HEAT-EXCHANGER NETWORK

1. Introduction

This article addresses the problem of heat-exchanger network design. However, it should be recognized that the design of such systems is no longer conducted in isolation from consideration of alternative means of energy conservation (eg, local application of heat pumps) and, most importantly, from consideration of how the design of the heat recovery system interacts and impinges on the design of the core of the process (reactors, separators etc.) and the utility system (steam levels, power generation system, refrigeration levels etc.). Today, many heat exchangers networks are designed by the use of software packages employing pinch technology. Placing too much reliance upon software is dangerous and one purpose of the present article is to move away from this situation.

Modern techniques for the design of energy intensive processes can be traced back to the pioneering work of Umeda and co-workers (1). This work called for a complete change in design philosophy. Prior to this work process design was considered to be a serial activity. First, the core of the process (usually the reactors) was designed. This work yielded the basic heat and mass balance for the plant. Given the heat and mass balance the heat recovery system was designed. Finally, having identified heater and cooler loads, the linkage to the site utility system could be identified. Umeda argued that this approach led to inefficient designs. He recognized that changes to the core of a process could result in large cost savings in both heat recovery system and utility system design. Similarly, he recognized that the design of the heat recovery system affected the cost of the utility system. Changes in network structure could allow the operator to make use of lower cost utility and thereby make large cost savings. Sometimes the ability to use lower cost utility could be brought about through changes to the core of the process itself. He therefore proposed a new design philosophy in which the interactions between process core, heat recovery system and utility system where recognized. This involved iterating across the three design interfaces.

The development of this philosophy was accompanied by the production of a number of new tools and techniques such as composite curves, pinch manipulation, algorithms for determining network area, total network cost analysis, heat supply and demand diagram, exergy composite curves and raffinate diagram.

The temperatures at which heat is being rejected by the *hot process* streams and demanded by the *cold process* streams control the opportunities for heat recovery. These temperatures are controlled by the design of the core process. Therefore, changes to the core process can result in increased heat recovery and reduced operating cost. Examples of process parameters that affect the temperature at which heat is demanded or rejected by the process are operating pressures (of reactors and distillation columns) and pump around flow rates. Changes in equipment configuration can also be used. For instance, intermediate condensers and reboilers can be used in distillation processes. The composite curves provide the engineer with a guide to the benefit process changes have for heat recovery system design. They are a key tool in iterating across the interface between the core of the process and the heat recovery network.

The benefits are not restricted to operating cost. The amount of surface area needed for heat recovery is dependant upon the shapes of the composite curves. Furthermore, it is possible to reduce the number of individual heat recovery matches by manipulating the basic heat and mass balance. So, changes to the core of the process can also result in reductions in the capital cost of the network.

Opportunities for improved overall process design extend to the utility system. Umeda and co-workers (2) introduced a graphical representation showing heating and cooling needs of a process as a function of temperature. They named this representation the *Heat Demand and Supply Diagram*. Linnhoff and co-workers (3) subsequently called this diagram the Grand Composite Curve.

This diagram is used for utility selection including the identification of optimal combined heat and power arrangements. Since, the shape of this curve is again dependant upon the core of the process it is sometimes possible to make changes to the core of the process that provide opportunity to make use of lower cost utility.

1.1. Grid Representation of Heat Recovery System. The design of a heat recovery network is aided by systematic representation of its structure. The *grid diagram* provides a picture that can contain much information. The hot and cold process streams are represented on the grid as a series of horizontal lines. The arrangement favored by the authors is hot streams placed at top of diagram and, in temperature terms, running from right to left; cold streams at bottom of diagram running from left to right (Fig. 1). This arrangement follows the direction of the two composite curves.

Information relating to the stream can be placed adjacent to the lines. For instance, in the example shown the stream name (H1), the heat capacity flowrate of the stream (20.2) and the heat transfer coefficient characterizing the stream (1000) are listed.

Heat recovery exchangers are represented by two circles (one on each of the stream involved) joined by a vertical line. The temperatures at inlet and exit are shown above the line, one each side of the circle. The heat load can be shown below the lower circle (see Fig. 2).

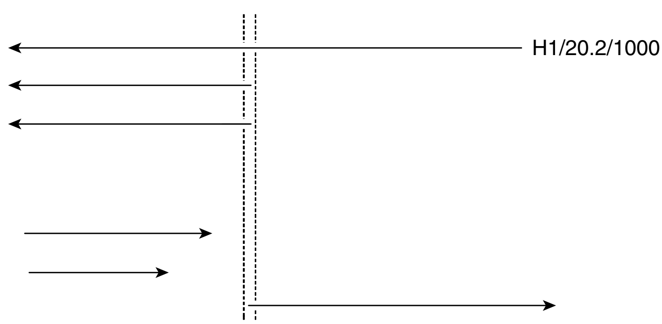

Fig. 1. Representation of hot and cold streams.

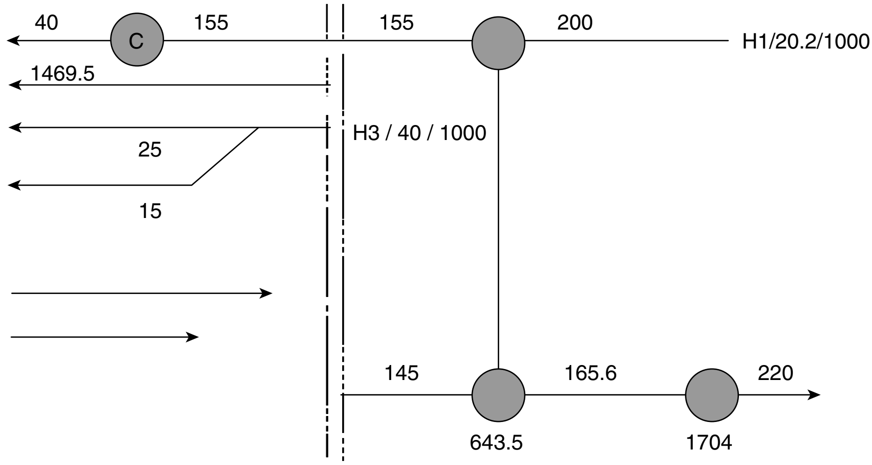

Fig. 2. Representation of heat recovery units coolers, heaters and stream splits.

Further information that can be included (but not shown in our example) could be terminal temperature differences (one each side of vertical line), the exchanger number (written above or inside upper circle), and exchanger area (written inside lower circle). Coolers are shown as single circles positioned on lines representing hot streams. Heaters are shown as single circles positioned on lines representing cold streams.

Finally, stream splits can also be represented on the diagram by lines branching off the process lines and running parallel with them. The heat capacity flow-rates of the split streams can be written below (thereby avoiding confusion with local temperature information) the relevant lines (Fig. 2).

2. Economics of Heat Recovery

Consider a very simple process system involving just one hot and one cold stream. Both streams need to be taken from a set *supply temperature* to a set *target temperature*. These process objectives can be achieved by using hot and cold utilities or by using heat recovery supplemented by the use of hot and cold utilities. How does the engineer identify the optimum design?

The logical way forward would be to set up the heat recovery system shown in Figure 3 and undertake cost optimization.

When there is no recovery (X equals zero), only the heater and cooler are present. The cold utility consumption equates with the heat rejected by the hot stream. The hot utility consumption equates with the heat demanded by the cold stream. Hence, the annual operating cost can be determined.

The sizes of the two utility exchangers can be estimated using the standard heat exchanger design equation. Assuming that the cost of each unit can be determined from a cost equation, the capital cost of the system can be determined. This cost number is multiplied by an annualization factor in order to set it on the same basis as the operating cost. Then the addition of the operating

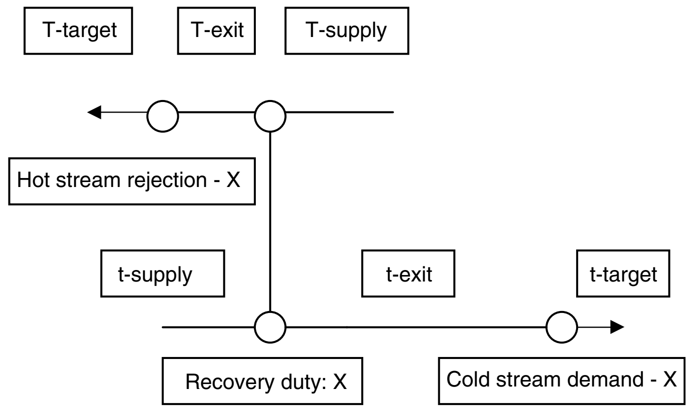

Fig. 3. Simple heat recovery system.

cost and the annualized capital cost yields the total annual cost of the system that provides the objective function to be minimized.

If heat recovery is now introduced (X is now finite) both heater and cooler loads are reduced by X and the total quantity of heat transferred within the system is reduced by X. This can mean that heat recovery leads to a saving in capital. However, at low heat recovery levels the capital cost increases because an additional heat exchanger is introduced. The reduced heater and cooler loads translate into reduced operating cost.

In the optimization the load on the heat recovery unit is named and the total annual cost of the system is determined. The minimum value is the economic heat recovery level.

2.1. Energy Demand. The process streams in the simple system described above is represented on a plot of temperature against heat load (Fig. 4).

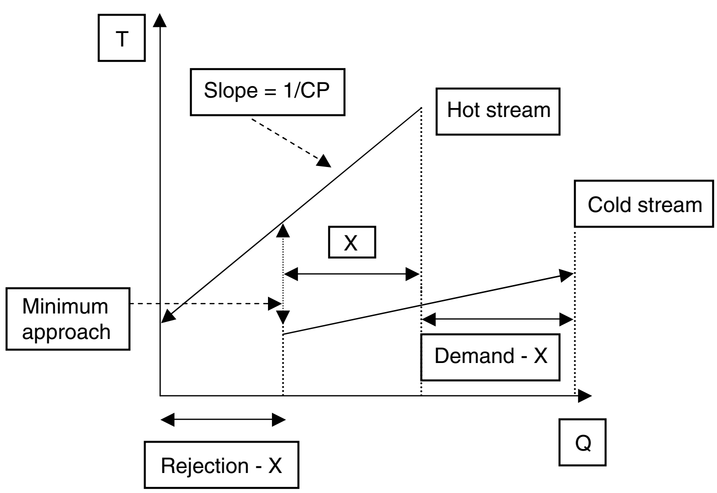

Fig. 4. T-Q Plot of simple process system.

If the heat capacity flow-rate of a stream does not vary with temperature it appears as a straight line. The slope of the line is the reciprocal of the heat capacity flow-rate of the stream. If the heat capacity flow-rate varies with temperature, it appears as a curved line. When the lines representing the hot stream are superimposed such that the overlap equates with a heat recovery level X, the distance the hot stream over-shoots the cold equates with the heat rejected to cold utility and the distance the cold stream overshoots the hot equates with the heat demanded from the hot utility, the two lines make a close approach on the temperature axis. This is called the *minimum temperature approach*.

When the heat capacity flow-rate of the cold stream exceeds that of the hot stream (which therefore has line of larger slope) the close approach occurs at the cold end of the unit (Fig. 4). When hot stream has the larger heat capacity flow-rate, the close approach occurs at the hot end of the unit. (As will become apparent later, these observations have relevance in design).

