Chapter 2.1 in Handbook of Process Integration

Heat Integration – Targets and Heat Exchanger Network Design

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Abstract ......................................................................................................................... 2
1 Introduction to Key Concepts and Major Topics ....................................................... 2
2 Stages in Design of Heat Recovery Systems ........................................................... 6
3 Data Extraction ............................................................................................................ 8
  3.1 Data required for basic Heat Integration ............................................................... 8
  3.2 Major Challenges in Data Extraction .................................................................... 9
4 Performance Targets .................................................................................................. 12
  4.1 Minimum External Heating and Cooling ............................................................... 14
  4.2 Minimum Number of Heat Exchangers ................................................................. 18
5 Process Modifications ................................................................................................ 21
  5.1 Plus/Minus Principle and Appropriate Placement ............................................... 22
  5.2 Grand Composite Curve and Correct Integration ................................................. 25
6 Network Design ......................................................................................................... 28
  6.1 The Pinch Design Method ..................................................................................... 29
  6.2 Developing an initial MER Design ....................................................................... 30
  6.3 A Strategy for Stream Splitting ............................................................................ 33
7 Design Evolution ........................................................................................................ 35
  7.1 A Three-way Trade-off in Network Design ......................................................... 36
  7.2 Using Loops and Paths to evolve Network Design ............................................. 37
8 Concluding Remarks ................................................................................................. 42
9 Sources of further Information .................................................................................. 43
References ..................................................................................................................... 43
Abstract
This chapter describes the basic steps of Pinch Analysis for heat recovery that made Process Integration a methodology employed by numerous designers and engineers worldwide and made industrial leaders in the 1980s claim these concepts to be the results of academic research with the largest impact on industrial thinking related to design and operation in the process industries. Key elements of the chapter are the Heat Recovery Pinch, Performance Targets ahead of design, and a step-wise and systematic procedure for Heat Exchanger Network Design. One of the main characteristics of Pinch Analysis is the extensive use of Graphical Diagrams and Representations that give the designer a good overview of even the most complex processes. These tools provide insight and ease the communication between designers and engineers.

Keywords: Heat Recovery Pinch; Composite Curves; Performance Targets; Pinch Design Method; Design Evolution

1 Introduction to Key Concepts and Major Topics
This section provides a brief introduction to the history, the major topics and the key concepts of Heat Integration, while subsequent sections will provide more details about performance targets, graphical diagrams and representations, as well as design procedures that form the basis of the methodology, with a few simple and illustrating examples to assist the explanations in the text. The section repeats and expands parts of Chapter 1.2 on Basic Terminology of Process Integration. Heat Integration was an early ("pioneering") activity and is still an important part of the "discipline" that today is referred to as Process Integration.

The term Heat Integration has two meanings. First, it refers to the physical arrangement of equipment, process sections, production plants, entire sites, and even the surroundings in case of district heating or district cooling. Second, it refers to an area of Process Synthesis, with methods and tools aiming at increased energy efficiency in industrial processes and energy plants. Such improved energy efficiency can be achieved by combining (i.e. integrating) heating and cooling demands and thereby reducing the need for external heating and cooling
utilities. Efficient use of equipment is of course also part of the scope, since energy efficiency only becomes interesting and will be implemented if it is economically feasible.

The birth of Process Integration as a Systems oriented design activity is related to the discovery of the concept referred to as the *Heat Recovery Pinch*. A methodology coined Pinch Analysis was developed in the late 1970s and early 1980s, which resulted in a departure from traditional design practice that had been based on extending and improving process technologies by the use of operating and engineering insight (i.e. following the “learning curve”) and to choose the best design from a set of case studies. Of course, there would occasionally be discoveries and breakthroughs that made step changes possible, but the uncertainty whether designs could be further improved and by how much, was still left with the designer. Most of the text in this chapter on Heat Integration will focus on increasing Energy Efficiency of processing systems, using Pinch Analysis as the main methodology.

The most important new feature in Pinch Analysis was the ability to establish *Performance Targets* ahead of design only based on information about the change in thermodynamic state for the process streams, described in table form and hereafter referred to as Stream Data. These targets were first developed for thermal energy (external heating and cooling) and have later been extended to mechanical energy (power or shaft work), number of heat exchangers, and total heat transfer area. In fact, by using analogies, the original Heat Pinch concept and the idea of targeting ahead of design have been re-used (as indicated in Chapter 1.2 and several other chapters of this Handbook) in other areas such as mass exchange processes (Mass Pinch), wastewater minimisation and distributed effluent treatment systems (Water Pinch), hydrogen management in oil refineries (Hydrogen Pinch), and oxygen consuming processes (Oxygen Pinch). Similar ideas have also been applied to study for example Supply Chains and Carbon Emission reductions.

The concept of a Heat Recovery Pinch was independently discovered and developed by Hohmann (1971), Huang and Elshout (1976), Linnhoff et al. (1978,1979) and Umeda et al. (1978). The basic idea is to draw separately the total heating and the total cooling requirements of a process in a cumulative manner in a Temperature – Enthalpy diagram, commonly referred to as *Composite Curves*. Similar drawings had already been used by designers of low temperature processes such as air separation (Linde AG in Germany) to
design multi-stream heat exchangers (Gundersen, 2000). The graphical construction of Composite Curves is explained with an example in Section 4.1.

Heat recovery between hot and cold streams is restricted by the shape of the Composite Curves and the fact that heat can only be transferred from higher to lower temperature. The minimum allowed temperature difference ($\Delta T_{\text{min}}$) is an economic parameter that indicates a near optimal trade-off between investment cost (heat exchangers) and operating cost (energy). The point of smallest vertical distance (equal to $\Delta T_{\text{min}}$) between the Composite Curves represents a bottleneck for heat recovery and is referred to as the Heat Recovery Pinch.

An alternative representation of the overall heating and cooling demands in a process is the Heat Cascade, a special case of the transshipment model from Operations Research. Here, hot streams (i.e. “sources” of heat) contribute to a set of temperature intervals (i.e. “warehouses” of heat), while cold streams (i.e. “sinks” of heat) draw heat from the same intervals. The temperature intervals are established on the basis of the supply and target temperatures of all process streams. A heat balance is made for each temperature interval, and any heat surplus in one interval is cascaded (thus the name) down to the next interval with lower temperatures. The Heat Cascade also forms the basis of the Grand Composite Curve (also referred to as the Heat Surplus Diagram), a very important tool for studying the interface between the process and the utility system (consumption and generation of various types of utilities, both load and level) and for evaluating heat integration of special equipment such as Distillation Columns, Evaporators, Heat Pumps and Heat Engines. The graphical construction of Heat Cascades and the Grand Composite Curve is explained with an example in Section 5.2.

