar to feed preheating and cooling. Therefore, appropriate feed stage location should be identified before targeting for any column modification.

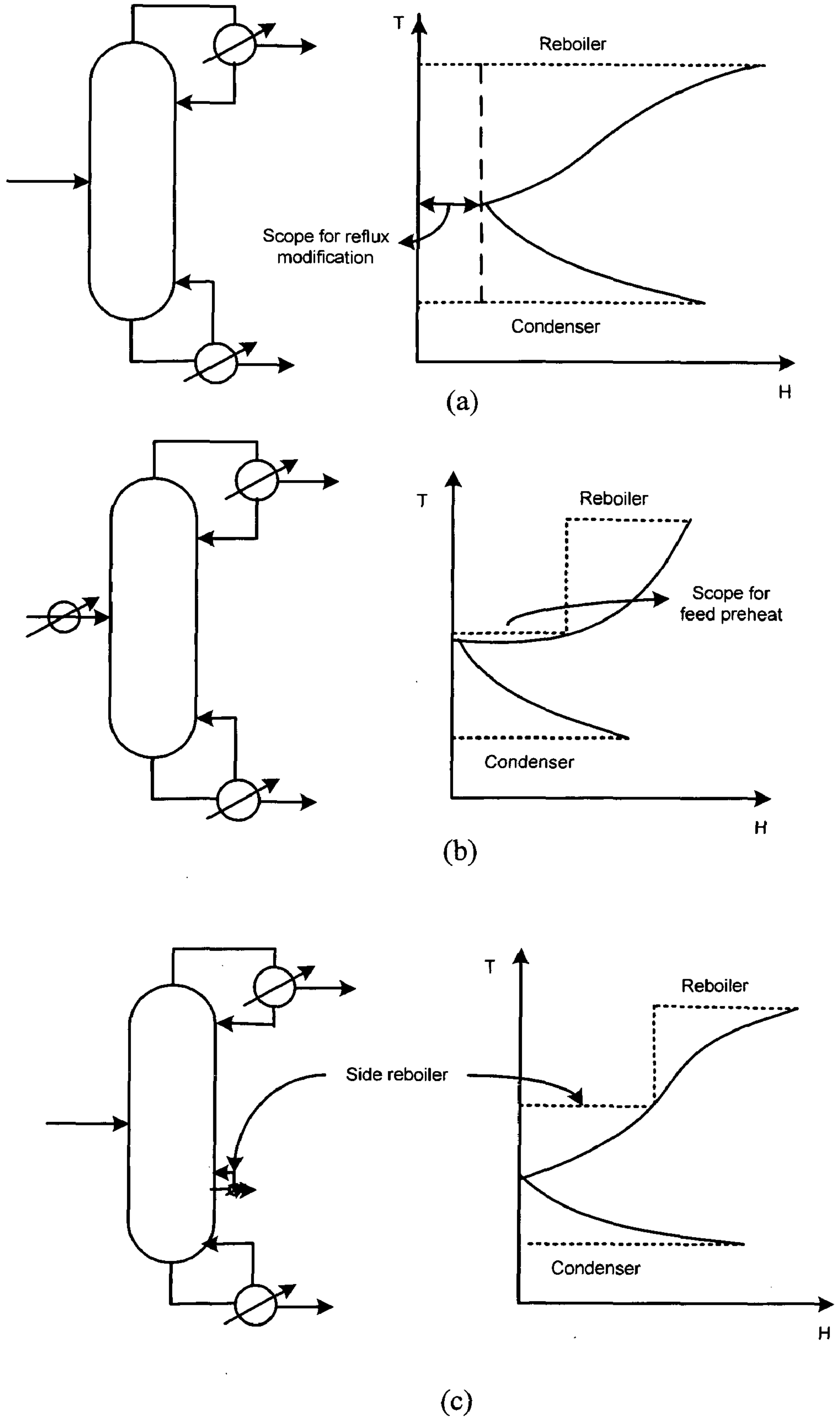


Fig. 2.9. Targeting for various column modifications using the CGCC. (a) Scope for Reflux Modification, (b) Scope for feed preheat and (c) Side Reboiler

.2.2.4 Column Composite Curve (CCC)

Modifications are aimed at reducing the excess driving forces in the column. Fig. 2.10a shows the driving force for a stage at temperature T_2. Vapor rises from the lower stage at temperature T_3, and mixes with the liquid coming from the previous stage at temperature T_1. The mixing results in the transfer of enthalpy from the vapor to the liquid stream and consequently loss of heat and mass transfer driving forces. The temperature of the mixture is T_2 which is the stage temperature. The construction is repeated for each stage. The overall construction is shown in Fig. 2.10b. This plot is termed as "Column Composite Curves" (CCC). The CCC depicts the distribution of stages in different sections of the column. The region between the vapor and the liquid composites represents combined heat and mass transfer loss in the column. As we increase the number of stages, the CCC will become tighter, reducing the heat and mass transfer loss in the column and vice versa when we reduce the number of stages. Thus, the CCC provides a link between driving forces and the number of stages in the column.

2.2.5 Relation between CGCC and CCC

The relationship is identical to that between pinch analysis Composite Curves and pinch analysis Grand Composite Curves. The horizontal distance between the CGCC and the vertical axis is the same as the horizontal distance between the CCC. The area between the CCC equals the area between the CGCC and the temperature axis (shown as a dotted area) and represents heat and mass transfer loss (Fig. 2.11). The CCC can be easily constructed from the CGCC. CGCC provides the energy targets for column modifications while CCC enables the designer to assess the effect of modifications on the number of stages, i.e. on capital cost. A modification in a section where the CCC show "tight" driving forces implies high capital cost penalty. The combined use of CGCC and CCC provide an assessment of both energy and capital cost implications of proposed column modifications in the targeting mode. Together, the CGCC and the CCC identify the most promising design options ahead of design.

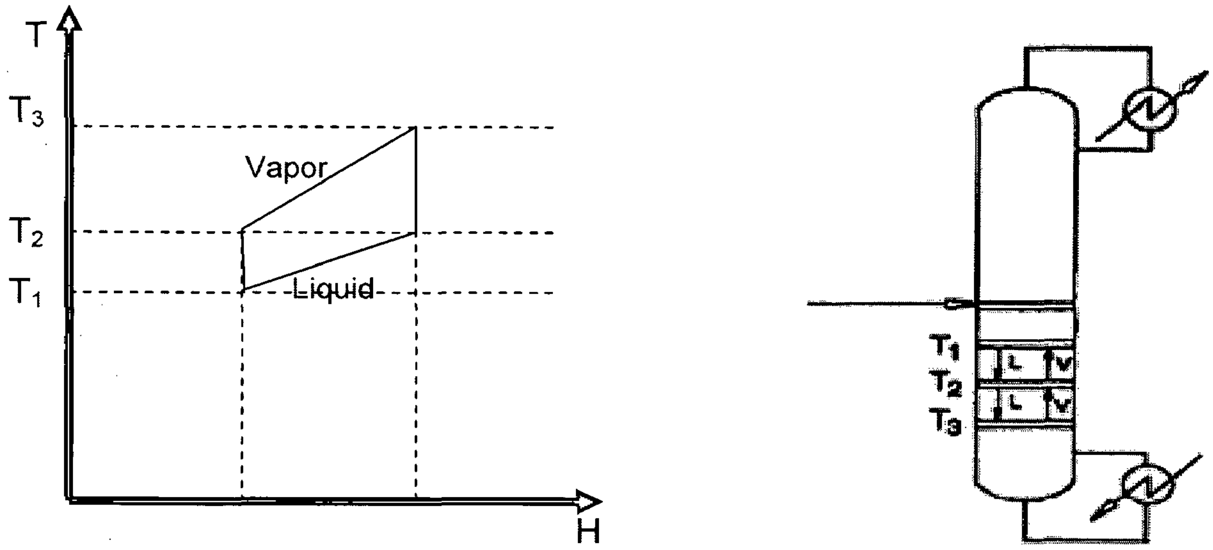


Figure 2.10a. Enthalpy change along a single stage.

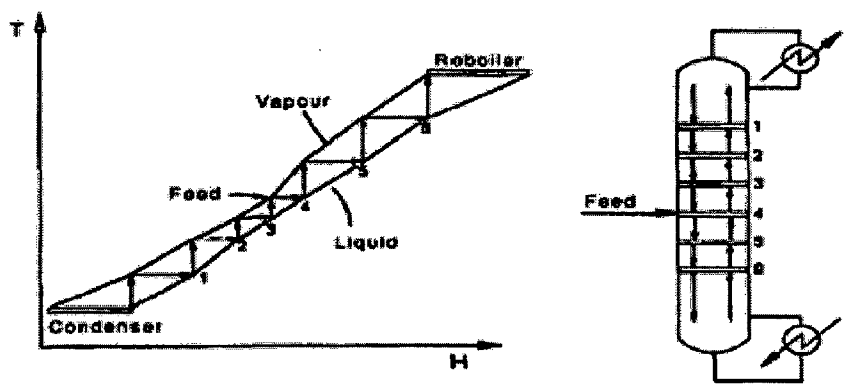
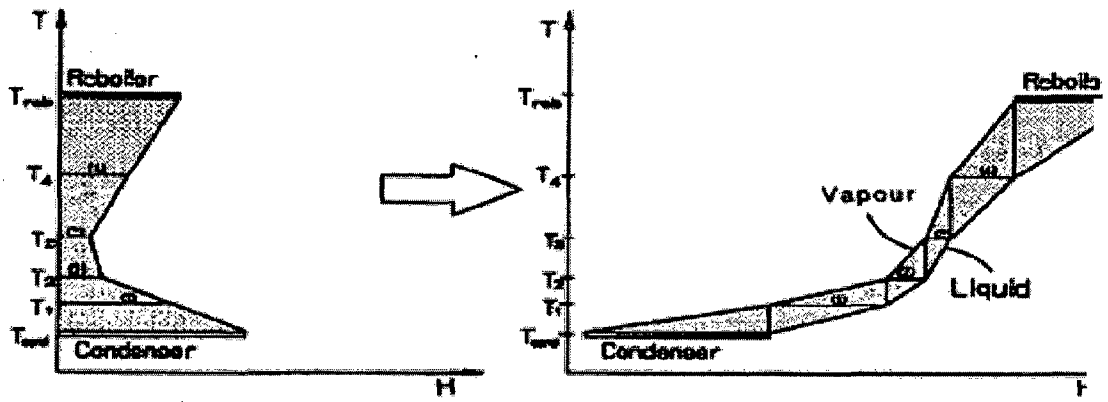


Figure 2.10b. Column Composite Curves (CCC).



(a) (b)

Figure 2.11. Relation between CGCC and CCC.

HEAT EXCHANGER OPTIMIZATION

2.3.1 Optimization Techniques

The optimum thermal design of a shell and tube heat exchanger involves appropriate selection of a large number of interacting design parameters from their feasible ranges and therefore, offers a considerably large feasible region to be searched making it a cumbersome task. Several combinations of discrete design parameters are possible, which can lead to an optimum or near optimum solution. Therefore, the designer needs an efficient strategy to quickly locate the design configurations which can offer minimum area or TAC of heat exchanger. As a result, the optimal design of heat exchanger can be posed as a large scale, discrete, combinatorial optimization problem (Chaudhuri et al., 1997).

The optimization technique helps in identifying a set of design parameters that maximizes or minimizes a given objective function. Though, the concept looks simple but the task of identifying the optimum geometry for a shell and tube heat exchanger (S&T-HE) for a given duty is not simple. Therefore, many investigators tried to search alternative routes for the determination of optimum geometry of a S&T HE. Smith, 1981, based on cost analysis, developed a set of equations that needs to be solved simultaneously for the estimation of economic velocity in heat exchanger. Rao et al., 1990, based on the cost computation of 9000 S&T-HE, developed a criterion that could narrow down design space which would lead to optimum configuration. According to them, 50% of the feasible designs are within the cost ratio (capital cost of S&T-HE to minimum capital cost of S&T-HE in the same set) of 2. The authors have pointed out that the cost/unit area of a S&T-HE tends to become constant after the area becomes more that 200 m^2. Steinmeyer, 1996, has provided a method based on pressure drop, temperature difference and cost parameters to achieve near optimal S&T-HE designs. Mukherjee, 1996, has provided useful tips for the selection of design parameters, which can lead to near optimal designs. The numbers of parameters available to the designer are large and their normal ranges are also substantial. Poddar et al., 2000, have listed the geometrical parameters and their valid ranges for optimization. Based on a tube count vs.

tube length, graph he suggested narrow ranges of design parameters that can produce optimal design. Thus it can be concluded that finding out an optimum design configuration based on classical design coupled with optimization is cumbersome. Any method which can develop a targeting procedure, based on a small set of equations, to find out minimum TAC of heat exchanger or minimum area of heat exchanger will be beneficial for process integration studies.

Muralikrishna and Shenoy, 2000, first demonstrated that the complex design space of S&T HE can be conveniently plotted on a two-dimensional plot of shell side vs. tube side pressure drop which can be reduced further using the constraints. Minimum -area and -TAC targets can be determined from this plot. They demonstrated their methodology using Kern's Method which is considered to be a simplified design procedure and does not take in to account the losses. Later, Choudhury et al., 2008, find out minimum -area and -TAC targets using a more realistic Bell's Method by accounting leakage, bypassing and flow through window and end zones in the shell side. However, they did not substantiate their optimal design by comparing it with standard optimal design obtained from established software like Aspen. In our recent work on equation based targeting of shell and tube heat exchanger for obtaiing minimum –area and –cost (Pednekar et al., 2008), the method proposed by Choudhury et al. (2008) with some refinements to find out minimum -area and -TAC targets. It included three case studies for the design of shell and tube heat exchangers, spanning the whole spectrum of operation, i.e., liquid-liquid, gas-liquid and gas-gas, to test the above methodology. Further, comparisons were also made between the results obtained by the present study and that from Aspen. The solution techniques adopted for the optimization of heat exchanger is explained in Section 4.3.

CHAPTER 3

PROBLEM DEFINITON

The problem for study is taken from Javid et. al., 2006. The process which is the kerosene pre-fractionation unit of a refinery consists of two simple continuous distillation columns, i.e. at first the light product is separating by the first column and the middle and heavy products are separated by the second column. The feed to the first column is first converted to the top-light product (Light) and its bottom-heavy product (Stream). The bottom heavy product enters into the second column, where is converted to light product at the top (Heart) and the heavy product at the bottom (Heavy) as shown in Fig. 3.1.

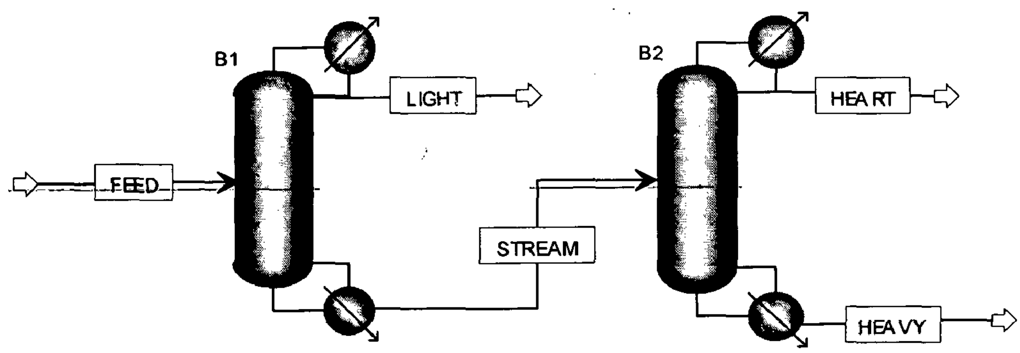


Figure 3.1: Existing distillation unit of case study

The existing heat exchangers besides the distillation columns consist of condensers and reboilers. The first column has 53 trays and the second column has 65 trays. In the first column the feed enters to the tray no.30 and the bottom product of this column feeds to the tray no.30 of the second column. Operational specifications and the thermophysical conditions of the existing distillation column is shown in Table 3.1 and 3.2 respectively. Table 3.3 shows the streams compounds of the existing process. Hot oil and cooling water are used as hot and cold utility respectively. Cooling water is available at 25^0C which can be heated upto 35^0C. Hot oil is available at 295^0C, 250^0C and 220^0C and the maximum outlet temperature is 275^0C, 230^0C and 200^0C. Heat exchanger specification is shown in Table 3.4. The required physical properties of process and utilities are given in Appendix E.

Only the boiling points of pseudo components were available in the paper (Javid et. al. 2006). To define a component API gravity and their corresponding boiling point are required. So API gravity for the given pseudo components were assumed from the relation between API gravity and their corresponding boiling points given by Eckert and Vanek (2005), which are shown in Table 3.3.

Table 3.1: Operational specifications of distillation columns

Column No.	Condenser heat duty (kW)	Reboiler heat duty (kW)	Top product flow rate (kg/hr)	Bottom product flowrate (kg/hr)
B1	3524.08	4281.4	3816.1	44865.3
B2	8055.90	7807.92	42371.3	2494.0

Table 3.1(cont): Operational specifications of distillation columns

Column No.	No. of trays	Condenser type	Reboiler type	Reflux Ratio	Boil up Ratio	Feed tray No.
B1	53	TOTAL	KETTLE	9.988	1.4662	30
B2	65	TOTAL	KETTLE	1.174	48.222	30

Table 3.1 (cont): Operational specifications of distillation columns

Column No.	Top Column output flow tray no.	Bottom column output tray no.	Top pressure (atm)	Bottom pressure (atm)
B1	1	53	0.44	0.74
B2	1	65	0.54	0.98

Table 3.2: Thermophysical conditions of existing process streams

Streams name	Liquid fraction	Vapor fraction	Temperature (^0C)	Pressure (atm)	Flowrate (kg/hr)
Feed	1	0	150	1.9	48681.4
Light	1	0	114.4	0.44	3816.1
Stream	1	0	177.9	0.74	44865.3
Heart	1	0	163.9	0.54	42371.3
Heavy	1	0	267.2	0.98	2494.0

Table 3.3: Streams compounds of existing process

Pseudo components	Boiling point (^0C)	API Gravity	Feed	Light	Stream	Heart	Heavy
Pc1	131.6	56	5.51	5.51	nil	nil	nil
Pc2	143.3	53	17.36	17.28	0.08	0.08	nil
Pc3	156.1	51	41.45	7.65	33.80	33.80	nil
Pc4	170	48.5	45.13	nil	45.13	45.13	nil
Pc5	182.7	46.5	46.15	nil	46.15	46.15	nil
Pc6	197.2	44	40.84	nil	40.84	40.84	nil
Pc7	211.1	42.5	37.17	nil	37.17	37.17	nil
Pc8	225.5	40.5	30.94	nil	30.94	30.94	nil
Pc9	239.4	39	21.65	nil	21.65	21.65	nil
Pc10	252.7	37.5	11.54	nil	11.54	11.54	nil
Pc11	266.1	36	10.82	nil	10.82	2.04	8.78
Pc12	275.5	35	3.17	nil	3.17	0.00	3.17

Table 3.4: Heat Exchanger specifications of existing process

Heat Exchanger Type	Tube Side	Shell Side	Hot stream flowrate (kg/hr)	Cold stream flowrate (kg/hr)
Column 1 condenser	Petroleum	Cooling water	41931.3	303506.7
Column 1 reboiler	Hot oil	Petroleum	376832.9	105521.9
Column 2 condenser	Petroleum	Cooling water	92115.2	693757.9
Column 2 reboiler	Hot oil	Petroleum	562122.7	122698.2

Table 3.4 (cont): Heat Exchanger specifications of existing process

Heat Exchanger Type	Hot stream input Temp. (^0C)	Cold stream input Temp. (^0C)	Hot stream output Temp. (^0C)	Cold stream output Temp. (^0C)	Exchanger heat duty (kW)
Column 1 condenser	116.6	25	114.5	35	3524.05
Column 1 reboiler	295	168.6	275	177.9	5233.79
Column 2 condenser	183.1	25	163.9	35	8055.30
Column 2 reboiler	295	266.5	275	267.2	7807.26

The present problem is analyzed and required column modifications are applied to show the possible savings that can be achieved using column targeting. The detailed solution techniques adopted is given in Chapter 4 and the possible savings that can be obtained is shown in Chapter 5.

CHAPTER 4

SOLUTION TECHNIQUES ADOPTED

4.1 GENERATING CGCC

. Dhole and Linnhoff, 1993, proposed a new approach that utilizes results from an already converged column simulation. Normally the outputs from simulations provide molar flows and compositions on a stage-by-stage basis. The enthalpies used in plotting the CGCCs are calculated at a given stage of the column by assuming that the equilibrium and operating lines coincide at this stage. This approximation takes into account the losses or inefficiencies introduced through practicalities of column design (such as pressure drops, multiple side-products, side strippers, etc.), while preserving the meaning of the CGCC. As a close approximation to PNMTC, we simultaneously solve the equilibrium and the operating line equations for the key components. Fig. 4.1 illustrates the operating line equations for light and heavy keys. For equilibrium compositions we use the stage-by-stage compositions as provided by the simulator output (asterisked in Fig. 4.1).