This picture of heat recovery can be extended to cover systems involving many hot and cold streams by using *composite curves*. Originally proposed by Huang and Elshout (4) a Cold Composite Curve shows the cumulative heat demanded by a process as a function of temperature. A Hot Composite Curve shows the cumulative heat rejected by process streams as a function of temperature. The hot curve can be superimposed upon the cold curve (Fig. 5) such that the two curves are separated by a minimum temperature approach. Where the

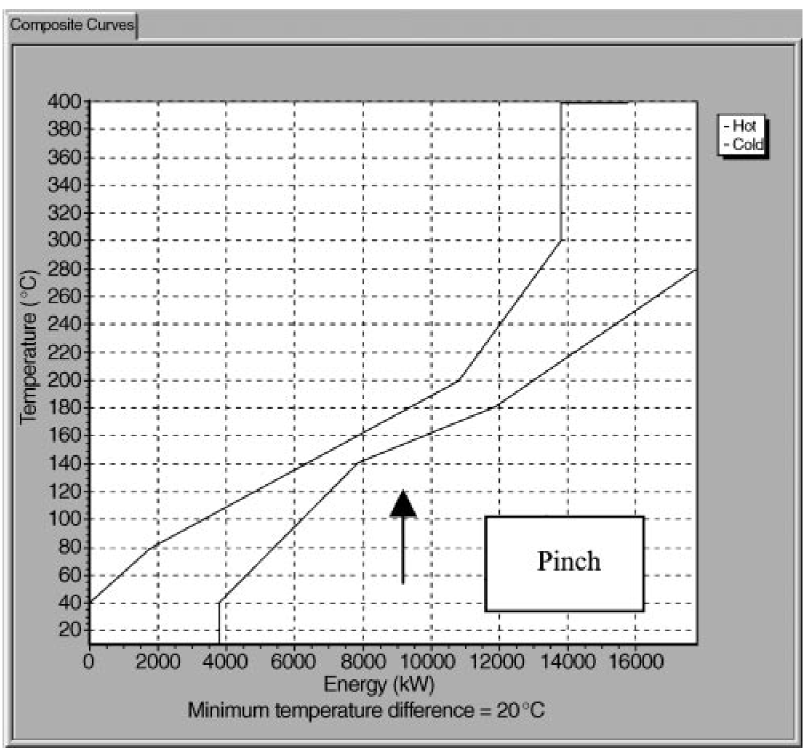

Fig. 5. Superimposed composite curves.

curves overlap heat can be transferred from the hot streams to the cold streams without infringing the minimum temperature approach. The over-shoot made by the cold composite indicates how much hot utility is required to drive the process. The over-shoot made by the hot composite indicates how much heat is rejected to cold utility.

It is common to find that the point of close approach occurs at temperature part way along the temperature span covered by the process streams. This point has been termed the Pinch Point.

However, before the design problem can be finalized it must be placed in the context of overall system design. Since the curves are derived from the basic heat and mass balance of the process, changes to the core of the process change the shape of the curves. Every feature on the curves can be related to an actual process parameter. For instance, the slope of the cold composite curve at any point is the reciprocal of the sum of the heat capacity flow-rates of the cold streams passing through that temperature. A change in slope means a change in the sum of these heat capacity flow-rates. This is generally associated with the arrival or departure of a stream from the temperature field and so corresponds with a supply or target temperature.

Process changes that increase the overlap of the two composites give rise to increased heat recovery. Changes that increase the vertical separation between the two composites result in increased temperature driving forces for the heat exchangers and thus result in capital cost savings.

2.2. Approximate Capital Cost. The capital cost of a heat exchanger network is controlled by a number of factors:

- The number of heat recovery matches made
- The number of individual exchangers used in these matches
- The size of the individual exchangers
- The cost of installation of the exchangers
- The cost of piping

Algorithms which allow the engineer to determine the minimum number of heat recovery matches and the number of exchangers required in these matches have been successfully developed. One very important aspect of these algorithms is that they do not depend on the design of the exchanger network. They are applied ahead of the design process and are used to guide the design process.

Determination of the size of individual exchangers presents a difficulty. Virtually all developers of process synthesis procedures assume that the capital cost of a heat exchanger is dependant upon the surface area of the exchanger. A power law relationship is often used to provide economy of scale. Unfortunately, the true cost is not a direct function of surface area. It is a strong function of exchanger diameter and a weak function of exchanger length. This means that a long thin unit can be substantially cheaper than a short fat unit containing much less area. The dimensions of an exchanger are set by pressure drop considerations as well as heat load and temperature driving force. No workers have actually incorporated algorithms for exchanger sizing into process synthesis procedures. Consequently, only an approximate sizing can be obtained without undertaking some exchanger design calculations.

The engineer should pay close attention to the number of heat recovery matches used in the system. The fixed cost released by the removal of a unit pays for a substantial increase in area of another unit.

Although the size of individual units cannot be determined ahead of design, an approximate determination of the total amount of surface the network will require can be made. Furthermore, a network must be structured in order to achieve a design that is close to this value. Total network area does not fully reflect the exchanger costs but it is a guide to those costs.

In the list of cost factors, piping costs are separated from local installation cost. The local installation cost covers factors such as foundations, insulation, land occupation. The reason for the separation is that piping costs can exceed exchanger costs. They are a function of network design and plant layout. Consideration of piping cost is *not* part of most established synthesis procedures. The engineer must not loose sight of the importance of these costs.

Number of Units. Consider the system illustrated in Figure 6**a**. The circles represent process and utility streams. The hot streams are placed along the top of the diagram (streams A-C). The heat loads associated with the streams is written above the circles. The cold streams are placed along the bottom

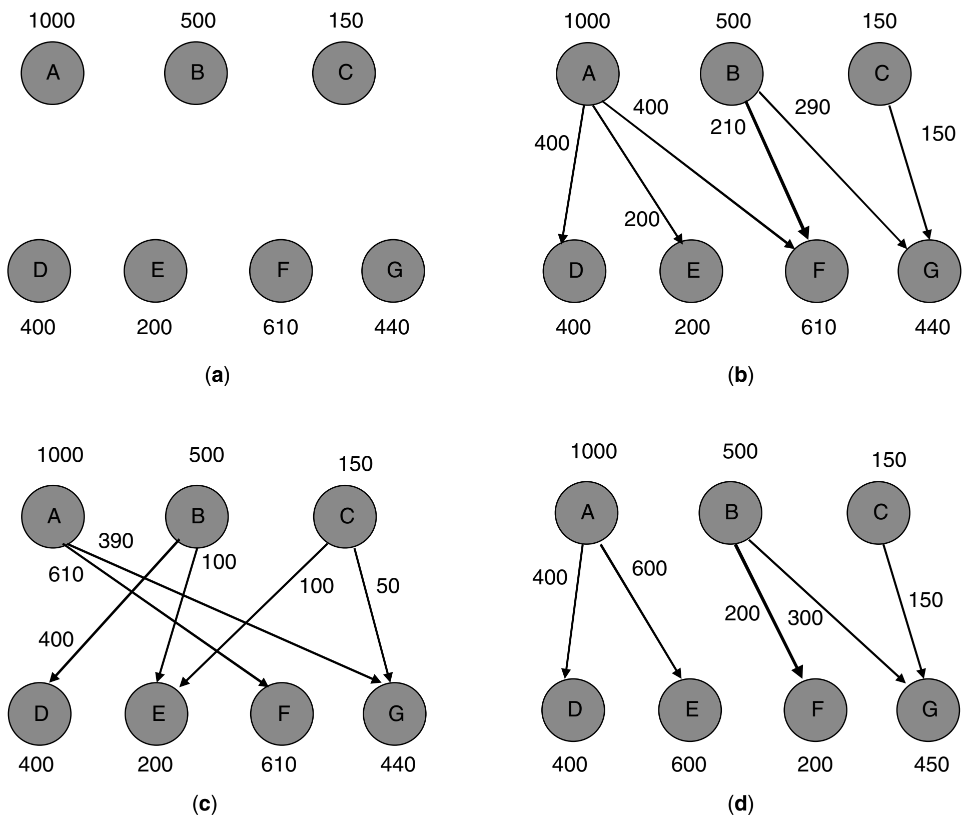

Fig. 6. (**a**) Set of heat balanced streams. (**b**) One possible match arrangement. (**c**) Alternative match arrangement. (**d**) Another problem. Two heat balanced system.

(streams D-G) with heat loads written below the circles. The system is in heat balance.

Now match the streams such that demands made by the cold streams are just satisfied. Use the minimum number of matches.

One result could be that shown in Figure 6**b**. The arrows represent individual matches. The heat load on each match is listed alongside the arrow. Six matches are used. An alternative is shown in Fig. 6**c**. Again six matches are used. The minimum number of matches is one less than the number of streams involved.

In Fig. 6**d** the solution to a different problem is presented. Here just five matches are required. However, this problem can be decomposed into two separate heat balanced systems. The first problem had just one heat balanced system.

Network Area. The condition required for the minimization of the amount of heat transfer surface within an individual heat exchanger is *pure counter-flow*. Any disruption of this arrangement results in waste of temperature driving force and the need for additional area in order to fulfill the required duty. The condition of pure counter-flow extends to thermal systems. In the context of heat exchanger networks it requires the individual streams to be matched such that hot and cold contact temperatures align vertically between the composite curves.

The system can be divided up into a series of *heat intervals* (as shown in Figure 7). Within each interval both the hot and the cold composites are single straight lines (indicating constant heat capacity flow-rate). The hot and cold temperatures at entry to and exit from each interval are fixed by the positioning of the composites (and can be estimated).

Heat Interval Example. Assume that three hot streams (H1-H3) and two cold streams (C1 and C2) flow through the interval. Further assume that the simple network shown in Figure 8 can be constructed such that all of the duties are satisfied.

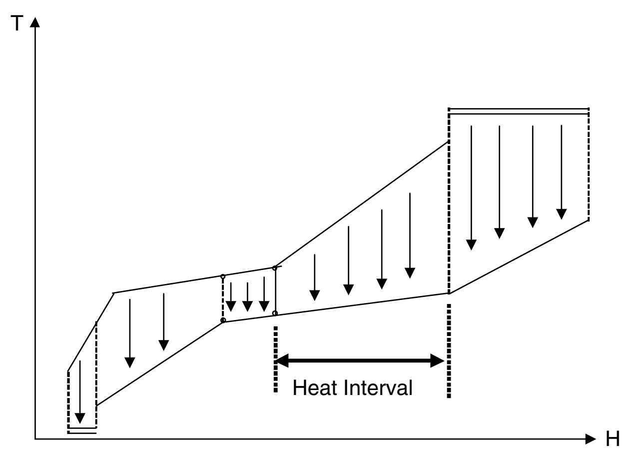

Fig. 7. Pure counter-flow in heat exchanger networks.

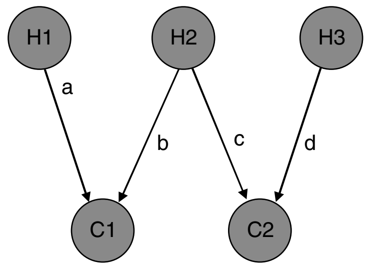

Fig. 8. Example interval structure.