Perhaps the most important property of the Heat Recovery Pinch is that it decomposes the process into a heat deficit region above Pinch and a heat surplus region below Pinch. There is not enough heat available in the hot streams above Pinch to satisfy the overall heating demand of the cold streams, and external heating is required. Below Pinch, there is not enough cooling available in the cold streams to satisfy the overall cooling demand of the hot streams, and external cooling is required. Based on the insight about this Pinch Decomposition, systematic, general and step-wise design procedures have been developed for heat exchanger networks with minimum energy consumption.
Pinch Decomposition also forms the basis of more general rules for design modifications as well as integration of special equipment, heat and power considerations and beyond. The **Plus/Minus Principle** states that any design modification aiming to reduce external heating and cooling should result in (1) an increase (“Plus”) in the duty of hot streams above Pinch or cold streams below Pinch, and/or (2) a decrease (“Minus”) in the duty of colds streams above Pinch or hot streams below Pinch, where (1) and (2) often can be combined. The **Appropriate Placement** concept is based on the same philosophy and can be stated in general terms as the placement (or integration) of heat sources above Pinch and heat sinks below Pinch.

As indicated above, Pinch Decomposition is the basis for the **Pinch Design Method** (PDM) adopted and extensively used in a large number of industrial sectors such as, but not limited to, oil & gas, chemical & petrochemical, pulp & paper, metal production, dairies, breweries and pharmaceuticals. Any heat leakage from the higher temperature region above Pinch (with heat deficit) to the lower temperature region below Pinch (with heat surplus) results in an increase of both hot and cold utilities; thus there is a **Double Penalty** associated with such heat transfer. The main philosophy behind the PDM (Linnhoff and Hindmarsh, 1983) is thus to avoid any cross Pinch heat transfer by designing separate heat exchanger networks above and below Pinch. PDM is a step-wise (sequential) design procedure that provides rules for selection of hot and cold streams to be matched in heat exchangers, the sequence and duty of the heat exchangers, the need for stream splitting, etc.

This chapter describes tools and methods for designing new plants, often referred to as **Grassroots Design**, while a much more common activity in industry is the improvement of existing plants by making investments that increase production volumes and/or process efficiency. The latter is often referred to as **Retrofit Design** where, as explained in Chapter 2.5, the main objective is to reduce energy consumption in existing process plants by increasing the level of heat recovery. The design project then focuses on investing in some new equipment, repiping or changing internals in heat exchangers, while making maximum use of existing units. Heat exchangers are tailor-made, thus there is no second hand market and equipment that is purchased and installed should be kept as part of the modified design.
2 Stages in Design of Heat Recovery Systems

As indicated above, there is a fundamental difference between designing heat exchanger networks for new designs (“grassroots”) and modifying heat recovery systems for existing plants (“retrofit”). At the same time, the design process contains the same stages for the two cases, however, with quite different content, as will be explained in Section 3.2 and Chapter 2.5. More specifically, the stages in design of heat recovery systems are:

- Data Extraction
- Performance Targets
- Process Modifications
- Network Design
- Design Evolution
- Process Simulation

**Data Extraction** means collecting and processing data about heating and cooling requirements as well as the need for evaporation and condensation of process streams, often referred to as Stream Data. Similar data must be collected for systems available for external heating and cooling, often referred to as Utility Data. Finally, Economic Data are needed, such as cost of heat exchangers and utilities, economic parameters, etc. For grassroots, consistent data are normally available from rigorous mass and energy balances established by process simulators, however, the quality of the data depends on whether or not the simulation model is a good representation of the actual process. For retrofit, data could also be taken from process simulations; alternatively data collected from measurements could be used. In the latter case, some data reconciliation is required after removing obvious (or “gross”) errors.

**Performance Targets** refer to establishing measures for best performance ahead of design only based on information available in the Stream Data, Utility Data and Economic Data (the type of data that is needed depends on the actual target(s) to be obtained). Typical targets for heat exchanger networks include minimum external heating \( Q_{H,\text{min}} \) and cooling \( Q_{C,\text{min}} \) demands, fewest number of heat exchangers \( U_{\text{min}} \), and minimum total heat transfer area \( A_{\text{min}} \). With multiple utilities, targets can also be established for the cost optimal utility mix.
**Process Modifications** refer to the consideration of making changes in the basic process (reactor system, separation system, recycle system, etc.) in such a way that the scope for heat recovery is improved. The indicator for the potential advantage of making such process changes is the Composite Curves established during the Performance Targets stage, and the tool for suggesting these changes is the Plus/Minus principle mentioned in Section 1.

**Network Design** in this chapter means establishing a network of heat exchangers that achieves the Performance Targets for energy consumption and number of heat exchangers, referred to as an MER design (Maximum Energy Recovery). The primary tool for network design is the Pinch Design Method (PDM) mentioned in Section 1, which is fundamentally based on the Pinch Decomposition principle. Separate networks are established above and below Pinch based on firm rules for matching hot and cold process streams in heat exchangers, rules for the sequence of these units, and a strategy for stream splitting when favorable or required.

**Design Evolution** primarily means to refine MER designs by removing small heat exchangers resulting from the use of the PDM, where separate networks are developed above and below Pinch. This activity is also referred to as Energy Relaxation, since removing units will require more utilities (energy) and possibly also more heat transfer area. The motivation for this stage is cost reduction as well as reduction in network complexity (fewer units, which is often followed by fewer stream splits). The tools are the degrees of freedom in the network referred to as Stream Splits, Heat Load Loops and Heat Load Paths that will be described in Section 7.1.

Finally, **Process Simulation** in this context means testing the feasibility of the heat exchanger network that has been designed and optimised in the previous stages of the design process. This stage may also involve switching from simplified models of pure counter-current heat exchangers to more practical and realistic design configurations. However, to describe the Process Simulation stage is regarded to be beyond the scope of this chapter.
3 Data Extraction

Data Extraction is a very time consuming and critical activity, since the quality and realism of the design solutions depend heavily on the correctness of the data. The saying “garbage in means garbage out” also applies here.

