CHAPTER II
LITERATURE REVIEW

2.1 Heat Exchanger Network Design Methodologies

The world's energy price has been increasing since 1970s and it has been considered one of important issues especially in chemical engineering processes. Before that period of time, most chemical plants mainly focused on capital investment cost and tried to keep the cost as low as possible. When energy crisis took place, many efforts have been made to reduce the energy consumption in the plants to reduce operational cost. There are many methods to improve energy efficiency in processes such as adjusting operating condition, changing or modifying equipment, using other chemicals, etc. One of commonly used techniques is installation of heat exchanger network (HEN) which can recover excess energy from heat source to heat sink.

Chemical plants generally involve with thermal processes because heat is a critical factor. For instance, it helps in catalyzing reactions or being used to heat up the products in separation unit corresponding to the boiling point. In HEN synthesis, streams are divided into two major types which are hot streams that need to be cooled and cold streams that need to be heated. A heat exchanger will match at least one hot stream and one cold stream in order to transfer excess heat from the hot stream (heat source) to the cold stream (heat sink). Thus, HEN is defined as a network that consists of many heat exchangers integrated in a system in order that all target temperatures are satisfied. The configuration can be generated in various possibilities such as heat exchanger in series, parallel, stream bypassing, stream splitting, etc. HEN will satisfy target temperatures of every stream incorporating with hot and/or cold utility.

The early methods for synthesizing HEN were based on thermodynamic principles and heuristic methods and then were developed to be more systematic with the implementation of mathematical computer programming (Verheyen and Zhang, 2006). To explain the evolution of HEN synthesis, the proposed methodologies will be categorized into three major groups which will be reviewed and compared in the next sections.

2.1.1 Pinch Analysis Concept

Pinch analysis was first introduced by Hohmann (1971) and then was refined and published by Linnhoff and Flower (1978). It is used for preliminary prediction of maximum energy recovery (or minimum utility required) and minimum number of heat exchangers corresponding to streams data (temperatures and heat capacity flowrates) and a chosen minimum temperature difference (ΔT_{min}). In HEN research area, pinch analysis is widely known as a thermodynamic analysis applying first and second laws of thermodynamics (Verheyen and Zhang, 2006).

A diagram called composite curve is used as a representation of all stream data. An example is shown in Figure 2.1. The upper and lower lines are hot and cold composite curves, respectively. The overlapping region represents the amount of energy recovered within the process. The non-overlapping region on both sides represents minimum hot (right) and cold utility (left) requirement at a chosen ΔT_{min}.

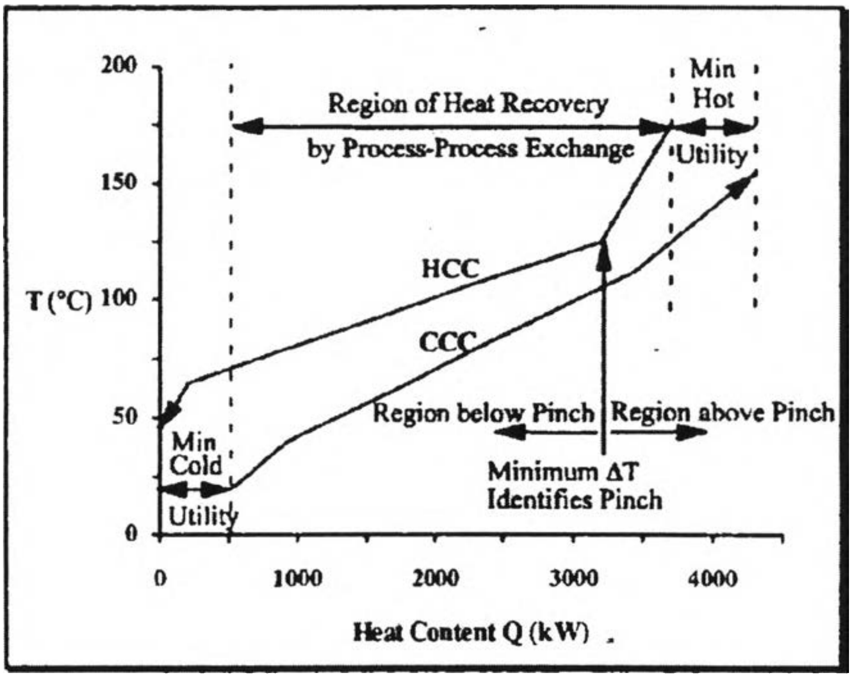

Figure 2.1 Hot and cold composite curves (Shenoy, 1995).

For a given ΔT_{min}, pinch is located on the diagram where the distance between hot and cold composite curves is narrowest and the temperature difference between two curves is equal to the ΔT_{min}. The composite curve is separated into two subsystems: above pinch and below pinch. The HEN design of each subsystem must be done separately starting from pinch location because there are more designing rules and constraints. Once we finish creating the network at pinch, the constraints are more relaxed and there is more flexibility to place a match depending on one's judgment.

Three pivotal rules of pinch analysis are described following:

- Heat transfer is not allowed across pinch.
- Each heat exchanger must have temperature difference larger than ΔT_{min}.
- Hot and cold utility are placed only at the end of the streams below and above pinch respectively in case target temperature is not reached.

As pinch design method creates a structure of HEN, some heat exchanger might have too large area, too small area, or higher number of exchangers than predicted one. Loop and path technique can be applied for energy relaxation in order to change heat exchanger area or remove some heat exchangers in case of unsatisfaction and high degree of complexity. However, it will result in increment of utility usage.

In reality, the objective of HEN synthesis is to minimize total annualized cost (TAC) which is the sum of utility cost per year and annualized capital (area and equipment) cost. If ΔT_{min} decreases (more energy recovery), the minimum utilities required will also decrease whereas the heat exchanger area needed is increased. Therefore, a trade-off between utility cost and area cost should be taken into account in order to find the optimized ΔT_{min} as shown in Figure 2.2. This is so-called supertargetting.

Pinch analysis is a concept that is easy to comprehend because it is a graphical method. Many sequential methods take advantage of this method for obtaining minimum utility and minimum number of units. The method, however, has some critical drawbacks. It does not deliberate heat transfer coefficients and heat exchanger areas properly; therefore, it might lead to ineffective solutions somehow (Verheyen and Zhang, 2006).

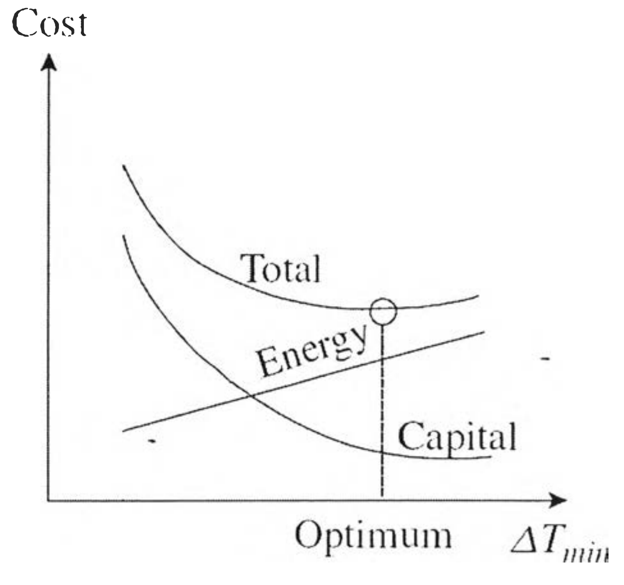

Figure 2.2 Economic trade-off between energy cost and capital cost (Shenoy, 1995).