The sizes of each of the individual exchangers can be estimated using the standard exchanger design equation. For instance, for exchanger b:

$$A_b = \left(\frac{Q_b}{Q_b + Q_c}\right) Q_{H2}\left(\frac{1}{h_{H2}} + \frac{1}{h_{C1}}\right)\Delta T_m = Q_{C1}\left(\frac{1}{h_{H2}} + \frac{1}{h_{C1}}\right)\Delta T_m$$

When the relationships between the heat flows and sum all of the areas are recognized, a result that is independent of the structure is obtained

$$A_{interval} = \left(\sum \frac{Q_H}{h_H} + \sum \frac{Q_C}{h_C}\right)\Delta T_m$$

The network area can be obtained by summing the results of all of the heat intervals in the system:

$$A_{Network} = \sum_{intervals} \Delta T_m \left(\sum_{hot-streams} \frac{Q_H}{h_H} + \sum_{cold-streams} \frac{Q_C}{h_C}\right)$$

This estimate of minimum network area is rigorous for cases in which the exchangers use pure counter-flow and both hot and cold streams have uniform heat transfer coefficients. In real networks neither of these conditions is met.

The extension of the methodology to handle exchangers that do not exhibit pure counter-flow (eg, shell-and-tube exchangers having multiple tube passes) has been reported by Ahmad and Smith (5). Their algorithm is incorporated into many computer programs.

The problem of differing stream heat transfer coefficients is difficult to resolve without resorting to some level of detailed design. However, the accuracy of the information on these heat transfer coefficients does not justify the use of a sophisticated approach.

The objectives of minimization of area and the use of a minimum number of units are not mutually compatible. A large number of units would be required in a network exhibiting minimum area. A significant amount of nonvertical matching must be introduced in order to obtain a network having the minimum number of units. This would increase the amount of surface needed. However, the increase can be partially compensated by making matches that exploit the

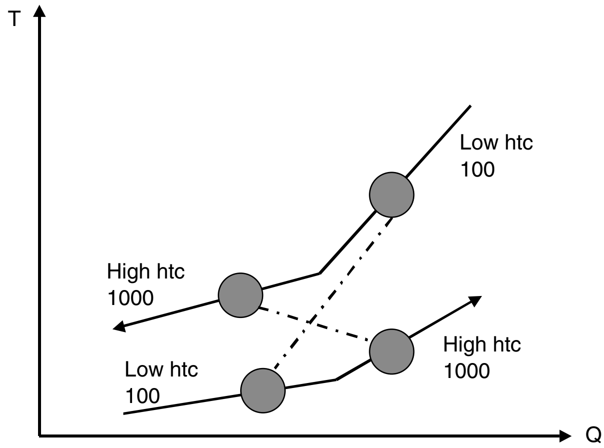

Fig. 9. Use larger temperature driving force on lower U.

differences in the individual stream heat transfer coefficients. Consider the options open to the designer in matching the streams shown in Figure 9.

The system has two hot and two cold streams. Vertical matching would result in two exchangers having low overall heat-transfer coefficients (91). One exchanger would have large mean temperature difference. The other would have a low value.

Nonvertical matching would result in one exchanger having a high U value [500] but lower mean temperature difference. The other exchanger would have a lower U value [50], but much higher mean temperature difference. The result is reduced area.

The combination of nonvertical matching and the exploitation of differences in stream heat transfer coefficients leads to a situation in which the network required for a practical solution utilizing a minimum number of units is generally just 10–15% greater than the minimum value determined from the equation presented above.

Finally, how are the values used for stream heat-transfer coefficients arrived at. In the case of streams undergoing phase change, values from tables of typical values can be used. The true coefficients encountered for these duties are usually significantly higher than those involving sensible heating or cooling.

In the case of sensible heat transfer, the stream coefficients can be calculated on the basis of a specified velocity or an allowable pressure drop using an algorithm developed by Polley and Panjeh-Shahi (6).

The above approach applies to systems in which the same cost equation applies to each exchanger. This is not the case where the process streams require differing materials of construction or differing exchanger design pressures. The estimation of capital costs in these situations has been considered by Jegede and Polley (7).

2.3. Economic Heat Recovery Level. Utility cost is an annual figure. Capital cost is an initial outlay. The two costs are brought together by multiplying the capital cost by an annualization factor to yield an annual capital cost and summing the two components to give a total annual cost. One task is to identify, ahead of any design activity, the *minimum total annual cost* that can be achieved.

Example. Consider a simple heat recovery problem involving just one hot and one cold stream. If the temperature-heat load lines are superimposed such that they touch (the minimum temperature approach is zero) the overlap of the lines shows the thermodynamically maximum heat recovery level. Since, the temperature approach is zero the heat recovery exchanger would require an infinite amount of surface. A finite temperature approach is needed in order to have a practical exchanger. The approach that can be accommodated depends on the type of heat exchanger used. For a shell-and-tube heat exchanger, the minimum practical approach is around 10°C. For a plate-and-frame unit it is around 5°C. If the lines are superimposed such that they have a vertical separation (a close approach) equating to the practical limit, the overlap of the lines yields the minimum practical heat recovery.

The practical limit is below the thermodynamic limit. However, the investment needed to achieve the recovery is now finite. As the vertical separation of the hot and cold lines increases the temperature driving forces increase and the size of heat recovery exchanger needed reduces. However, the overlap of the two lines reduces. So, the capital cost reduction is accompanied by a reduction in heat recovery and increase in operating cost.

The minimum total cost of the system can be determined by systematically varying the temperature approach and, for each value, determining the operating cost and annualized capital cost for the system. Examination of a curve of total annual cost against temperature approach identifies the optimum approach. This can then be translated into hot and cold utility consumption and system capital cost.

The same procedure can be applied to heat exchanger networks (8). Here the minimum temperature approach is systematically varied between the hot and cold composite curves and the utility needs and network capital cost are determined. These costs are the combined to give a total annual cost for the network.

3. Theory of the Pinch Design Method

The costs associated with a heat-exchanger network have two components: operating costs (dominated by utility consumption) and capital cost. These costs can be combined to yield a total annual cost. In developing a design it is necessary to control costs as a network structure is progressively built up. The problem can be approached by controlling utility cost, controlling capital cost, or controlling total annual cost. The Pinch Design Method (9) uses a staged approach. First, the heuristics described above are used to determine an economic heat recovery level. This sets the optimum minimum temperature approach. Then, an initial network structure is developed using a systematic procedure which both ensures that the minimum temperature approach constraint will be met and the network will only use the identified amount of hot and cold utility. Finally, this network structure is subjected to optimization aimed at minimizing its total annual cost.

The Pinch Design Method is systematic and elegant but it does have pitfalls. The procedure dictates which stream matches should be made. Sometimes the matches are impractical. No guidance is given on how alternatives should be identified. The method is centred on the use of the grid diagram. This diagram does not contain information on where the streams originate and where they are going. The result can be a network design that incurs a great deal of expense in terms of piping and layout related costs. The method is easy and straightforward to apply in problems of (typically) up to ten streams. However, some processes have in excess of forty hot and cold streams. With this size of problem the method is not always easy to apply and layout considerations become very important. Finally, the results of the optimization applied at the end of the design process are dependent upon the starting structure. Different structures yield different final designs. Most importantly, the presence of small and unnecessary design elements can prevent the identification of significant structural changes. So, the optimization procedure does not guarantee a result that is truly optimal.

3.1. Pinch Principle. Consider the situation illustrated in Figure 10. A set of composite curves is shown. A vertical part is positioned way along the heat load axis. The line divides the problem into two systems each of which is in heat balance. To the left of the line, the heat rejected by the hot streams equates to that demanded by the cold streams plus the cold utility requirement. To the right of the line, the heat demanded by the cold streams equates with that rejected by the hot streams plus the hot utility requirement. A design is obtained that only consumes the identified utility requirements by simply designing two self-contained systems and ensuring that no hot utility is used in the system that balances with the cold utility and no cold utility is used in the system that balances with the hot utility.

The above observation is the key to the Pinch Design Method. One factor controlling network capital cost is the minimum temperature approach between

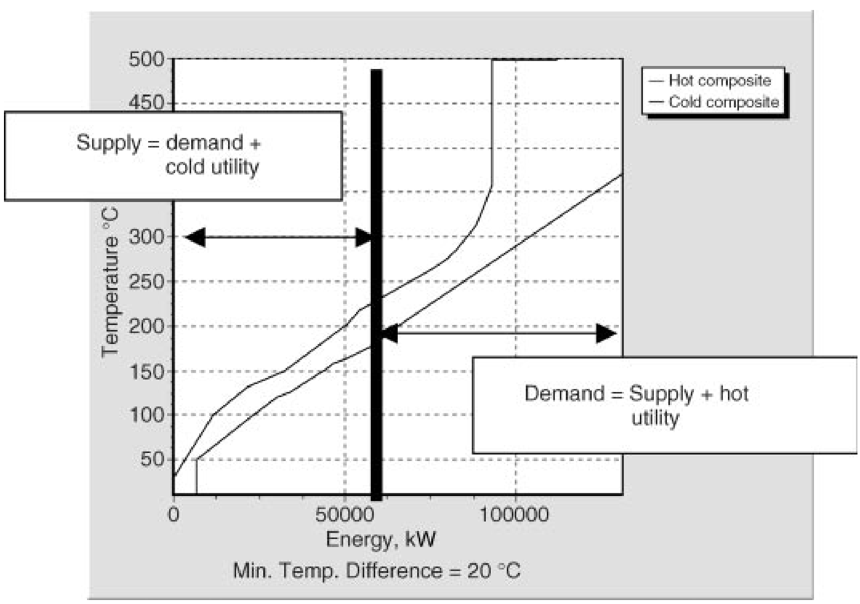

Fig. 10. Problem division.

the composites. The point of close approach has been named the *Pinch*. The logical place to sub-divide the system is at the Pinch, for this provides a means of controlling the minimum temperature approach.

So, to implement the Pinch Design Method:

- divide the problem at the Pinch.
- develop design for streams positioned below the Pinch making sure that no hot utility is used.
- develop design for streams positioned above the Pinch making sure that no cold utility is used.
- merge the two subnetworks.
- subject the resultant structure to optimization.

3.2. Making Stream Matches at the Pinch. In developing the subnetworks, start by making stream matches at the Pinch. Aim to ensure that the minimum temperature approach is not infringed.

Consider the network being developed for the system below the Pinch. When matches are made at the Pinch, the minimum temperature approach occurs at the hot end of the unit. In order to prevent infringement of the approach constraint within the heat recovery unit, the heat capacity flow-rate of the hot stream involved in the match must be greater or equal to that of the cold stream.

A mirror situation exists for the network developed for the system above the Pinch. The minimum temperature approach occurs at the cold end of the heat recovery units positioned at the Pinch. In order to prevent infringement of the approach constraint within the heat recovery unit the heat capacity flow-rate of the cold stream involved in the match must be greater or equal to that of the hot stream.

It is considered that below the Pinch the hot streams travel into the Pinch (or, leave Pinch) and the cold streams travel away from the Pinch. Above the Pinch the situation is reversed. These two observations can be combined to yield a single match rule.

A convenient way of identifying the matches is that obey these rules is the *CP Matrix*. Here the streams leaving the Pinch are listed along the top of the matrix in order to increasing heat capacity flow-rate. Those entering the Pinch are listed down the side of the matrix in order of increasing heat capacity flow-rate. The matrix elements can then be made up of the differences between the out stream and in stream values. A zero or positive number indicates an acceptable match. A negative number indicates an unacceptable match.

A few examples follow.

1. Consider the following matrix:

Out:			
In	10	25	40
8	+2	+17	+32
21	−11	+4	+19
33	−23	−8	+7

Looking down the diagonal one observes that all of the numbers are positive. These matches can be made without infringing the temperature approach constraint. Looking at the numbers below the diagonal, all of the numbers are negative. None of these matches is suitable. So, the design has only one option: the matches shown down the diagonal.