3.1 Data required for basic Heat Integration

The type of data required for Heat Integration projects obviously relate to the need for heating, cooling, evaporation and condensation in the process. In short, what is needed is a quantification of the required enthalpy changes of the process streams. From thermodynamics, the change in the total enthalpy flow, $H$ (kW), that a process stream undergoes when changing conditions can be obtained from [3.1].

$$\Delta H = \int m \cdot dh$$  \hspace{1cm} [3.1]

where $m$ is mass flowrate (kg/s) and $h$ is specific enthalpy (kJ/kg), giving change in enthalpy flow the units of (kJ/s = kW). Enthalpy is in general a complicated function of stream pressure, temperature and composition. In Heat Integration, a process stream is defined as one that does not change mass flowrate or composition. Whenever such changes take place, a new process stream is introduced. If we assume constant mass flowrate and stream composition, and ignore the effect of pressure on enthalpy, then [3.1] can be simplified to [3.2].

$$\Delta H = m \cdot \int c_p \cdot dT$$  \hspace{1cm} [3.2]

where $c_p$ is the specific heat capacity at constant pressure (kJ/kgK). In order to replace numerical integration by simple summation, the assumption of a constant $c_p$ or a piece-wise linear relation between temperature and enthalpy has been extensively used in Pinch Analysis. If $c_p$ is assumed constant, and the supply and target temperatures of a process stream are denoted $T_s$ and $T_t$ respectively, then [3.2] is simplified even further to [3.3].

$$\Delta H = m \cdot c_p \cdot \int_{T_s}^{T_t} dT = CP \cdot (T_t - T_s)$$  \hspace{1cm} [3.3]
where $CP$ is a lumped parameter (the product of $m$ and $c_p$) that is referred to as the “heat capacity flowrate” with units (kW/K). For process streams changing phase, information about the latent heat of such phase changes would be required. It should be emphasized that in this chapter, no formal sign convention is made, rather common sense is applied. This means that both hot and cold streams have changes in enthalpy flow that are positive as indicated in [3.4], where subscripts $h$ and $c$ refer to hot and cold process streams respectively. By definition, a hot process stream is one that is being cooled and/or condensed, while a cold process stream is one being heated and/or evaporated.

$$\Delta H_h = CP_h \cdot (T_{S,h} - T_{T,h}) \quad \text{and} \quad \Delta H_c = CP_c \cdot (T_{T,c} - T_{S,c})$$  \[3.4\]

If the relationship between temperature and enthalpy is non-linear, either caused by a non-constant $c_p$ or a phase change, improved accuracy in the analysis can be obtained by dividing process streams into piece-wise linear substreams often referred to as stream segments. This improves the targeting part (Section 4), while it adds complexity to the network design (Section 6), since one needs to keep track of the relationship between stream segments and the original process streams when suggesting heat exchangers in the network.

The data described above are sufficient to establish targets for minimum external heating and cooling in the process, as soon as a value for the minimum allowed temperature difference mentioned in Section 1, $\Delta T_{min}$, is established. When information about the utility system (the number of different utility types for external heating and cooling) is available, targets for the minimum number of heat exchangers (often referred to as units) can be established. In order to calculate targets for minimum total heat transfer area in the heat recovery systems, data for film heat transfer coefficients, $h$ (kW/m$^2$K), for process streams and utilities are also required.

3.2 Major Challenges in Data Extraction

There are two very different types of challenges related to data extraction for a Heat Integration project:

(i) To establish the most correct set of data related to flowrates and thermodynamic conditions of process streams used as input to heat recovery analysis and design.
(ii) To represent the heating, cooling, evaporation and condensation needs of the process streams in such a way that the degrees of freedom are kept open for network design.

While activity (i) is fairly straightforward (but involves a lot of work), activity (ii) requires skills and experience. It has often been stated that data extraction (in particular the second part) is more art than science, thus most of the attempts to provide procedures and guidelines for this activity has failed, including the development of knowledge based systems (also referred to as expert systems). Some of the commercial general purpose process simulators have features for automatic stream data extraction on the basis of a converged steady state mass and energy balance calculation. While these procedures enable easy generation of Composite and Grand Composite Curves, they do not keep the degrees of freedom open.

Despite the importance of data extraction, the topic has not been much discussed in the literature beyond text-books on Process Integration, such as Linnhoff et al. (1982), Smith (2005) and Kemp (2007). The topic is also thoroughly covered in the recent book by Klemeš et al. (2010). Interestingly, rather detailed literature on data extraction has been provided in the form of reports from research institutes (such as CANMET, 2003), software vendors (such as AspenTech, 2009) and consulting companies (such as Linnhoff March, 1998), which again illustrates the importance of proper data extraction for a successful heat integration project.

For manual data extraction, the following guidelines can be useful:

a) Do not copy all features of the conceptual flowsheet or an existing design.

b) Do not mix streams at different temperatures.

c) Do not include utilities as stream data.

d) Do not accept the prejudice of colleagues against heat integration.

e) Do not ignore true practical constraints.

f) Distinguish between soft and hard stream data.

Rule (a) refers to the issue of keeping the degrees of freedom open in order not to overlook promising solutions for heat recovery systems. Rule (b) involves several aspects and should be discussed in more depth. First, a mixer can act as a heat exchanger, thus saving capital cost, however, mixing streams with different composition is only an option if the streams are entering the same unit operation, such as a chemical reactor. Second, mixing streams at
different temperatures introduces exergy losses and should be avoided. Third, mixing streams may eliminate potential heat recovery solutions. Finally, mixing streams may be required from a practical point of view, such as adding steam to hydrocarbon streams to avoid coking inside pipes and equipment, or it may be forbidden from a safety point of view, such as mixing oxygen rich streams with hydrocarbon streams. Rule (c) is rather obvious, since the goal of the exercise is to establish minimum utility requirements, however, there are cases where it is not so easy to distinguish whether a stream is a process stream or acts as a utility.

Rule (d) relates to the common reluctance in the process industries to accept heat integration solutions from an operational point of view, however, it is a fact that most industrial processes are heavily integrated, and rather than focusing on maximum heat recovery, one should focus on correct or appropriate heat integration. In addition, it should be mentioned that when the economical potential of heat integration is established and well documented, it is often easier to get acceptance for such projects. Rule (e) means that even though one should try to keep the degrees of freedom open, obviously one should not forget that some practical constraints cannot be ignored. One example is related to metal dusting, a severe form of corrosive degradation of metals that happens in some temperature range when CO is present. This is a problem in synthesis gas production, and in order to keep the metal temperature at a sufficiently low level, the boiler is placed upstream of the steam superheater, which is not the best solution from a thermodynamic point of view as conveyed in Pinch Analysis.

Finally, rule (f) is quite important in the sense that some stream data must be considered as hard specifications, while others can be adjusted if that improves or simplifies the heat recovery system (as discussed in Chapter 1.2 on Basic Terminology of Process Integration). An inlet temperature to a reactor or distillation column must often be regarded as a hard specification, while the target temperature of a process stream going to some sort of storage is an example of soft process data. Specifying a low target temperature for a hot product stream going to storage in order to increase the heat recovery potential will only result in increased need for external cooling if the target temperature is below the Pinch temperature. Instead, this cooling could have been taken care of by nature itself through convective heat losses to the environment.
Returning to activity (i) of the data extraction exercise there are two distinctly different situations. For grassroots design, there is normally a simulation model available for the process providing stream data as part of a steady state material and energy balance calculation. The advantage in this case is that the data are consistent. As an example, the hot and cold side of a heat exchanger will always be in balance for a converged simulation. The quality of the data, however, depends on to what extent the simulation models describes the behavior of the real process.