2.1.2 <u>Sequential Approaches</u>

When mathematical programming initially caught considerable attention from researchers, the existing computers did not have adequate performance; in addition, optimization techniques had not been developed well enough. The procedures, as a result, were decomposed into several subproblems and then solved step by step. This method is known as sequential approach. In general, HENs optimization is mostly decomposed to these three subproblems (Biegler *et al.*, 1997):

- Minimum utility cost
- Minimum number of units
- Minimum total investment cost

Papoulias and Grossmann (1983) proposed a transshipment model. There are two main steps. The first step is linear programming (LP) problem which subjects to minimum utility cost. Second, the number of units is minimized using mixed integer linear programming (MILP). The concept of transshipment model is to distribute heat source (hot streams and hot utilities) to heat sink (cold streams and cold utilities). All streams are divided into temperature intervals. From Figure 2.3, each temperature interval (warehouse) will receive heat from higher temperature interval or hot utilities and then distribute to every cold stream in the same interval. The heat remained will be cascaded to lower temperature interval. Because the amount of heat source, heat sink, and temperatures are fixed, heat residuals which are passed to the next interval is the only one design variable to be optimized by LP model.

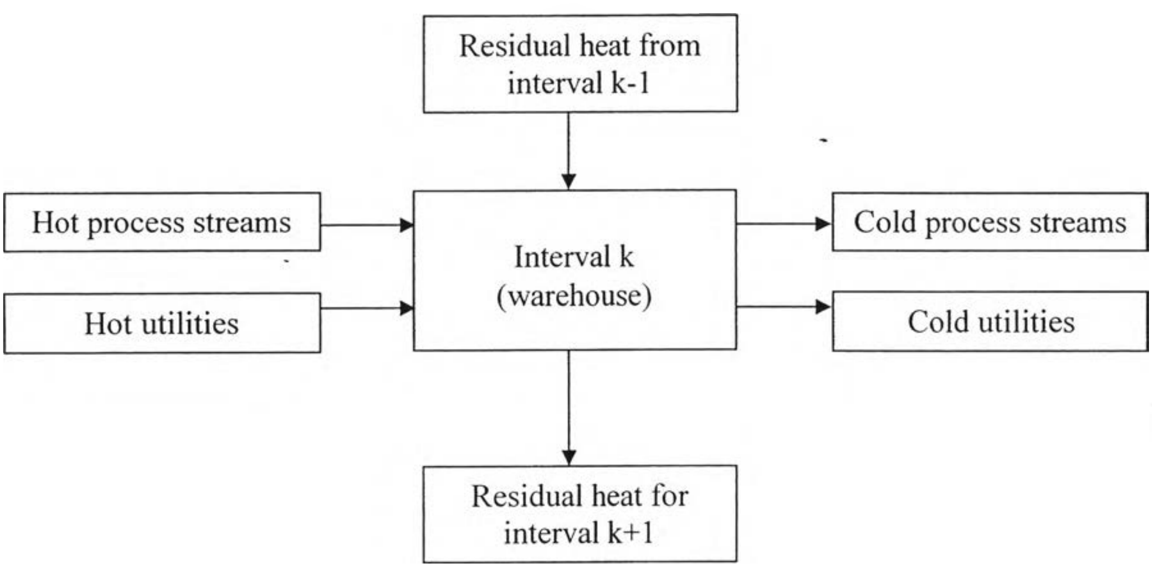

Figure 2.3 Heat flows in interval k (Biegler *et al.*, 1997).

For the MILP of transshipment model which is subject to minimum number of units, the formulation is analogous to LP model except that binary integer is used and denoted as the existence of a heat exchanger.

The extension of the transshipment model was addressed by Floudas *et al.* (1986). The HEN configurations will be generated automatically by using computer programming. Network synthesis is derived starting with LP transshipment model to predict minimum utility cost and pinch point will divide temperature range into subnetworks. Then the fewest number of exchangers is minimized by MILP transshipment model. The solution also provides the amount of heat exchanged in each match. Next step, a superstructure is derived for each subnetwork corresponding to those matches which are predicted by MILP transshipment model. The superstructure embeds various alternative configurations such as stream splitting, bypassing, matches in series, matches in parallel, matches in series-parallel, matches in parallel-series, etc. The superstructure is modified by NLP formulation featuring minimum investment cost. Note that the heat loads are treated as fixed parameter but flowrates and temperatures are design variables. However, as this methodology is a sequential method, the solutions of HENs might have been led to sub-optimal solutions because some good feasible solutions might not be included in search space of another subproblem.

A new decomposition method was introduced by Zhu (1995) for the purpose of automated synthesis of HENs using block decomposition and heuristic rules. The concept is to simplify a problem by decomposing composite curve into a number of blocks. In each block, two straight lines (for hot and cold composite curves) represent an enthalpy interval. Those two lines are called quasi-composite curves. Then, the design is performed using area targeting and newly heuristic rules for match selection. The final design with cost optimization is obtained via NLP model.

2.1.3 Simultaneous Approaches

As time progressed, a number of sequential approaches had been proposed, many researchers attempted to solve those decomposed problems simultaneously accompanying with modern computer technology and higher performance of optimization techniques. Since the simultaneous techniques consider all trade-offs, i.e. area targeting, number-of-unit targeting, and utility cost, their formulations are mostly MINLP model. They usually give more preferable results more than sequential methods, but they have more complexity which is the major problem of these approaches. In other words, they are mostly nonlinear, non-convex, and non-continuous that can mislead to locally optimal solutions. Therefore, some assumptions should be made in order to reduce the complexity and size of models.

Ciric and Floudas (1991) mentioned that the decomposition approach can lead to uncertainty of the optimality of final HEN. That means globally optimal solution is not guaranteed and most cases showed that the solutions tended to be local optima. From those reasons, they proposed a HEN synthesis method using MINLP simultaneous technique. The model includes hyperstructure of Floudas and Ciric (1989) and modified transshipment model by Papoulias and Grossmann (1983). Many different configurations of HEN are embedded in the hyperstructure as shown in Figure 2.4. The authors illustrated two cases of HEN synthesis which began with different assumptions so that the designs are differentiated by pinch point. One is strict-pinch design, another one is pseudo-pinch design. The comparison between two designs demonstrated that a pseudo-pinch approach which allows heat to transfer across pinch point leads to more desirable HENs.