2. Now consider the following matrix:

Out:			
In	18	25	40
8	+10	+17	+32
15	+3	+10	+25
33	−15	−8	+7

Here there is a positive number positioned below the diagonal. This indicates that there is more than one possible structure at the Pinch.

3. Next, consider the following matrix:

Out:			
In	10	30	40
12	−2	+18	+28
18	−8	+12	+22
33	−23	−3	+7

Here there is a negative number on the diagonal. This means it is necessary to split one of the out streams in order to generate an acceptable solution. Obviously, the stream that must be split must have a higher heat capacity flow-rate. Examination of the matrix shows that there is no alternative match for the in stream having the highest heat capacity flow-rate. It must be matched with the out stream having the highest heat capacity flow-rate. However, the middle stream has an excess of 12 following its match diagonal match. It can therefore be used to satisfy the first ‘in’ stream. The resultant matrix is:

Out:				
In	10	a-12	b-18	40
12	−2	0		
18	−8		0	
33	−23			+7

This is a 4×3 matrix. One matches down the diagonal of the right-hand square matrix.

3.3. Setting the Heat Load of Individual Matches. Consider the situation illustrated in Figure 6**b**. This is a solution used to lead to an equation for determining the minimum number of matches required in a network. The match between streams A and D fully absorbs the heat demand made by D.

The match between A and E fully absorbs the demand made by E. The heat that remains in A is then fully absorbed by F. What remains of demand made by stream F is then satisfied using heat from stream B. The heat remaining in B is then fully absorbed by G.

Each match results in fully satisfying either a supply or a demand made by one of the streams involved. In other words, *the heat load is maximized on individual matches*. If this rule is applied in design, a network that has the minimum number of matches (a key aspect in minimizing capital expenditure) is automatically derived.

3.4. Completing Initial Design. It is necessary to complete the design by making heat recovery matches that complete the duties remaining for the streams already used in the Pinch region design and fulfill the duties for the streams which are not present in the Pinch region.

Good design can generally be achieved through inspection. However, in making these matches the designer must keep in mind how the plant will be started up, controlled and shutdown. The use of heaters and coolers is important in this context. Since, the purpose of the network is heat recovery, the novice engineer tends to concentrate upon this aspect of design. This can result in networks that contain heat recovery units in preference to utility units. Provided the overall heat balance is maintained it is possible to achieve required energy efficiency where using a significant number of utility units.

3.5. Principles Behind Final Optimization. Having correctly observed the procedure described above, the designer has two networks that can simply be merged together. Each of the networks elements will use the minimum number of units associated with the number of streams involved in the subproblem. However, the summation of these numbers will not generally equate with the minimum number of units required for the overall problem. This is because it is common to find that a number of streams appear on both sides of the Pinch.

Consider the system of matches shown in Figure 11.

This system has seven matches. The minimum number is six and the structure has a significant feature. It is possible to move up and down the lines connecting the nodes and trace a path by which it is possible to return to a starting position (bold lines). This feature is called a loop and is present whenever the minimum number of units is exceeded.

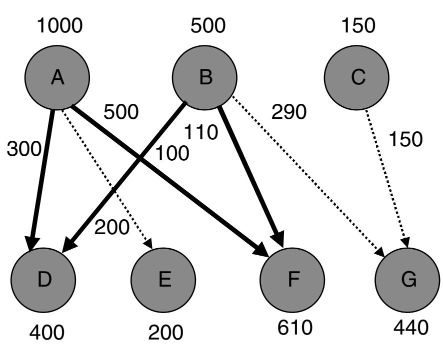

Fig. 11. System with more than minimum number of matches—a loop.

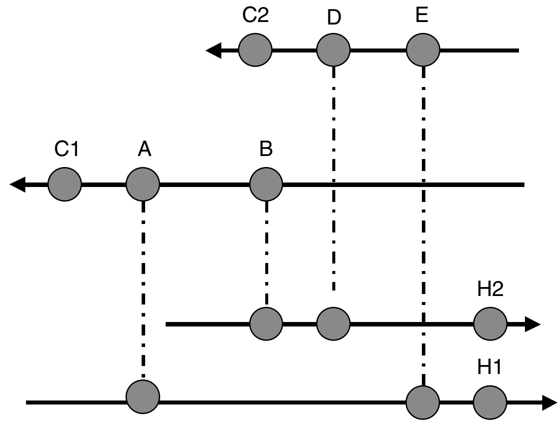

Fig. 12. Identifying loops on grid diagram.

Loops can be mapped out on the network grid diagram. An example is shown in Figure 12. Starting at node A, one move down the exchanger, along the cold stream to the next unit, up to node E, along the cold stream to node D, down the exchanger, along the hot stream to the previous exchanger, up the exchanger to node B, which neighbors the starting node (A) on the cold stream. So, one can move back along the stream to the starting point.

Loops can also involve utility units. For instance, returning to Figure 12 a route between the two coolers (C1-A-B-D-C2) can be traced as can a route between the two heaters (H1-B-A-H2).

In order to reduce the number of units a loop must be eliminated. This is done by removing one of the exchangers by transferring its duty to either another heat recovery unit (by reducing the allowable temperature approach) or by transferring the duty to existing heater and cooler. Of course, the two options can be used in combination.

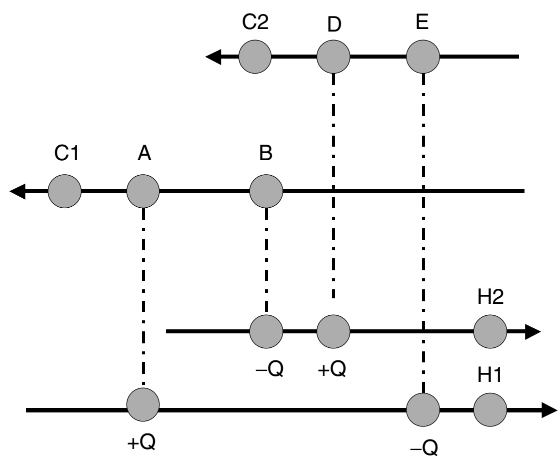

Fig. 13. Heat load adjustments associated with removal of Unit B.

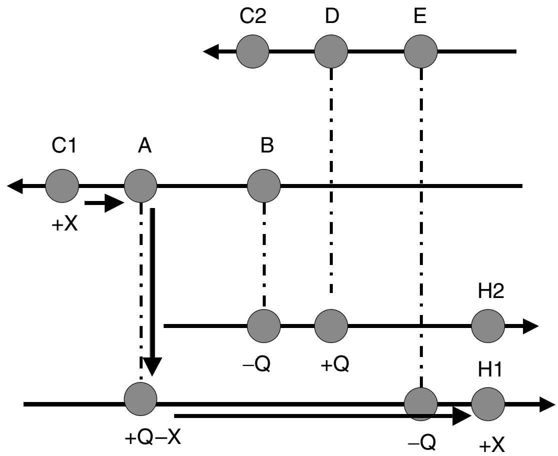

Fig. 14. Optimization of effect of removal of Unit B on sizing of Unit A.

Whatever choice is used, there is a trade-off between costs. This trade-off can be between the cost of a unit and the energy costs of the process or a trade-off between capital associated with a full unit and that associated with increased size of a unit.

The network shown in Figure 12 has seven units. The minimum number required is five. Assume that the load on unit B is small (and has a value of Q). As illustrated in Figure 13, this unit can be removed by transferring the cooling duty to unit A and the heating duty to unit D.

When the exit temperatures for these exchangers are redetermined for the changed heat load, it may be found that the minimum temperature approach has been infringed or that the new values are infeasible (due to temperature cross). The problem can be solved by taking up part or all of the changed duty in the utility units. For instance, assume that the changed duty would result in infeasible temperatures in unit A. Feasibility can be restored by taking up part of the cooling duty in cooler C1 and part of the heating duty in heater H2 (see Figure 14). Once feasibility is restored, there is an optimization problem. The utility load can be adjusted until the combined cost of changed utility consumption and differential annualized capital cost of unit A is minimized.

Removal of unit B also affects the feasibility and economics of unit D. The optimization of this unit involves the other cooler and the other heater (Figure 15).

4. Worked Example of Pinch Design Problem

A convenient example is the “Simplified Aromatics Plant” reported in the *IChemE User Guide on Process Integration* (10). The process flow diagram for the plant is presented in Figure 16.

The stream data for the problem is given in Table 1. The duty heat transfer coefficients (that is the stream film heat transfer coefficient after allowance for

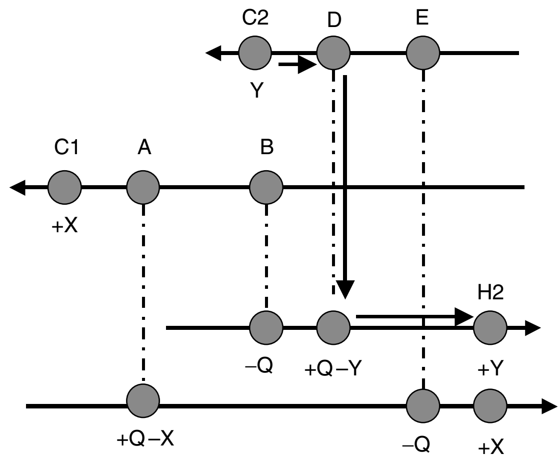

Fig. 15. Optimization of effect of removal of Unit B on sizing of Unit D.

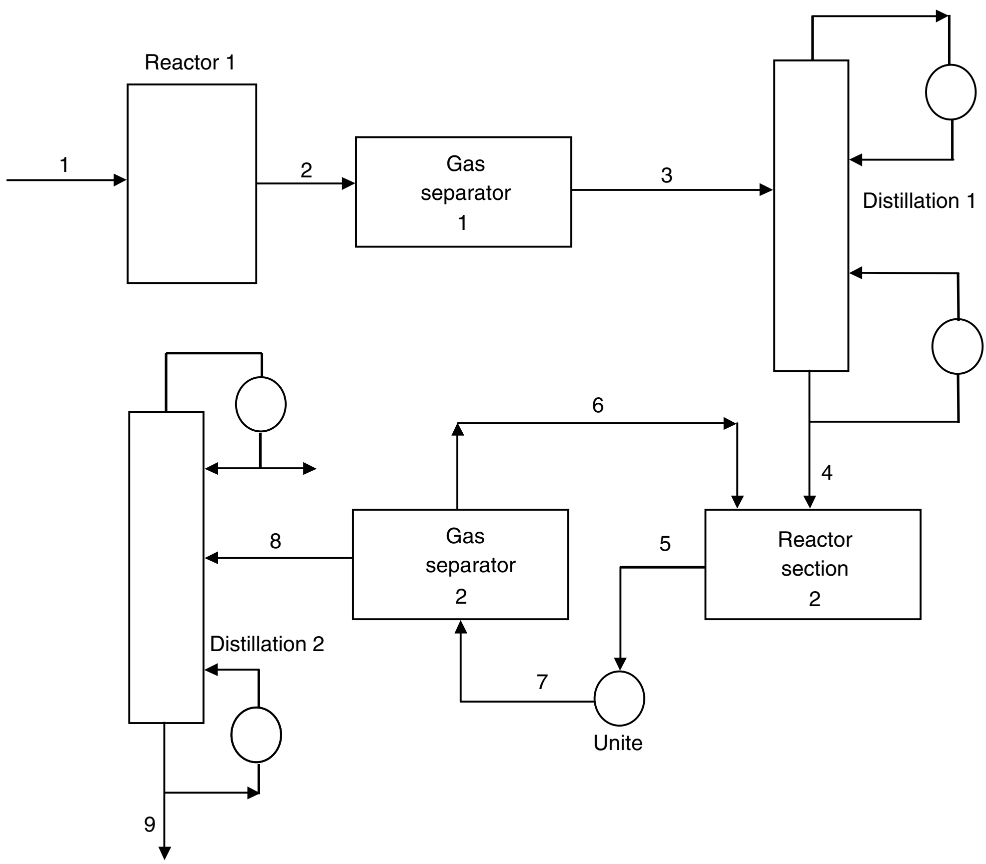

Fig. 16. Process flow diagram for aromatics plant.