For retrofit design, in addition to using a simulation model if available, one could resort to the original specification sheets for the process, or one could use measurements from the plant. However, the plant may have been modified several times since its start-up, and flowsheets and specification sheets are not always updated. Regarding the use of measurements, the typical situation is that some measurements are missing, and instruments may either not be functioning at all, or they may give incorrect readings. In such cases, the task of data reconciliation can be enormous, and a key to success is to work very close with operators and plant engineers.

4 Performance Targets

In order to illustrate the different stages in the design of heat recovery systems, consider the very simple process example in Fig. 4.1, where two feed streams (A and B) are heated before entering a chemical reactor where a product and a by-product are produced. The main product is then recovered in the bottoms stream from a distillation column, while the by-product and traces of unreacted raw material (feed) is taken from the top of the same distillation column. Fig. 4.1 also indicates the supply and target temperatures of the process streams as well as the lumped parameter CP (“heat capacity flowrate”) in brackets with units (kW/K). Notice that heat exchangers are not included, since these will be the result of the Heat Integration study.

It is important at this stage to emphasize that “heat capacity flowrate” \(CP = mCp\) is different from mass flow rate \(m\). This is why the reactor effluent \(CP = 100\) is less than the sum of the reactor inlets \(CP = 50 + 150 = 200\). The explanation in this case is that feed stream B is subject to evaporation before entering the reactor. Since the outlet temperature from the reactor is higher than the inlet temperature, the chemical reactions taking place must be
exothermic. Based on the flowsheet in Fig. 4.1 and the given data for temperatures, heat capacity flow rates and duties for the distillation column reboiler and condenser, the stream data resulting from data extraction (Section 3) are listed in Table 4.1.

![Figure 4.1 Simple Process used as an Illustrative Example](image)

**Table 4.1 Stream Data for the Simple Process in Fig. 4.1**

<table>
<thead>
<tr>
<th>ID</th>
<th>Description</th>
<th>$T_s$ (°C)</th>
<th>$T_R$ (°C)</th>
<th>$CP$ (kW/K)</th>
<th>$\Delta H$ (kW)</th>
</tr>
</thead>
<tbody>
<tr>
<td>H1</td>
<td>Reactor Effluent</td>
<td>220</td>
<td>130</td>
<td>100</td>
<td>9,000</td>
</tr>
<tr>
<td>H2</td>
<td>Main Product</td>
<td>130</td>
<td>50</td>
<td>90</td>
<td>7,200</td>
</tr>
<tr>
<td>C1</td>
<td>Feed A</td>
<td>40</td>
<td>150</td>
<td>50</td>
<td>5,500</td>
</tr>
<tr>
<td>C2</td>
<td>Feed B</td>
<td>80</td>
<td>150</td>
<td>150</td>
<td>10,500</td>
</tr>
<tr>
<td>CON</td>
<td>Column Condenser</td>
<td>120</td>
<td>120</td>
<td>n.a.</td>
<td>3,000</td>
</tr>
<tr>
<td>REB</td>
<td>Column Reboiler</td>
<td>130</td>
<td>130</td>
<td>n.a.</td>
<td>3,000</td>
</tr>
<tr>
<td>ST</td>
<td>Steam for heating</td>
<td>250</td>
<td>250</td>
<td>n.a.</td>
<td>variable</td>
</tr>
<tr>
<td>CW</td>
<td>Cooling Water</td>
<td>20</td>
<td>30</td>
<td>n.a.</td>
<td>variable</td>
</tr>
</tbody>
</table>

For distillation columns, the heating and cooling requirements are normally given as the duties of the reboiler and the condenser. The reflux and boil-up are circulating internal streams in the column and therefore normally not measured. This means that the $CP$ values...
are not available (n.a.) as indicated in Table 4.1. The duties of steam and cooling water are listed as “variable” in Table 4.1, which is obvious since these are the unknowns that will be found as a result of the energy targeting exercise.

4.1 Minimum External Heating and Cooling

As mentioned in Section 1, targets for heat recovery systems depend on the specification of a minimum allowed temperature difference for heat transfer, $\Delta T_{\text{min}}$, which is an economic parameter for the trade-off between investment cost (heat exchangers) and operating cost (energy). Given a value for this parameter, targets for minimum external heating, $Q_{H,\text{min}}$, and minimum external cooling, $Q_{C,\text{min}}$, can be obtained by graphical or numerical methods established in the early period of Pinch Analysis. The graphical representations are referred to as Composite and Grand Composite Curves, while there are several numerical methods such as the Problem Table Algorithm (Linnhoff and Flower, 1978) and the Heat Cascade (Linnhoff, 1979). Actually, Hohmann (1971) was the first to provide a systematic way to obtain energy targets by his Feasibility Table. The Heat Cascade will be used in this chapter since it (i) provides a nice illustration of the heat flows and decompositions in heat recovery systems and (ii) provides the necessary information to construct the Grand Composite Curve.

Optimisation techniques such as Linear and Mixed Integer Programming (LP and MILP) can also be used to obtain targets for minimum external heating and cooling as well as targets for the fewest number of heat exchangers, especially in more complicated situations such as for example when there are forbidden matches between hot and cold process streams. These optimisation techniques will be briefly discussed in Chapter 2.5.

**Composite Curves**

Composite Curves have been described and applied by a number of authors, such as Huang and Elshout (1976), Umeda et al. (1978) and Linnhoff et al. (1982). The Composite Curves ($T$-$H$ diagram) are constructed by dividing the temperature axis into intervals based on the supply ($T_s$) and target ($T_r$) temperatures of the process streams, and to add together the enthalpy contributions (hot streams) and requirements (cold streams) in each temperature interval. Finally, these enthalpies are drawn in a cumulative manner against the corresponding temperatures, resulting in one curve for the hot streams and one curve for the cold streams.
These curves are then positioned relative to each other in such a way that the Hot Composite Curve (the cooling curve) is always above the Cold Composite Curve (the heating curve). In this way, heat can be recovered in the overlapping region of the Composite Curves. This “positioning” in the $T-H$ diagram is obtained by shifting the two curves horizontally. Moving the curves closer together means increased heat recovery, and the economic limit is when the smallest vertical distance between the curves becomes equal to $\Delta T_{\text{min}}$, while the thermodynamic limit is when this vertical distance becomes zero. The point where the vertical distance between the Composite Curves is at its minimum (and equal to $\Delta T_{\text{min}}$) acts as a bottleneck against increased heat integration and has therefore been referred to as the Heat Recovery Pinch.