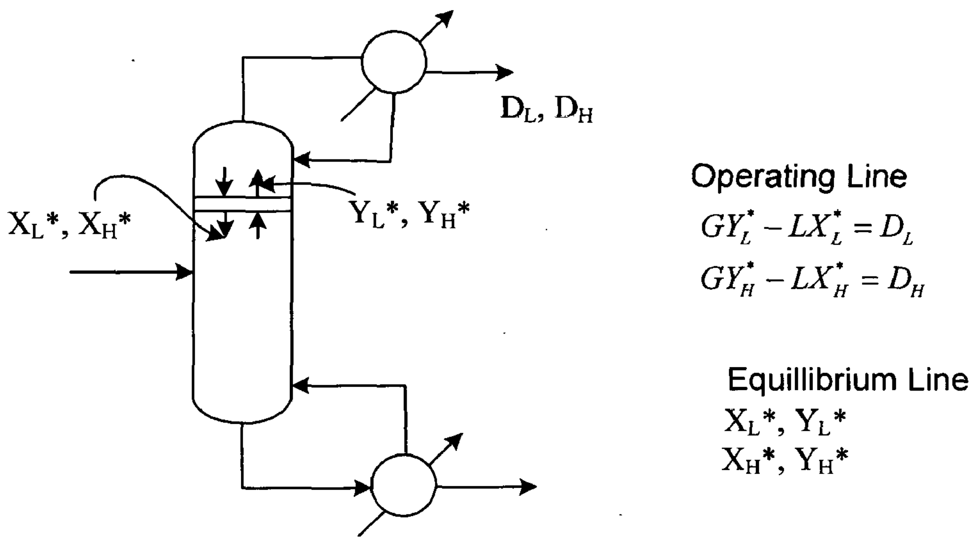


Figure 4.1: Dhole and Linnhoff, 1992, approach

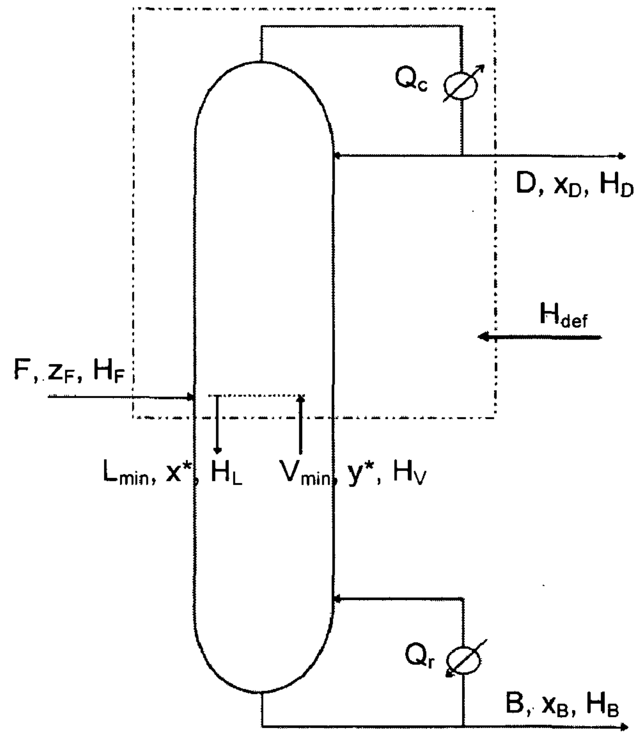


Figure 4.2a: Envelope for CGCC generation from condenser side

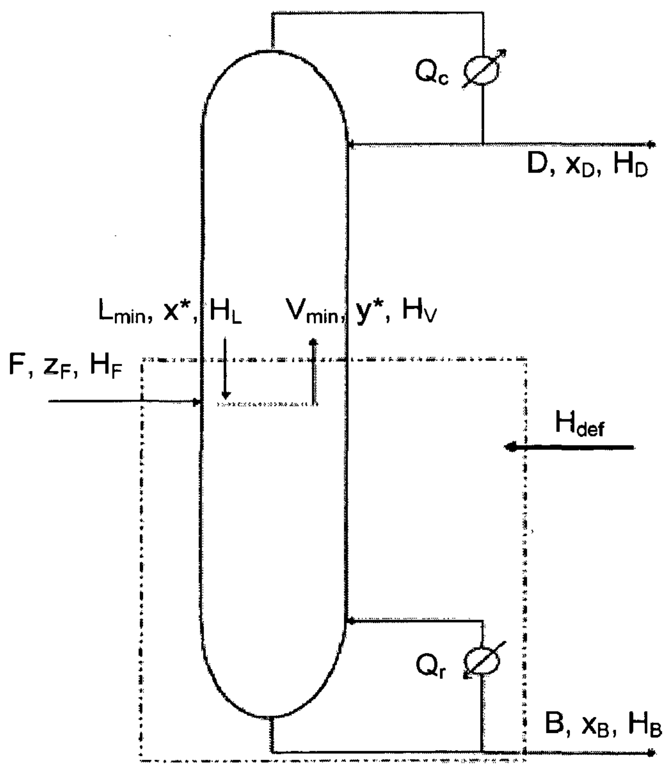


Figure 4.2b: Envelope for CGCC generation from reboiler side

The equation used for obtaining H_{CGCC} can be derived as follows.

From Fig. 4.2a,

Overall material balance and the component balance

$$V_{\min} + F = L_{\min} + D \tag{1}$$

$$V_{\min} y^* + F z_F = L_{\min} x^* + D x_D \tag{2}$$

Solving Eq. 1 and 2. we get,

$$L_{\min} = \left[D(x_D - y^*) - F(z_F - y^*)\right]/(y^* - x^*) \tag{3}$$

$$V_{\min} = \left[D(x_D - x^*) - F(z_F - x^*)\right]/(y^* - x^*) \tag{4}$$

Enthalpy balance:

$$V_{\min} H_V + F H_F + H_{def} = L_{\min} H_L + D H_D \tag{5}$$

The stage enthalpy deficit H_{def} may be added to the condenser duty (Q_c) to obtain the enthalpy values for plotting the CGCC (H_{CGCC})

$$H_{CGCC} = Q_c + H_{def}$$

$$\begin{aligned} H_{CGCC} = Q_c &+ D\{H_D + [(H_L - H_V)x_D + (H_V x^* - H_L y^*)]/(y^* - x^*)\} \\ &- F\{H_F + [(H_L - H_V)z_F + (H_V x^* - H_L y^*)]/(y^* - x^*)\} \end{aligned} \tag{6}$$

Similarly, we can also start considering the envelope from reboiler side (Fig. 4.2b). We get similar equation:

$$\begin{aligned} H_{CGCC} = Q_r &- B\{H_B + [(H_L - H_V)x_B + (H_V x^* - H_L y^*)]/(y^* - x^*)\} \\ &+ F\{H_F + [(H_L - H_V)z_F + (H_V x^* - H_L y^*)]/(y^* - x^*)\} \end{aligned} \tag{7}$$

Consider the top envelope for obtaining H_{def} above the feed plate and consider the bottom envelope to obtain H_{def} below the feed plate. Thus starting from 1st plate Enthalpy deficit (H_{def}) at each plate is calculated (Fig. 4.3). The values of H_L, H_V, x^* and y^* for each plate are obtained from converged simulation.

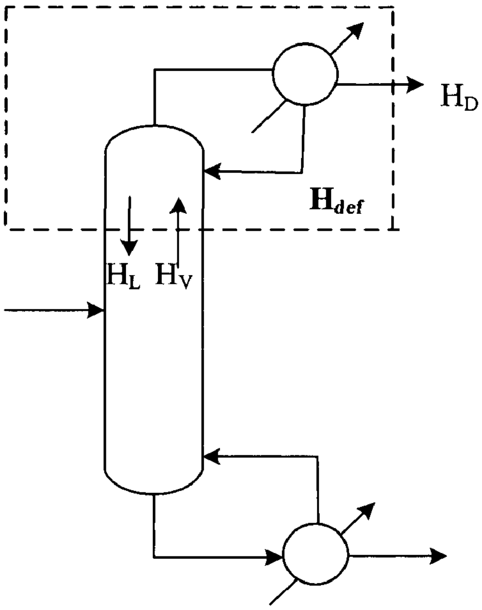


Figure 4.3a: Evaluating enthalpy deficit at a stage.

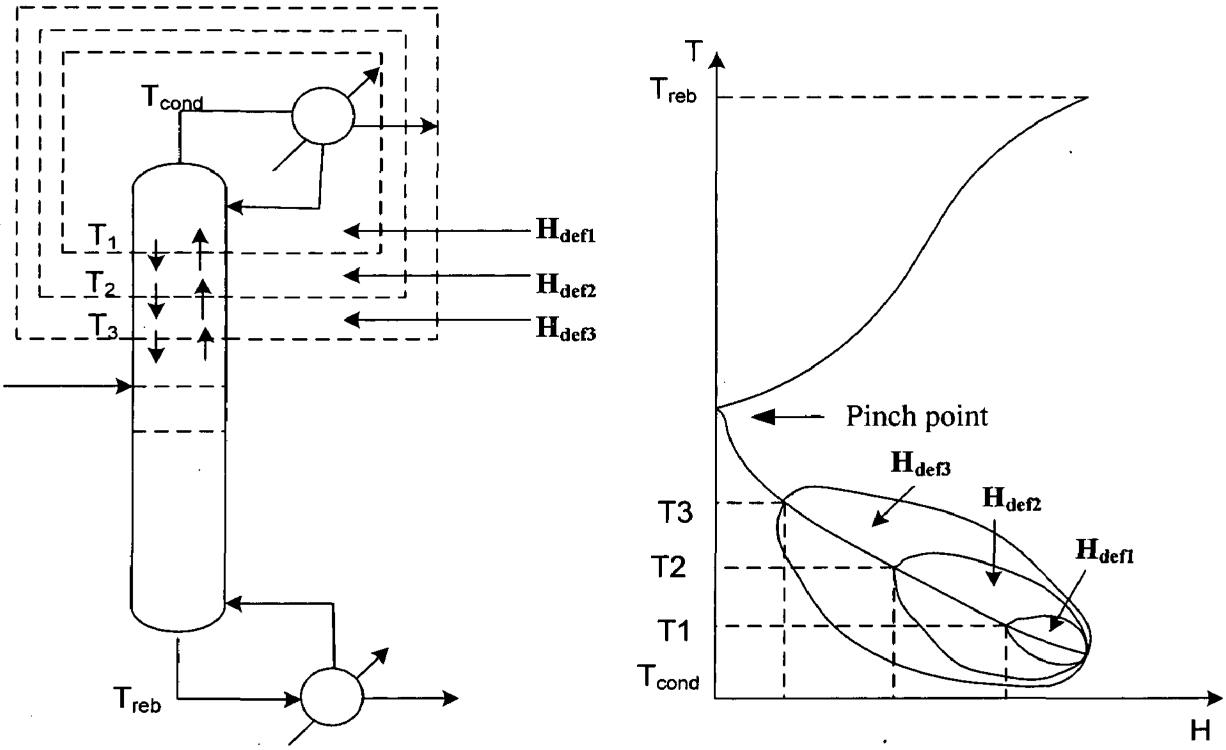


Figure 4.3: Constructing the CGCC from stagewise enthalpy deficits

Fig. 4.3b demonstrates how the individual enthalpy deficits are cascaded to construct the CGCC. The values of the stage temperatures (T_1, T_2, etc.) and the corresponding heat deficits [H_{def1}, H_{def2}, etc.] are plotted in the T-H dimension as shown. The algorithm used for developing the cascade is identical to the problem table algorithm introduced by Linnhoff et. al. (1994). The feed enthalpy strongly influences the shape of the CGCC near the feed stage. The CGCC usually shows a pinch point near the feed stage.

The algorithm for obtaining CGCC can be summarized as follows

1. Start with a converged simulation and obtain the minimum vapor and liquid flows, at each stage temperature, by solving the two simultaneous equations (Eqns. 1 and 2).
2. Express these minimum flows as equivalent enthalpy flows to obtain the net heat deficit at the stage temperature.
3. The CGCC can be plotted as Stage No. vs enthalpy or Temperature vs. Enthalpy (Fig. 4.3b).

The procedure described above is for a simple column involving a single feed and two products. It is easily extended for columns involving multiple feeds and products and for complex column arrangements (Dhole, 1991). As regards number of components, the description here considered a single choice of light and heavy keys. The procedure can be applied to different choices of key components in different sections of the column or to several components grouped together as light and heavy keys. Grouped components are particularly useful for refinery columns. The separating light and heavy key components can be identified from the stagewise compositions (obtained from simulation output). The grouped light and heavy key compositions are evaluated by summing the light and heavy key compositions, respectively. These grouped key compositions are used in Eqn. 1. to obtain G_{min} and L_{min} values, the remaining procedure remains unchanged.

4.2 ABOUT ASPEN PLUS COLUMN TARGETING

The Aspen Plus Column Targeting tool generates the CGCCs based on PNMTC described above. It uses the simulated results and obtains the values of H_L, H_V, x^* and y^* for each plate. Then using Eqn. 6 and Eqn. 7, it calculates H_{def} at each plate and gives a plot of Temperature vs. Enthalpy and Stage Number vs. Enthalpy i.e. the required CGCC curve. The Aspen Plus Column Targeting tool has a built-in capability to select light key and heavy key components for each stage of the column.

Selection of Key Components

Results of the column targeting analysis depend strongly on the selection of light key and heavy key components. The Aspen Plus Column Targeting tool provides the following four methods for judicious selection of these key components:

1. **User defined :** Allows you to specify the light key and heavy key components.
2. **Based on component split fractions :** Selects the light key and heavy key components on the basis of component split-fractions in column product streams. This method is best suited for sharp or near-sharp splits.
3. **Based on component Kvalues:** Selects the light key and heavy key components on the basis of component K-values. This method is best suited for sloppy splits.
4. **Based on column composition profiles:** Selects the light and heavy key components on the basis of composition profiles. In principle, this method is similar to the K-value based method. It is best suited for sloppy splits and it is, in general, inferior to the K-value based method.

Models available for Column Targeting

1. **RadFrac:** Rigorous two or three phase fractionation for single column
2. **MultiFrac:** Rigorous fractionation for complex columns involving strippers and absorbers
3. **PetroFrac:** Rigorous fractionation for petroleum refining applications

The CGCCs can be plotted using the PlotWizard. The thermal analysis results provide a practical approach to identifying and implementing potential modifications to the column design. In our present study, the column is simulated using Aspen Plus - RadFrac Model and selection of key component is based on component split fractions.

4.3 COLUMN MODIFICATION

Column modifications that are based on inspection of the CGCCs are:

1. Feed location (appropriate placement)
2. Reflux ratio modification (reflux ratio vs. number of stages)
3. Feed conditioning (heating or cooling)
4. Side condensing or reboiling

Let us briefly discuss each modification with the help of a distillation column with the help of example problem that separates a mixture of n-heptane and n-octane from heavier hydrocarbons (n-nonane, n-decane, and n-pentadecane). Feed and product specification for the example problem is given in Table 4.1. Table 4.2 gives the parameters value for existing case (Design 1)

Table 4.1 Feed and product specification for the example problem

Properties	Feed	Top Product	Bottom Product
Molar flow (kmol/hr)	1000	398	602
Pressure (kPa)	200	200	200
Temperature (^{0}C)	100	137	205
Vapor fraction	0	0	0
Mole fractions			
n-heptane	0.2	0.502	0
n-octane	0.2	0.473	0.019
n-nonane	0.2	0.021	0.319
n-decane	0.2	0.004	0.330
n-pentadecane	0.2	0	0.332

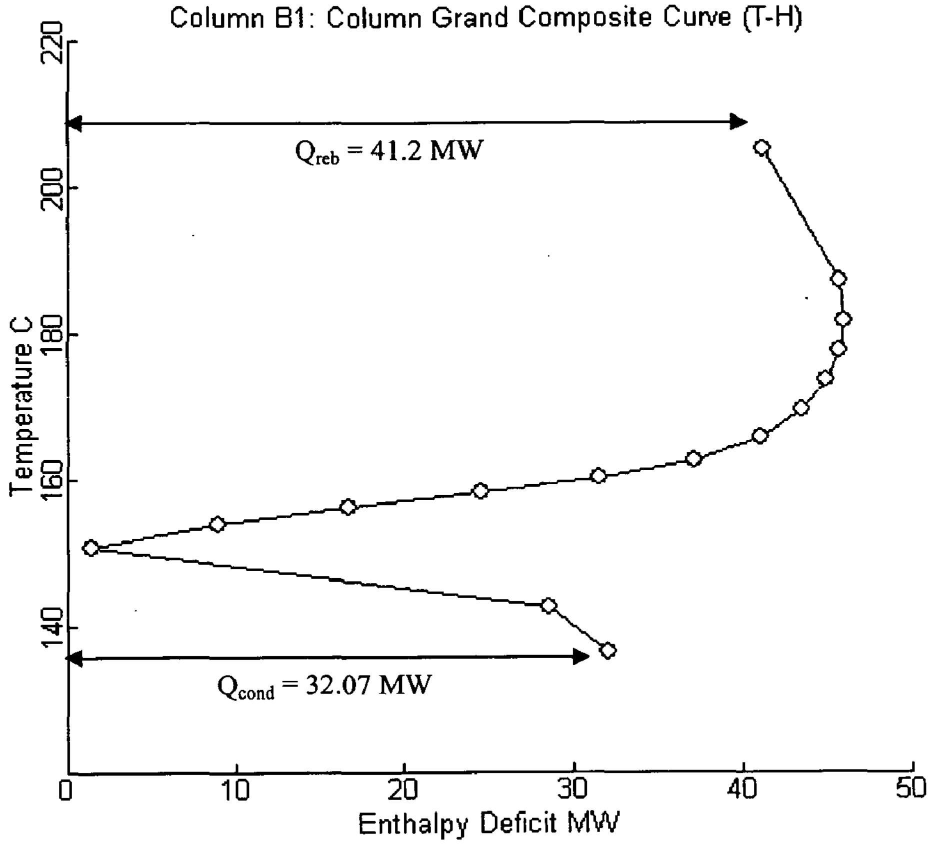
Table 4.2.Design 1 (Existing Case)

Parameter	**Design 1**
No. of stages	15
Reflux ratio	8
Feed location	3
Feed temperature (^0C)	100
Condenser duty (MW)	32.07
Condenser temperature (^0C)	136.5
Reboiler duty (MW)	41.20
Reboiler temperature (^0C)	205.1

4.3.1 Feed Location

Inspection of the CGCC can identify any anomalies or distortions due to inappropriate feed placement. Normally, such distortions will be apparent as significant projections at the feed location (pinch point) on the Stage-H CGCC. This is due to a need for extra local reflux to compensate for the inappropriate feed placement. A feed introduced too high up in the column will show a sharp enthalpy change on the condenser side on the Stage-H CGCC and should be moved down. Similarly, a feed introduced too low in the column will show a sharp enthalpy change on the reboiler side on the Stage-H CGCC and should be moved up the column. A correctly placed feed not only removes the distortions in the Stage-H CGCC but also results in reduced condenser and reboiler duties.

The Stage-H CGCC for Design 1 of our distillation column is shown in Fig. 4.4. It clearly shows a distortion on the condenser side at the pinch point (stages 2 and 3). Therefore, the feed must be moved down the column. The Fig. 4.5 shows the CGCC for Design 2, where the feed is moved down to stage 7 which shows a significant increase in the scope for reflux modification. Table 4.3 shows a comparison between Design 1 and Design 2



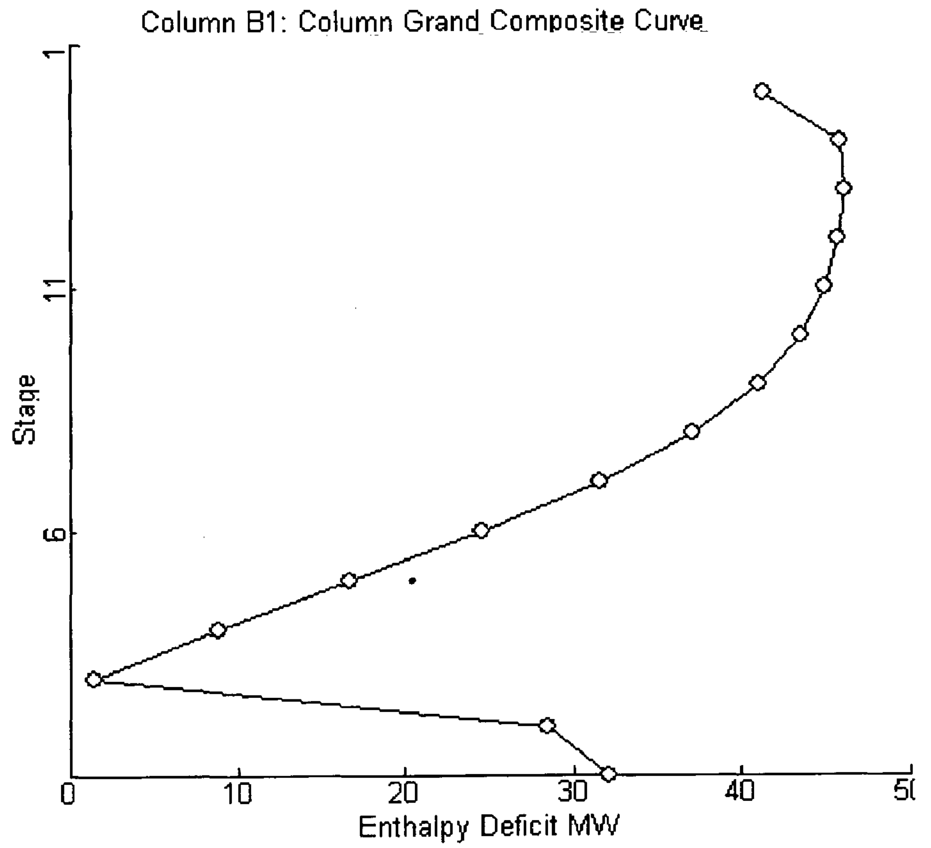
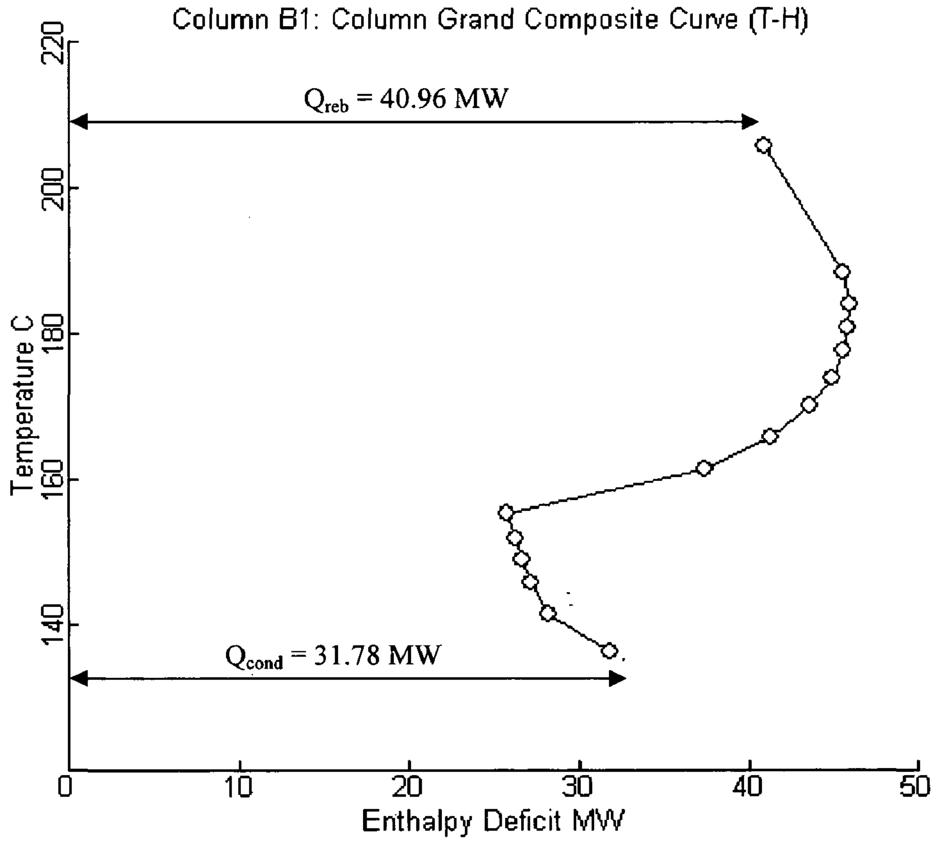