Table 1. **Stream Data for Standard Problem**

Stream	Name	Supply temp.	Target temp.	Heat capacity flowrate
1	R1 feed	100	300	100
2	R1 product	327	40	100
3	D1 feed	35	164	70
4	R2 feed	140	300	200
5	R2 product	220	160	160
6	R2 recycle	85	138	350
7	Sep. feed	160	60	400
8	D2 feed	60	170	60
9	D2 product	220	60	60

fouling has been made) for each of the process streams is set at 500 W/.m^2.K. The duty coefficient for the hot utility (which is recirculating hot oil) is set at 1000 W/.m^2. K. That of the cold utility (which is cooling water) is set at 2000 W/.m^2. K.

The temperature of the hot utility is set at 330°C (and, as a simplification, it is assumed to behave isothermally). The temperature of the cold utility is set at 15°C (and again it is assumed to act isothermally).

The cost of the hot utility is assumed to be 0.012 \$/kWh. While that of the cold utility is 0.001 \$/kWh. The plant is assumed to operate for 8000 hours/year.

The capital cost of installed heat exchangers is assumed to be given by the equation:

$$C = 16000 + 3200A^{0.7}$$

The interest rate is set at 10%. The plant life is set at five years.

Range targeting indicates that the minimum temperature approach for the network design should be set at 27°C. The targets for the design are listed in Table 2.

4.1. Solution Using Pinch Design Method. Given a minimum temperature approach of 27°C the process Pinch is found to be located at a hot stream temperature of 127°C and a cold stream temperature of 100°C.

The division of the problem is therefore that illustrated in Figure 17.

Starting with the design of a subnetwork for the system positioned above the Pinch, the streams positioned immediately above the Pinch are:

Hot (which are also the "in") streams	:	1 3 6 and 8
Cold (or "out") streams	:	2 7 and 9

Table 2. **Targets for Standard Problem**

Hot utility MW	Area m^2	No. of units	No. of shells	Capital cost, M\$	TAC, M\$[a]
25.7	9727	15	23	5.44	4.11

[a] TAC = Total annual cost.

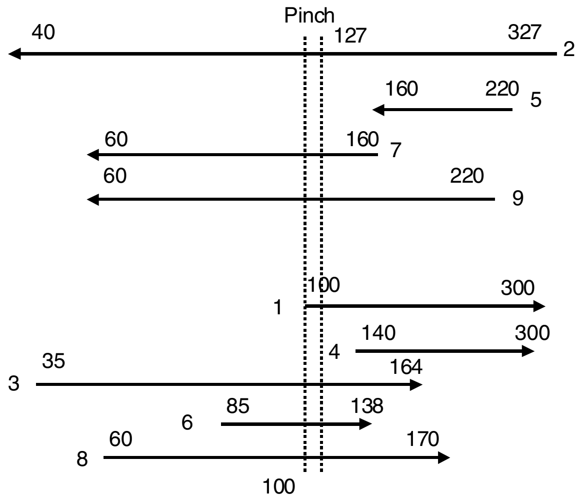

Fig. 17. Pinch division.

The CP Matrix for this population is given in Table 3.

Examination of the CP Matrix indicates a need to split stream 7. Looking at the values of the out stream CPs, a suitable split would be one stream having a CP of 340 (that can be matched with stream 5) and one having a CP of 60 (that can be matched with either stream 8 or stream 3). The resultant CP Matrix is shown in Table 4.

Now examine the heat loads associated with these streams. These are listed in Table 5.

Table 3. **CP Matrix for Above Pinch System**

Out:	8	3	1	5
In	60	70	100	350
9 60	/	/	/	/
2 100	X	X	/	/
7 400	X	X	X	X

Table 4. **CP Matrix Incorporating Stream Split**

Out:	8	3	1	5
In	60	70	100	350
7a 60	/	/	/	/
9 60	/	/	/	/
2 100	X	X	/	/
7b 340	X	X	X	/

Table 5. **Pinch Match Heat Loads**

Stream	CP	Supply temp.	Target temp.	Load
1	100	100	300	20.0
3	70	100	164	4.48
6	350	100	138	13.3
8	60	100	170	4.2
2	100	327	127	20.0
9	60	220	127	5.58
7a	60	160	127	1.98
7b	340	160	127	11.22

Table 6. **Remaining Duties for Above Pinch Design**

Stream	Supply temp.	Target temp.	Load
9 D2 Product	220	197	1.38
5 R2 Product	220	160	9.60
4 R2 Feed	140	300	32.00
6 R2 Recycle	132.06	138	2.08
3 D1 Feed	128.3	164	2.50

Table 4 indicates that it is necessary to match streams 7a (split of separator feed) and 5 (R2 product) and streams 2 (R1 product) and 1 (R1 feed).

Table 5 indicates which matches are feasible. Table 6 indicates what the heat load on each of these matches should be.

The choice here is to either match stream 7a with stream 8 (D2 feed) and stream 9 (D2 product) with stream 3 (D1 feed) or match 7a with 3 and stream 9 with 8. Given the locations of these streams the second arrangement (this gives local integration around column D2) is choosen.

The first part of the network design is shown in Figure 18.

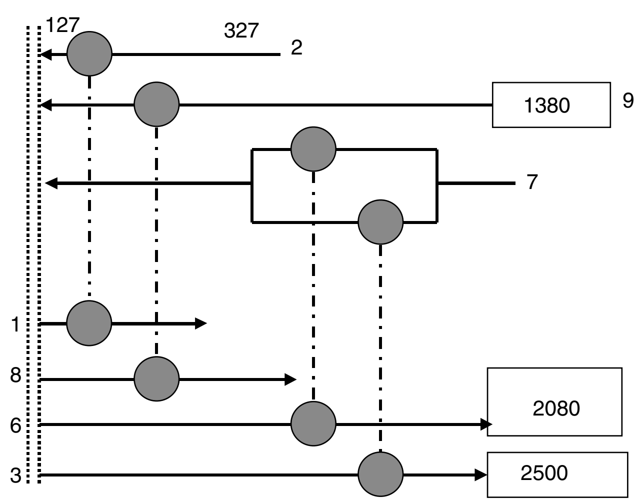

Fig. 18. Matches made immediately above Pinch.

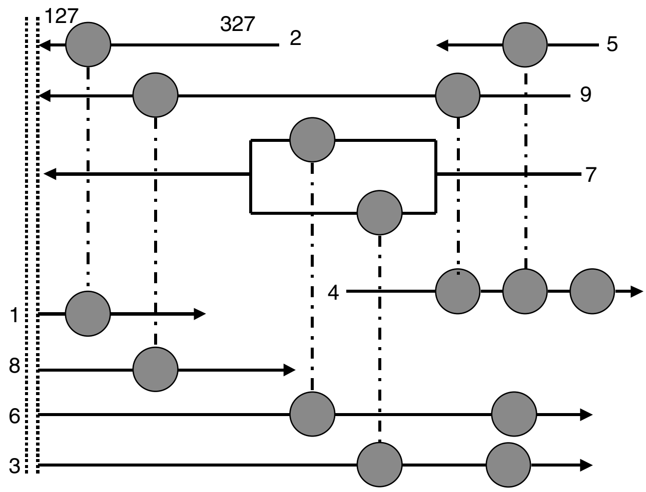

Fig. 19. Above Pinch design.

The remaining duties are listed in Table 6.

As a check, the loads on the hot streams (which comes to 10.98 MW) are totaled and compared with the sum of the loads on the cold streams (which is 36.58 MW). The difference is 25.6 MW which equates with the predicted hot utility demand.

The remaining matches can now be made. Given the location of the matches, stream 5 (R2 product) would match with stream 4 (R2 feed) and stream 9 (D2 product) with stream 6 (R2 recycle). The heaters would be used to complete the duties on streams 3, 4 and 6.

The final arrangement for the above Pinch network is shown in Figure 19.

Now the synthesis of a network dealing with the system positioned below the Pinch are described. The streams present are shown in Figure 20.

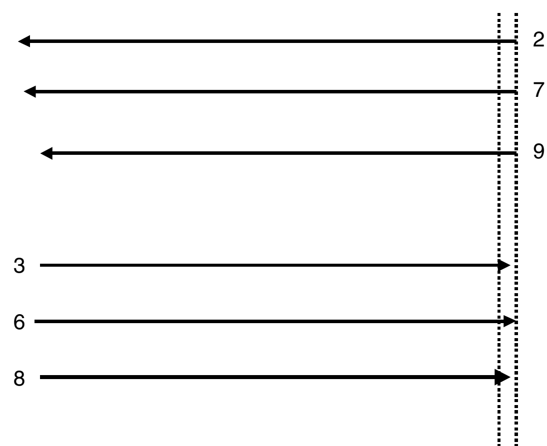

Fig. 20. Stream population below Pinch.

Table 7. **CP Matrix: Below Pinch System**

Out	9	2	7
In	60	100	400
8 60	/	/	/
3 70	X	/	/
6 350	X	X	/

Table 8. **Below Pinch Heat Loads**

Stream	Supply temp.	Target temp.	Load
2	127	40	8.7
7	127	60	26.8
9	127	60	4.02
3	35	100	4.55
6	85	100	5.25
8	60	100	2.40

All of the streams are present immediately below the Pinch. The CP Matrix is given in Table 7.

It is seen from this matrix that just one system of matches avoids the use of stream splits (the matches formed down the matrix diagonal).

The stream loads are given in Table 8.

In all of the matches it is the cold stream load that is taken up. The below Pinch network design is straight forward and is shown in Figure 21.

The two designs are now merged and the result is shown in Figure 22.

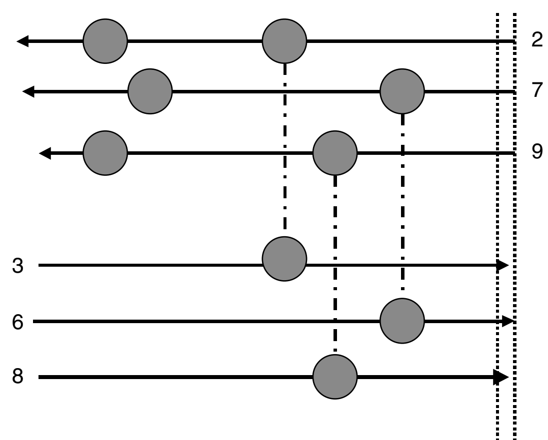

Fig. 21. Below Pinch design.