Composite Curves for the stream data in Table 4.1 are shown in Figure 4.2 for $\Delta T_{\text{min}} = 20^\circ\text{C}$.

Key information about the heat recovery system can be obtained from this graphical diagram, such as the process Pinch, maximum heat recovery, and the corresponding minimum external heating and cooling requirements. Reading accurate information from such diagrams can be somewhat difficult, which is why numerical methods are often preferred. The real advantage of such graphical diagrams, however, is that they provide an overview of the system and they contribute strongly to the understanding of the problem. One such insight is that any heat leakage from the region above Pinch to the region below Pinch (cross Pinch heat transfer) will result in increased need for both hot and cold utilities (i.e. double penalty).
Based on the Composite Curves in Fig. 4.2, approximate targets for minimum external heating and cooling seem to be in the range 3.0-3.5 MW. As indicated above, there are also a number of numerical methods that can be applied to obtain these targets more accurately. Fig. 4.3 shows the Heat Cascade for the same illustrative example with stream data in Table 4.1. The reboiler (REB) and the condenser (CON) of the distillation column adds two rather strange temperature intervals to the heat cascade, since it is assumed that condensation and evaporation in these units take place at constant temperature. This means that the streams with sensible heat (H1, H2, C1, and C2) do not deliver or extract any heat from these intervals that are marked with a “+” and a “-“ for the corresponding temperatures.

**Heat Cascade**

The heat cascade is an example of the Transhipment Model from Operations Research, with sources, warehouses and sinks. The hot streams are drawn as sources of heat on the left hand side of the cascade, with corresponding hot stream temperatures. The cold streams are drawn as sinks of heat on the right hand side of the cascade, with corresponding cold stream temperatures. All supply and target temperatures of the process streams should be represented as interval temperatures in the cascade; in addition there will be “corresponding” temperatures on the “opposite” side obtained by adding $\Delta T_{\text{min}}$ to the supply and target temperatures of cold process streams and subtracting $\Delta T_{\text{min}}$ from the supply and target temperatures of hot process streams. This means that the specification of a minimum temperature difference is built into the heat cascade. Any heat exchange taking place in the temperature intervals of Fig. 4.3 will be feasible and satisfy the requirement that $\Delta T \geq \Delta T_{\text{min}}$.

As indicated in Fig. 4.3, the hot streams provide heat to the temperature intervals according to their cooling requirements, and the cold streams extract heat from the temperature intervals according to their heating requirements. Heat balances are established for each interval, and any surplus of heat in one interval is cascaded (thus the name “heat cascade”) as a heat residual ($R_k$) to the next interval with lower temperatures. Since none of these residuals can be negative (would indicate transfer of heat from lower to higher temperatures which is infeasible with heat exchangers), the minimum external heating requirement can be identified as the minimum heat needed to make these residuals non-negative.
A simple, yet powerful way to establish values for $Q_{H,\text{min}}$ and $Q_{C,\text{min}}$ is to start by assuming $Q_H = 0$ kW in Fig. 4.3. The heat residuals can then be obtained for the entire heat cascade in a sequential manner as follows:

\[
\begin{align*}
R_1 &= Q_H + 5,000 = +5,000 \text{ kW} \\
R_2 &= R_1 - 2,000 = +3,000 \text{ kW} \\
R_3 &= R_2 - 3,000 = 0 \text{ kW} \\
R_4 &= R_3 - 2,000 = -2,000 \text{ kW} \\
R_5 &= R_4 - 1,100 = -3,100 \text{ kW} \\
R_6 &= R_5 + 3,000 = -100 \text{ kW} \\
R_7 &= R_6 - 2,200 = -2,300 \text{ kW} \\
R_8 &= R_7 + 1,600 = -700 \text{ kW} \\
Q_C &= R_8 + 900 = 200 \text{ kW}
\end{align*}
\]

\[R_5\]

Figure 4.3 Heat Cascade for the Illustrative Example

As mentioned above, negative residuals are infeasible when designing a system of heat exchangers, and residual $R_5$ has the largest negative value of the entire heat cascade. This
residual then becomes the bottleneck (i.e. Heat Recovery Pinch), and minimum external cooling is found to be \( Q_{H,\text{min}} = 3,100 \text{ kW} \). The corresponding minimum external heating is then \( Q_{C,\text{min}} = 3,300 \text{ kW} \), and the process Pinch is defined by the temperatures 120°C (for hot streams) and 100°C (for cold streams). Since the residual \( R_3 \) actually is related to the temperatures 120°C and 100°C, it means that the column condenser (CON) is supplying heat below Pinch. The implications of this will be discussed in Sections 5.1 and 5.2.

4.2 Minimum Number of Heat Exchangers

The next logical step in a Heat Integration project is to establish targets for the fewest number of heat exchangers, also referred to as units. This is done by the so-called \((N - 1)\) rule, used for the first time by Hohmann (1971). Linnhoff et al. (1979) explained that the \((N - 1)\) rule is a simplification of Euler’s Rule from graph theory \((U = N + L - S)\). The analogy between graphs and heat exchanger networks is that nodes represent streams, while edges represent heat exchangers. Thus \(N\) is the total number of process streams and utility types, \(U\) is the number of heat exchangers (units), \(L\) is the number of independent loops, and \(S\) is the number of sub-networks (or subgraphs). Since the objective is to establish a target for the number of units ahead of design, network related features such as loops and sub-networks are not known. This is overcome by setting \(L = 0\) (loops can be removed as will be shown in Section 7.2) and \(S = 1\) (conservative, since the presence of sub-networks reduces the number of units). As a result, Euler’s Rule reduces to \(U = N - 1\). As pointed out in Section 1, however, separate networks must be designed above and below Pinch in order to achieve the targets for minimum external heating and cooling. Thus, the \((N - 1)\) rule has to be applied separately above and below Pinch in order to have a target for the number of units that is compatible with the heating and cooling targets. A targeting formula for minimum number of units in a heat exchanger network achieving Maximum Energy Recovery (MER) is then given by [4.1].

\[
U_{\text{min,MER}} = (N_{\text{above}} - 1) + (N_{\text{below}} - 1) \tag{4.1}
\]

As will become evident in Section 7.2, the fewest number of units in heat recovery systems that relax the MER requirement (do not decompose at the Pinch) is another useful target value and is given by [4.2].
Another invention from the pioneering period of Pinch Analysis is the Grid Diagram, which is an alternative way to draw heat exchanger networks. While the grid diagram is a departure from standard ways to represent process flowsheets, it has the important advantage that it mimics the desirable counter-current flow of heat exchangers and thereby makes it easy to implement Pinch decomposition in heat exchanger networks as well as to study cross Pinch heat transfer. Fig. 4.4 shows the grid diagram for the illustrative example presented in Figs. 4.1-4.3. Contrary to the first version of this representation proposed by Linnhoff (1979) the Handbook follows the more logical form, where temperatures increase from left to right, as they would do in any xy diagram. This then defines the directions of the hot and cold process streams in the diagram. It should also be emphasized that no linear temperature scale is applied; rather the main focus is on whether a stream is present above, across or below Pinch.