Figure 4.4: CGCC for Design 1



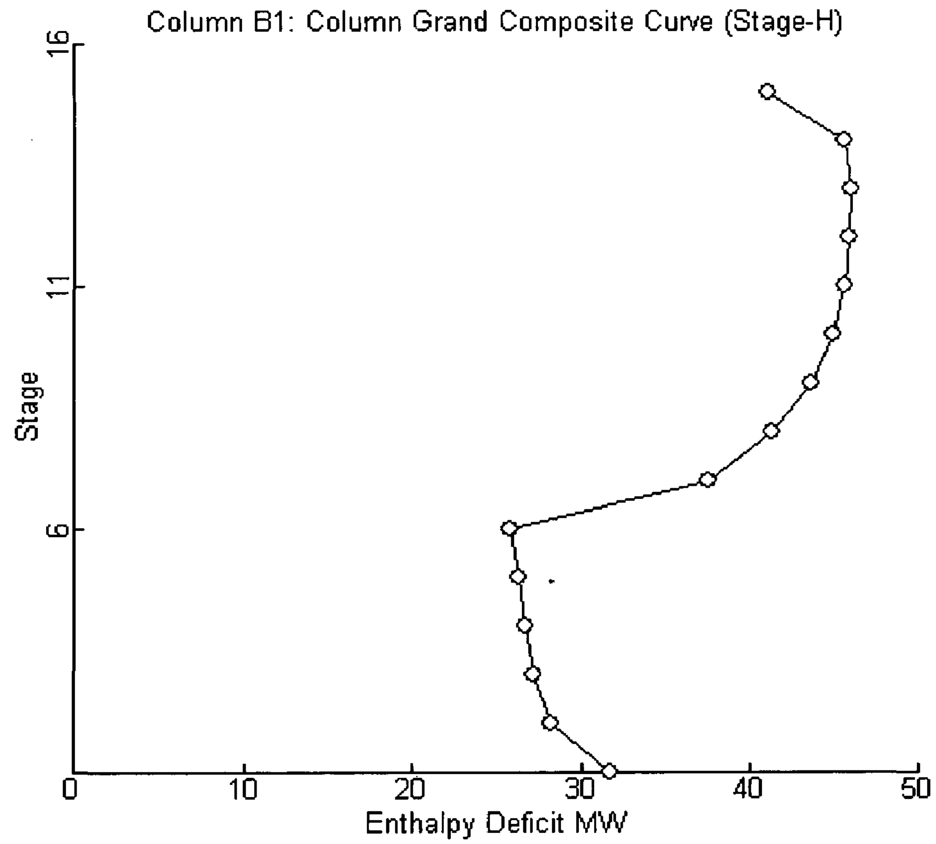


Figure 4.5: CGCC for Design 2

Table 4.3 Comparison of the design parameters for Design 1 and 2

Parameter	Design 1	Design 2
No. of stages	15	15
Reflux ratio	8	8
Feed location	3	7
Feed temperature (^{0}C)	100	100
Condenser duty (MW)	32.07	31.78
Condenser temperature (^{0}C)	136.5	136.3
Reboiler duty (MW)	41.20	40.96
Reboiler temperature (^{0}C)	205.1	205.8
Scope for reflux modification (MW)	1.5	25.7

4.3.2 Reflux Ratio Modification

The horizontal gap between the T-H CGCC pinch point and the ordinate represents the scope for reduction in heat duties through reduction in reflux ratio. As the reflux ratio is reduced (while increasing the number of stages to preserve the separation), the CGCC will move towards the ordinate, thus reducing both the condenser and reboiler loads. The T-H CGCC for Design 2 of our distillation column is shown in the following figure. This figure also identifies the scope for reduction in condenser and reboiler duties by reducing the reflux ratio.

It must be noted that, as the reflux ratio is reduced, the number of stages required to achieve the desired separation increases. In order to make a judicious choice for the reflux ratio, the increase in the capital cost due to the increase in the number of stages should be traded-off against the savings in the operating costs due to reduced condenser and reboiler loads. For our distillation column, if we reduce the reflux ratio to 1.4 (Design 3), we have to use 20 stages to preserve the separation. The T-H CGCC for Design 3 is shown Fig. 4.6. Comparison between the design parameter of Design 2 and Design 3 reveals the energy saving achieved by reducing the reflux.

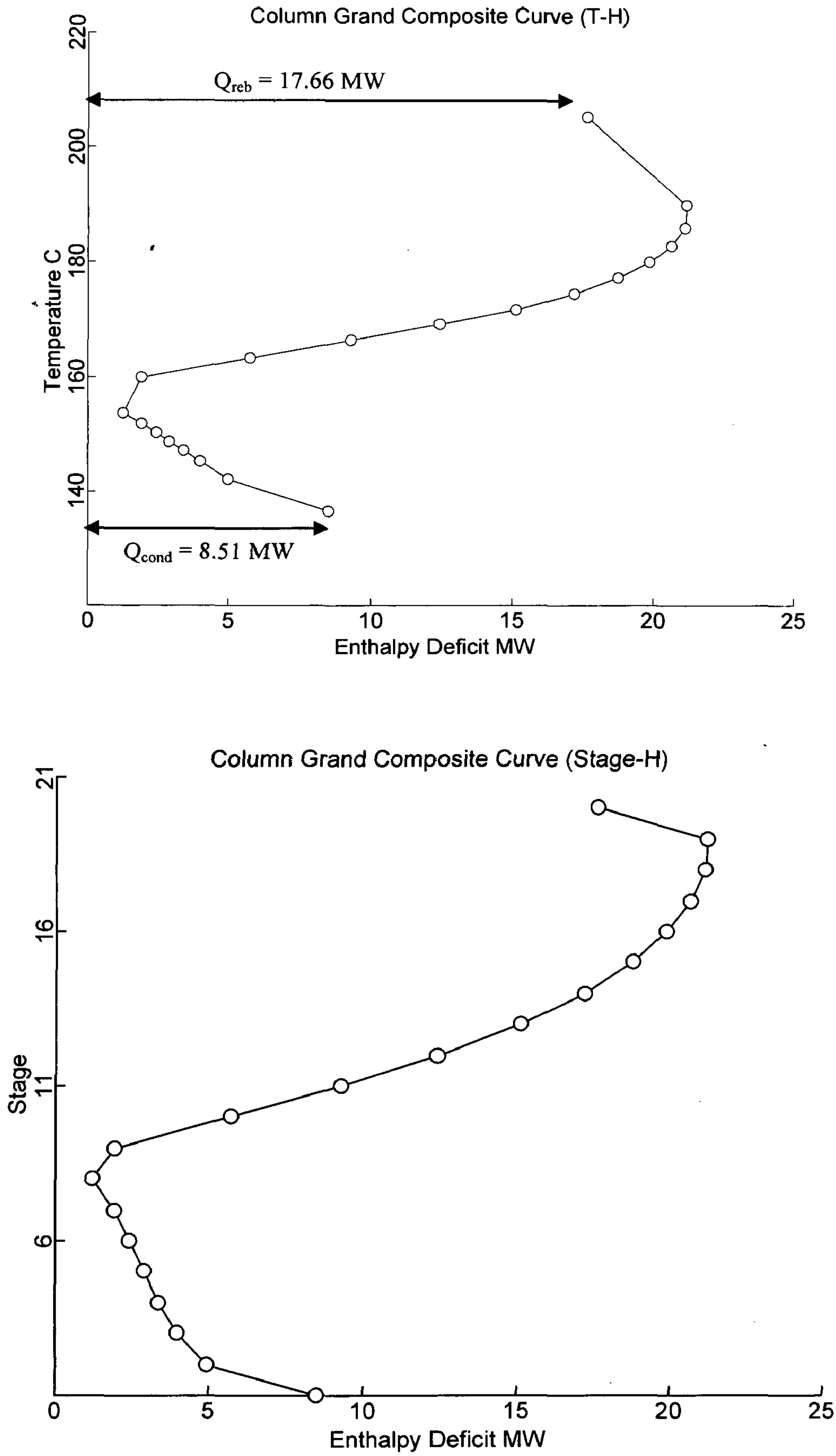


Figure 4.6: CGCC for Design 3

Table 4.4 Comparison of the design parameters for Design 2 and 3

Parameter	**Design 2**	**Design 3**
No. of stages	15	20
Reflux ratio	8	1.4
Feed location	7	9
Feed temperature (^{0}C)	100	100
Condenser duty (MW)	31.78	8.51
Condenser temperature (^{0}C)	136.3	136.44
Reboiler duty (MW)	40.96	17.66
Reboiler temperature (^{0}C)	205.8	205.4

4.3.3 Feed Conditioning

Scope for adjustment of feed quality can be identified from sharp enthalpy changes on the Stage-H or T-H CGCC. A feed that is excessively sub-cooled will show a sharp enthalpy change on the reboiler side of the CGCC. The extent of this change determines the approximate feed-heating duty required. Similar arguments also apply for feed cooling. Changes in the heat duty of feed pre-heaters or pre-coolers will lead to similar duty changes in the column reboiler or condenser loads, respectively. The Stage-H CGCC for Design 3 of our distillation column is shown in the following figure. The enthalpy change on the reboiler side is noticeably sharper. Therefore, our design can benefit from addition of a feed pre-heater.

Design 4 adds a feed preheater with a duty of 3.07 MW. The comparison between Design 3 and Design 4 is shown in Table 4.5. Note that feed preheating not only reduces the reboiler duty but also reduces the temperature levels at which the hot utility (for the reboiler and for the pre-heating the feed) needs to be supplied. CGCC for Design 4 is shown in Fig. 4.7.

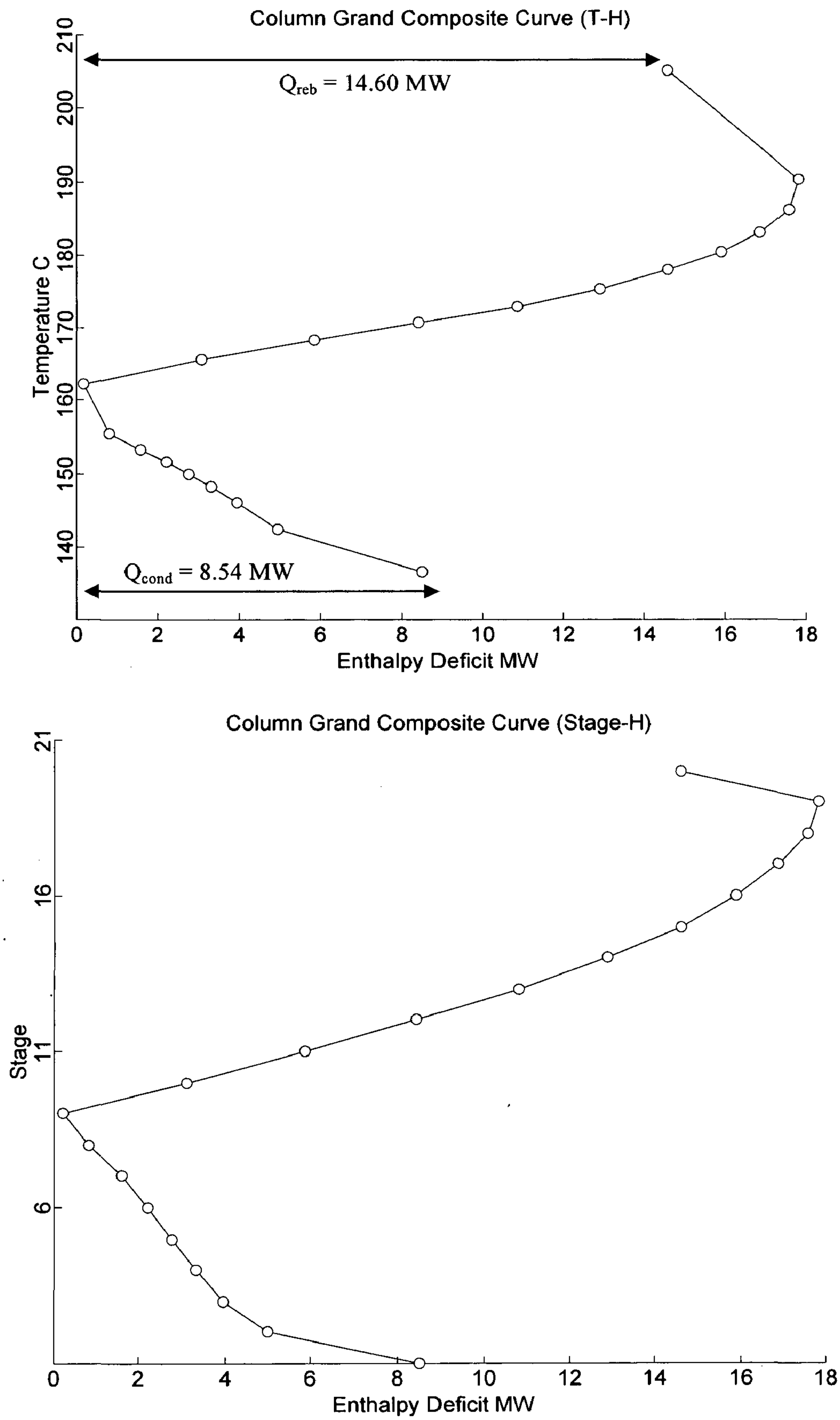


Figure 4.7: CGCC for Design 4

Table 4.5: Comparison of the design parameters for Design 3 and 4

Parameter	Design 3	Design 4
No. of stages	20	20
Reflux ratio	1.4	1.4
Feed location	9	9
Feed temperature (^{0}C)	100	130
Condenser duty (MW)	8.51	8.54
Condenser temperature (^{0}C)	136.4	136.5
Reboiler duty (MW)	17.66	14.60
Reboiler temperature (^{0}C)	205.4	205.1

4.3.4 Side Condensing or Reboiling

Feed conditioning is normally preferred to side condensing or side reboiling, as such modifications are external to the column and potentially at a more convenient temperature level. The scope for side condensing or side reboiling can be identified from the area beneath or above the CGCC pinch point (area between the ideal and actual enthalpy profiles). If a significant area exists, say below the pinch, a side-condenser can be placed at an appropriate temperature level. This allows heat removal from the column using a cheaper cold utility. A similar converse argument applies to scope for placing a side reboiler.

We can reduce the area on the reboiler side of the CGCC by using a side reboiler at stage 12 with a duty of about 5 MW (Design 5). The T-H CGCC for Design 5 is shown in the Fig. 4.8. Note that, the addition of the side reboiler, not only reduces the main reboiler duty but also reduces the temperature levels at which the hot utility (for the main reboiler and for the side reboiler) needs to be supplied. The Comparison between Design 4 and Design 5 is shown in Table 4.6.

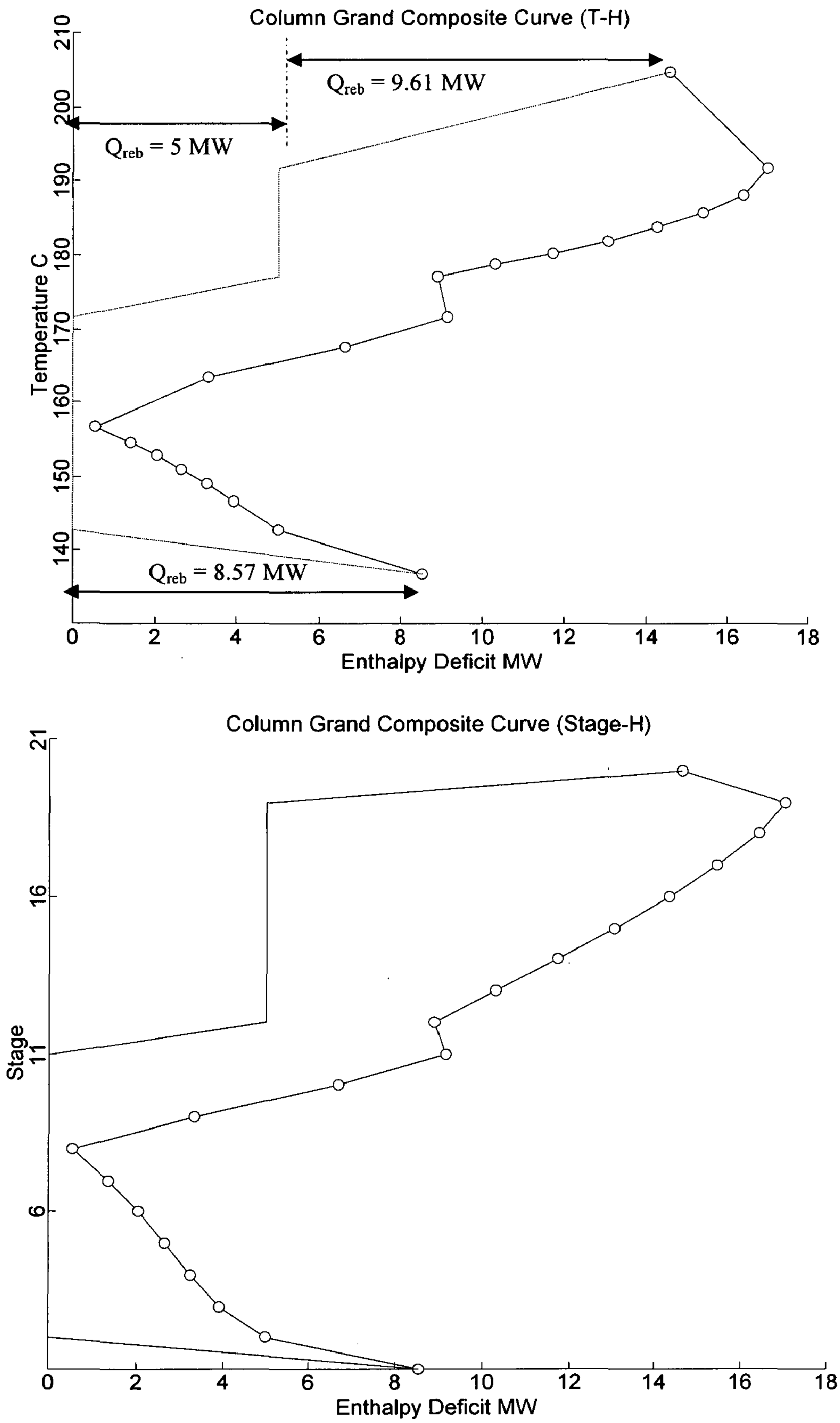


Figure 4.8: CGCC for design 5

Table 4.6: Comparison of design parameters for Design 4 and 5

Parameter	**Design 4**	**Design 5**
No. of stages	20	20
Reflux ratio	1.4	1.4
Feed location	9	9
Feed temperature (^{0}C)	130	130
Condenser duty (MW)	8.54	8.57
Condenser temperature (^{0}C)	136.5	136.6
Reboiler duty (MW)	14.60	9.61
Reboiler temperature (^{0}C)	205.1	204.7
Side-Reboiler duty (MW)	-	5
Side-Reboiler temperature (^{0}C)	-	177
Side-Reboiler location	-	12

4.3 HEAT EXCHANGER OPTIMIZATION - APPLICATION IN THE PRESENT CONTEXT

4.3.1 Heat Exchanger Optimization - Design Equations

The equations used for the design of S&T-HE using Bell's method for the proposed equation based optimization technique is given below. The equations are the simplified version of Sinnott (2005) which are taken from Choudhury et al. (2008).

Shell diameter can be expressed as:

$$D_s = a + b d_o N_t^{1/n} \quad (8)$$

where,

$$b = \frac{(m+1)}{K_1^{1/n}} \quad (9)$$

The area of heat exchanger with number of shell side pass equal to *NS* can be given as:

$$A = \pi d_o L(NS)\left(\frac{D_s - a}{bd_o}\right)^n \tag{10}$$

Baffle spacing computed from the mass flow rate is:

$$L_{bc} = R_{bs} D_s = \frac{M_s}{G_s D_s \left(1 - \frac{d_o}{p_t}\right)} \tag{11}$$

The equation relating R_{bs}, D_s and L can be given as:

$$L(D_s - a)^n = C_5\left[\frac{D_s^{1.2} R_{bs}^{0.6}}{C_4 \times corfact} + R_{bs1} + C_7\right] \tag{12}$$

where,

$$R_{bs1} = \frac{L^{1/3}(D_s - a)^{n/3}}{C_6} \qquad \text{for } R_{et} \leq 2100 \tag{13a}$$

$$R_{bs1} = \frac{1}{C_8\left(1 + \left(\frac{d_i}{L}\right)^{2/3}\right)\left(\frac{C_9}{(D_s - a)^{2n/3}} - 125\right)} \qquad \text{for } 2100 < R_{et} \leq 10000 \tag{13b}$$

$$R_{bs1} = \frac{(D_s - a)^{0.8n}}{C_{10}} \qquad \text{for } R_{et} > 10000 \tag{13c}$$

$$corfact = F_n F_w F_b F_L \tag{13d}$$

where F_n is the tube row correction factor, F_w is the window zone correction factor, F_b is the bypass correction factor and F_L is the leakage correction factor.

For tube side pressure drop, the expression given by Muralikrishna and Shenoy, 2000, is:

$$\Delta P_t = C_1 \frac{L}{(D_s - a)^{n(2+mt)}} + C_2 \frac{1}{(D_s - a)^{2n}} \tag{14}$$

The total shell side pressure drop can be expressed as:

$$\Delta P_s = \Delta P_i F_b'\left(\frac{2(N_{wv}+N_{cv})}{N_{cv}} + F_L'(N_b-1)\right) + F_L'(2+0.6N_{wv})\frac{\rho u_z^2}{2}N_b \tag{15}$$

where, ΔP_i is the ideal tube bank pressure drop, F_b' is the bypass correction factor for shell side pressure drop, N_{wv} is the number of restrictions for cross flow in window zone, N_{cv} is the number of constrictions, tube rows encountered in cross flow zone, F_L' is the leakage correction factor for the pressure drop, N_b is the number of baffles, u_z is the geometric mean velocity defined as:

$$u_z = \sqrt{u_w u_s} \tag{16}$$

where, u_w is the shell side velocity in window zone

Heat duty equation is given by:

$$Q = UAF(LMTD) \tag{17}$$

Some of the equations of Choudhury et al., 2008, are modified to improve accuracy

The tube row correction factor, F_n, is:

$$F_n = 0.8810498 + 0.0191251 \times N_{cv} - 9.775\times10^{-4}\times N_{cv}^2 + 2.448\times10^{-5}\times N_{cv}^3 \quad \text{for } R_{es}>2000 \tag{18a}$$

$$F_n=1 \quad \text{for } R_{es}>100 \text{ to } 2000 \tag{18b}$$

$$F_n \propto (N_c')^{-0.18} \quad \text{for } R_{es}<100 \tag{18c}$$

Window correction factor, F_w, is a function of R_w and is given as:

$$F_w = 1.003403 + 3.8145548\times R_w - 31.568113\times R_w^2 + 111.56454\times R_w^3 - 205.77385\times R_w^4 + 188.43572\times R_w^5 - 67.651795\times R_w^6 \tag{19}$$

j_f is the function of shell side Reynold number (R_{es}) and is expressed as

$$j_f = 10(mm\times\log \mathrm{Re}_s + cc) \tag{20}$$

where the constants used in equation are given in Table 4.7.