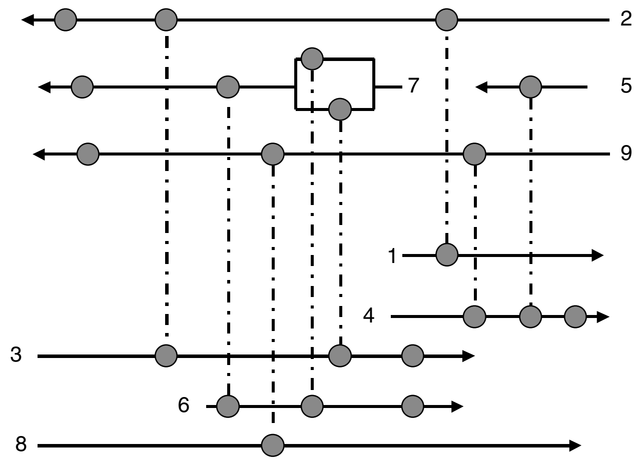

Fig. 22. Initial design.

The costs of this design are listed and compared with the original targets in Table 9.

The initial design is seen to have a total annual cost that is just 14.6% above the predicted target.

The targeting algorithm assumes that the minimum number of units equates with the summation of the two subnetwork values. However, identical matches (between streams 9 and 8) appear adjacent to each other, one on each side of the Pinch. These are merged into a single unit. Hence, the initial design has one fewer units than predicted by the cost algorithm.

The total number of units used in the design is fourteen. The minimum number required for a system having nine process streams and two utilities is ten. Hence, the scope for an energy/capital trade-off or a capital (units)/capital (area) trade-off is obtained.

Table 9. **Solution Derived Using Pinch Design Method**

Element	No. of shells	No. of units	Area	Capital cost	Hot utility	TAC[a]
2/1	8	1	3,837	2,055,000	0	542,000
9/8	6	2	1,190	872,000	0	242,000
5/4	8	2	3,461	1,921,000	21.0	2,525,000
7a/6	5	2	1,765	1,084,000	2.1	492,000
7b/3	2	2	388	313,000	2.5	330,000
9/4	2	1	86	132,000	0	35,000
2/3	3	2	815	533,000	0	174,000
7/6	4	2	1,485	869,000	0	370,000
Total	*38*	*14*	*13,027*	*7,779,000*	*25.6*	*4,710,000*
Target	*23*	*15*	*9,727*	*5,440,000*	*25.7*	*4,110,000*
Difference	*+15*	*−1*	*+33.9%*	*+43%*		*+14.6%*

[a] TAC = Total annual cost.

Examination of the structure indicates that most of the loops are associated with the presence of heaters or coolers. Since, these units will be required for start-up and provide an effective means of controlling the plant, their removal is not advised.

The obvious change that could be made to the heat recovery structure would be to either eliminate one of the matches involved in the stream split on stream 7 or move the match between stream 7 and 6 so that it can be merged with a similar match on the split stream. The nature of stream 7 is discussed in section 5.

5. Shortcomings of Pinch Design Method

The Pinch Design Method appears to be both simple and elegant. However, it contains a number of significant dangers for the unwary. The methodology should therefore be used in as part of a design approach that avoids its pitfalls.

The methodology has four major shortcomings:

1. It ignores practical aspects of exchanger design and application.
2. It make any consideration of plant layout.
3. Its application becomes more complex as number of streams involved increases.
4. The optimization result is dependant upon both the starting structure and the order in which heat recovery matches are eliminated.

Considered now is how exchanger practicality can affect the validity of the network design methodology. The individual synthesis steps are considered. First, is Pinch division. The division of the synthesis problem into two parts can result in a need for more than one heat exchanger handling a given stream. This is not always practical or desirable. Next, is the development of network structure at the Pinch. Here the engineer follows a set of rules relating to stream splitting. It is not always possible to split process streams. The procedure can indicate that there is only one acceptable stream matching arrangement at the pinch. If one of the matches is impractical no guidance is given regarding a feasible structure. Next, the designer is required to maximize the heat loads on a match. This can again give rise to the need for more than one exchanger on a given stream. Finally, having dealt with the matches in the Pinch region, the designer is asked to just finish the problem. Now some duties have a maximum temperature difference (eg, many forms of vaporizer) as well as a minimum temperature difference. After the Pinch matches have been set and loads maximized, the temperature differences for the remaining streams can be so high that this constraint is infringed.

The worked example given in section 4 demonstrates this type of problem. The derived network structure (Figure 22) has stream 7 passing through four separate exchangers. It is also involved in a stream split. Examination of the process flow diagram (Fig. 16) indicates that this stream flows from unit E (which is not defined) to a gas separator. This suggests that stream 7 is a two-phase stream. Such streams are difficult to handle for the liquid and gas phases will separate in piping branches, within tube bundles and in exchanger headers. They should not be split and the number of exchangers using such streams should be minimized.

Failure to consider plant layout is potentially serious. It can result in the engineer developing structures that are expensive, impractical or unsafe. For instance, only liquid streams can be cheaply transported over large distances. The costs of piping gases, vapors or two-phase flows over long distances are prohibitive in terms of both capital and power costs. The transportation of process streams between differing operating areas can result in operators finding plant operation difficult to understand.

The solution to the above problems is systematic problem decomposition. Here, the overall problem is broken down into a number of self contained problems set by a consideration of both practical and layout issues. Since, this decomposition leads to a number of small problems, it avoids the difficulties inherent in applying the methodology to a large number of streams. Since, it results in simple structures containing elements that have been subjected to optimization, the approach removes the need for complex final optimization.

6. Systematic Problem Decomposition

One approach that can be adopted for the analysis of large problems is to divide the overall problem into a number of smaller ones and to evaluate the consequences of treating the sub-problems individually: applying localized integration rather than plant wide integration (11).

The procedure is outlined in Figure 23 and is as follows. A self-contained zone is identified (usually through examination of the process flow diagram). The total cost analysis described above is then applied to the overall problem, to

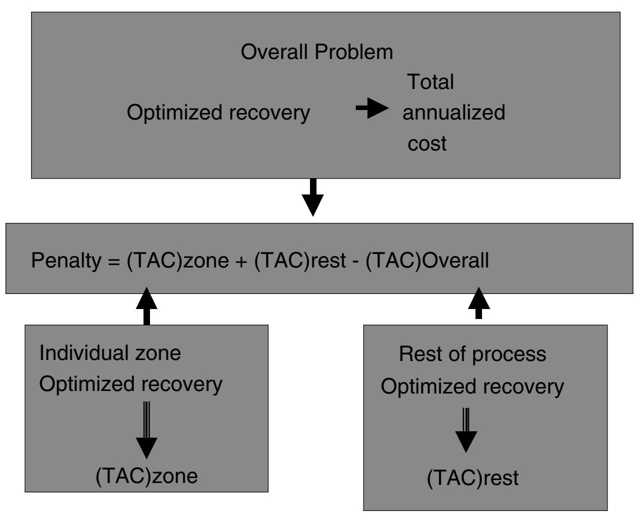

Fig. 23. Problem decomposition analysis.

the identified zone and to the rest of the problem (the overall problem minus the streams included in the zone). Predictions for optimum utility consumption, capital cost and total annual cost for the zone are obtained for the rest of the problem and for the overall problem. The problem decomposition can be evaluated using any of these criteria. The most informative criterion is usually the total annual cost. Add the total annual cost predicted for the zone to that predicted for the rest of the problem. Then, by comparing the value predicted for the overall problem, the possible overall economic consequence of only applying localized integration within the zone is obtained.

If the penalty is large this usually indicates a significant reduction in heat recovery. (This can be quickly confirmed by examining the predicted utility consumptions).

If there is an energy penalty the Pinch point for the zone will differ to that for the rest of the problem. If the Pinch point for the zone is at a higher temperature than that for the rest of the problem the energy efficiency can be improved by either transporting a hot stream (or, streams) at its Pinch Temperature out of the zone or transporting a cold stream (or, streams) at the Pinch temperature identified for the rest of the process into the zone (Fig. 24).

If capital cost is the dominant factor, add the capital cost predicted for the zone with that for the rest of the problem. If there is a large capital cost penalty, this generally indicates a tightening of the separation between the composite curves. One can identify where this is likely to be occurring through examining the composite curves for the three systems.

This type of decomposition can be conducted in stages. Large problems are decomposed into zones with identified linkages. These zones are then further divided to identify smaller self-contained zones.

In some cases the approach can be taken all the way to the identification of individual heat recovery matches. It is not necessary to subject the final network structure to final optimization for energy-capital trade-off calculations are embedded in the procedure.

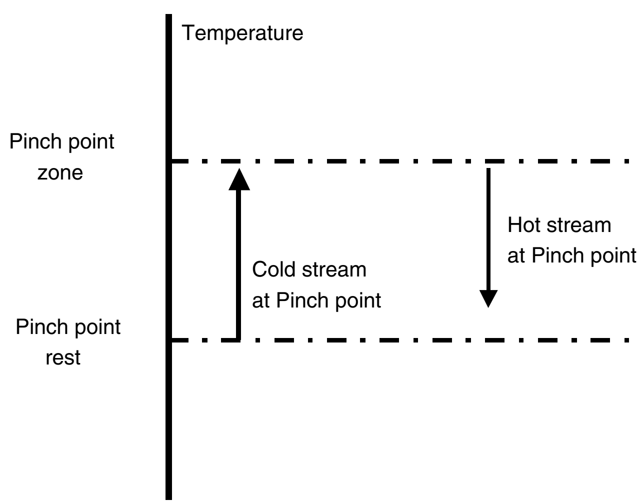

Fig. 24. Pinch points identify interzonal transfers.

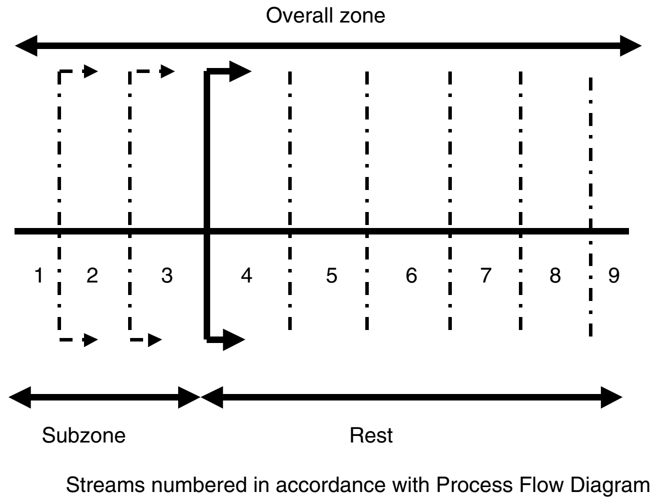

Fig. 25. Systematic application of analysis.

The process flow diagram should be used to guide the problem decomposition. One simple level of computerization is to list the streams in accordance with the process flow diagram (where possible interspersing hot and cold streams) and then systematically moving a boundary between the streams and applying the analysis to the identified systems (Fig. 25). Then, ranking the results and selecting a self contained zone. The procedure is applied again to the remainder of the problem (Figure 26).

The procedure is repeated until acceptable subdivision can no longer be identified. The Pinch Design Method can then be applied to obtain solutions for each of the identified zones.

6.1. Solution Using Problem Decomposition. Considered again is the example problem introduced in section 4.

The streams have already been number in accordance with the process flow diagram (Fig. 16).

Applying decomposition analysis, using the procedure illustrated in Figure 25, the results in Table 10 are obtained.