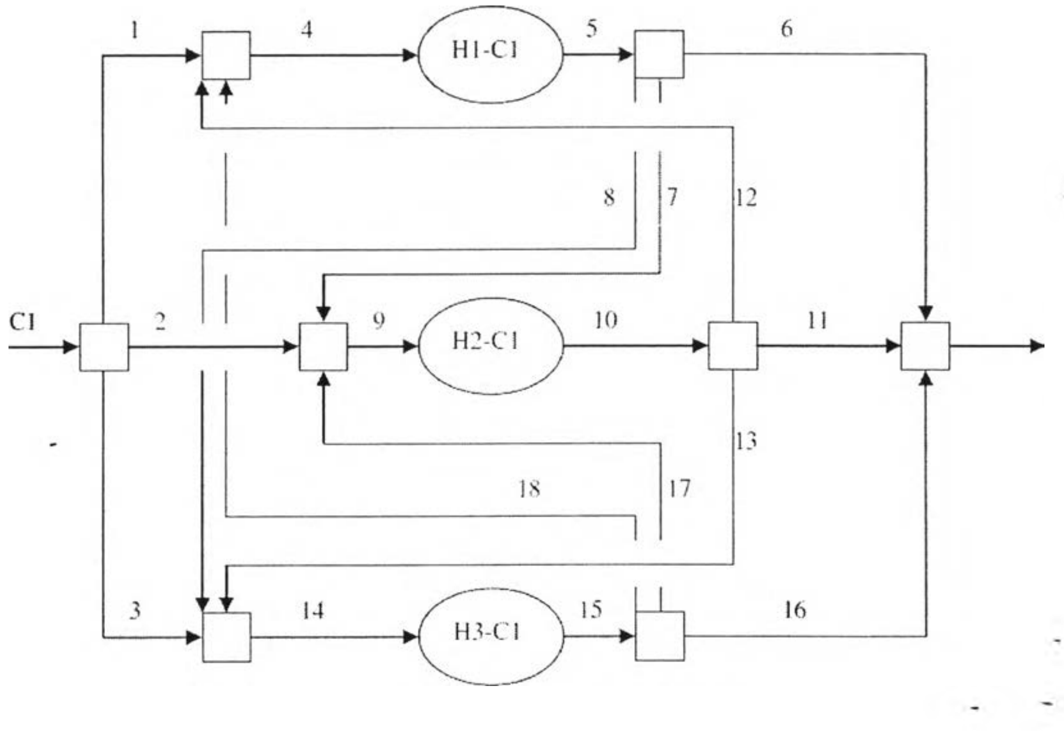

Figure 2.4 Hyperstructure of Floudas and Ciric (1989) (Verheyen and Zhang, 2006).

In the meantime, a stage-wise simplified superstructure was developed by Yee *et al.* (1990). The schematic is shown in Figure 2.5 but its detailed formulations will be written in the next section. Unlike transshipment model, the superstructure neither relies on pinch design method nor division into temperature intervals. For the superstructure, all constraints will be linear which results in rigorous model since the model was simplified by making following assumptions:

- Isothermal mixing; temperatures in location k must be equal in every split stream and mixed together before entering to the next stage.
- no split stream flowing through more than one heat exchanger
- hot and cold utilities are placed at the end of stream
- no stream bypass

The idea of the simplified superstructure model is to partition the structure into a number of stages. The illustration of two hot and two cold streams system is shown in Figure 2.5. In each stage, hot streams are split up into a number of cold streams and cold streams are similarly split into a number of hot streams. Every hot and cold split stream will be paired in all possible matches. One match represents a heat exchanger which is expressed by a binary integer and a set of matches will be chosen during optimization.

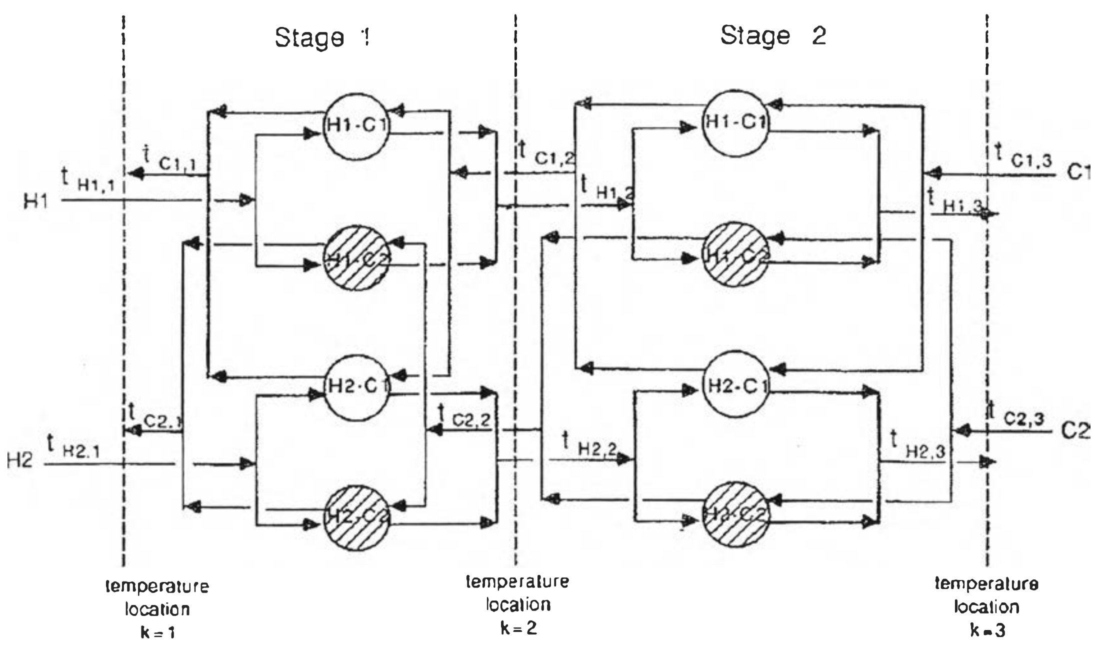

Figure 2.5 Stage-wise superstructure model (Yee *et al.*, 1990).

As the number of stages is concerned, it can be arbitrarily selected by designers; however, there is a rule of thumb stating that it is commonly chosen to be equal to either maximum number of hot and cold streams. Anyway, increasing of the number of stages may give better value of objective function or it may cause no effect on that.

Floudas (1995) had observed some weaknesses of superstructure. He pointed out that there are several network configurations which are excluded from the model. As can be seen in Figure 2.6, one branch cannot have more than one heat exchanger in series. Moreover, bypassing from one branch to another is not allowed in a stage and the combination of those two features is not available either.

In spite of its limitations, an important strength of the simplified superstructure is that all equations and constraints are linear except the objective function which is nonlinear due to the area calculation terms. As a result, the model was later extended in a number of studies.