Table 4.7. Value of constants used in Eq.11.

Type of pitch	Square pitch		Triangular pitch	
Re_s	mm	cc	mm	cc
<1000	-.62108	0.68932	-0.60206	0.80618
1000 to 10000	0.019023	-1.231	-0.18708	-0.43874
>10000	-0.27203	-0.0667	-0.1054267	-0.76538

For equations of D_b, H_b, N_{cv}, N_w, N_c, R_w, A_w, R_a, A_{tb}, A_{sb}, A_b, h_i, h_{oc}, h_s, F_L, ΔP_i, u_z, u_w, *C1, C2, C4, C5, C6, C7, C8, C9* and *C10* refer Choudhury et al. (2008).

4.3.2 Total Cost

The total cost consist of five components: capital cost of the exchanger and two pumps (one for the tube-side and another for the shell-side), and the annual operating (power) cost for these two pumps.

$$\text{Capital Cost} = C_a + C_b A^c + C_e + C_f\left(M_t \Delta P_t / \rho_t\right) + C_e + C_f\left(M_s \Delta P_s / \rho_s\right) \qquad (21)$$

$$\text{Operating Cost/year} = C_{pow} H / \eta\left[M_t \Delta P_t / \rho_t + M_s \Delta P_s / \rho_s\right] \qquad (22)$$

The values of above constants are given in Appendix D.

Data for remaining constants of Eq.9 is taken from Muralikrishna and Shenoy, 2000, which are as follows.

Capital cost of pump (\$): $C_e + C_f \times \left(M_t \Delta P_t / \rho_t\right)^g = 2000 + 5\left(M_t \Delta P_t / \rho_t\right)^{0.68}$

Cost of Power (\$/kWh): $C_{pow} = 0.05$

Pump efficiency: $\eta = 70\%$

Plant operation (h/yr): $H = 7200$

Rupees/Dollar = 40

4.3.3 Pressure Drop Diagram and Feasible Region

Three main equations, namely the heat duty; Eq. 17, tube side pressure drop, Eq. 14 and shell side pressure drop, Eq. 15, include five variables namely ΔP_t, ΔP_s, L, D_s and R_{bs}. Thus degree of freedom is two. This means that if two of the five variables are specified the rest can be found out. Therefore a point on the pressure drop diagram which represents values of two variable, ΔP_t and ΔP_s, are when fixed the values of remaining three variables can be computed from the three equations. Since the equations are non-linear in nature, each point on the pressure drop diagram does not give a unique design. In fact, it offers more than one set of variables that could satisfy a point on the pressure drop diagram.

Feasible region: Although, every point on the pressure drop diagram yields a possible design, every design is not acceptable because the design of a heat exchanger is typically governed by a number of constraints. If the geometrical constraints, four in numbers namely maximum shell diameter, maximum tube length and maximum and minimum ratio of baffle spacing to shell diameter and six operating constraints, e.g., maximum allowable tubeside and shellside pressure drop and maximum and minimum velocity of shellside and tubeside fluid are plotted on the pressure drop diagram, then a well defined region, shown by shaded area in Fig. 4.9 as feasible region can be obtained.

Any design which falls in this region can be considered as feasible design. The upper limits to the allowable tubeside and shellside pressure drops appear as vertical and horizontal lines on ΔP_t vs. ΔP_s plot (Fig. 4.9). Thus, the pressure drop constraints define a rectangle on the pressure drop diagram as the region of feasible designs which obeys the pressure drop constraints. This region is further reduced using other constraints. The maximum pressure drops ($\Delta P_{t,max}$ and $\Delta P_{s,max}$) represent the limiting abilities of the external pumps to transport the fluids along the exchanger. Mathematically,

$\Delta P_t \leq \Delta P_{t,max}$ (23a)

and $\Delta P_s \leq \Delta P_{s,max}$ (23b)

There are also upper and lower bounds to the tube side and shell side velocities.

$u_{t,min} \leq u_t \leq u_{t,max}$ (24a)

and $u_{s,min} \leq u_s \leq u_{s,max}$ (24b)

In terms of primary geometrial constraints, there are upper limits to shell diameter and tube length, i.e.,

$D_s \leq D_{s,max}$ (25a)

$L \leq L_{max}$ (25b)

Furthermore, close baffle spacing leads to higher heat transfer coefficient but at the expense of pressure drop whereas wide baffle spacing results in bypassing and reduced cross flow; hence, there is a decrease in the heat transfer coefficient. Rbs is a constraint which follows:

$R_{bs,min} \leq R_{bs} \leq R_{bs,max}$ (26)

Fig. 4.9 shows the feasible region after these constraints are plotted on pressure drop diagram.

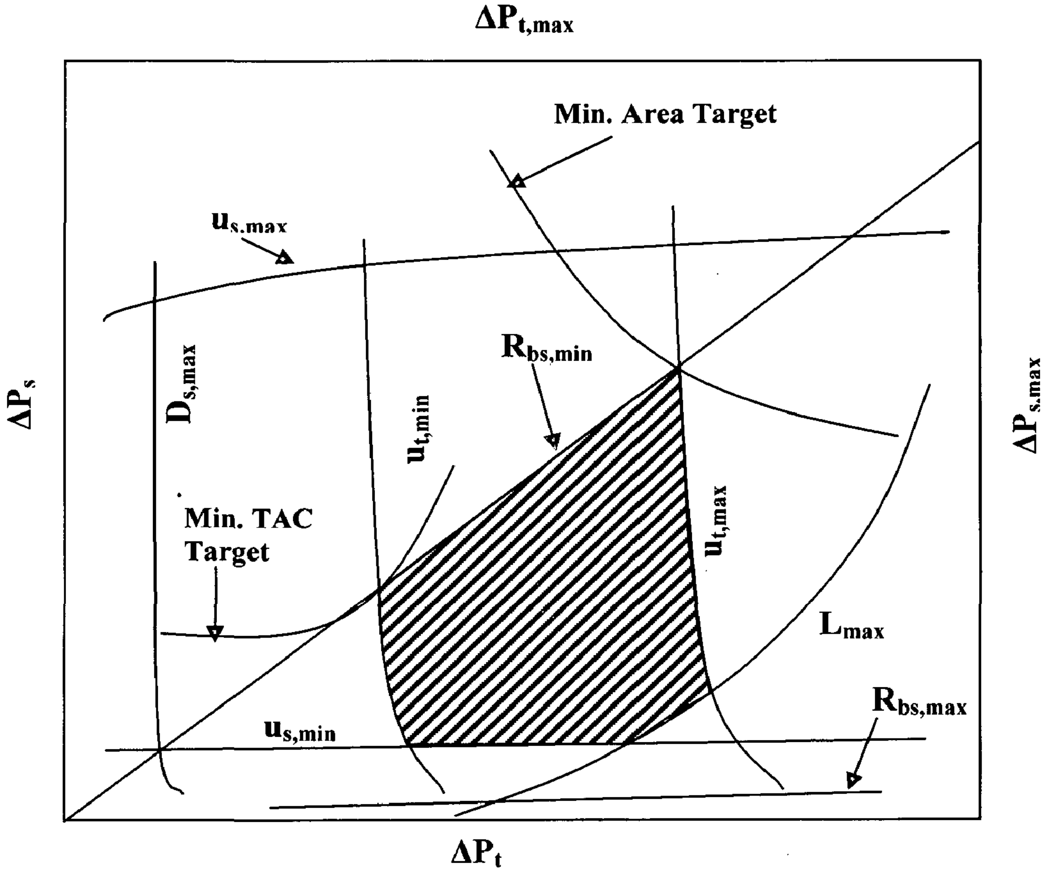


Figure 4.9. Geometrical and operating constraints define the feasible region for design on pressure drop diagram

4.3.4 Stepwise procedure for obtaining the feasible region

The stepwise procedure for obtaining the feasible region in pressure drop diagram is given in Table 4.8.

Table 4.8. Stepwise procedure for obtaining the feasible region

Specified variable	**Step 1**	**Step 2**	**Step 3**	**Step 4**	**Step 5**
$u_t = u_{t,min}$ or $u_t = u_{t,max}$	Compute D_s (Eq. 8)	Choose ΔP_t and vary	Compute L (Eq. 14)	Compute R_{bs} (Eq. 12)	Compute ΔP_s (Eq. 15)
$u_s = u_{s,min}$ or $u_s = u_{s,max}$	Choose D_s and vary	Compute R_{bs} (Eq. 11)	Compute L (Eq. 12)	Compute ΔP_t (Eq. 14)	Compute ΔP_s (Eq. 15)
$D_s = D_{s,max}$	Choose ΔP_t and vary	Compute L (Eq. 14)	Compute R_{bs} (Eq. 12)	Compute ΔP_s (Eq. 15)	
$L = L_{max}$	Choose D_s and vary	Compute ΔP_t (Eq. 14)	Compute R_{bs} (Eq. 12)	Compute ΔP_s (Eq. 15)	
$R_{bs} = R_{bs,max}$ or $R_{bs} = R_{bs,min}$	Choose D_s and vary	Compute L (Eq. 12)	Compute ΔP (Eq. 14)	Compute ΔP_s (Eq. 15)	

4.3.5 Application in present context

Optimization of heat exchanger can result in minimum –area/ -total annual cost. In our case study we recommend to use preheater in order to reduce the reboiler load. Now, if this preheater is optimized a further saving can be established thereby reducing the capital cost of the exchanger, hence the pay back period. The detailed solution is discussed in Section 5.3.

CHAPTER 5

RESULTS AND DISCUSSION

5.1 EXISTING CASE

Column Targeting techniques as discussed in Chapter 4 are applied to the problem under study defined in Chapter 3. The targeting results for the existing case are given in Appendix A. The Column Grand Composite Curve (CGCC) which is the plot of Enthalpy deficit vs. Stage Number or Enthalpy deficit vs. Temperature which is obtained for the existing process is shown in Fig. 5.1. The condenser and reboiler duty required for Column B1 are 3524.05 kW and 4281.39 kW respectively, whereas for Column B2 it is 8055.90 kW and 7807.92 kW respectively. From Fig. 5.1, the scope for reflux modification is 902 kW for Column B1 and 1125 kW for Column B2.

The approach to the problem will be to consider each column separately and for each column applying the column modification in the sequence as discussed in Section 4.3. But after solving and analyzing, the new order that must be used for column modification is proposed. The sequence is:

1. Feed conditioning (heating or cooling)
2. Feed location (appropriate placement)
3. Reflux ratio modification (reflux ratio vs. number of stages)
4. Side condensing or reboiling

The reason for this is, suppose that a column is modified according to earlier sequence, feed location is changed and the optimum feed location is found; now condition of feed is changed by placing a pre-heater. Due to change in the condition of feed, the optimum feed location is changed and hence, again we need to find the optimum feed location, as it depends on the condition of feed.

Two options are considered, one in which no capital cost is incorporated i.e. no pre-heater or side condenser/reboiler, and the second one in which all the possible modification.

5.2 OPTION A: NO ADDITIONAL CAPITAL INVESTMENT

5.2.1 Case 1: Column B1-Feed plate location

For Column B1, feed plate location is 30 in the existing case with scope for reflux modification of 902 kW. As feed plate location is increased, the scope for reflux modification decreases whereas with the decrease in the feed plate location the scope for reflux modification increases till feed plate number 23 after which the scope decreases. The scope for reflux modification at feed plate 23 is 925 kW and is shown in Fig. 5.2.

5.2.2 Case 2: Column B1-Reflux Ratio Modification

As the reflux ratio is reduced, the CGCC moves towards the ordinate, thus reducing both the condenser and reboiler loads. The CGCC for Case 1 as shown in the Fig. 5.2 also identifies the scope for reduction in condenser and reboiler duties by reducing the reflux ratio. It must be noted that, as the reflux ratio is reduced, the desired separation is bound to get affected. Therefore, the change in the desired separation should be noted. If the change is appreciably high then we need to increase number of stages. In our case the difference is less than 0.022 mole fraction which we assumed as acceptable (Appendix B). The reflux ratio is reduced to 7.2, we get the condenser and reboiler duty as 2630.04 kW and 3385.83 kW, hence resulting in saving of 894.01 kW and 895.56 kW of condenser and reboiler duty respectively. The CGCC for Case 2 is shown in the Fig. 5.3.

5.2.3 Case 3: Column B2-Feed plate location

For Column B2, feed plate location is 30, with reflux modification of 1125 kW. Also for Column B2, as the feed plate location is increased, the scope for reflux modification decreases and with the decrease in the feed plate location the scope for reflux modification increases till feed plate number 22 after which the scope decreases. The scope for reflux modification at feed plate 22 is 1146 kW and is shown in Fig. 5.4.

5.2.4 Case 4: Column B2-Reflux Ratio Modification

The CGCC for Case 3 as shown in the Fig. 5.4 also identifies the scope for reduction in condenser and reboiler duties by reducing the reflux ratio. If we reduce the reflux ratio to 0.87, the condenser and reboiler duty are 6932.82 kW and 6684.86 kW hence resulting in saving of 1123 kW in condenser and reboiler duty each. For this case too the change in the composition is less than 0.022 mole fraction which we assumed as acceptable. The CGCC for Case 4 is shown in the Fig. 5.5.

5.2.5 Savings

The total savings for Option A is given in Table 5.1.

Table 5.1: Savings for Option A

Parameter	Unit	Before targeting	After targeting	Savings
Condenser heat duty	kW	11579.95	9562.86	2017.09
Reboiler heat duty	kW	12089.31	10070.69	2018.62
Water consumption in condensers	kg/hr	997316.3	823595.6	173720.7
Hot Oil-1 consumption in reboilers	kg/hr	939178.2	782358.3	156819.9

5.3 OPTION B. WITH ADDITIONAL CAPITAL INVESTMENT

5.3.1: Case 5: Column B1-Feed Conditioning

From the Fig. 5.1a, the enthalpy change on the reboiler side is noticeably sharper. Therefore, our design can benefit from addition of a feed pre-heater. A S&T-HE is added prior to distillation column which increases feed temperature to column to 190^0C. CGCC for Case 5 is shown in Fig. 5.6. The reboiler duty is reduced to 2939.34 kW resulting in saving of 1342.05 kW at the cost of an exchanger.

5.3.2 Case 6: Column B1-Feed Plate location and Reflux Ratio Modification

Feed is properly placed at plate number 30. Changing feed location does not yield any energy saving. From Fig. 5.6, it is clear that there is no scope for reflux modification.

5.3.3 Case 7: Column B1-Side Condensing/Reboiling

The scope for side condensing or side reboiling can be identified from the area beneath or above the CGCC pinch point (area between the ideal and actual enthalpy profiles). As seen from Fig. 5.6 a significant area exists below the pinch, a side condenser can be placed at an appropriate temperature level in order to reduce the condenser load. Fig. 5.7 shows the CGCC for Column B1 after using Side Condenser with condenser duty of 1500 kW. Although the load on condenser is reduced to 2024.27 kW, but this would not be beneficial in present case as the cold utility available is only cooling water. The cooling required in the side condenser will also be given by cooling water resulting in the same amount of water to be used at the cost of extra side condenser. Therefore employing a side condenser will not yield any benefits under present condition. It should be noted that, if any of the process fluid is available at the temperature level shown in Fig. 5.7, that is required to be heated then it is possible to use that as a cooling utility in side condenser.

5.3.4 Case 8: Column B2-Feed preheating, Feed plate location and Reflux Ratio Modification

From the Fig. 5.1b, the enthalpy change on the reboiler side is not sharp. Due to the flatness of the curve above the pinch point there is no scope for feed preheating or in other words, feed preheating will not yield any significant savings. Then the present case becomes similar to case 3. Feed plate location is changed to 22 (Case 3, Fig 5.4) and reflux ratio changed to 0.85 (Case 4, Fig. 5.5). After this case the condenser and reboiler duties are 6932.82 kW and 6684.86 kW respectively, hence resulting in saving of 1123 kW each in condenser and reboiler duty.

5.3.5 Case 9: Column B2- Side Condensing/Reboiling

After observing the CGCC in Fig 5.5, it is clear that there is a scope for using side reboiler. As hot utility (Hot oil-2) at 250^{0}C is available, therefore we can employ a side reboiler of 3000 kW at plate number 28. Due to this the curve slightly shifts towards right , hence there exist a scope for reflux modification. The reflux ratio is changed to 0.7. Resulting CGCC is shown in Fig. 5.8. The composition of the simulated results of each streams are checked and the difference in mole fractions of respective component is less than 0.04 mole fraction which as specified earlier is assumed as acceptable. The load on this side reboiler is 3000 kW, the reboiler load is reduced to 3057.39 kW. Due to the change in the reflux ratio the condenser duty also reduced to 6305.26 kW.

5.3.5 Savings

The total savings for Option B is given in Table 5.2. Fig. 5.9 and 5.10 gives the summary of energy saving for Option A and Option B respectively.

Table 5.2: Savings for Option B

Parameter	Unit	Before targeting	After targeting	Savings
Condenser heat duty	kW	11579.95	9829.53	+ 1750.42
Reboiler heat duty	kW	12089.31	5996.73	+ 6092.58
Side Reboiler heat duty (Case 9)	kW	-	3000	- 3000.00
Feed Pre-heater heat duty (Case 5)	kW	-	1342.05	- 1342.05
Water consumption in condensers	kg/hr	997316.3	846562.4	+ 150753.9
Hot Oil-1 consumption in reboilers	kg/hr	939178.2	455255.8	+ 483922.4
Hot Oil-2 consumption in side-reboiler (Case 9)	kg/hr	-	245677.9	- 245677.9
Hot Oil-3 consumption in feed pre-heater (Case 5)	kg/hr	-	117266.5	- 117266.5