![Grid Diagram](image)

Figure 4.4 Grid Diagram for the Illustrative Example

While the grid diagram will act as a “drawing board” for network design in subsequent sections, the first use of this representation is to establish targets for minimum number of heat exchangers.
exchangers. In this respect, heat exchangers (or units) refer to process-to-process heat exchangers as well as utility exchangers (such as steam heaters and water coolers). The dashed lines in Fig. 4.4 indicate the “position” of the process Pinch, and all streams are drawn relative to these Pinch temperatures. Thus, similar to the heat cascade, the grid diagram also keeps track of both hot and cold stream temperatures. Utilities could also have been included in the grid diagram (particularly useful with multiple utilities), but has been left out for the sake of simplicity. It should be emphasized that targeting for units takes place after targeting for energy, thus the types and amounts of different utilities needed are known and fixed.

The grid diagram can be used to set targets for the fewest number of units as follows: For an MER design, Pinch decomposition must be obeyed, and [4.1] can be applied. Energy targeting established the need for both steam (3,100 kW) and cooling water (3,300 kW). Above the Pinch, there are 2 hot streams (H1 and H2), 3 cold streams (C1, C2 and REB), and 1 hot utility (ST). Below the Pinch, there are 2 hot streams (H2 and CON), 2 cold streams (C1 and C2), and 1 cold utility (CW). The MER target for minimum number of units then becomes:

\[ U_{\text{min,MER}} = (N_{\text{above}} - 1) + (N_{\text{below}} - 1) = (2 + 3 + 1 - 1) + (2 + 2 + 1 - 1) = 5 + 4 = 9 \]

Similarly, the target for the fewest number of units when strict Pinch decomposition is relaxed, can be found by [4.2]:

\[ U_{\text{min,global}} = (N_{\text{total}} - 1) = (3 + 3 + 2 - 1) = 7 \]

This mans that the penalty for Maximum Energy Recovery (and strict Pinch decomposition) is that two (i.e. 9–7) more heat exchangers are likely to be required in the network. This issue will be further discussed in Section 7.2

Targeting methods have also been developed for minimum total heat transfer area, fewest number of heat exchanger shells (rather than units) for situations where shell & tube exchangers are dominating, and total annual cost. The latter can be used to identify a reasonable value for \( \Delta T_{\text{min}} \) (also referred to as pre-optimisation or SuperTargeting). These more advanced targets will be presented in Chapter 2.5. Further targets that will be discussed in other chapters of the Handbook involve mechanical energy, such as shaftwork targets, and targets for total sites (both heat and power).
5 Process Modifications

With Composite Curves established and Performance Targets calculated, it makes sense to consider options for Process Modifications before continuing with Network Design. The shape of the Composite Curves indicates whether potentials for increased heat recovery exist, and what actions are needed to reduce external heating and cooling for the process. Ideally, the Composite Curves should be as parallel as possible, since this allows for a high level of heat recovery. In reality, the Composite Curves will have kinks or “knees” that act as bottlenecks (Pinch points), which will limit heat recovery. One way to make the Composite Curves more parallel is to move some of these kinks to other temperatures or to remove some of the kinks completely, with main focus on the near Pinch region of the process.

In addition, the fundamental feature of having a heat deficit region above the process Pinch and a heat surplus region below the process Pinch provides guidelines for how the process should be modified to increase the potential for heat recovery. These guidelines, later coined the Plus/Minus Principle, have been discussed again in various ways by Umeda et al. (1979), Linnhoff and Parker (1984) and Linnhoff and Vredeveld (1984). The Plus/Minus Principle suggests that above Pinch, one should try to increase the amount of heat provided by hot streams (+) or decrease the amount of heat required by cold streams (−). Likewise, below Pinch, one should try to increase the amount of heat required by cold streams (+) or decrease the amount of heat provided by hot streams (−). This means that if a hot stream (or part of it) is moved from below to above Pinch or a cold stream (or part of it) is moved from above to below Pinch, the situation improves in both regions. Such “moves” can be realized by changing the temperature of streams, which in some cases results from changing the stream pressure. Another possibility is to increase or reduce the enthalpy change of a stream.

Some obvious examples of process modifications include:

- Decrease the pressure and thus the boiling point temperature of an evaporator (cold stream) to move the operation from above to below Pinch
- Split an evaporator into several stages in series (i.e. multi-effect)
- Decrease the pressure of a distillation column to move the reboiler (a cold stream) from above to below Pinch
- Increase the pressure of a distillation column to move the condenser (a hot stream) from below to above Pinch
- Change the reflux of a distillation column
- Change the feed preheating or precooling of a distillation column
- Change the operating conditions of a reactor

5.1 Plus/Minus Principle and Appropriate Placement

The Plus/Minus Principle also provides guidelines for integration (or Appropriate Placement) of special equipment such as distillation columns (Linnhoff et al., 1983), evaporators (Smith and Linnhoff, 1988) heat pumps and heat engines (Linnhoff and Townsend, 1982). The background is that such equipment should not be integrated unless there are considerable energy savings involved that will compensate for additional operating problems as well as any increase in investment cost. The term Correct Integration has also been used, and the issue can be addressed simply by considering the different units as sources and sinks of heat. For obvious reasons, energy savings will only be made if one integrates a source of heat with a sink of heat, and simple thermodynamic principles require that the source must have a higher temperature than the sink. The following classification can be made for some of the typical equipment mentioned above as well as for the “background” (or remaining) process:

- The background process as represented by its heat cascade is a heat sink above the process Pinch and a heat source below the same Pinch
- A distillation column represents a heat sink in the reboiler and a heat source in the condenser
- An evaporator represents a heat sink in the evaporation stage (boiler) and a heat source in the condenser
- A heat pump represents a heat sink at lower temperature (evaporator) and a heat source at higher temperature (condenser), while consuming mechanical energy
- A heat engine represents a heat sink at higher temperature and a heat source at lower temperature, while producing mechanical energy

Correct use of a heat pump to reduce external heating in the background process then means to integrate the source of the process below Pinch with the sink of the heat pump (evaporator) as well as to integrate the source of the heat pump (condenser) with the sink of the process.
above Pinch. In short, this means that the heat pump should be integrated across Pinch in the sense that it utilizes heat from the surplus region below Pinch and supplies it to the deficit region above Pinch. Such integration (or use) of a heat pump reduces both hot and cold utility consumption. If a heat pump is integrated entirely above Pinch, the only result from an energy point of view is that mechanical energy usage is converted into thermal energy savings on a 1:1 basis, which is not a good idea, since mechanical energy has a higher value. If a heat pumps is integrated below Pinch, however, a much worse scenario can be drawn: Use of mechanical energy in the heat pump only results in an increased consumption of cold utility.