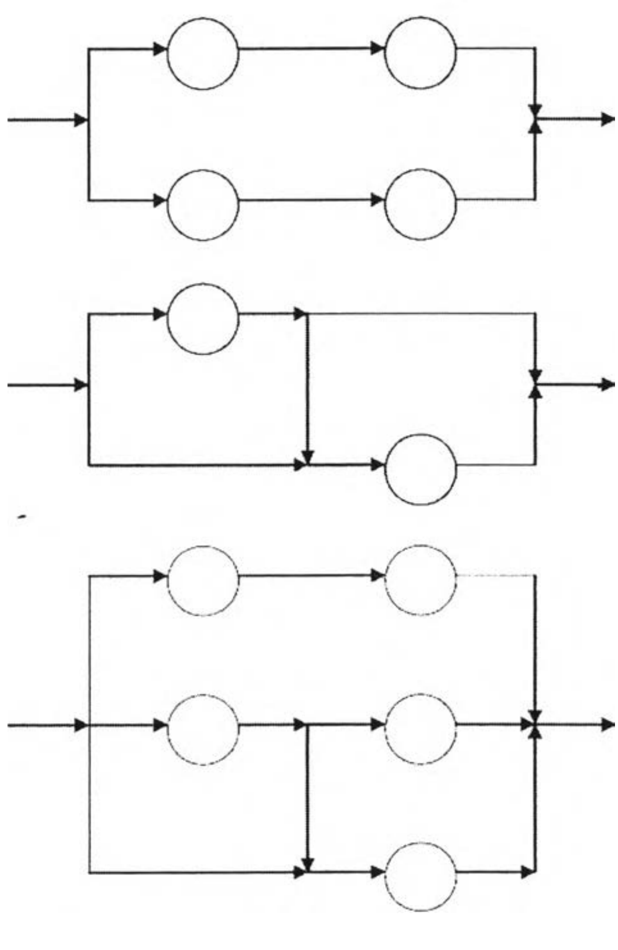

Figure 2.6 Excluded HEN configuration of simplified superstructure (Verheyen and Zhang, 2006).

One of the extensions of Synheat (another name of stage-wise superstructure) model was developed by Björk and Westerlund (2002). The main purposes of this work are elimination of the isothermal mixing assumption, which leads to a significant increase in number of variables and constants, together with using global optimization technique. Rather than using branching procedure, the strategy of this technique is try to convexify the non-convex terms in area equations so that when all equations in the problem are convex, global optimization can be obtained.

2.2 Multiperiod Heat Exchanger Network

In the past, most HENs were synthesized for only a fixed condition without taking into account the changes in parameters. Such HENs might cause deviation from optimal solution, or even no longer be operable or stay out of feasible region. The possible changes in operating conditions arise from two major reasons. First, uncertainties or unintentional changes around one nominal value, this kind of change is known as a resilient problem such as malfunction of process control system. Second, periodic changes can be either seasonal changes or intentional changes such as multiple feeds and a need for higher temperature in operating condition due to deactivation of catalyst.

2.2.1 Degree of Flexibility

As multiperiod HENs had been studied since 1980s, some efforts to create an index indicating flexibility of such HEN designs were introduced. For example, Saboo *et al.* (1985) proposed the resilient index (RI). It is a quantitative index which is used to compare between different HENs and guide the most potential candidate. At the same time, a flexibility index was proposed by Swaney and Grossmann (1985). The flexibility index indicates maximum deviation of uncertain variable and also lying in the operable or feasible region. Moreover, it provides information of critical points that restrict the design. From Figure 2.7, the point at the center of feasible region denotes as the nominal values of uncertain parameters. Each rectangular represents the maximum deviation of each parameters (θ_1, θ_2) while remaining in the feasible region (R).

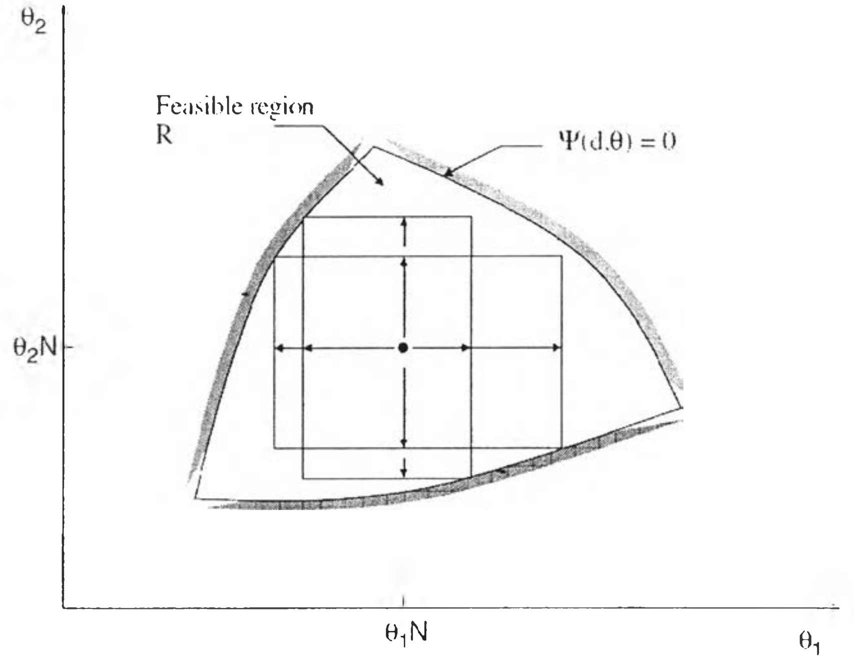

Figure 2.7 Feasible region of operation (Verheyen and Zhang, 2006).

2.2.2 Review of Multiperiod HEN Designs

There has been an increased interest in multiperiod features of chemical process designs including HENs since 1980s. One of renowned efforts was made by Floudas and Grossmann (1986). They took advantage of their work dealing with fixed conditions by using the same principles to apply to multiperiod problems. To recall the concept, LP transshipment model will be solved to predict minimum utility requirement and then MILP transshipment model is used to synthesize HEN configuration for each period featuring minimum utility cost and fewest number of units. The final network that satisfies all operational periods is obtained by assembling each topologies of each period manually. However, there are two major drawbacks for this method: sizing heat exchangers and bypassing around heat exchanger are not performed.

The model discussed above was improved by Floudas and Grossmann (1987) to overcome those limitations. They adopted the strategy proposed by Floudas *et al.* (1986), i.e. LP/MILP transshipment model is carried out, then the superstructure based on topology from transshipment model is derived and NLP formulations is run to improve the design afterwards. In case of multiperiod design, the only one difference is that all procedures have to be done for each period and the final solution is generated by integration of each subnetwork. In addition, when solving an NLP model, there are actually a lot of variables and constraints because all possible interconnections for the matches are taken into account. Hence, graph representation was introduced in order to reduce the problem size.

Iyer and Grossmann (1996) proposed an NLP model to find multiperiod HENs design with initially fixed configuration. The algorithm for global optimization by Quesada and Grossmann (1993) was used. Briefly, the objective function of the model is discrete function due to Max operation; thus, it is not guaranteed that the solution will be global optimum. To solve the problem, some constraints must be added using concept of convex underestimators to alter from non-convex objective function to the convex one.