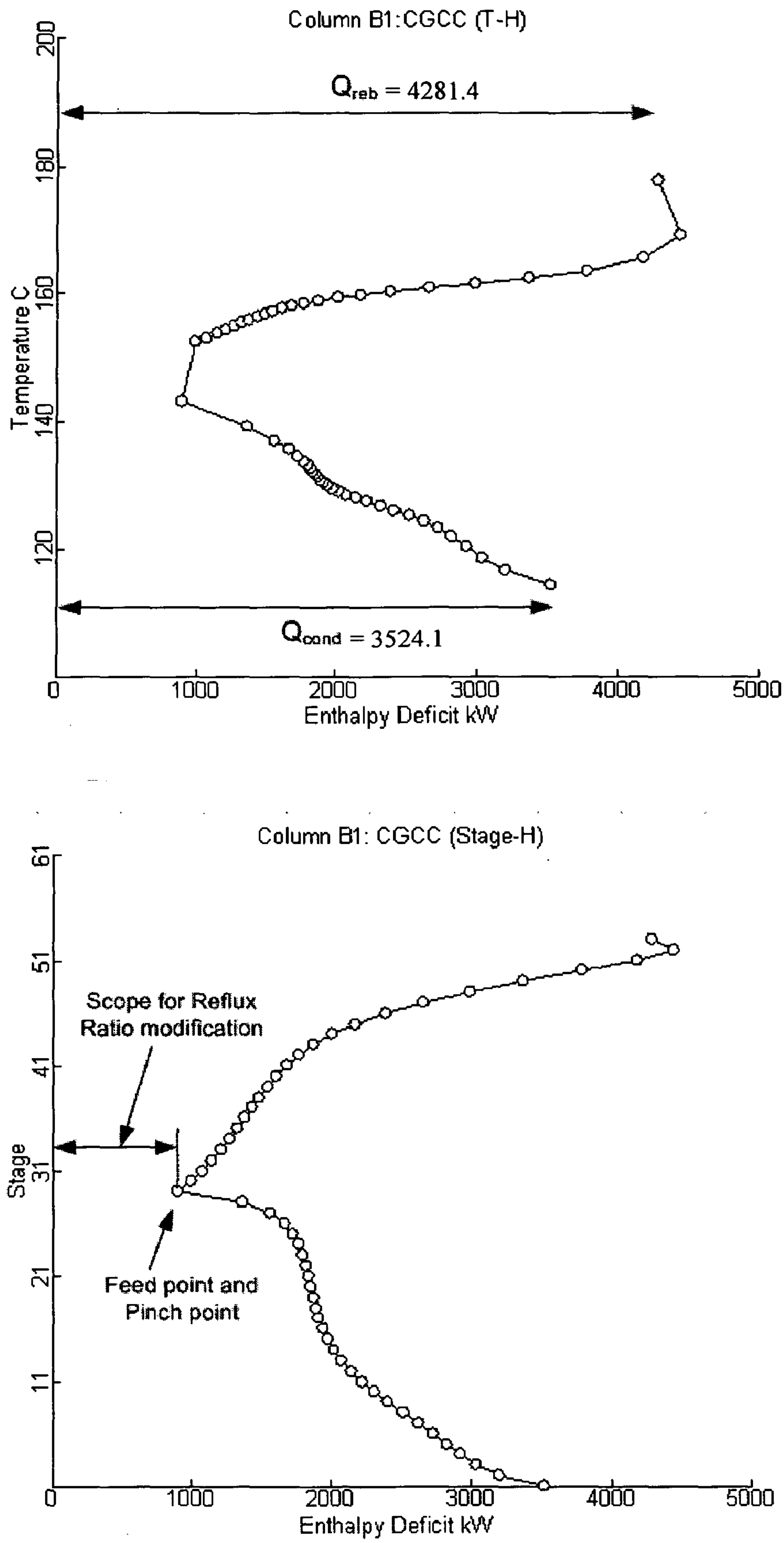


Figure 5.1a. Column B1- CGCC for existing process

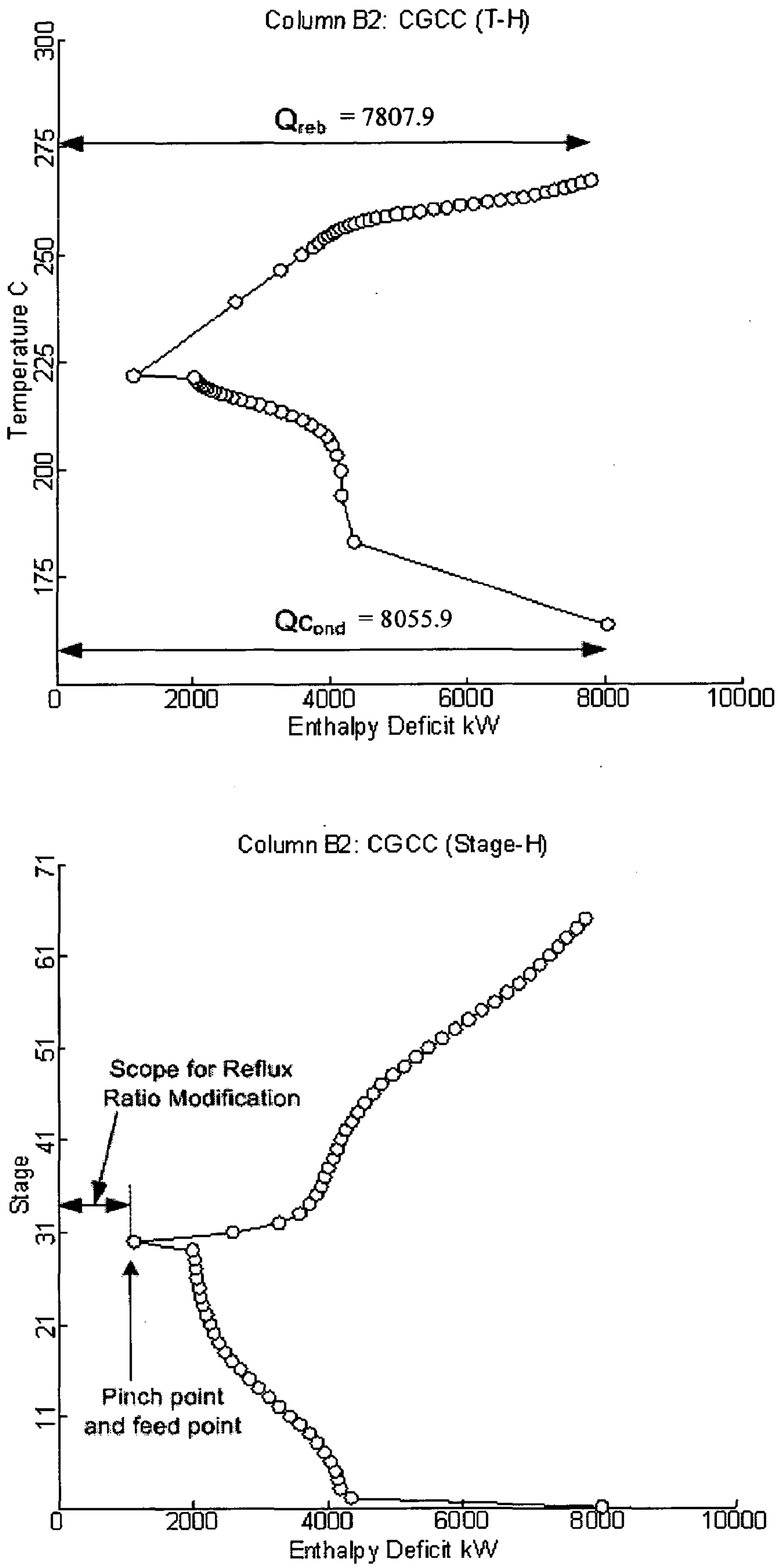
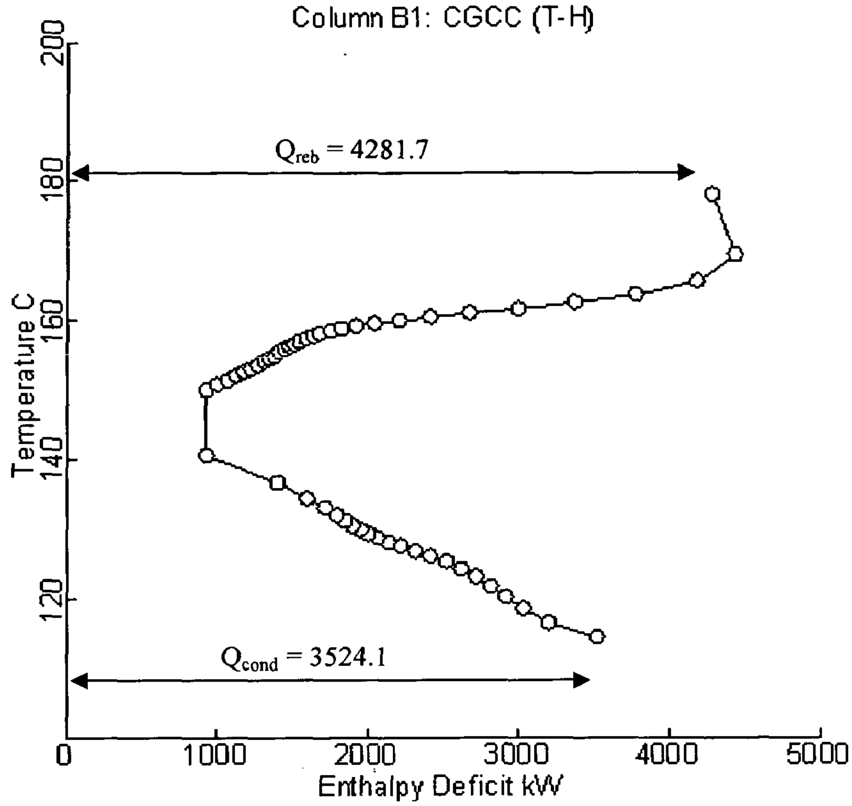


Figure 5.1b. Column B2- CGCC for existing process



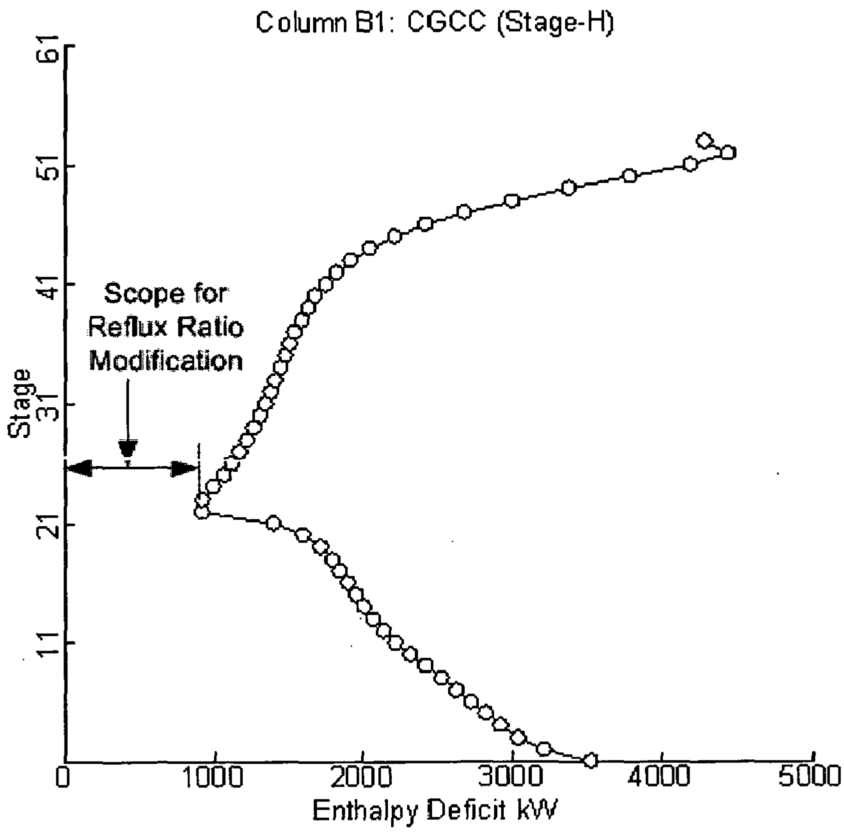
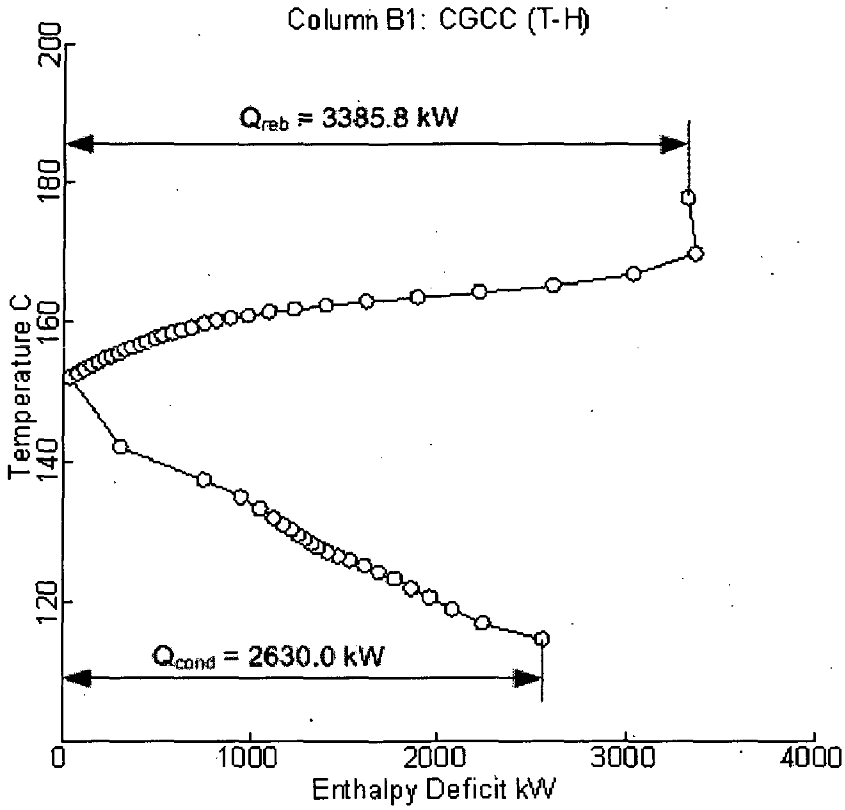


Figure 5.2.Column B1-CGCC after changing the feed plate location (Case 1)



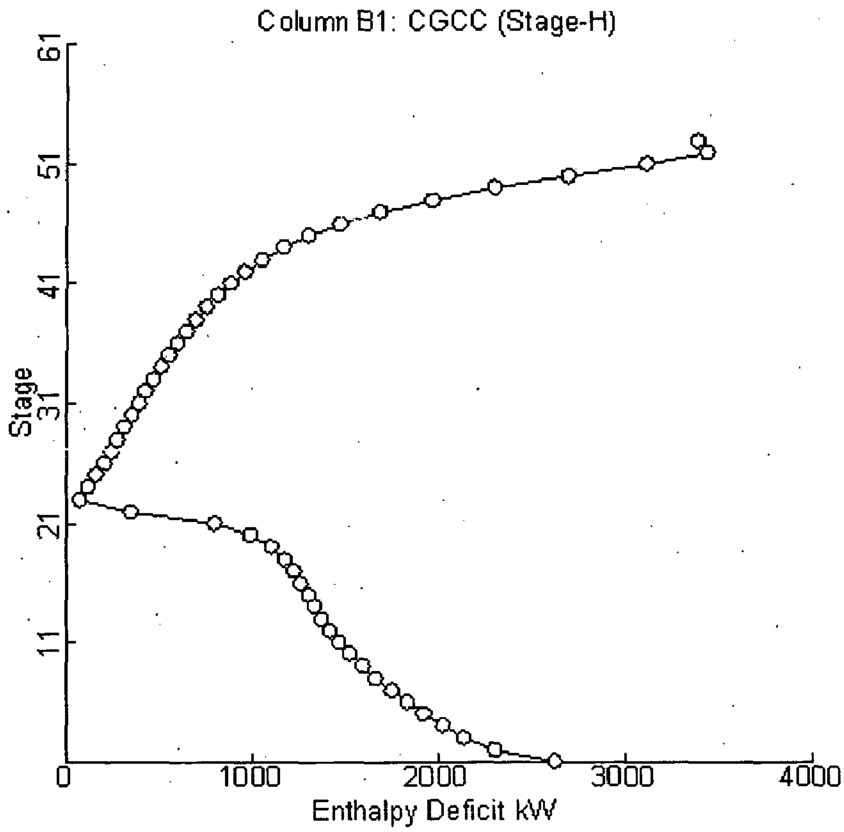


Figure 5.3. Column B1-CGCC after Reflux Ratio Modification (Case 2)

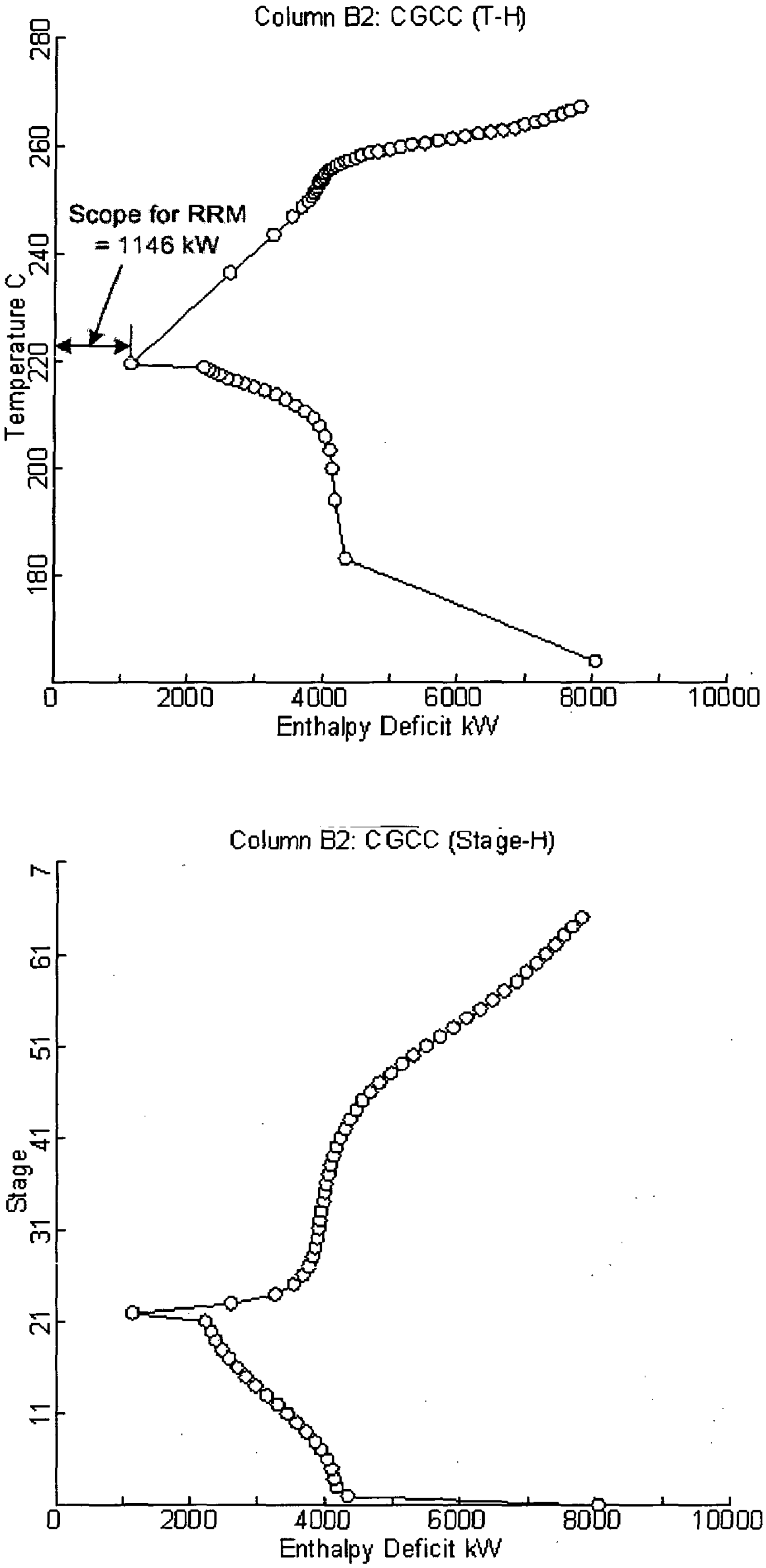


Figure 5.4. Column B2-CGCC after changing the feed plate location (Case 3)

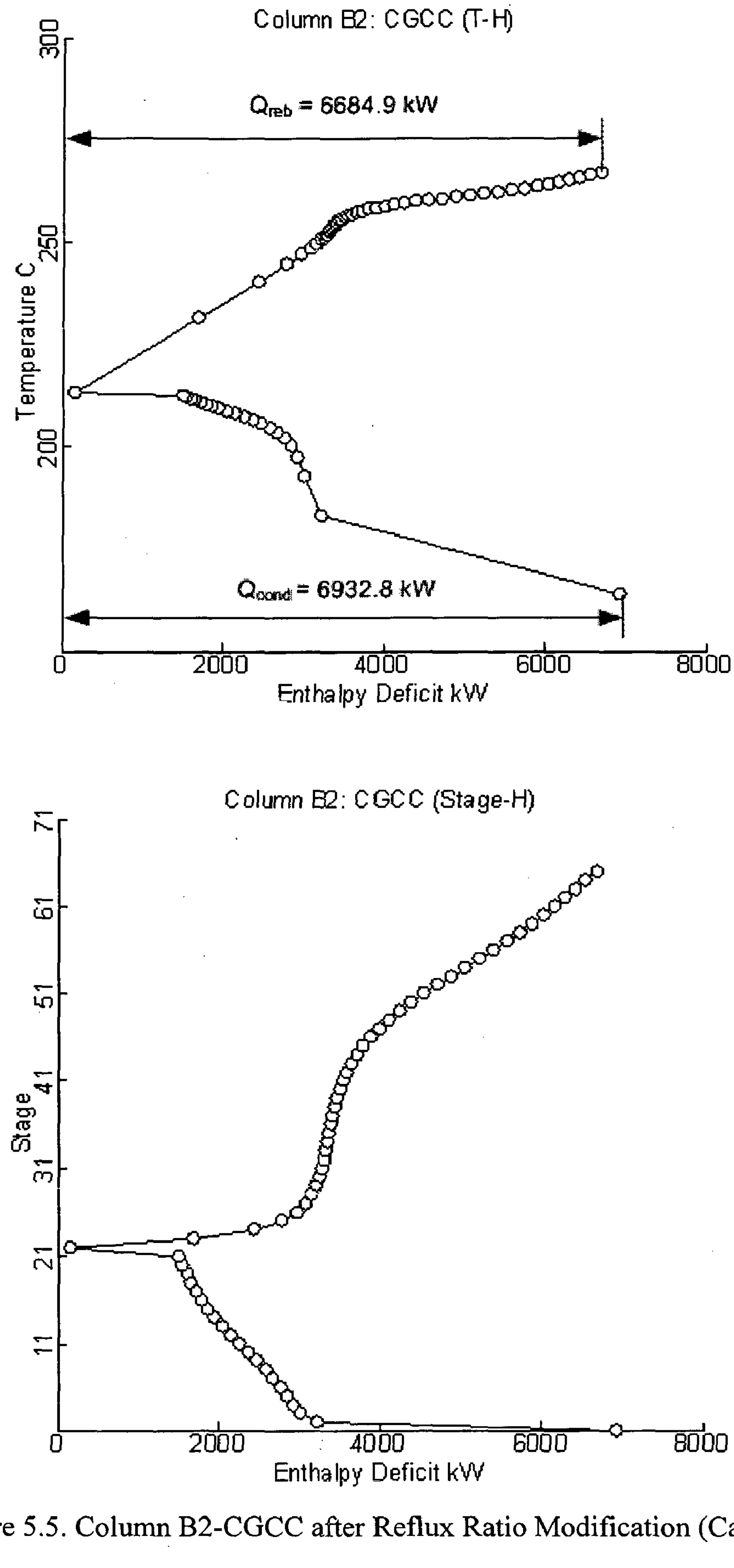


Figure 5.5. Column B2-CGCC after Reflux Ratio Modification (Case 4)