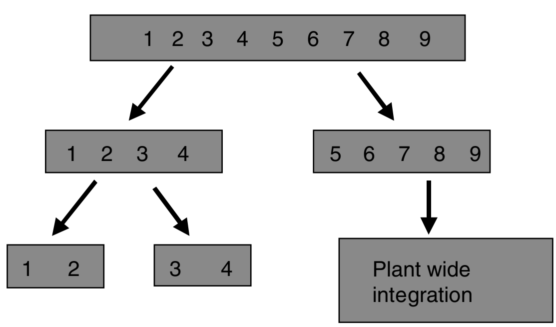

Fig. 26. Progessive decomposition.

Table 10. **Level 1 Analysis**[a]

Zone/Rest	TAC (zone)	TAC(rest)	TAC(total)
1–9	4.112	–	4.112
1-5+6-9	3.331	0.853	4.184
1-7+8-9	4.034	0.165	4.199
1-3+4-9	0.890	3.352	4.242
1-2+3-9	0.622	3.698	4.320

[a]TAC = Total annual cost.

Table 11. **Level 2 Analysis (Zone A)**[a]

Zone/Rest	TAC(zone)	TAC(rest)	TAC(total)
1-5	3.331		3.331
1-3+4-5	0.890	2.557	3.447
1-2+3-5	2.022	1.691	3.713

[a]TAC = Total annual cost.

The total annual cost predicted for the overall problem is 4.112 M$. The combined costs for a system having a zone incorporating streams 1 to 5 and a zone containing streams 6 to 9 is predicted to be 4.184 M$. The difference between the two predictions is very small. The difference of 0.072 M$ is just 1.75 % of the overall target. This problem division is accepted.

The analysis is repeated for each of the identified zones. In the case of the first zone, the data in Table 11 are obtained. The separation of streams 4 and 5 from the problem lead to a total annual cost penalty of 0.116 M$. This is just 2.8% of the value predicted for the overall problem whereas, as seen earlier, the accuracy on this target is likely to be +/− 15%. The arrangement is now down to the individual match level. This match is accepted.

The analysis for the second zone yields the results shown in Table 12. Here it is found that by making two individual matches (one between streams 6 and 7, the other between streams 8 and 9) yields a design that has a total annual cost that has a cost penalty of 0.127 M$ (for the zone). This is 3.1 % of the original target. These matches are accepted.

Three heat recovery matches (4/5, 6/7 and 8/9) have been quickly identified and the size of the problem has been reduced to just three stream (1-3). Since, the cost analysis involves the trading-off of utility and capital costs, the heat recovery levels on these matches have all been optimised. All of the matches have been made locally – so none of the streams are being transported large distances. No

Table 12. **Level 2 Analysis (Zone B)**[a]

Zone/Rest	TAC (zone)	TAC (rest)	TAC (total)
6-9	0.853		0.853
6-7+8-9	0.815	0.165	0.980

[a]TAC = Total annual cost.

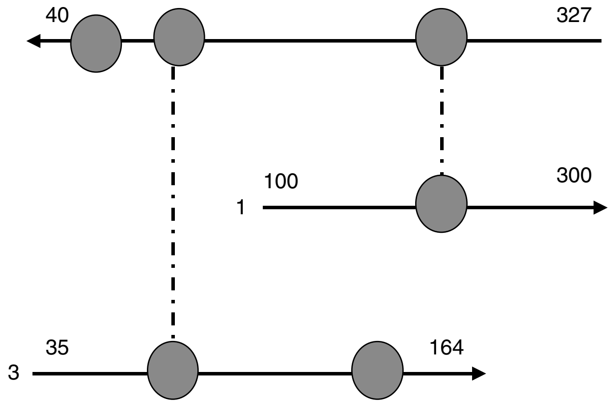

Fig. 27. Structure for remaining three streams.

stream splits have been used and all of the matches have involved the minimum number of exchangers.

Finally attention is turned to the streams 1 to 3. The temperature ranges spanned by these streams suggests that the structure illustrated in Figure 27 should be used.

Like stream 7, stream 2 is possibly a two-phase stream. If this is the case, its use in more than one heat recovery match may be undesirable. In this case, it is the match with stream 3 that should be rejected.

The obvious question is how does this effect the problem decomposition? If stream 2 is only to be used in a single match would one of the other cold streams be a better option. Examination of the location of the Pinch for the overall problem indicates that the match with stream 1 absorbs all of the heat available from the part of the stream situated above the Pinch. The total annual cost of the match is 0.497 M$.

If the second heat recovery match is acceptable, energy-capital trade-off calculations indicate an optimum heat recovery of 3.64 MW and a total annual cost for the design element (heat recovery unit, heater and cooler) of 0.602 M$.

The design is now complete and is summarized in Table 13.

Table 13. **Summary of Design Derived Using Problem Analysis**

Element	No. shells	No. units	Area	Capital cost	Hot utility	TAC[a]
6/7	10	2	4434	2.44	0	0.815
8/9	3	2	819	0.535	0	0.165
4/5	6	2	2281	1.542	22.4	2.557
3/2	5	3	1285	0.858	3.6	0.602
1/2	8	1	3937	2.055	0	0.542
Total	*32*	*10*	*13356*	*7.43*	*26*	*4.681*
Target	*23*	*15*	*9727*	*5.44*	*25.7*	*4.112*
Difference	*+9*	*–5*	*+37%*	*+37%*	*+1.3%*	*+13.8%*

[a] TAC = Total annual cost.

The Table also shows the 'targets' derived from the application of the pinch analysis to the overall problem. The difference between the targeted and final utility consumption is small (1.3%). Major differences are observed for the factors affecting capital cost. The number of individual exchangers used is large. This is due to the close temperature approaches made. The network area is much greater (+38%) than the target value. The result is much higher capital cost (+38%). Despite these quite large differences, the results are marginally better than those obtained using the Pinch Design Method (where capital cost is 43% higher than initial prediction).

The comparison highlights shortcomings in the capital cost models used in Pinch Technology.

7. Practical Design Considerations

Examination of Table 13 immediately shows one problem associated with the use of shell-and-tube exchangers: at close temperature approaches the designer needs to use a number of shells in series (in the case of the 6/7 match the need is for ten shells distributed over a single heat recovery exchanger and a cooler) in order to achieve the required performance.

In some instances the need for a series of shells can be removed through the use of a unit with an internal longitudinal baffle (named an F shell). However, a better approach may be to use a plate-and-frame exchanger rather than a shell-and-tube unit.

Plate-and-frame heat exchangers have a major cost and weight advantage over shell-and-tube units. However, in the context of network design they have a further distinct advantage. In Figure 28 (taken from Ref. 12) one sees the plant arrangement necessary if shell-and-tube exchangers are used on a simple heat

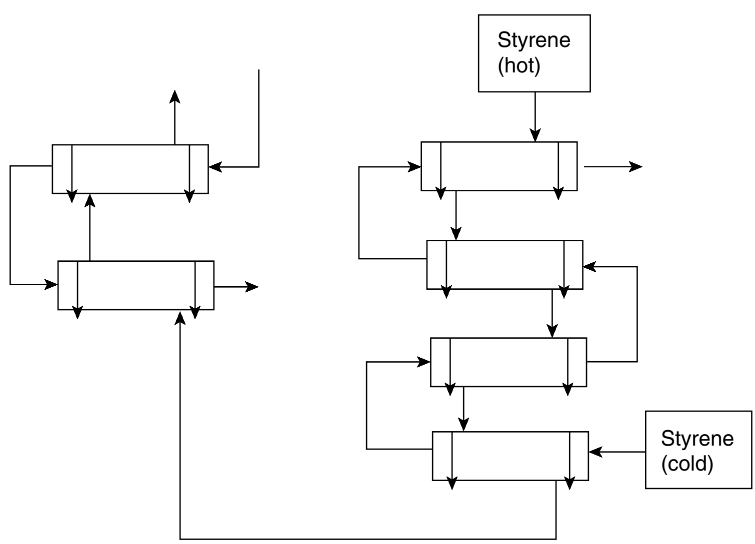

Fig. 28. Shell-and-tube design for heat recovery/trim cooling system.

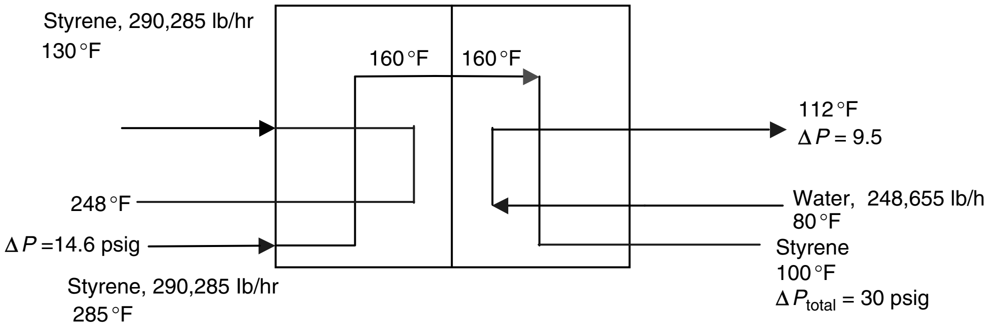

Fig. 29. Multistream plate-and-frame design for heat recovery/trim cooling system.

recovery/trim cooling system. The recovery unit requires four individual exchangers. The trim cooling requires two exchangers. Plate-and-frame exchangers can handle more than one hot and one cold stream in a given unit. This results in simpler plants and significant savings in piping costs. In Figure 29 (also taken from Ref. 12) a plate-and-frame exchanger (set up as a multi-stream exchanger) that can be used to satisfy the duty performed by the system shown in Figure 28 is shown. The benefit of pure counter flow removes the need for exchangers in series in order to maintain a good temperature driving force. The ability to multistream eliminates the need for two separate units.

This brings up the question of when exchanger selection should be undertaken.

The design of a process plant is undertaken in stages. The first stage is often called *Process Synthesis*. The objective is the generation of the basic flowsheet for the process, the derivation of the heat and mass balance for the process and the generation of specifications for the design of individual items of equipment.

The process synthesis stage is generally the province of the process engineer and the number of personnel employed on a project is low. However, once approval has been given, the project enters the detailed design stage where the number of personnel involved increases markedly as the range of engineering functions increases. During this stage a number of activities are undertaken simultaneously.

In most cases exchanger selection is not considered during the process synthesis stage. This situation is being reinforced by the linkage that is being made between process simulators and heat exchanger programs that solely handle shell-and-tube heat exchanger analysis.

When exchanger selection is left until the later stages of design the full benefit of moving towards an alternative technology cannot be taken for many parameters have already been set on the assumption that shell-and-tube exchangers will be used. The work of other engineers is based on these parameters. The result is that selection is based upon purchase prices and not total cost.

Compact heat exchangers offer the following benefits: large reduction in size and weight; pure counter-flow arrangement; and ability to handle more than one hot and one cold stream in a single unit.

The installed cost of a heat exchanger is very dependant upon exchanger weight and space requirements. Since, the installed cost is up to four times the original purchase price, making decisions on the basis of purchase price alone is false economy. Many of the benefits relating to weight and space savings can only be realized if the decision to use a compact design has been made before plant layout is fixed.

Current design procedures are based on the use of shell-and-tube technology. Applying the cost relationships for this technology results in much lower heat recovery than using costs for plate-and-frame exchangers (13). Since, the structure required for a heat recovery network can change as the heat recovery level changes the choice of exchanger technology affects network design.