A simultaneous MINLP model was developed by Aaltola (2002) based on superstructure of Yee and Grossmann (1990) which does not rely on pinch point. The objective function includes utility cost, area cost, and capital cost of units where assumption of average area of all periods is ruled to maintain linearity of the objective function. LP/NLP search algorithm is applied in the next step to improve the system. There are four main purposes of this step: trade-offs between utility and area cost, eliminating bypass streams which increase complexity of the network, removing the impractically average area assumption, and eliminating an isothermal mixing assumption.

From the model proposed by Aaltola (2002), Verheyen and Zhang (2006) observed its weaknesses and presented a new improved model. The new model comprises of a simultaneous MINLP model with maximum area formulation in the objective function and an improved NLP model in which slack variables and weighed parameter are not included.

Chen and Hung (2004) also studied simultaneous synthesis of multiperiod HENs and adopted the flexibility test using flexibility index (Swaney and Grossmann, 1985) to examine the solution whether it is qualified or not. The MINLP model based on Yee and Grossmann (1990) and extended model by Aaltola (2002) were applied using finite number of extreme operating conditions, i.e. the conditions that have tendency to demand the heat exchanger area as large as possible. The network that can be used for those operating conditions is tested by flexibility index to check ability of full-range operations. If the network is qualified, the synthesis procedure will be terminated. But if the network is unqualified, one more iteration has to be performed while some constraints are added to prevent from attaining the same disqualified network or, in other words, to reduce search space.

Recently Ma *et al.* (2008) pointed out that the models of Aaltola (2002) and Verheyen and Zhang (2006) have difficulty when solving more complex problems such as increase in number of periods. Such models give over-synthesized networks for all operational periods. Two-stage method was introduced for multiperiod operation. In the first stage, temperature-enthalpy (T-H) diagram is used to synthesize an over-synthesized HEN based on the stream pseudo-temperature. Compared with MINLP superstructure model, this method has less complexity and smaller size. Moreover, it can guarantee the feasibility of the initial solution to be used in the second stage. In the second stage, the over-synthesized HEN is improved. An area which is less than maximum area from the initial solution will be optimized. The main idea is that instead of using the maximum areas of each period, the optimal areas which are not satisfied in some periods, but the insufficient required area will be compensated by utilities. Genetic/simulated annealing (GA/SA) algorithm is also applied to guarantee global optimization at high probability.

2.2.3 Simultaneous MINLP Model

In this work, a model based on stage-wise simplified superstructure by Yee and Grossmann (1990) and also the extension to multiperiod version of Verheyen and Zhang (2006) will be applied because it is a rigorous model without decomposition and can provide good results (Verheyen and Zhang, 2006). The overall concept of the simplified superstructure has been explained in previous section. Isothermal mixing assumption will be applied initially because it significantly helps decrease a number of variables such as temperatures and heat capacity flowrates of outlet each heat exchanger in each branch. A single period model actually resembles a multiperiod model except that one more index, p, referring to period of operation is introduced.

2.2.3.1 Area Calculations

The heat transfer area for both process-process heat exchanger and utility-process heat exchanger can be calculated using these following information: heat load of each match, inlet and outlet temperatures of both hot and cold streams, and overall heat transfer coefficient for each match. The variables are illustrated in Figure 2.8. Area calculation is performed using Eq. 2.1.

$$A(i,j,k,p) = \frac{q(i,j,k,p)}{LMTD(i,j,k,p).U(i,j)} \tag{2.1}$$

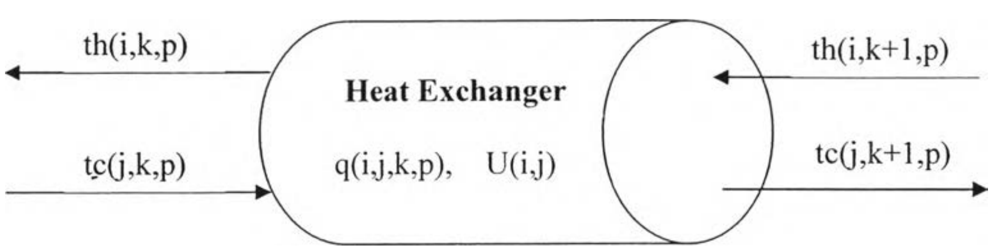

Figure 2.8 Illustration of variables involving in a heat exchanger.

The log mean temperature difference is calculated as Eq. 2.2:

$$LMTD(i,j,k,p) = \frac{[th(i,k,p) - tc(j,k,p)] - [th(i,k+1,p) - tc(j,k+1,p)]}{\ln \dfrac{[th(i,k,p) - tc(j,k,p)]}{[th(i,k+1,p) - tc(j,k+1,p)]}} \tag{2.2}$$

This formulation is generally used when hand calculation is performed. However, the difficulties probably come up in case of using mathematical programming. The reason is that there might be one or more set of variables in search space in which the temperature difference on both sides of heat exchanger are the same values. In consequence, the division by zero value will take place and cause the set of variables cannot be used due to undefined solution, but actually it is operable in realistic. For this reason, several approximations for LMTD were proposed as follows:

- Average LMTD
- Paterson approximation
- Chen approximation

The average LMTD is a simple one. The temperature differences of both sides are just averaged as shown in Eq. 2.3:

$$LMTD(i,j,k,p) = \frac{[th(i,k,p) - tc(j,k,p)] + [th(i,k+1,p) - tc(j,k+1,p)]}{2} \tag{2.3}$$

This approximation is very rough; therefore, it results in large error when comparing with the actual value of LMTD. The error will be amplified as the temperature differences of both sides are not in the same order of magnitude. Then the predicted area will be under-estimate.

The second approximation was proposed by Paterson (1984) as shown in Eq. 2.4:

$$\begin{aligned} LMTD(i,j,k,p) \\ &= \frac{2}{3}.\left[(th(i,k,p) - tc(j,k,p)).(th(i,k+1,p) - tc(j,k+1,p))\right]^{0.5} \\ &+ \frac{1}{6}\left[(th(i,k,p) - tc(j,k,p)) + (th(i,k+1,p) - tc(j,k+1,p))\right] \end{aligned} \tag{2.4}$$

This approximation gives slightly over-estimate LMTD or under-estimate heat transfer area.

Lastly, the approximation was introduced by Chen (1987) as formulated in Eq. 2.5:

$$LMTD(i,j,k,p) = \left[(th(i,k,p) - tc(i,k,p)).(th(i,k+1,p) - tc(j,k+1,p)).\frac{(th(i,k,p) - tc(j,k,p)) + (th(i,k+1,p) - tc(j,k+1,p))}{2} \right]^{1/3} \quad (2.5)$$

The LMTD value of Chen approximation tends to be underestimated or gives underestimated area. A good point of this approximation is that in case the temperature differences on both sides of heat exchanger are zero, the approximation equation will give a zero value. The accuracy of each proposed approximation is illustrated compared to actual LMTD value in Figure 2.9. It demonstrates that the average LMTD greatly deviates from the actual value of LMTD while Paterson's and Chen's approximations are much more accurate. In this work, Chen's approximation is selected because not only it predicts LMTD precisely, but also the over-estimated heat exchanger area can be considered as reserved area to assure capability of operation according to the designed networks.