Correct Integration of a distillation column with the background process requires integration of the consender (heat source) with cold streams above Pinch (heat sink) or integration of the reboiler (heat sink) with hot streams below Pinch (heat source). Since the reboiler temperature is always higher than the condenser temperature, only one of these options are available for a single column. For a condenser integrated above Pinch, the maximum savings in both hot and cold utilities are equal to the duty of the condenser, $Q_{\text{cond}}$. Likewise, for a reboiler integrated below Pinch, the maximum savings in both hot and cold utilities are equal to the duty of the reboiler, $Q_{\text{reb}}$. In summary, there are 3 distinct cases with very simple rules for integration:

1) $T_{\text{reb}} > T_{\text{cond}} > T_{\text{Pinch}}$ : The condenser of the distillation column should be integrated with the background process above Pinch, and the savings are $\Delta Q_H = \Delta Q_C \leq Q_{\text{cond}}$.

2) $T_{\text{reb}} > T_{\text{Pinch}} > T_{\text{cond}}$ : The distillation column should not be integrated with the background process, since no energy savings will be obtained.

3) $T_{\text{Pinch}} > T_{\text{reb}} > T_{\text{cond}}$ : The reboiler of the distillation column should be integrated with the background process below Pinch, and the savings are $\Delta Q_H = \Delta Q_C \leq Q_{\text{reb}}$.

Returning to the simple process example in Fig. 4.1, with corresponding Composite Curves in Fig. 4.2 and Heat Cascade in Fig. 4.3, the principles of Appropriate Placement can be applied to the distillation column that is part of the process. Both Figs. 4.2 and 4.3 indicate that the process Pinch is caused by the condenser of the distillation column. According to the simple rules mentioned above, the illustrative process example falls into category (2) which means that one should not integrate the distillation column in this case. However, the background process can be established and analyzed by removing the column reboiler and condenser from the stream data. The resulting heat cascade for the background process is given in Fig. 5.1. By
analyzing the temperature intervals in this case, it is obvious that the third interval will be the limiting one, thus minimum external heating is given by the need to make $R_3$ non-negative.

For the background process, the energy targeting exercise provides the following results:

- Process Pinch is given by $R_3 = 0$, $T_{Pinch} = 100{°C}/80{°C}$ (for hot/cold streams)
- Minimum external heating is given by $Q_{H,min} = 3,300 + 4,000 - 5,000 = 2,300$ kW
- Minimum external cooling is given by $Q_{C,min} = 1,600 + 900 = 2,500$ kW

![Figure 5.1 Heat Cascade for the Background Process](image)

These results are surprising and should be analyzed and explained. The simple rule applied above indicated that no savings would be obtained by integrating the distillation column. The background process needs 2,300 kW of external heating, while the reboiler of the column needs 3,000 kW. In the non-integrated case, the total need for hot utility is then $2,300 + 3,000 = 5,300$ kW. The assumption behind the Composite Curves (Fig. 4.2) and the total Heat Cascade (Fig. 4.3) is that the distillation column is integrated with the background process. The background for this statement is that the condenser and the reboiler are included in the stream data in Table 4.1, which is the basis for Figs. 4.2 and 4.3. In the integrated case, the calculated minimum external heating requirements are 3,100 kW, which means that despite the predictions of the simple analysis above, heat integration of the column actually saves $5,300 - 3,100 = 2,200$ kW. The reason for this apparent contradiction is that the Pinch for the
background process (100°C/80°C) is different from the case when the distillation column is integrated (120°C/100°C). In fact, distillation columns often cause Pinch points due to their large duties at near constant temperature, a feature that results in marked “knees” on the Composite Curves. With reference to the Pinch of the background process, the distillation column operates entirely above Pinch and the example falls into category (1) above.

Another important result is that the savings are less than the duty of the integrated condenser ($2,200 < 3,000 = Q_{cond}$). Explaining this result requires use of another graphical diagram; the Grand Composite Curve. While the simple rules can be used in a qualitative way to establish whether Appropriate Placement is feasible or not, the Grand Composite Curve provides quantitative information about the amount of heat that can be correctly integrated, and thus how much heating and cooling that can be saved by Correct Integration.

### 5.2 Grand Composite Curve and Correct Integration

As mentioned when introducing the Heat Cascade in Section 1, one of its advantages is that it provides the necessary information (or data) to construct the Grand Composite Curve (Linnhoff et al., 1982). This is a diagram that shows the net accumulated heat surplus and heat deficit in the process, and provides an excellent interface between the process and the utility system. It can also be used to evaluate the integration of special equipment such as distillation columns, heat pumps, etc. While the Composite Curves show two independent curves for hot and cold process streams using real temperatures, the Grand Composite Curve (which is another $TH$ diagram) shows the residual of heat in the Heat Cascade as a single curve. Thus, there is a need for a common temperature scale that can be used for both hot and cold process streams, which is why the so-called “modified” temperatures have been introduced. The simplest way to establish such modified temperatures is to subtract half of $\Delta T_{min}$ from hot stream temperatures and likewise add half of $\Delta T_{min}$ to cold stream temperatures.

Considering the Heat Cascade in Fig. 5.1, this means that the average values of the hot and cold interval temperatures will be used as the new temperature scale. A more advanced way to establish modified temperatures is to introduce individual stream contributions to $\Delta T_{min}$. This allows for a more realistic approach to industrial problems where film heat transfer coefficients may vary by one or two orders of magnitude. In such cases, the use of a single
global value for $\Delta T_{\text{min}}$ is a gross over-simplification. Modified temperatures for hot streams ($i$) and cold streams ($j$) are established by [5.1], where $\Delta T_i$ and $\Delta T_j$ are individual contributions for hot and cold streams. The simple approach used in this chapter, however, is that $\Delta T_i = \Delta T_j = 0.5 \cdot \Delta T_{\text{min}}$.

\[
\begin{align*}
T^*_i & = T_i - \Delta T_i \\
T^*_j & = T_j + \Delta T_j
\end{align*}
\]  

[5.1]

**Figure 5.2 Grand Composite Curve for the Background Process**

The Grand Composite Curve for the background process is shown in Fig. 5.2. This diagram provides the same fundamental information as the Composite Curves (i.e. location of the Pinch and minimum external heating and cooling), but it also hides information related to process-to-process heat transfer. As a net Heat Surplus/Deficit Curve, the only process-to-process heat transfer shown in the Grand Composite Curve, is the transfer of surplus heat from one interval to another interval at lower temperature with heat deficit. This is referred to as heat “pockets”, and the Grand Composite Curve in Fig. 5.2 has one such pocket.