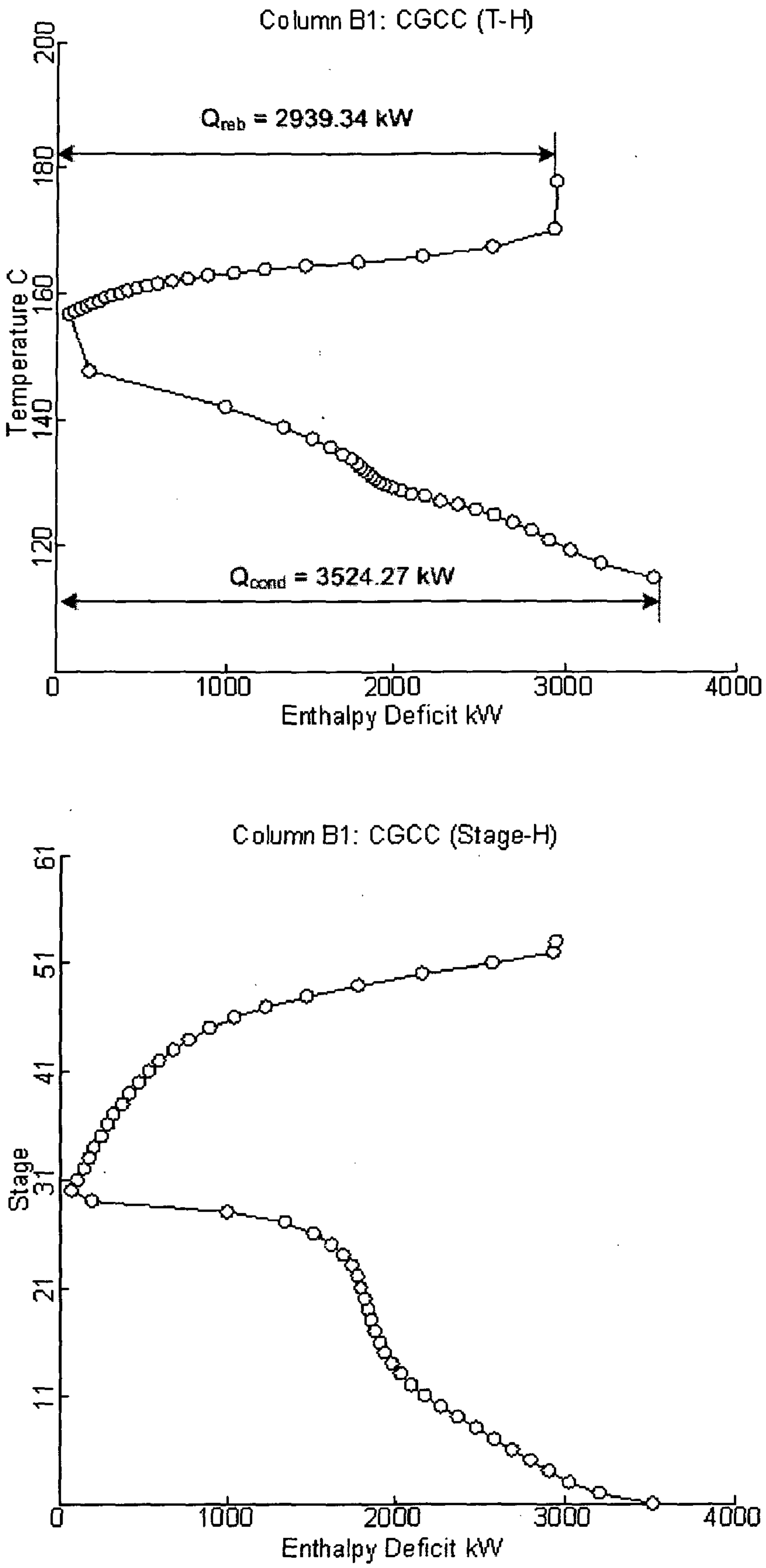
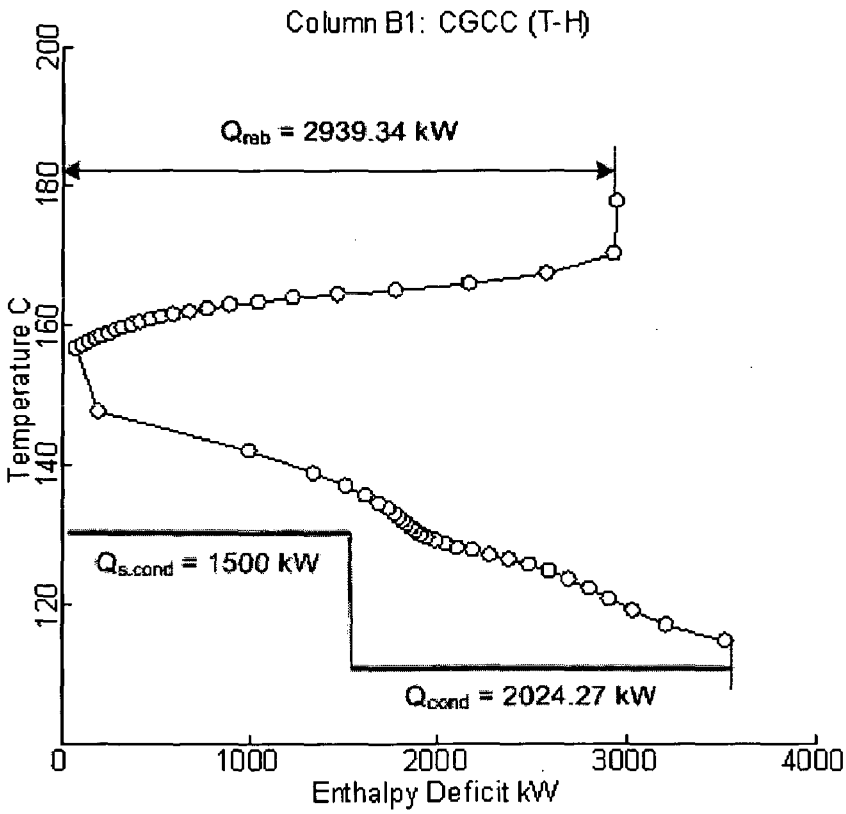


Figure 5.6. Column B1-CGCC after Feed preheating (Case 5)



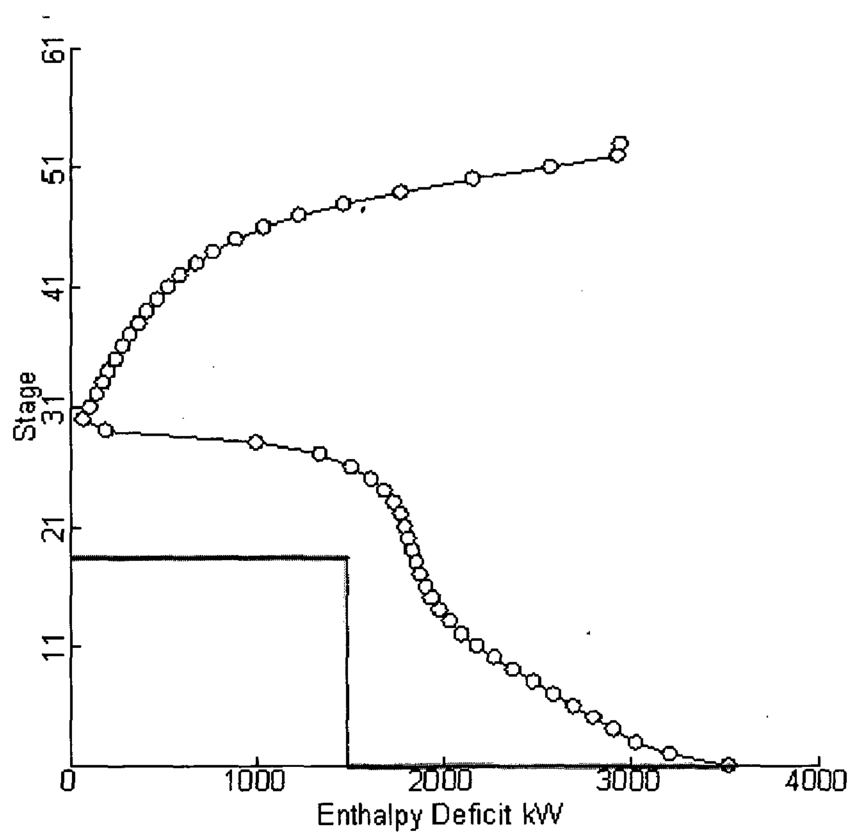


Figure 5.7: Column B1-CGCC after using Side Condenser (Case 7)

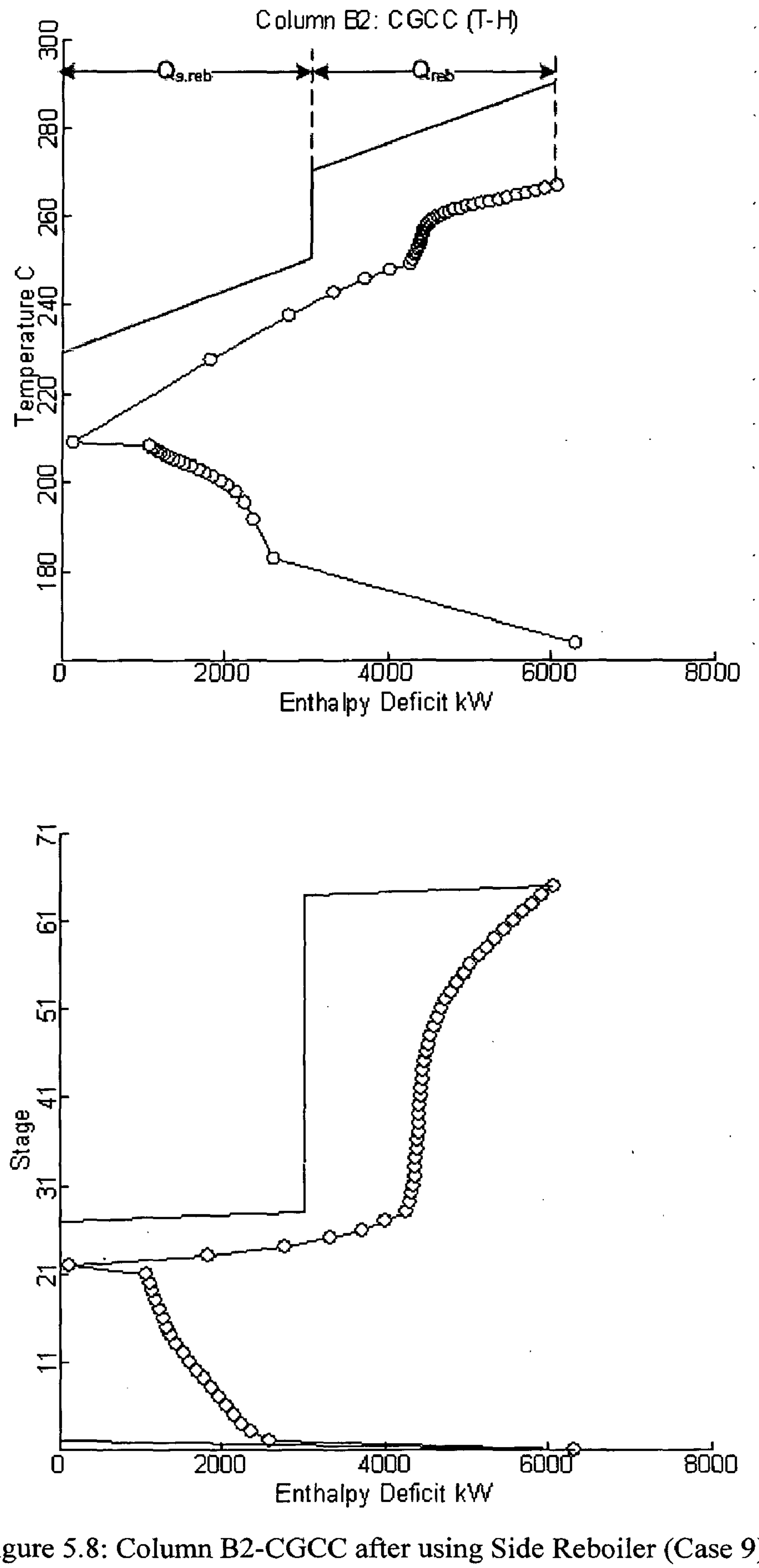


Figure 5.8: Column B2-CGCC after using Side Reboiler (Case 9)

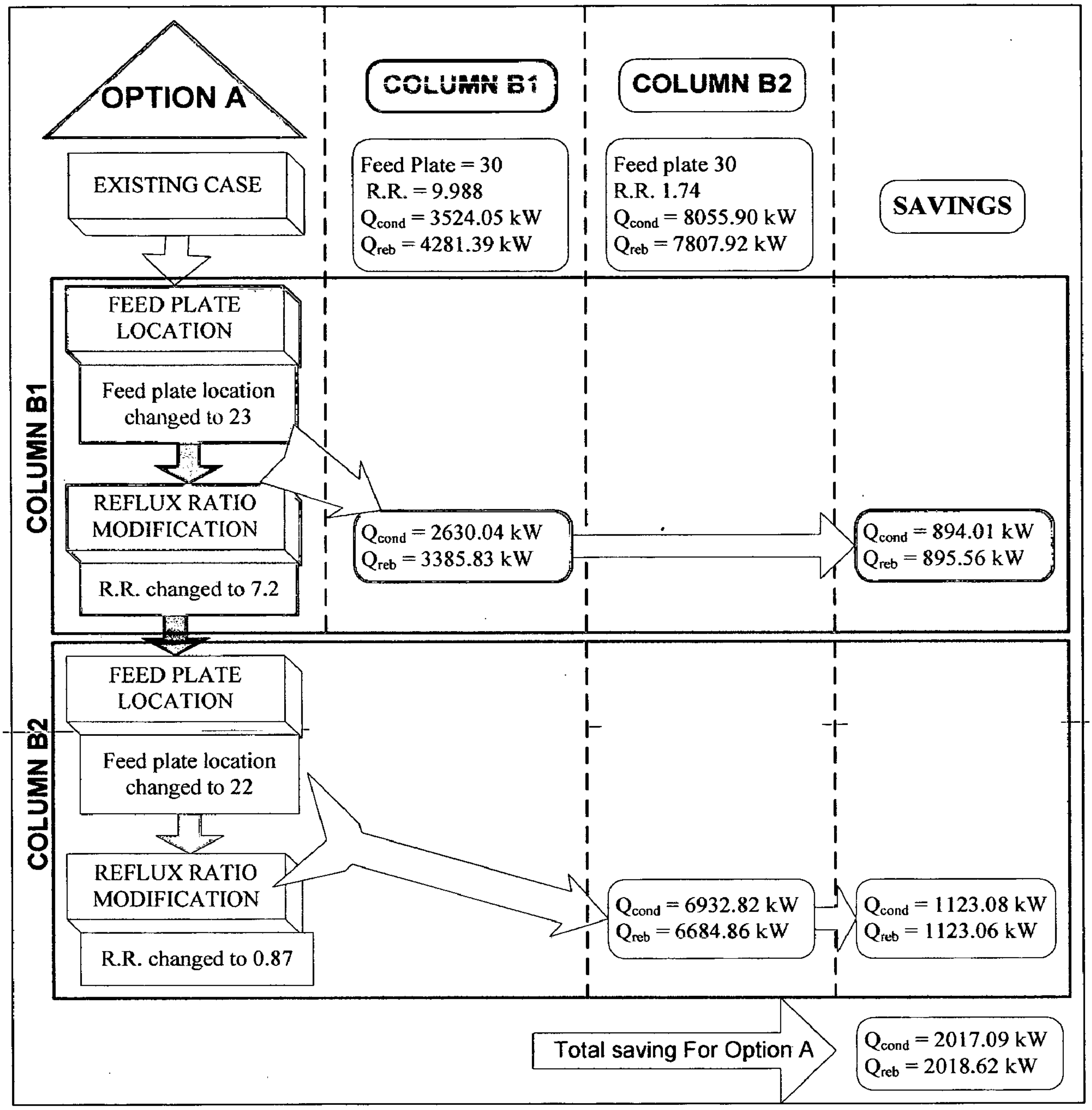


Figure 5.9: Summary of energy saving for Option A

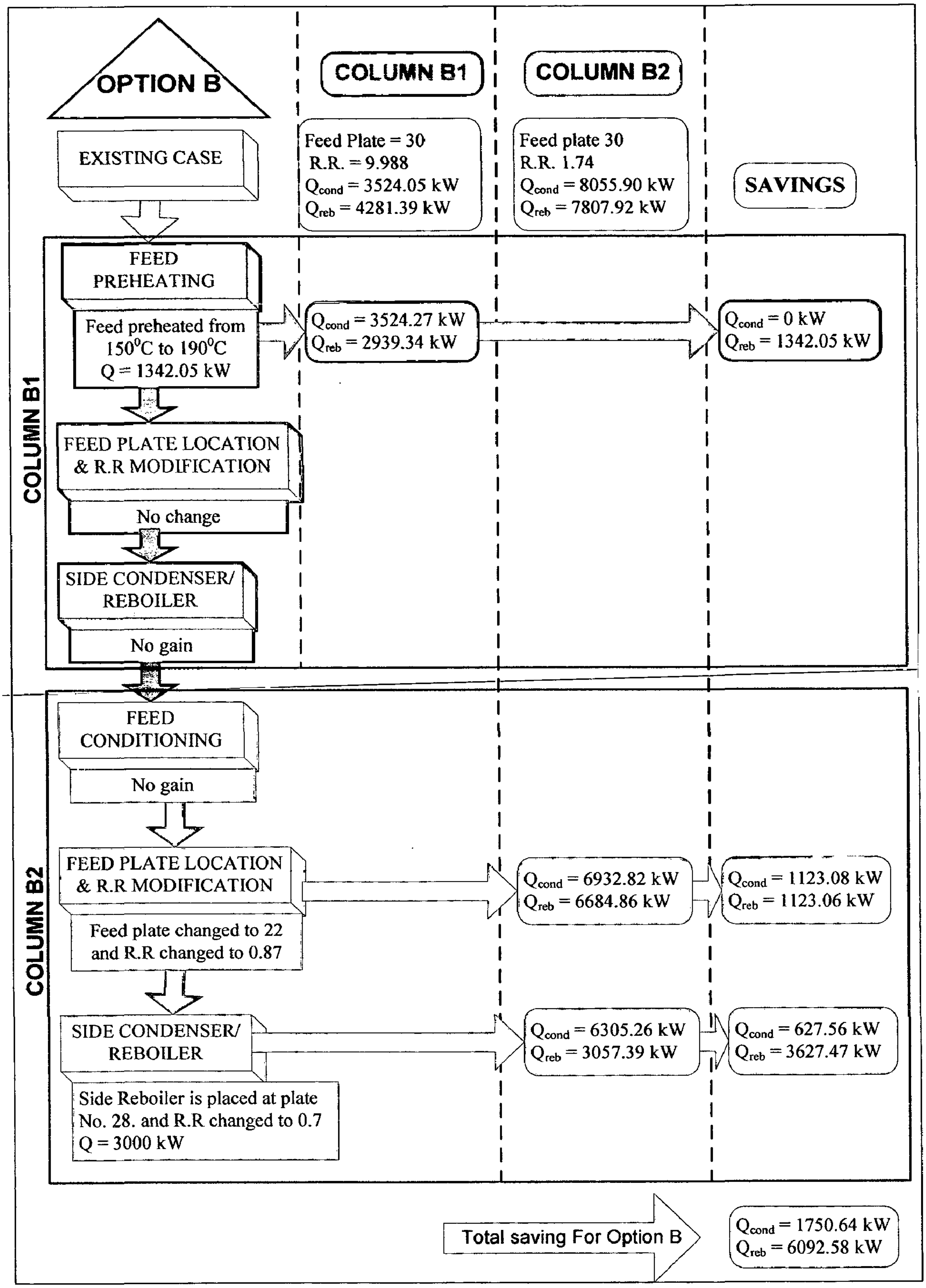


Figure 5.10: Summary of energy saving for Option B

5.4 COSTING

Basis for the utility cost data is given in Appendix E.

5.4.1 Costing for Option A

Table 5.3 shows the savings in terms of Rs/yr for Option A.

Table 5.3: Savings in terms of Rs/yr for Option A

Cases	Process utility	Savings (kg/hr)	Savings (Rs/yr)	Total Savings (Rs/yr)
After Column 1 Modification (Case 1 and 2)	Hot Oil-1 saved	69573.1	110203741	112033167
	Cooling water saved	76996.1	1829427	
After Column 2 Modification (Case 3 and 4)	Hot Oil-1 saved	87246.8	138198907	140497083
	Cooling water saved	96724.6	2298176	
Total (Rs/yr)				252530251

With Option A, we get an annual saving of Rs. 25,25,30,251, without any additional investment.

5.4.2 Costing for Option B

Option B includes pre-heater and side reboiler, therefore total operating cost is given as:

Total Annual Operating cost (Rs/yr) = Savings in Rs/yr in (Cooling water + Hot Oil-1) – Consumption in Rs/yr in (Hot Oil-2 + Hot Oil-3) – (Operational cost of pre-heater and side reboiler)

Operating cost of pre-heater and side reboiler mainly consist of pumping cost of both side fluids, where the pumping cost is given as:

$$\text{Pumping Cost} = \frac{(Rs/kWh)\,H}{\eta}\left(\left(\frac{\Delta P_t M_t}{\rho_t}\right)+\left(\frac{\Delta P_s M_s}{\rho_s}\right)\right) \tag{5.1}$$

Assuming following operating cost data (Muralikrishna and Shenoy, 2000)

Pumping cost (Rs/kWh) = 4

Hours of operation per year (hr/yr) = H = 7200

Pump Efficiency (η) = 70 %

Pressure drop on both tube side and shell side is assumed as 0.5 atm.

$$\text{Pumping cost of pre-heater (Rs/yr)} = (4/1000) \times 7200/0.7 \times ((50663.5 \times 13.52/772) + (50663.5 \times 32.7205/916))$$
$$= 110963.$$

Therefore pumping cost is found to be approximately Rs. 1.1 lakhs per annum. Assuming overall annual operating cost of pre-heater to be Rs. 1.5 lakhs.

Also limited data for side-reboiler is available we assume the same overall annual operating cost of side reboiler as Rs. 1.5 lakhs. Table 5.4 shows the savings in terms of Rs/yr for Option B.

Table 5.4: Savings in terms of Rs/yr for Option B

Cases	**Process utility**	**Savings (kg/hr)**	**Savings (Rs/yr)**	**Total Savings (Rs/yr)**
After Column 1 Modification (Case 5 and 6)	Hot Oil-1 saved	104259.4	165146869	114337739
	Hot Oil-3 spent	- 117266.5	- 50659130	
	Preheater Op.cost	-	- 1500000	
After Column 2 Modification (Case 8 and 9)	Hot Oil-1 saved	369052.8	584579678	367051940
	Cooling water saved	150772.8	3582362	
	Hot Oil-2 spent	- 245677.9	- 221110100	
	Side-Reb. Op. Cost	-	- 1500000	
Total				481389679

Capital cost calculation of feed pre-heater and side-reboiler is done in Appendix D.

Capital cost of feed preheater = Rs. 23,30,324.

Capital cost of side reboiler = Rs. 74,13,749.

When compared Table 5.1 and Table 5.2, when we opt for Option B

For Column 1 modification, we get additional profit of Rs. 23,04,572, with an additional capital investment of Rs. 23,30,324.

Assuming Salvage value as zero and heat exchanger life as 10 years, with straight line depreciation we get,

Depreciation /yr = (Cost of exchanger – salvage value)/ No. of years

For feed pre-heater

Depreciation /yr = (2330324 – 0)/10 = 233032.4

Payback Period (yr) = Additional capital investment / (Annual Profit + Depreciation/yr)

= 2330324/(114337739+233032.4)

= 0.0203

Payback Period (days) = 7.4

For Column 2 modification, we get additional profit of Rs. 22,65,54,857, with an additional capital investment of Rs. 74,13,749.

Depreciation /yr = (7413749 – 0)/10 = 741374.9

Payback Period (yr) = 7413749/ (367051940+741374.9)

= 0.0202

Payback Period (days) = 7.4

Overall Payback Period(yr)= Total Capital Investment/(Total Profit/yr+ Depreciation/yr)

= 9744073/ (481389679 + 974407.3)

= 0.0202

Overall Payback Period (days) = 7.4

5.5 FURTHER SAVING AFTER HEAT EXCHANGER OPTIMIZATION

As discussed in Section 4.4, heat exchanger optimization can further decrease the capital cost of heat exchanger. Capital cost of feed pre-heater is Rs. 23,30,324. We will apply our heat exchanger optimization technique that will reduce the capital cost of pre-heater. Data for the prehater is shown in Table 5.5.