2.2.3.2 Multiperiod MINLP Model Formulation

In this section, all equations in multiperiod MINLP model by an optimization program called GAMS (General Algebraic Modeling System) are presented. Those equations are not able to be put in the program directly, but they are needed to be converted into GAMS's language appropriately. The objective function of the model is to minimize total annualized cost (TAC) comprised of utilities cost, heat exchanger areas cost, and capital cost of heat exchangers. Binary variable will be used to determine the existence of each match for heat exchanger. The design variables are temperatures at every location in stage model and heat loads of each heat exchanger.

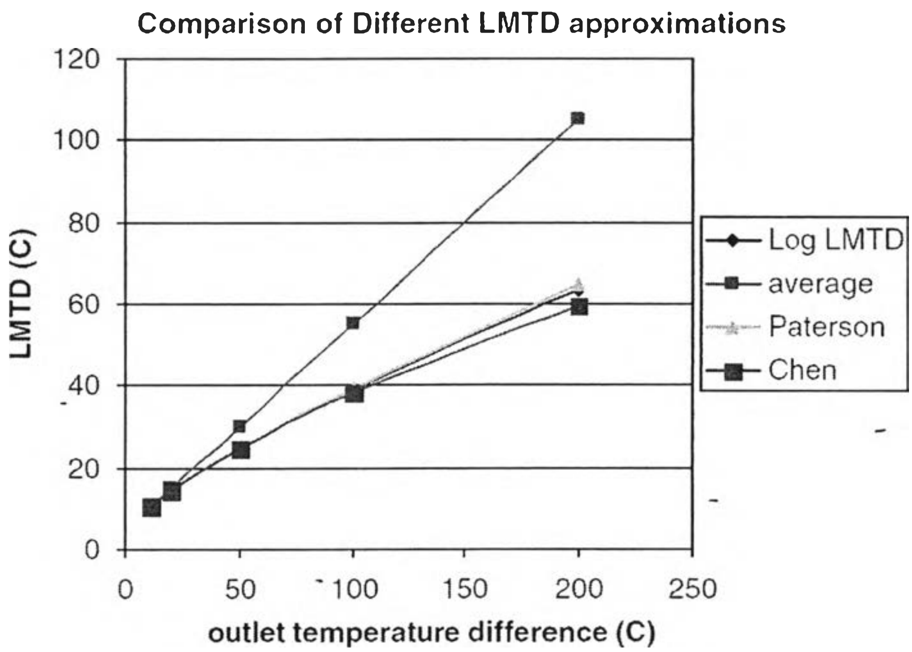

Figure 2.9 LMTD Approximation (Verheyen and Zhang, 2006).

Overall stream heat balances are performed to make sure that the total heat load required for each stream is sufficient in each period. The heat balance equations for both hot and cold streams are shown in Eq. 2.6 and Eq. 2.7.

$$[Th_{in}(i,p) - Th_{out}(i,p)].FCph(i,p) = \sum_{k\in ST}\sum_{j\in CP} q(i,j,k,p) + q_{cu}(i,p) \qquad i \in HP, p \in PR \tag{2.6}$$

$$[Tc_{in}(j,p) - Tc_{out}(j,p)].FCpc(j,p) = \sum_{k\in ST}\sum_{i\in HP} q(i,j,k,p) + q_{hu}(i,p) \qquad j \in CP, p \in PR \tag{2.7}$$

According to the isothermal mixing assumption, temperatures at each location for both sides of heat exchangers within a stage have to be equal. They are calculated from total amount of heat added or rejected in each stream as shown in Eq. 2.8 and Eq. 2.9 for hot and cold streams, respectively.

$$[th(i,k,p) - th(i,k+1,p)].FCph(i,p) = \sum_{j \in CP} q(i,j,k,p) \qquad k \in ST, i \in HP, p \in PR \tag{2.8}$$

$$[tc(j,k,p) - tc(j,k+1,p)].FCpc(j,p) = \sum_{i \in HP} q(i,j,k,p) \qquad k \in ST, j \in CP, p \in PR \tag{2.9}$$

The target temperatures of all streams in each period will be assigned to the first location (location = 1) for hot streams and the last location (location = NOK+1) for cold streams as formulated as Eq. 2.10 and Eq. 2.11.

$$Th_{in}(i,p) = th(i,1,p) \qquad i \in HP, p \in PR \tag{2.10}$$

$$Tc_{in}(j,p) = tc(i,NOK+1,p) \qquad j \in CP, p \in PR \tag{2.11}$$

In the stage-wise superstructure model, there must be monotonic increase or decrease in temperature. In other words, temperatures at the left-side location will always be greater than the right-side for every stream. That means the temperature of hot stream decreases continuously until it reaches the target temperature at the outlet. For cold streams, in other way round, temperature increases continuously because of heat received from hot stream until it reaches the desire temperature at the outlet. The formulations are shown in Eq. 2.12 and Eq. 2.12.

In case the temperature at the last location (for hot stream) or the first location (for cold stream) does not reach its target temperature, cold utility or hot utility have to be utilized respectively. Therefore, for hot streams, the outlet temperature at last location will be greater than or equal to target temperature. For cold streams, the outlet temperature at the first location will be less than or equal to target temperature. The equations are shown in Eq. 2.13 and Eq. 2.14.

$$th(i,k,p) \geq th(i,k+1,p) \qquad k \in ST, i \in HP, p \in PR \tag{2.11}$$

$$tc(j,k,p) \geq tc(i,k+1,p) \qquad k \in ST, j \in CP, p \in PR \tag{2.12}$$

$$th(i,NOK+1,p) \geq Th_{out}(i,p) \qquad i \in HP, p \in PR \tag{2.13}$$

$$TC_{out}(j,p) \geq tc(j,1,p) \qquad j \in CP, p \in PR \tag{2.14}$$

-Energy balances are also performed to find utility loads required to make temperatures of each stream reach its target temperature. The following equations are used. The equations are shown in Eq. 2.15 and Eq. 2.16.

$$[th(i,NOK+1,p) - Th_{out}(i,p)].FCph(i,p) = qcu(i,p) \qquad i \in HP, p \in PR \tag{2.15}$$

$$[Tc_{out}(j,p) - tc(j,1,p)].FCpc(j,p) = qhu(j,p) \qquad j \in CP, p \in PR \tag{2.16}$$

As binary variables is used to represent the existence of matches. There has to be logical constraints which are conducted to determine values of binary variables. If a match takes place or there is heat load, binary variable (z) is forced to be unity and heat load (q) is controlled to be less than its upper bound. But if there is no heat load (q=0), the value of z can be either 0 or 1. In fact, it should be only 0 because it is impossible to have heat exchanger with no heat load. However, it may be 0 since the model tries to minimize overall number of exchangers. Such constraints for process-process exchangers and utility exchangers are shown as Eq. 2.17, Eq.2.18, and Eq. 2.19.