Exchanger selection must be undertaken at the process synthesis stage if the most economic heat recovery level is to be properly identified.

7.1. Factors to be Considered for the Design of a Heat-Exchanger Network.

1. In the eagerness to maximize heat recovery, it is possible to overdo it. It must always be borne in mind that heat recovery is justifiable only to the extent it is economical. As argued earlier, the cost of energy recovered must always be balanced against the equipment cost to recover energy.
2. The heat-exchanger network must be workable from a process viewpoint, especially for different operations. This means that various temperature constraints or requirements such as for pump-around streams must be fulfilled at all operating scenarios (such as different feedstocks). If a hot stream goes from a heat-exchanger network to a reboiler, it has to be ensured that it is hot enough for proper reboiler operation under all operating conditions. In order to achieve these goals, partial or total by-passing of one or more heat exchangers has to be anticipated in advance.
3. Practical equipment operability and maintenance should be ensured. For example, stream velocities must be maintained sufficiently high so as to minimize fouling under all plant operations, notably processing different feedstocks and turndown operation. In order to achieve this goal, heat exchangers have often to have multiple units in parallel so that at times of lower throughout, one or more of the multiple shells can be by-passed.

7.2. Software. A range of software tools is available. The degree of complexity varies from source to source. The methods described above can be employed using simple targeting software. In the hands of someone who understands the technology, these methods yield better results than those from any of the automated methods. The reader is steered away from automated method by describing some of the pitfalls of the fixed approach. While specific to the Pinch Design Method, the arguments also apply to automated methods. The developers of these methods have not truly incorporated the practical considerations the designer needs to make.

Data to be Furnished to the Software.

1. Flow rates of all hot and cold streams.
2. Inlet temperatures and target outlet temperatures of all hot and cold streams.

3. Specific heat values of all hot and cold streams. As the heat capacity of any stream will vary significantly over a large temperature range, some programs permit the subdivision of streams in order to accommodate changes in properties. This is also the case for heat transfer coefficients.
4. Assumed individual stream heat transfer coefficients for all streams.

   Individual stream values are used not only for the prediction of network area but in some programs, they are used in the determination of the overall heat-transfer coefficient for any heat exchanger in the network.

   Stream heat-transfer coefficients can be specified in a number of ways. The original approach used experience values. These assumed values depend principally upon the flow rates and viscosities of the various streams. The more authentic these values, the more authentic will be the network that will be synthesized by the program.

   An alternative that has recently been introduced is linkage of process integration and software package, such that rigorous thermal rating of the individual exchangers can be undertaken. However, this can only be done after the process flowsheet has been generated and, of course, the construction data have to be keyed in. Consequently, the approach requires some level of iteration between design functions.

   Finally, the algorithm proposed in Ref. 6 can be used to derive film heat-transfer coefficients from values of allowable pressure drop or specified velocity. It can also be used to determine likely pressure drop from specified heat transfer coefficient or velocity. All these can be done at the targeting stage. No iteration between design functions is necessary.
5. Fouling resistance of each stream.
6. Wall resistance of the tube material. As tube wall resistance invariably contributes only 2–4% to the overall heat transfer resistance, it is not very significant.
7. The maximum heat transfer area that can be incorporated in a single shell. This is based upon the bundle-pulling limitation for removable-bundle heat exchangers. Whenever the required heat transfer area of a certain heat exchanger in the network being synthesized exceeds this limit, the program will consider multiple shells, usually in series, as this gives an MTD advantage when compared to multiple shells in parallel. Even otherwise, whenever there is a temperature cross (cold stream heated to a temperature greater than the outlet temperature of the hot stream), the program will select the required number of multiple shells in series.
8. Some software also allow the user to specify the minimum heat duty that should be considered for any match. This is because if a network incorporates heat exchangers handling small heat duties (with small heat transfer areas), the number of shells may be inordinately high, as a result, the total initial cost of all the heat exchangers will be very high.
9. A lower ΔT_{min} will translate into a higher heat recovery between the hot and the cold streams and thereby, lower hot and cold utility requirements. However, as the MTDs will be lower, this will be at the expense of higher heat transfer area and thereby initial cost of all the heat exchangers.

Thus, as described earlier, there is an optimum $\Delta T_{\min}$ that depends upon the nature of the process unit, the cost of energy and the cost of heat transfer equipment. The latter would mean the cost of not only the heat exchangers themselves but also that of piping, foundations and instrumentation.

When the cost of energy tends to be relatively high, the $\Delta T_{\min}$ will tend to be lower, and vice versa, hi order to determine the optimum $\Delta T_{\min}$, some quick runs (with the program working on an automatic basis) will have to be taken for various values of $\Delta T_{\min}$ to determine the total heat recovered and the requisite total heat transfer area. Thereafter, a differential payout period can be applied, starting the highest value of $\Delta T_{\min}$ considered and working down to the lowest, to determine the optimum value. This optimum value of $\Delta T_{\min}$ in most units in the chemical process industries will usually vary between 10°C and 14°C.

How the Software Packages Work. As a first step, the hot and cold composite curves and the grand composite curve are generated. The problem is decomposed into two distinct areas, one above the pinch and the other below the pinch, and matches are made Progressively as per the principles of Pinch elaborated in the earlier section. For the section above the Pinch, the matches are started from the Pinch and move away from it on the T-H plot while for the section below the pinch, the matches are started at the Pinch and once again move away from it on the T-H plot. All systems need not have both sections.

Most software packages work in the automatic mode but some work in the interactive mode as well. In the automatic mode, the software package makes the best matches in order of thermodynamic efficiency. However, in the interactive mode, the software packages indicate all the feasible (legal) matches (again in order of thermodynamic efficiency) at each point in the network. Thermodynamic efficiency really means the heat duty transferred per unit area of heat transfer. This feature is incorporated so that the designer has the flexibility of ignoring certain matches which are forbidden or undesirable (for example, by virtue of incompatibility in flow rates or difficulty in control or long pipe runs due to physical location) or are very small (with respect to heat duty and/or heat transfer area) and thereby uneconomical.

Once the heat-exchanger network is generated and all process-to-process heat exchange maximized to the extent viable, the feasible MP and LP steam generators are identified from the Grand Composite Curve (GCC) and the steam pressure levels. It should be noted that since process-to-process heat exchange has already been maximized, such steam generation is not at the expense of process heat exchange. Thus, the grand composite curve is independent of integration scheme. The usual practice is to use the GCC to identify steam generation opportunities. Then use range targeting to optimize these. The identified opportunity is then incorporated into the stream set as process stream.

7.3. Practical Application. *Crude Preheat Train.* The crude preheat train in an oil refinery represents a very common and yet a very special application of a heat-exchanger network, with just one cold stream (crude oil) to be heated through a very large temperature range by several hot streams having overlapping temperature spans. As such, it is worthwhile to bring out certain special features for the design of these preheat trains.

Special Features.

1. Looking at typical hot and cold composite curves (see Fig. 5) it can be observed that instead of the presence of a distinct and clear Pinch point, the curves are close together over quite a large temperature range.
2. There are three distinct sections in a crude preheat train: (*1*) a cold section from storage to the desalter, (*2*) a middle section from the desalter to the preflash vessel or column and finally (*3*) a hot section from the preflash vessel or column to the furnace. The operating temperature of both the desalter and the preflash vessel or column can only be varied over a very small temperature span as these are very important to the efficient operation of the unit. Furthermore, there is a temperature drop across the desalter (usually 5°C) and a flow rate and temperature change across the preflash vessel or column.
3. Because of steps 1 and 2 above, the conventional Pinch Design Method is not suitable for use in designing crude preheat trains. Firstly, one has a pinch region as opposed to a distinct Pinch point. The desalter and preflash temperatures generally occur within this region. These processing temperatures are more significant than the precise Pinch point. It would be better to basing the design on these temperatures than on the Pinch location.

   The usual practice is to start at the hot end of the preheat train and then move away towards the preflash, the desalter and eventually to the cold end of the train. Matches are made on the basis of both the flow rate and the temperature levels of the hot streams. The hottest crude will evidently be heated by the hottest hot streams as temperature difference is the driving force for heat transfer. Sometimes, however, this can lead to very high wall temperatures which promote fouling. Therefore, while maximizing preheat, the selection of services should be done carefully so that wall temperatures are within limits.

   The hot stream flow rate is also very important, as a larger flow rate will be able to sustain a larger extent of crude preheat. The temperature difference between the hot streams and the crude is not allowed to go below the $\Delta T_{\min}$.
4. Crude is almost always split into two or sometimes even more parallel streams for the following reasons: (*1*). The number of heating streams being large and the flow rates of many of them being relatively low compared to that of crude oil, a single train would require a very large number of relatively small shells. This is because the heat duty of any given shell would tend to be low as the hot stream temperature will drop rapidly when exchanging heat with the crude oil having a much higher flow rate. By splitting the crude oil into two or three parallel legs, the flow rates of the hot streams are much better balanced against the crude oil flow rate, thereby increasing heat exchange capacity of the heat exchangers. This finally results in a much smaller number of heat exchangers. (*2*) By varying the split of crude oil into the parallel legs of the network, it is possible to have a good degree of control and flexibility in a network having multiple

legs in parallel, something which would be totally missing in a single-leg network.

5. As the number of hot streams is large and their temperature ranges overlapping, a number of interspersed repetitive matches (sometimes called serial matches or multilevel heat transfer) is an usual feature of crude preheat trains. Some hot streams exchange heat with crude in three or even four interspaced matches. The reason serial matching occurs is that the temperature spans of the hot streams cross the important process temperatures (de salter, pre flash). The heat recovery from these streams must therefore be distributed between the differing sections of the pre-heat train. Within a given section stream splits are used in order to make good use of temperature driving force.

7.4. Process Limitations.

1. Oil refineries invariably process crude oils from several sources or even their blends. With variation in the source and thereby the composition of the various crude oils, the flow rates, starting temperatures and viscosities of the different product and pump-around streams vary significantly. The final preheat train configuration will have to be such that it represents good heat recovery across the entire spectrum of operations. Evidently, this calls for an iterative approach.
2. Pump-around streams represent fixed heat duty sources. The inlet temperature being fixed, the flow rates and outlet temperatures of these streams can be varied. The benefit of a higher flow rate is that we get a higher T-Q profile. The result is a larger temperature driving force for exchanger design and lower capital cost. However, the fouling consideration must be kept in mind.
3. The crude oil temperature at the desalter and at the prefractionator are precise values for good operation which can be varied within a very small temperature range.

In the final analysis, the crude preheat train should have sufficient flexibility to achieve economical heat recovery over entire plant operating range after meeting all the process and operating constraints.

7.5. Overall Design Methodology. As pointed out earlier, crude preheat trains have to be designed for processing varying crude oils with varying flow rates, temperatures and physical properties of the hot streams. Consider the case where two disparate crude oils are to be handled in a refinery. Two approaches are possible. In the first, a network is synthesized for one crude and then checked for the other. Evidently, there would be severe limitations which would have to be overcome. The modified network would then have to be checked for the first crude. This entire process would have to be repeated until an acceptable network evolved.

In the second method, two different networks are synthesized for the two cases. Evidently, these would be very different. From the two networks, a "hybrid" network can be developed, incorporating the important features of each as best possible. This hybrid network would then have to checked rigorously for each case with further modifications incorporated until a satisfactory network evolved. It is generally better to follow this latter method as it is less cumbersome.

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