For each constraints, the program is run to get value of ΔP_s for corresponding value of ΔP_t (Appendix C). These values are plotted on the pressure drop diagram and a feasible region is obtained (Fig. 5.11). At each vertex of this feasible region the Total Annual Cost (TAC) and Area are calculated. In the present case the intersection of $R_{bs,min}$ and $u_{s,max}$ gives the minimum TAC as well as minimum Area. The results obtained are given in Table 5.7

Table 5.5 Data for preheater

	Tube side		Shell side	
	Liquid (cold stream)		Liquid (hot stream)	
Inlet temperature (^{0}C)	150		220	
Outlet temperature (^{0}C)	190		200	
Flow rate (kg/s)	13.52		32.71	
Heat capacity (J(kg^{0}C)$^{-1}$)	2492		2060	
Density (kg m^{-3})	916		777	
Viscosity (kg(m s)$^{-1}$)	0.788 x 10^{-3}		0.417 x 10^{-3}	
Thermal conductivity (W/m^{0}C)	0.12		0.122	
Outer diameter of tube (m)	0.02	Inner diameter of tube (m)	0.016	
Thermal conductivity of tube (W/m^{0}C)	36	Segmental baffle cut	25 %	
TEMA Type : BEM				

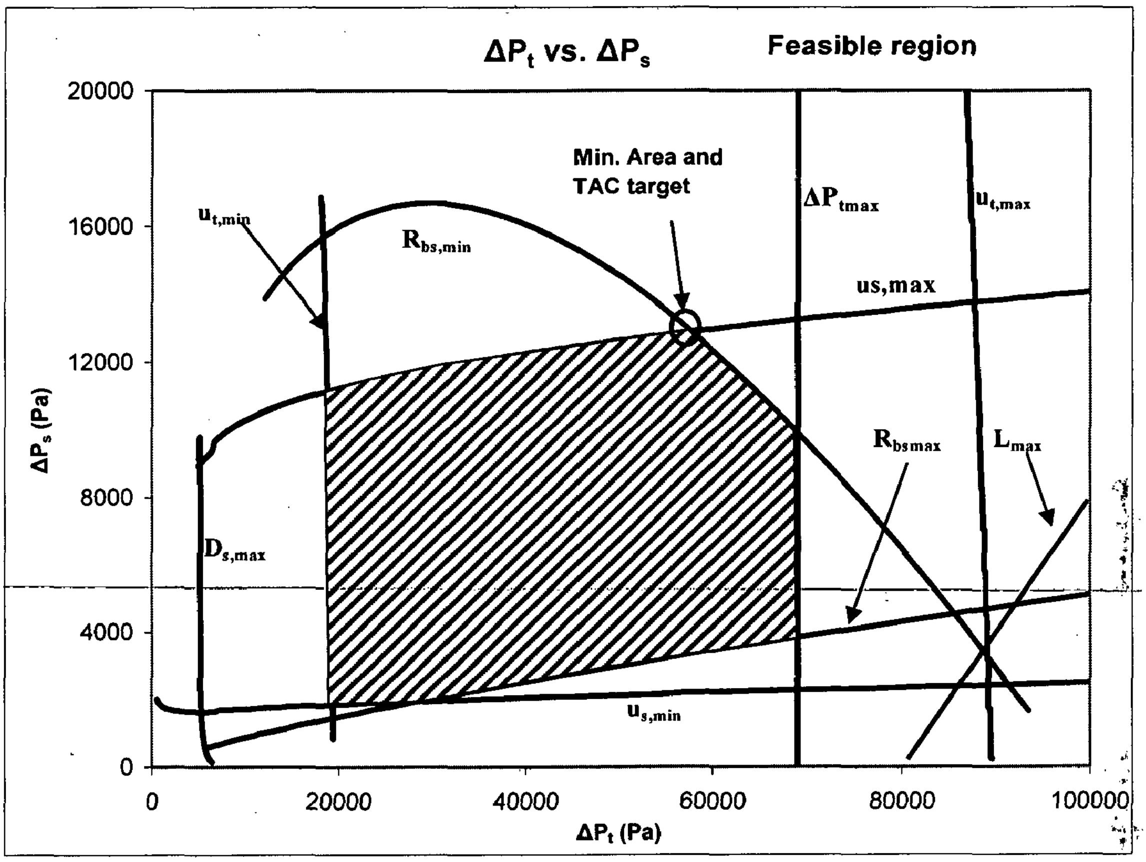


Figure 5.11: Geometrical constraints and operating constraints for pre-heater on pressure drop diagram

From heuristics, following values of constrained parameters are assumed.

Table 5.6 Values of constrained parameters

Parameter	**Value**	**Parameter**	**Value**
$\Delta P_{t,max}$ (Pa)	69000	**$\Delta P_{s,max}$ (Pa)**	69000
$u_{t,min}$ (m/s)	1	**$u_{t,max}$ (m/s)**	2
$u_{s,min}$ (m/s)	0.3	**$u_{s,max}$ (m/s)**	0.8
$R_{bs,min}$	0.2	**$R_{bs,max}$**	1
$D_{s,max}$ (m)	1.5	**L_{max} (m)**	6

Table 5.7 Results of pre-heater design.

Parameter	Minimum Area and TAC Target
ΔP_t (Pa)	57591
ΔP_s (Pa)	12905
D_s *(m)*	0.675
L (m)	4.274
R_{bs}	0.49
L_{bc}*(m)*	0.331
u_t *(m/s)*	1.67
u_s *(m/s)*	0.8
A (m^2)	84.087
Capital cost (Rs.)	**1206732**
TAC (Rs./yr)	284766

After heat exchanger optimization, the capital cost for feed pre-heater reduced to Rs. 12,06,732

Depreciation/yr = (1206732 – 0)/10 = 120673.2.

Payback period (yr) = 1206732 / (114337739 + 120773.2)

= 0.0105

Payback Period (days) = 3.8

Overall investment (Rs) = 1206732 + 7413749 = 8620481

Overall Payback Period (yr) = 8620481 / (481389679 + 862048.1)

= 0.0179

Overall Payback Period (days) = 6.5

Therefore, by adopting the optimization techniques developed in the present work the payback period of preheater is reduced from 7.4 days to 3.8 days. Hence, the overall payback period for additional equipments can be reduced from 7.4 days to 6.5 days. When Options A and B are compared, we find option B to be best.

CHAPTER 6

CONCLUSION AND RECOMMENDATION

6.1 CONCLUSION

From the study of problem undertaken, the following salient conclusions can be drawn.

1. With both the options considered, i.e., **Option A**- With no capital investment offers an annual profit of Rs. 25,25,30,251 and **Option B**- With capital investment offers an annual profit of Rs. 48,13,89,679 which involves capital investment of Rs. 86,20,481. Option B appears to be best of the two options as it gives higher profits with payback period of 6.5 days

2. **Feed location** plays an important role in optimizing the column and the energy required can be considerably reduced by properly locating the feed point. In the present case study, the feed location was near optimum; however, by bringing it to an optimal position 21 kW of heat duty was saved in reboiler as well as in the condenser.

3. **Reflux Ratio Modification** can reduce the heat duty on both reboiler and condenser side, but it changes the outlet composition. Thus numbers of trays are to be modified to compensate for this change in reflux ratio. In the present study, a decrease in R.R. from 1.74 to 0.85 offered a substantial saving of about 1000 kW each in reboiler and condenser duty. The composition of outlet streams (in terms of mole fraction) after modification when were compared with base case it was observed that maximum difference in any component was not more than 0.04. This was acceptable to the process.

4. Proper **feed conditioning** can reduce load on reboiler/condenser at the cost of an extra exchanger, hence the trade off between the extra exchanger and reduced load

should be studied. In the presented case study, by introducing a preheater, about 1342 kW of reboiler duty could be saved with a payback period of 3.8 days.

5. If the hot/cold utilities are available at suitable temperatures then installation of **side-reboiler/ side condensers** are good options for further reduction of load on reboiler and condenser respectively. In the presented case study, only cooling utility as cooling water was available at the lowest temperature, hence side-condenser did not yield any saving. But a hot utility at low temperature was available, so a side reboiler is used which reduced the reboiler duty by 3000 kW at the cost of an extra exchanger.

6. Equation based **Heat Exchanger optimization** technique which relies on Bell's Method helps to arrive at an optimum design targeted at minimum-area and –total annual cost of exchanger. Due to the high profits in column targeting, the benefits obtained from optimization appear to be negligible. Never the less, it reduces the overall payback period from 7.4 days to 6.5 days.

RECOMMENDATION

The Column Targeting technique studied in the present work involves a lot of human intervention. Software can be developed that generates all the possible options, so that more options can be tried.

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APPENDIX A
COLUMN TARGETING RESULTS

Table A1: Column Targeting Results for Column B1

Stage	Existing Case		Case 1		Case 2		Case 5	
	Temp. (C)	Enthalpy deficit (kW)	Temp. (C)	Enthalpy deficit (kW)	Temp. (C)	Enthalpy deficit (kW)	Temp. (C)	Enthalpy deficit (kW)
1	114.5	3524.1	114.4	3524.1	114.7	2630.0	114.8	3524.3
2	116.6	3203.4	116.6	3203.3	116.9	2309.4	117.1	3203.7
3	118.6	3035.3	118.6	3035.8	118.9	2142.4	119.2	3030.8
4	120.4	2919.9	120.3	2921.2	120.5	2027.9	120.9	2908.2
5	121.9	2820.9	121.9	2823.0	122.0	1932.1	122.4	2800.7
6	123.2	2723.6	123.2	2726.5	123.2	1842.9	123.7	2694.7
7	124.4	2623.1	124.3	2626.5	124.2	1756.9	124.8	2586.1
8	125.3	2519.5	125.3	2523.1	125.1	1675.0	125.7	2476.1
9	126.1	2416.1	126.1	2419.2	125.8	1599.0	126.5	2368.5
10	126.9	2317.0	126.9	2318.7	126.5	1530.9	127.1	2267.9
11	127.5	2226.3	127.5	2225.7	127.2	1471.7	127.7	2178.0
12	128.1	2146.8	128.2	2142.5	127.7	1421.0	128.3	2101.0
13	128.6	2079.7	128.7	2070.0	128.3	1377.7	128.8	2037.2
14	129.1	2024.7	129.3	2007.3	128.9	1339.6	129.3	1985.7
15	129.5	1980.6	129.9	1952.3	129.5	1304.2	129.7	1944.8
16	130.0	1945.7	130.5	1901.5	130.2	1268.2	130.2	1912.2
17	130.4	1918.1	131.2	1850.2	131.0	1227.7	130.6	1886.0
18	130.8	1895.7	132.0	1792.0	131.9	1176.7	131.0	1863.8
19	131.3	1876.9	133.0	1716.2	133.2	1105.8	131.5	1843.9
20	131.7	1860.0	134.4	1602.3	134.8	996.0	132.0	1824.2
21	132.2	1843.4	136.6	1398.7	137.4	799.0	132.5	1802.5
22	132.6	1825.1	140.6	926.6	141.9	349.7	133.1	1776.3
23	133.2	1802.8	150.1	924.6	152.0	76.5	133.8	1742.0
24	133.8	1773.4	150.8	1001.0	152.6	122.6	134.6	1694.3
25	134.6	1731.4	151.4	1067.6	153.1	165.3	135.6	1624.7
26	135.6	1667.7	152.0	1126.3	153.6	205.4	137.0	1517.7

27	137.0	1562.7	152.6	1178.6	154.0	243.9	139.0	1339.7
28	139.1	1365.9	153.1	1225.7	154.5	281.4	142.1	1001.0
29	143.1	901.1	153.6	1268.6	155.0	318.6	147.7	192.8
30	152.5	994.1	154.1	1308.1	155.4	356.0	156.8	73.4
31	153.2	1075.8	154.5	1344.9	155.8	393.9	157.2	107.5
32	153.8	1148.1	155.0	1379.6	156.3	432.7	157.7	141.7
33	154.4	1213.1	155.4	1413.0	156.7	472.8	158.1	175.7
34	154.9	1272.7	155.8	1445.7	157.1	514.3	158.5	210.2
35	155.5	1328.6	156.2	1478.4	157.6	557.6	158.9	246.5
36	156.0	1382.3	156.6	1512.0	158.0	603.1	159.3	284.2
37	156.4	1435.5	157.0	1547.4	158.4	651.2	159.7	324.7
38	156.9	1489.9	157.3	1586.7	158.8	702.5	160.1	368.1
39	157.3	1547.7	157.7	1630.9	159.3	758.0	160.5	415.6
40	157.7	1611.5	158.1	1682.8	159.7	818.8	160.9	468.6
41	158.1	1684.7	158.5	1745.6	160.1	887.0	161.3	528.0
42	158.6	1771.8	158.8	1823.6	160.5	965.2	161.7	596.1
43	159.0	1878.5	159.2	1922.2	160.9	1057.2	162.1	676.6
44	159.4	2011.9	159.6	2048.5	161.4	1168.2	162.5	772.7
45	159.9	2180.8	160.0	2210.8	161.8	1305.2	162.9	891.0
46	160.3	2394.4	160.5	2418.5	162.3	1477.3	163.4	1040.1
47	160.9	2661.5	161.0	2680.1	162.8	1695.0	163.9	1229.8
48	161.6	2987.1	161.7	3000.6	163.4	1968.7	164.4	1473.2
49	162.4	3367.3	162.5	3376.3	164.1	2304.8	165.1	1780.5
50	163.6	3782.2	163.7	3787.5	165.2	2697.5	166.0	2153.9
51	165.6	4180.7	165.6	4183.4	166.8	3111.8	167.5	2570.8
52	169.2	4439.2	169.3	4440.3	170.0	3438.2	170.4	2931.2
53	177.9	4281.4	177.9	4281.7	177.9	3385.8	177.8	2939.3

Table A2: Column Targeting Results for Column B2

Stage	Existing Case		Case 3		Case 4 & 8		Case 9	
	Temp. (C)	Enthalpy deficit (kW)	Temp. (C)	Enthalpy deficit (kW)	Temp. (C)	Enthalpy deficit (kW)	Temp. (C)	Enthalpy deficit (kW)
1	163.9	8055.9	163.9	8059.9	163.9	6932.8	163.8	6305.3
2	183.1	4350.4	183.1	4352.5	183.1	3225.4	183.1	2596.3
3	194.0	4180.1	194.0	4182.8	192.7	3014.9	191.7	2360.6
4	199.8	4153.5	199.8	4156.3	197.4	2933.8	195.6	2245.4
5	203.3	4109.0	203.3	4111.7	200.1	2854.1	197.8	2144.0
6	205.8	4043.0	205.8	4045.5	202.0	2768.9	199.3	2047.8
7	207.7	3957.9	207.7	3960.3	203.4	2677.4	200.5	1954.0
8	209.2	3855.1	209.2	3857.1	204.6	2579.3	201.4	1861.4
9	210.6	3735.3	210.6	3736.9	205.7	2475.3	202.3	1770.1
10	211.7	3600.6	211.7	3601.6	206.6	2367.2	203.0	1681.3
11	212.7	3454.0	212.8	3454.3	207.4	2257.6	203.7	1596.2
12	213.6	3299.9	213.6	3299.2	208.1	2149.4	204.3	1516.2
13	214.4	3143.6	214.4	3141.6	208.8	2045.5	204.8	1442.3
14	215.1	2990.5	215.1	2986.9	209.4	1948.2	205.4	1375.2
15	215.7	2845.5	215.8	2840.0	209.9	1859.2	205.8	1315.0
16	216.3	2712.5	216.3	2704.8	210.4	1779.4	206.3	1261.5
17	216.8	2593.9	216.9	2583.5	210.9	1709.0	206.7	1214.2
18	217.3	2490.6	217.4	2477.0	211.4	1647.4	207.1	1172.3
19	217.8	2402.3	217.8	2384.8	211.8	1593.7	207.5	1135.1
20	218.2	2328.0	218.3	2305.8	212.2	1546.7	207.9	1101.6
21	218.6	2265.8	218.7	2238.0	212.6	1505.3	208.3	1071.0
22	219.0	2214.1	219.6	1145.9	213.4	154.4	209.1	127.0
23	219.4	2171.0	236.3	2617.9	231.7	1679.7	227.8	1829.2
24	219.7	2134.7	243.5	3265.1	240.6	2435.8	237.8	2783.8
25	220.1	2103.8	246.8	3544.5	244.9	2790.0	243.0	3334.9
26	220.5	2077.0	248.5	3685.0	247.3	2976.4	246.0	3712.2
27	220.8	2053.0	249.7	3767.3	248.8	3087.8	248.0	4009.7
28	221.2	2031.0	250.5	3821.5	249.9	3161.2	249.4	4264.3
29	221.5	2010.1	251.2	3860.5	250.8	3213.0	250.4	4298.8
30	221.9	1125.0	251.8	3890.7	251.5	3251.6	251.3	4324.4
31	239.1	2620.0	252.4	3915.9	252.1	3282.1	252.0	4343.9

32	246.5	3292.6	252.9	3938.4	252.7	3307.4	252.6	4359.1
33	249.9	3590.5	253.3	3960.0	253.2	3329.7	253.2	4371.3
34	251.7	3746.7	253.8	3981.9	253.7	3350.5	253.7	4381.4
35	252.9	3845.2	254.2	4005.4	254.2	3371.3	254.3	4390.1
36	253.7	3917.6	254.6	4031.4	254.6	3392.9	254.7	4397.7
37	254.4	3977.9	255.1	4061.1	255.0	3416.6	255.2	4404.9
38	255.0	4033.5	255.5	4095.6	255.5	3443.1	255.6	4411.9
39	255.5	4089.0	255.9	4135.9	255.9	3473.6	256.1	4419.2
40	256.0	4147.4	256.3	4183.4	256.3	3509.2	256.5	4427.0
41	256.5	4211.5	256.7	4239.4	256.7	3550.9	256.9	4435.7
42	256.9	4283.3	257.0	4305.2	257.1	3599.9	257.3	4445.8
43	257.3	4364.9	257.4	4382.3	257.4	3657.6	257.7	4457.6
44	257.7	4458.0	257.8	4471.9	257.8	3725.1	258.1	4471.6
45	258.1	4564.0	258.2	4575.4	258.2	3803.6	258.4	4488.4
46	258.5	4684.1	258.6	4693.6	258.6	3894.2	258.8	4508.6
47	258.9	4819.1	259.0	4827.0	259.0	3997.9	259.2	4532.9
48	259.3	4968.9	259.3	4975.7	259.4	4114.9	259.6	4562.0
49	259.7	5132.9	259.7	5138.9	259.7	4245.4	260.0	4596.8
50	260.1	5309.7	260.1	5315.1	260.1	4388.7	260.3	4637.9
51	260.5	5497.1	260.5	5502.1	260.5	4543.4	260.7	4686.3
52	260.9	5692.2	260.9	5696.8	260.9	4707.6	261.1	4742.4
53	261.3	5891.4	261.3	5895.8	261.3	4878.6	261.5	4806.9
54	261.7	6091.1	261.7	6095.4	261.7	5053.3	261.9	4879.8
55	262.1	6287.7	262.1	6291.8	262.1	5228.6	262.3	4961.0
56	262.5	6477.9	262.5	6482.0	262.5	5401.3	262.7	5049.8
57	262.9	6659.2	262.9	6663.2	262.9	5568.7	263.1	5145.3
58	263.4	6829.8	263.4	6833.8	263.4	5728.7	263.5	5246.3
59	263.8	6989.1	263.8	6993.1	263.8	5880.3	264.0	5351.3
60	264.3	7137.5	264.3	7141.4	264.3	6023.2	264.4	5459.2
61	264.8	7276.2	264.8	7280.2	264.8	6158.3	264.9	5569.2
62	265.4	7407.9	265.4	7411.9	265.4	6287.6	265.4	5681.5
63	265.9	7536.5	265.9	7540.4	265.9	6414.7	266.0	5797.5
64	266.5	7667.2	266.5	7671.2	266.5	6544.6	266.5	5920.7
65	267.2	7807.9	267.2	7811.9	267.2	6684.9	267.2	6057.4

APPENDIX B

STREAM MOLE FRACTIONS

Table B1: Stream Mole Fractions for Existing Case

Pseudo Components	FEED	LIGHT	STREAM	HEART	HEAVY
1	0.018	0.181	nil	nil	nil
2	0.056	0.568	nil	nil	nil
3	0.133	0.251	0.120	0.126	nil
4	0.145	nil	0.160	0.168	nil
5	0.148	nil	0.164	0.171	nil
6	0.131	nil	0.145	0.152	nil
7	0.119	nil	0.132	0.138	nil
8	0.099	nil	0.110	0.115	nil
9	0.069	nil	0.077	0.080	nil
10	0.037	nil	0.041	0.043	nil
11	0.035	nil	0.038	0.008	0.735
12	0.010	nil	0.011	0.000	0.265

Table B2: Stream Mole Fractions after Case 2

Pseudo Components	FEED	LIGHT	STREAM	HEART	HEAVY
1	0.018	0.181	nil	nil	nil
2	0.056	0.546	0.003	0.003	nil
3	0.133	0.273	0.118	0.123	nil
4	0.145	nil	0.160	0.168	nil
5	0.148	nil	0.164	0.171	nil
6	0.131	nil	0.145	0.152	nil
7	0.119	nil	0.132	0.138	nil
8	0.099	nil	0.110	0.115	nil
9	0.069	nil	0.077	0.080	nil
10	0.037	nil	0.041	0.043	nil
11	0.035	nil	0.038	0.008	0.735
12	0.010	nil	0.011	nil	0.265

Table B3: Stream Mole Fractions after Case 4

Pseudo Components	FEED	LIGHT	STREAM	HEART	HEAVY
1	0.018	0.181	nil	nil	nil
2	0.056	0.546	0.003	0.003	nil
3	0.133	0.273	0.118	0.123	nil
4	0.145	nil	0.160	0.168	nil
5	0.148	nil	0.164	0.171	nil
6	0.131	nil	0.145	0.152	nil
7	0.119	nil	0.132	0.138	nil
8	0.099	nil	0.110	0.115	nil
9	0.069	nil	0.077	0.080	nil
10	0.037	nil	0.041	0.043	nil
11	0.035	nil	0.038	0.008	0.735
12	0.010	nil	0.011	nil	0.265

Table B4: Stream Mole Fractions after Case 8

Pseudo Components	FEED	LIGHT	STREAM	HEART	HEAVY
1	0.018	0.181	nil	nil	nil
2	0.056	0.529	0.005	0.005	nil
3	0.133	0.289	0.116	0.121	nil
4	0.145	nil	0.160	0.168	nil
5	0.148	nil	0.164	0.171	nil
6	0.131	nil	0.145	0.152	nil
7	0.119	nil	0.132	0.138	nil
8	0.099	nil	0.110	0.115	nil
9	0.069	nil	0.077	0.080	nil
10	0.037	nil	0.041	0.043	nil
11	0.035	nil	0.038	0.008	0.735
12	0.010	nil	0.011	nil	0.265

Table B5: Stream Mole Fractions after Case 9

Pseudo Components	FEED	LIGHT	STREAM	HEART	HEAVY
1	0.018	0.181	nil	nil	nil
2	0.056	0.529	0.005	0.005	nil
3	0.133	0.289	0.116	0.121	nil
4	0.145	nil	0.160	0.168	nil
5	0.148	nil	0.164	0.171	nil
6	0.131	nil	0.145	0.152	nil
7	0.119	nil	0.132	0.138	nil
8	0.099	nil	0.110	0.115	nil
9	0.069	nil	0.077	0.080	nil
10	0.037	nil	0.041	0.043	nil
11	0.035	nil	0.038	0.008	0.735
12	0.010	nil	0.011	nil	0.265