$$q(i,j,k,p) - Q_{up}.z(i,j,k) \leq 0 \qquad i \in HP, j \in CP, k \in ST, p \in PR \tag{2.17}$$

$$qcu(i,p) - Q_{up}.zcu(i) \leq 0 \qquad i \in HP, p \in PR \tag{2.18}$$

$$qhu(j,p) - Q_{up}.zhu(j) \leq 0 \qquad j \in CP, p \in PR \tag{2.19}$$

Where $z(i,j,k), zcu(i), zhu(i) \in \{0,1\}$

Driving force for heat transfer is temperature difference. To assure their feasibility of both sides, i.e. cold stream temperature should not be greater than hot stream and the temperature difference should be high enough so that the heat exchanger area will not be too large, following constraints must be included as shown in Eq.2.20, Eq.2.21, and Eq.2.22. When a match exists, the temperature difference is forced to be higher than a certain value which usually equal to exchanger minimum approach temperature (EMAT). Therefore, Eq. 2.23 must be included.

$$dt(i,j,k,p) \leq th(i,k,p) - tc(j,k,p) + DT_{up}.\left(1 - z(i,j,k)\right)$$
$$i \in HP, j \in CP, k \in ST, p \in PR \qquad (2.20)$$

$$dt(i,j,k+1,p) \leq th(i,k+1,p) - tc(j,k+1,p) + DT_{up}.\left(1 - z(i,j,k)\right)$$
$$i \in HP, j \in CP, k \in ST, p \in PR \qquad (2.21)$$

$$dtcu(i,p) \leq th(i,NOK+1,p) - Tcu_{out} + DT_{up}.\left(1 - zcu(i)\right)$$
$$i \in HP, p \in PR \qquad (2.22)$$

$$dt(i,j,p) \geq EMAT \qquad (2.23)$$

To identify required areas of each match that is used in all periods, several constraints are included. The aim of these constraints is to find maximum area among all considered periods in which it can be operable for all periods. The formulations are shown in Eq. 2.24, Eq. 2.25, and Eq. 2.26.

$$Amax(i,j,k) \geq \frac{q(i,j,k,p)}{LMTD(i,j,k,p).U(i,j)} \qquad i \in HP, j \in CP, k \in ST, p \in PR$$
$$(2.24)$$

$$Ahumax(j) \geq \frac{qhu(j,p)}{LMTD(j,p).U(hu,j)} \qquad j \in CP, p \in PR \qquad (2.25)$$

$$Acumax(i) \geq \frac{qcu(i,p)}{LMTD(i,p).U(i,cu)} \qquad i \in HP, p \in PR \qquad (2.26)$$

Finally, the objective function is formulated to calculate the total annualized cost (TAC) consisting of all fixed costs (capital cost and maximum area cost) and operating costs (hot and cold utility costs) as shown in Eq. 2.27. The objective function will be asked to minimize by varying design variables within search space.

$$\begin{aligned} Objective\ function &= minimize\ TAC \\ &= AF[\sum_{i \in HP} \sum_{j \in CP} \sum_{k \in ST} C_f.z(i,j,k) + \sum_{i \in HP} \sum_{CU} C_f.zcu(i) \\ &\quad + \sum_{j \in CP} \sum_{HU} C_f.zhu(j)] + AF.\sum_{i \in HP} \sum_{j \in CP} \sum_{k \in ST} C.Amax(i,j,k)^B \\ &\quad + AF.\sum_{j \in CP} C.Ahumax(j)^B + AF.\sum_{i \in HP} C.Acumax(i)^B \\ &\quad + \sum_{p \in PR} \frac{DOP(p)}{NOP}.\sum_{i \in HP} C_{cu}qcu(i,p) \\ &\quad + \sum_{p \in PR} \frac{DOP(p)}{NOP}.\sum_{j \in CP} C_{hu}qhu(j,p) \qquad (2.27) \end{aligned}$$

2.3 Heat Integration in Crude Distillation Unit (CDU)

2.3.1 Crude Oil Refinery

Petroleum refinery process begins with crude oil distillation, and then the products will be treated in subsequent processes, such as hydrotreating, catalytic cracking, etc., to recover more valuable products as much as possible. It produces variety of petroleum products such as light gas, light naphtha, heavy naphtha, kerosene gasoil, and residue. There are mainly two types of distillation column which are atmospheric distillation unit (ADU) and vacuum distillation unit (VDU). Additionally, one of important equipment is desalter. It is considered as a part of atmospheric distillation unit facilities. It is used for removing salts, solids, and water from crude oil before entering to the distillation column to prevent damages. In Figure 2.10, the intermediate petroleum products will be separated based on different boiling points. The heavy products, i.e. atmospheric residue, will be sent further to VDU.

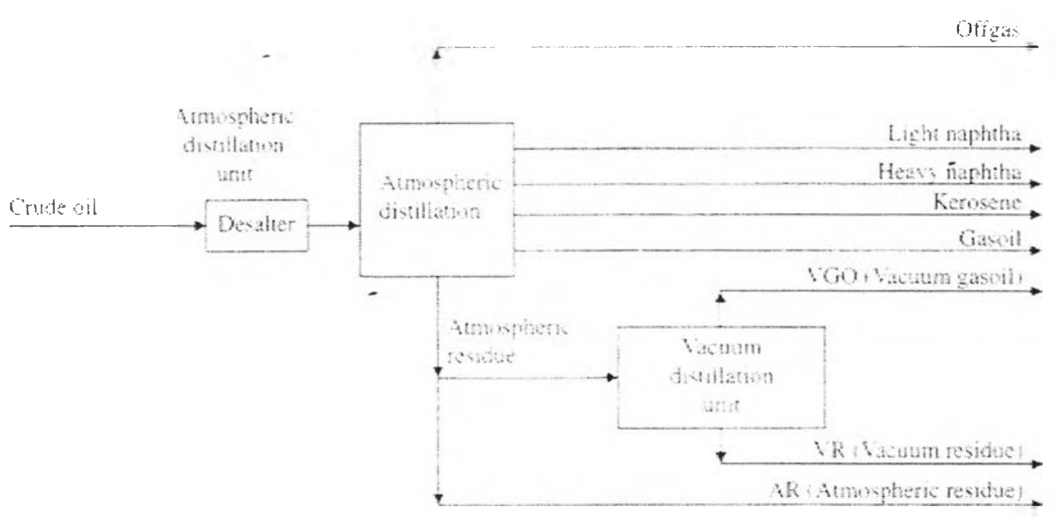

Figure 2.10 A crude distillation complex (Petroleum, 2000).