APPENDIX C

COMPUTER PROGRAM FOR OBTAINING FEASIBLE REGION ON PRESSURE DROP DIAGRAM AIMED AT OPTIMISING A SHELL AND TUBE HEAT EXCHANGER USING MATLAB

% Function data (data for problem under study)

```
Tt1=150;Tt2=190;Ts1=220;Ts2=200;
R=(Ts1-Ts2)/(Tt2-Tt1); P=(Tt2-Tt1)/(Ts1-Tt1);
F=(R*R+1)^0.5/(R-1)*log((1-P)/(1-P*R))/log((2/P-1-R+(R*R+1)^0.5)/(2/P-1-R-
(R*R+1)^0.5));
Mt=13.52;Ms=32.7105; Cpt=2492;Cps=2060; dent=772;dens=916;
vist=.000788;viss=.000417; kt=0.122;ks=0.12;
utmin=1;utmax=2; usmin=0.3;usmax=0.8; delptmax=80000;delpsmax=8000;
Rbsmin=0.2;Rbsmax=1; do=0.02;di=0.016; Lmax=6; Dsmax=1.5;
k=36;Rd=0.0001; Bc=0.25;thetab=2.1; Ct=0.0008;Cs=0.0048;
Pt=1.25*do; NS=1; NT=6; a=0.0444;m=0.028;
i=0;j=0; alfa=1.35;alfadash=4; vistw=vist; vissw=viss;
Pttype=1;%square pitch Pttype=1, triangular pitch Pttype =2
if Pttype==1, Ptdash=Pt;
else Pttype==2, Ptdash=0.877*Pt;
end
if Pttype==1
    if NT==1, K1=.215;n=2.207;
    elseif NT==2, K1=.156;n=2.291;
    elseif NT==4, K1=.158;n=2.263;
    elseif NT==6, K1=.0402;n=2.617;
    else NT==8, K1=.0331;n=2.643;
    end
else Pttype==2
    if NT==1, K1=.319;n=2.142;
    elseif NT==2, K1=.249;n=2.207;
    elseif NT==4, K1=.175;n=2.285;
    elseif NT==6, K1=.0743;n=2.499;
    else NT==8, K1=.0365;n=2.675;
    end
end
b=(m+1)/K1^(1/n); Prt=Cpt*vist/kt; Prs=Cps*viss/ks;
LMTD=((Ts2-Tt1)-(Ts1-Tt2))/log((Ts2-Tt1)/(Ts1-Tt2));
de=4*(Pt*Ptdash-pi/4*do*do)/(pi*do);
Q=Mt*Cpt*(Tt2-Tt1);
Kt=0.046; mt=-0.2; ac=0.14; Ks=0.4475; ms=-0.19;
C3=2*Ks*de^(ms-1)*NS*Ms^(2+ms)/(viss^ms*dens*(viss/vissw)^.14*(1-do/Pt)^(2+ms));
C4=0.36*ks/de*(Cps*viss/ks)^(1/3)*(viss/vissw)^0.14*(de/viss)^0.55*Ms^0.55/(1-
do/Pt)^0.55;
C5=Q*(b*do)^n/(pi*do*NS*F*LMTD);
C6=1.86*kt/do*(vist/vistw)^.14*(Cpt*vist/kt*4*Mt*NT/pi/vist)^(1/3)*(b*do)^(n/3);
C7=Rd+do*log(do/di)/(2*k);
C8=0.116*kt/do*(vist/vistw)^0.14*(Cpt*vist/kt)^(1/3);
```

```
C9=(4*Mt*NT)^(2/3)*(b*do)^(2*n/3)/(pi*di*vist)^(2/3);
C10=kt*0.023/do*(vist/vistw)^0.14*(Cpt*vist/kt)^(1/3)*(4*Mt*NT)^0.8*(b*do)^(0.8*
n)/(pi*di*vist)^0.8;
```

% Function after_Ds (Calculation after Ds)

```
Db=do*(Nt/K1)^(1/n); Hb=Db/2-Ds*(0.5-Bc);
Bb=Hb/Db; Radash=-0.005+0.55*Bb+Bb*Bb;
Nw=Nt*Radash;
Atb=Ct*pi*do*(Nt-Nw)/2; Asb=Cs*Ds*(2*pi-thetab)/2;
AL=Atb+Asb; Rw=2*Nw/Nt;
Fw= 1.003403 +3.8145548*Rw + -31.568113*Rw^2 + 111.56454*Rw^3 + -205.77385 *Rw^4
+ 188.43572  *Rw^5 + -67.651795 *Rw^6;
Ncv=(Db-2*Hb)/Ptdash; Ns=0.2*Ncv;
Fb=exp(-alfa*Pt*(Ds-Db)*(1-(2*Ns/Ncv)^(1/3))/((Pt-do)*Ds));
Fbdash=exp(-alfadash*Pt*(Ds-Db)*(1-(2*Ns/Ncv)^(1/3))/((Pt-do)*Ds));
Nwv=Hb/Ptdash;
```

% Function after_Res (Calculation after Res)

```
if Res>2000,
Fn = .88105+.0191251*Ncv-9.77*10^-4*Ncv^2+2.448*10^-5*Ncv^3-2.349*10^-7*Ncv^4;
else, Fn=1;
end
if Ptdash==1
      if Res>10000, mm=-.27203; cc=-0.0667;
      elseif Res<1000, mm=-.62108; cc=.68932;
      else, mm=.019023; cc=-1.231;
      end
else
      if Res>10000, mm=-.1054267; cc=-0.76538;
      elseif Res<1000, mm=-0.60206; cc=.80618;
      else, mm=-0.18708; cc=-.43874;
      end
end
jf=10^(mm*log10(Res)+cc);
```

% Function calc_FL (To calculate FL)

```
alas=AL/As;
if (alas<0.1), betaL= -8.6869*alas^2 + 2.1687*alas+ 0.02;
else betaL= 0.4444*alas + 0.1056;
end
if (alas<0.2), betaLdash= 63.333*alas^3- 29*alas^2 + 5.1417*alas + 0.0015;
else betaLdash = 0.5625*alas+ 0.2625;
end
FL=1-betaL*(Atb+2*Asb)/AL;
FLdash=1-betaLdash*(Atb+2*Asb)/AL;
```

% Function calc_delps (To calculate delps)

```
delptt(i)=delpt;
Nb=L/Lbc-1;
Ra=-0.0367493+0.748932*Bc+0.704894*Bc*Bc;
Aw=(pi*Ds^2*Ra/4)-(Nw*pi*do*do/4);
uw=Ms/Aw/dens; uz=(uw*us)^0.5;
delps(i)=delpi*Fbdash*(2*(Nwv+Ncv)/Ncv+FLdash*(Nb-1))+FLdash*(2+.6*Nwv)*dens*
uz^2*dens*uz^2*Nb/2;
```

% Program to plot delpt vs. delps for utmin/utmax

```
data;
constraint=input('Enter constraint 1 for utmin and 2 for utmax:')
if constraint==1, ut=utmin;
else ut==utmax;
end
Gt=dent*ut; Ret=di*ut*dent/vist;
if Ret<=2100, Kt=16;mt=-1;ac=0.35;
else Kt=0.046; mt=-0.2; ac=0.14;
end
C1=(2*Kt*NT^(3+mt)*NS*(4*Mt)^(2+mt)*(b*do)^(n*(2+mt)))/(dent*vist^mt*(vist/vistw
)^ac*pi^(2+mt)*di^(5+mt));
C2=(20*NT^3*NS*Mt^2*(b*do)^(n*2))/(dent*pi^2*di^4);
Nt=4*Mt*NT/(pi*Gt*di^2);
Ds=a+b*do*Nt^(1/n);
after_Ds;
for delpt = 10:10:90000
i=i+1;
Rbs=0.01;Rbs3=100;
while abs(Rbs-Rbs3)>0.001
    Rbs=Rbs+0.0001;
    Lbc=Rbs*Ds;
    Rbs=Lbc/Ds;
    As=(Pt-do)*Ds*Lbc/Pt;
    us=Ms/dens/As;
    Res=do*us*dens/viss;
    after_Res;
    delpi=8*jf*Ncv*dens/2*us^2*(viss/vissw)^0.14;
    calc_FL;
    L=(Ds-a)^(n*(2+mt))/C1*(delpt-C2/(Ds-a)^(2*n));
    if Ret<=2100, Rbs1=L^(1/3)*(Ds-a)^(n/3)/C6;
    elseif Ret>10000, Rbs1=(Ds-a)^(0.8*n)/C10;
    else Rbs1=1/(C8*(1+(di/L)^(2/3))*(C9/(Ds-a)^(2*n/3)-125));
    end
    Rbs3=(((L*(Ds-a)^n/C5)-C7-Rbs1)*Fn*Fb*Fw*FL*C4/Ds^1.2)^(1/0.6);
    if (Rbs>2), break, end
end
calc_delps;
end
plot(delptt,delps)
delpts=[delptt',delps'];
save delpts.xls delpts -ascii;
```

% Program to plot delpt vs. delps for Rbsmin/Rbsmax

```
data;
constraint=input('Enter constraint 1 for Rbsmin and 2 for Rbsmax:')
if constraint==1, Rbs=Rbsmin;
else Rbs==Rbsmax;
end
for Ds=0.1:0.001:Dsmax
i=i+1;
Lbc=Ds*Rbs;
Nt=((Ds-a)/b/do)^n;
At=pi/4*di*di*Nt/NT;
ut=Mt/dent/At;
Ret=di*ut*dent/vist;
if Ret<=2100, Kt=16;mt=-1;ac=0.35;
else Kt=0.046; mt=-0.2; ac=0.14;
end
C1=(2*Kt*NT^(3+mt)*NS*(4*Mt)^(2+mt)*(b*do)^(n*(2+mt)))/(dent*vist^mt*(vist/vistw
)^ac*pi^(2+mt)*di^(5+mt));
C2=(20*NT^3*NS*Mt^2*(b*do)^(n*2))/(dent*pi^2*di^4);
after_Ds;
As=(Pt-do)*Ds*Lbc/Pt;
us=Ms/dens/As;
Res=do*us*dens/viss;
after_Res;
delpi=8*jf*Ncv*dens*us^2*(viss/vissw)^0.14/2;
calc_FL;
if Ret<=2100, Rbs1=L^(1/3)*(Ds-a)^(n/3)/C6;
elseif Ret>10000, Rbs1=(Ds-a)^(0.8*n)/C10;
else Rbs1=1/(C8*(1+(di/L)^(2/3))*(C9/(Ds-a)^(2*n/3)-125));
end
L=C5/(Ds-a)^n*(Ds^1.2*Rbs^0.6/C4/(Fn*Fb*Fw*FL)+Rbs1+C7);
delpt=C1*L/(Ds-a)^(n*(2+mt))+C2/(Ds-a)^(2*n);
calc_delps;
end
plot(delptt,delps)
delpts=[delptt',delps'];
save delpts.xls delpts -ascii
```

```
% Program to plot delpt vs. delps for usmin/usmax

data;
constraint=input('Enter constraint 1 for usmin and 2 for usmax:')
if constraint==1, us=usmin;
else us==usmax;
end
Gs=dens*us;
Res=do*us*dens/viss;
As=Ms/dens/us;
for Ds=0.1:0.001:Dsmax
i=i+1;
Lbc=Ms/Gs/Ds/(1-do/Pt);
Rbs=Lbc/Ds;
Nt=((Ds-a)/b/do)^n;
At=pi/4*di*di*Nt/NT;
ut=Mt/dent/At;
Ret=di*ut*dent/vist;
if Ret<=2100, Kt=16;mt=-1;ac=0.35;
else Kt=0.046; mt=-0.2; ac=0.14;
end
C1=(2*Kt*NT^(3+mt)*NS*(4*Mt)^(2+mt)*(b*do)^(n*(2+mt)))/(dent*vist^mt*(vist/vistw
)^ac*pi^(2+mt)*di^(5+mt));
C2=(20*NT^3*NS*Mt^2*(b*do)^(n*2))/(dent*pi^2*di^4);
after_Ds;
after_Res;
delpi=8*jf*Ncv*dens*us^2*(viss/vissw)^0.14/2;
calc_FL;
if Ret<=2100, Rbs1=L^(1/3)*(Ds-a)^(n/3)/C6;
elseif Ret>10000, Rbs1=(Ds-a)^(0.8*n)/C10;
else Rbs1=1/(C8*(1+(di/L)^(2/3))*(C9/(Ds-a)^(2*n/3)-125));
end
L=C5/(Ds-a)^n*(Ds^1.2*Rbs^0.6/C4/(Fn*Fb*Fw*FL)+Rbs1+C7);
delpt=C1*L/(Ds-a)^(n*(2+mt))+C2/(Ds-a)^(2*n);
calc_delps;
end
plot(delptt,delps)
delpts=[delptt',delps'];
save delpts.xls delpts -ascii
```

% Program to plot delpt vs. delps for Dsmax

```
data;
Ds=Dsmax;
Nt=((Ds-a)/b/do)^n;
At=pi/4*di*di*Nt/NT;
ut=Mt/dent/At;
Ret=di*ut*dent/vist;
if Ret<=2100, Kt=16;mt=-1;ac=0.35;
else Kt=0.046; mt=-0.2; ac=0.14;
end
C1=(2*Kt*NT^(3+mt)*NS*(4*Mt)^(2+mt)*(b*do)^(n*(2+mt)))/(dent*vist^mt*(vist/vistw
)^ac*pi^(2+mt)*di^(5+mt));
C2=(20*NT^3*NS*Mt^2*(b*do)^(n*2))/(dent*pi^2*di^4);
after_Ds;
for delpt =10:10:delptmax
i=i+1;
Rbs=0.01;Rbs3=100;
while abs(Rbs-Rbs3)>0.001
Rbs=Rbs+0.0001;
Lbc=Rbs*Ds;
As=(Pt-do)*Ds*Lbc/Pt;
us=Ms/dens/As;
Res=do*us*dens/viss;
after_Res;
delpi=8*jf*Ncv*dens*us^2*(viss/vissw)^0.14/2;
calc_FL;
L=((Ds-a)^(n*(2+mt))/C1)*(delpt-C2/(Ds-a)^(2*n));
if Ret<=2100, Rbs1=L^(1/3)*(Ds-a)^(n/3)/C6;
elseif Ret>10000, Rbs1=(Ds-a)^(0.8*n)/C10;
else Rbs1=1/(C8*(1+(di/L)^(2/3))*(C9/(Ds-a)^(2*n/3)-125));
end
if (Rbs>2), break, end
end
calc_delps;
end
plot(delptt,delps)
delpts=[delptt',delps'];
save delpts.xls delpts -ascii
```

```
% Program to plot delpt vs. delps for Lmax

data;
L=Lmax;
for Ds=0.1:0.001:Dsmax
i=i+1;
Nt=((Ds-a)/b/do)^n;
At=pi/4*di*di*Nt/NT;
ut=Mt/dent/At;
Ret=di*ut*dent/vist;
if Ret<=2100, Kt=16;mt=-1;ac=0.35;
else Kt=0.046; mt=-0.2; ac=0.14;
end
C1=(2*Kt*NT^(3+mt)*NS*(4*Mt)^(2+mt)*(b*do)^(n*(2+mt)))/(dent*vist^mt*(vist/vistw
)^ac*pi^(2+mt)*di^(5+mt));
C2=(20*NT^3*NS*Mt^2*(b*do)^(n*2))/(dent*pi^2*di^4);
after_Ds;
Rbs=0.01;Rbs3=100;j=0;
while abs(Rbs-Rbs3)>0.001
Rbs=Rbs+0.0001;
Lbc=Rbs*Ds;
As=(Pt-do)*Ds*Lbc/Pt;
us=Ms/dens/As;
Res=do*us*dens/viss;
after_Res;
delpi=8*jf*Ncv*dens*us^2/2*(viss/vissw)^-0.14;
calc_FL;
if Ret<=2100, Rbs1=L^(1/3)*(Ds-a)^(n/3)/C6;
elseif Ret>10000, Rbs1=(Ds-a)^(0.8*n)/C10;
else Rbs1=1/(C8*(1+(di/L)^(2/3))*(C9/(Ds-a)^(2*n/3)-125));
Rbs3=(((L*(Ds-a)^n/C5)-C7-Rbs1)*Fn*Fb*Fw*FL*C4/Ds^1.1)^(1/0.55);
if (Rbs>2), break, end
end
calc_delps;
end
plot(delptt,delps)
delpts=[delptt',delps'];
save delpts.xls delpts -ascii
```

Table C1: Output values for constraints u_{tmin}, u_{tmax}, R_{bsmin} and R_{bsmax}

u_{tmin}		u_{tmax}		R_{bsmin}		R_{bsmax}	
ΔP_t	ΔP_s	ΔP_t	ΔP_s	ΔP_t	ΔP_s	ΔP_t	ΔP_s
18625	11763	88040	12513	93480	1662	82910	4427
18640	11762	88050	11519	84187	5043	75249	4109
18654	11761	88060	11518	75986	7787	68413	3819
18674	11392	88070	11517	68729	9999	62302	3553
18692	11004	88080	11515	62291	11770	56827	3311
18709	11001	88090	11514	56566	13171	51914	3089
18728	10591	88100	11513	51463	14264	47497	2885
18748	10160	88110	11512	46904	15100	43519	2697
18764	10156	88120	11511	42823	15722	39931	2525
18784	9701	88130	11510	39161	16165	36688	2366
18804	9695	88140	11508	35870	16460	33753	2219
18824	9216	88150	11507	32905	16630	31093	2084
18848	9207	88160	11506	30231	16698	28678	1958
18875	8699	88170	11505	27813	16681	26483	1842
18896	8167	88180	11504	25625	16594	24484	1735
18917	8157	88190	10470	23639	16449	22662	1635
18950	7591	88200	10468	21836	16259	20999	1543
18978	6998	88210	10467	20196	16030	19479	1456
19001	6985	88250	10461	18701	15772	18088	1376
19032	6358	88260	10460	17337	15490	16813	1301
19053	5707	88290	10456	16091	15190	15644	1232
19074	5692	88300	10455	14951	14876	14570	1167
19100	5003	88310	10453	13906	14552	13582	1106
19127	4981	88400	9365	12948	14221	12673	1049
19151	4258	88490	8233	12068	13887	10350	901
19185	4229	88790	7014			7992	744
19218	3459	89110	4472			7288	684
19251	2650	89300	1751			6854	654
19275	2626	89460	1705			6451	625
19316	1766	89560	257			6077	599
19380	836	89650	229			5729	574

Table C2: Output values for constraints u_{smin}, u_{smax}, L_{max} and D_{smax}

u_{smin}		u_{smax}		L_{max}		D_{smax}	
ΔP_t	ΔP_s	ΔP_t	ΔP_s	ΔP_t	ΔP_s	ΔP_t	ΔP_s
81102	2365	81969	13617	41093	2084	5100	9158
73136	2309	81128	13595	38678	1958	5110	9305
66082	2258	80296	13573	36483	1842	5120	9467
59822	2210	79475	13551	34484	1735	5130	9556
54256	2167	78665	13529	32662	1635	5140	9634
49295	2127	77864	13507	30999	1543	5150	9699
44866	2090	77072	13485	29479	1456	5160	9725
40902	2056	76291	13464	28088	1376	5170	9736
37349	2025	75518	13442	26813	1301	5180	9748
34159	1996	74756	13421	25644	1232	5190	9759
31288	1969	74002	13400	24570	1167	5200	9771
28701	1944	73257	13379	23582	1106	5210	9783
26365	1921	72522	13358	22673	1049	5220	9794
24253	1900	71795	13337	21836	996	5230	9806
22340	1880	71077	13317	21063	947	5240	2452
20605	1862	70368	13296	20350	901	5250	2355
19028	1844	56170	12858	19691	858	5260	2255
17594	1828	44863	12459	19081	818	5270	2169
16287	1812	34639	12040	18516	780	5280	2088
15095	1797	28718	11757	17992	744	5290	2011
14006	1783	23404	11463	17288	684	5300	1930
13010	1769	18509	11140	16854	654	5310	1861
12098	1755	10784	10398	16451	625	5320	1796
11261	1741	7432	9828	16077	599	5420	1290
4975	1629	6152	9278	15729	574	5630	740
2439	1672	5566	9116			5860	446
1591	1734	5019	8935			6060	304
1133	1805					6240	222
836	1888					6390	171
660	1967					6500	141
584	2015						

APPENDIX D

COST OF HEAT EXCHANGER

Capital Cost of heat exchanger (Rs.) = 40(3000+ 750$A^{0.81}$) (Shenoy, 1995)

Captital Cost Estimation for problem under study

Assuming overall heat transfer coefficient (U) as 170 W/(m^2 K) (Kern, 1950)

For Feed Preheater

Hot fluid in Temperature (^{0}C) = 220

Hot fluid out Temperature (^{0}C) = 200

Cold fluid in Temperature (^{0}C) = 150

Cold fluid out Temperature (^{0}C) = 190

For feed preheater, Q = 1342.05 kW

LMTD = 39.15 ^{0}C

A = Q/(U (LMTD)) = 202 m^2

Capital cost of feed preheater = Rs. 23,30,324.

For Side Reboiler

Hot fluid in Temperature (^{0}C) = 250

Hot fluid out Temperature (^{0}C) = 230

Cold fluid in Temperature (^{0}C) = 210

Cold fluid out Temperature (^{0}C) = 230

For side reboiler, Q = 3000 kW

LMTD = 20 ^{0}C

A = Q/(U (LMTD)) = 882 m^2

Capital cost of side reboiler = Rs. 74,13,749

APPENDIX E

PROCESS FLUID AND UTILITY DATA

Table E1: Physical properties of process fluid and utilities: (Aspen Plus Database)

Fluids	Temp. (^0C)	Specific heat (kJ/(kg K)	Viscosity (cp)	Thermal Conductivity (W/(m K))	Density (kg/m^3)
Process Fluid	170	2.492	0.788	0.122	772
***Hot Oil-1**	285	2.317	0.226	0.109	824
****Hot Oil-2**	240	2198	0.296	0.114	869
*****Hot Oil-3**	210	2.060	0.417	0.12	916
Cooling water	30	4.192	0.988	0.60	998

*Hot Oil-1: Hot oil available at 295^{0}C

**Hot Oil-2: Hot oil available at 250^{0}C

**Hot Oil-3: Hot oil available at 220^{0}C

Table E2: Utility cost data

Fluids	Cost (Rs/kg)
Hot Oil-1	0.22
Hot Oil-2	0.125
Hot Oil-3	0.06
Cooling water	0.0033