2.3.2 Crude Oil Assay

Crude oil is complex mixture of hydrocarbon composed of millions of compounds, most of which cannot be identified. Only light compounds such as methane, ethane, propane, and benzene can be identified. It also has some impurities, for instance, sulfur, oxygen, nitrogen, and metal. These compositions and other properties of particular crude oil are declared in a crude assay. One of important characteristics that affect the design of ADU is true boiling point (TBP) distillation (Petroleum, 2000). This property is specific for each crude oil as shown in Figure 2.11. It illustrates the product yield in percentage of cumulative volume distilled according to its TBP cut range determined by product specifications and market demand. Moreover, there are other properties reported in a crude assay such as API gravity (specific gravity used in petroleum industry), flash point, and sulfur content.

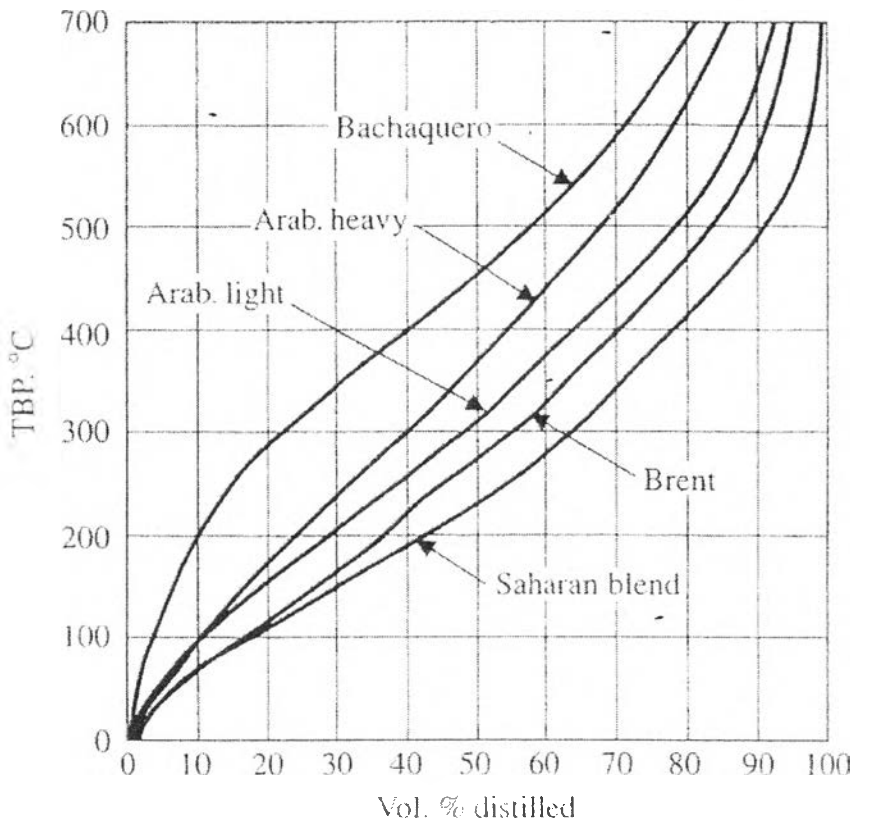

Figure 2.11 TBP distillation curve of different crude oils (Petroleum, 2000).

2.3.3 <u>Heat Integration in Crude Distillation Unit</u>

Atmospheric distillation unit is one of the largest energy consuming units in petroleum refinery plant. Generally, cold streams are crude oil feed and hot streams are the intermediate products leaving from the distillation column. After the crude oil is preheated, it will be sent to furnace to be heated up to the desired temperature around 310-370°C (Petroleum, 2000). The example for heat integration in atmospheric distillation unit process flow diagram is shown in Figure 2.12. Because crude oil characteristics are spontaneously uncertain in each reservoir, it causes temperatures and heat capacity flowrates varying over the period. Therefore, multiperiod heat exchanger network design is essential so that the HEN will be flexible and can be operable efficiently for every type of crude oil.

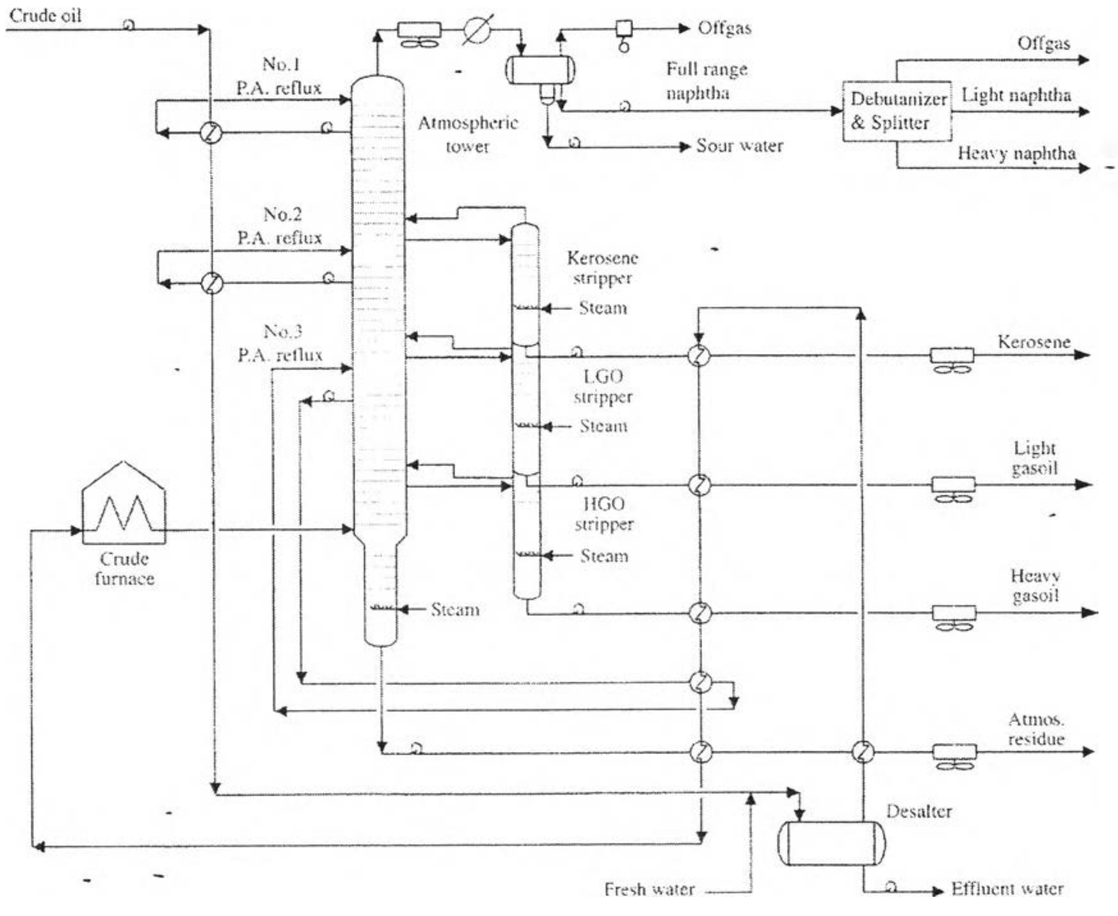

Figure 2.12 CDU process flow diagram (Petroleum, 2000).