The Grand Composite Curve not only shows the required external heating and cooling, it also shows at what temperatures such external heating and cooling is required. This combination of load and level is of course important information for utility placement, and can be used to identify near-optimal consumption and possible production of various utility types. In addition, as mentioned above, the Grand Composite Curve can be used to quantify how much energy savings can be made by integrating distillation columns, evaporators, heat pumps and
heat engines with the background process. Focusing on the illustrative example, Fig. 5.2 shows that the need for external heating is in the range from 90°C to 110.9°C (found by interpolation) in modified temperatures, which means that the hot utility must have temperatures between 120.9°C and 100°C to satisfy the specified $\Delta T_{min}$ of 20°C. Using steam at 250°C as indicated in Table 4.1 would be a real waste of energy quality in this case, since for example very low pressure steam at 121°C or hot water (or hot oil) operating between 121°C and 100°C would be sufficient.

As already stated, the Grand Composite Curve can be used to quantitatively address the issue of Correct Integration. Since the simple process example in Fig. 4.1 includes a distillation column, it will be used to illustrate how the Grand Composite Curve can be used to identify the scope for integrating the column with the background process. A distillation column can be plotted as a box in $TH$ diagrams, where temperature profile and duty are plotted for the condenser and the reboiler. In the illustrative example, condensation (120°C) and evaporation (130°C) take place at constant temperature, and the duty of these units are equal (3 MW). The box representation in a $TH$ diagram then becomes a simple rectangle. When plotted together with the Grand Composite Curve, modified temperatures must be used, which means that for $\Delta T_{min} = 20°C$, the condenser (hot stream) should be plotted at 110°C, while the reboiler (cold stream) should be plotted at 140°C. Fig 5.3 shows the Grand Composite Curve (GCC) for the background process and the distillation column plotted in the same $TH$ diagram.

![Figure 5.3](image)

Figure 5.3  Background Process GCC and Distillation Column
The solid rectangle in Fig. 5.3 is in conflict with the GCC, and feasible heat transfer (meaning $\Delta T \geq \Delta T_{\text{min}}$) between the condenser and the background process is limited to 2,200 kW (found by interpolation in the GCC for a modified temperature of 110°C). If, however, the pressure of the column is increased slightly, the condenser and reboiler temperatures will increase accordingly, and the enthalpy box representing the column would fit into the heat pocket of the GCC. Increasing pressure will in most cases make the separation more difficult, thus there is a need for more equilibrium stages in the column or more reflux (with a balanced trade-off, this normally means an increase in both number of stages and reflux). This is indicated in Fig. 5.3 by making the dashed rectangle for the column after pressure increase somewhat wider along the enthalpy axis, since reboiler and condenser duties are proportional to the reflux.

In summary for this example, the background process requires 2,300 kW of external heating while the distillation column needs 3,000 kW heating in the reboiler, a total of 5,300 kW. Total cooling is correspondingly $2,500 + 3,000 = 5,500$ kW. By integrating the column condenser with cold streams above the Pinch, it is possible to utilize 2,200 kW of the condenser duty. This means that 2,200 kW of savings are made in external heating (41.5%) as well as cooling (40%). However, using process modifications, in this case a slight increase in the column pressure, makes it possible to integrate the entire condenser and save 3,000 kW of external heating (56.6%) and cooling (54.6%). The disadvantage with this last scheme that saves an additional 800 kW, is the need to integrate both the condenser (totally) and the reboiler (partly) with corresponding operational challenges related to the control of the column.

6 Network Design

With performance targets for energy and units, the next step is the actual design of the heat exchanger network. One of the most significant features of Pinch Analysis is that the insight obtained in establishing performance targets ahead of design actually forms the core of the design methodology. The discovery of the Heat Recovery Pinch in the early and late 1970s was followed by the understanding that in order to design heat exchanger networks with minimum external heating and cooling requirements, decomposition at the Pinch and the development of two independent networks were absolute requirements.
6.1 The Pinch Design Method

The Pinch Design Method (PDM) was outlined by Linnhoff and Turner (1981) and also by Linnhoff et al. (1982) before it was comprehensively described by Linnhoff and Hindmarsh (1983). The PDM provides a strategy for developing the network in a sequential manner deciding on one heat exchanger at a time, with rules for matching hot and cold streams for these heat exchangers. The method also indicates when and how stream splitting should be applied. The key elements of a simplified version of the PDM are the following design actions and rules:

- Decompose the heat recovery problem at the Pinch
- Develop separate networks above and below Pinch, starting at the Pinch
- Start network design immediately above and immediately below Pinch, since this is where the problem is most constrained (small driving forces) and thus where the degrees of freedom to match hot and cold process streams are most limited
- Assign Pinch Exchangers first (units that bring hot streams to Pinch temperature above Pinch and cold streams to Pinch temperature below Pinch), then assign the other process-to-process units, and finally install utility exchangers where required to reach the target temperatures for the streams
- Use the CP rules (see [6.1]) to decide on the matching between hot and cold process streams in the Pinch Exchangers
- Whenever the CP rules cannot be applied or the topology rules (see [6.2]) are broken, stream splitting has to be considered
- For each accepted match, maximise the duty of the heat exchanger to increase the probability of reaching the target for fewest number of units (i.e. the “tick-off” rule)

Pinch Exchangers will have minimum allowed driving forces ($\Delta T_{min}$) in the cold end (units above Pinch) or the hot end (units below Pinch) of the heat exchanger. Since the driving forces are at the minimum, there can be no further reduction, and the CP rules assure exactly that. Also, there has to be at least one stream (or branch of a stream) of the opposite type to bring the hot streams (above Pinch) and cold streams (below Pinch) to the Pinch temperature. The resulting CP rules and topology rules are fundamental when applying the PDM:
If either the topology rules [6.2] are not satisfied for the entire set of streams, or the CP rules [6.1] are not satisfied for each and all of the Pinch Exchangers, then stream splitting is required, and a detailed on this topic discussion is provided in Section 6.3.

When the Pinch Exchangers are installed, the next important task is to utilize any remaining heat in the hot streams above Pinch and any remaining cooling in the cold streams below Pinch. This can be achieved by adding more process-to-process units, and in this case the matches are no longer restricted by the CP rules, since the driving forces have opened up when moving away from the Pinch. Whenever a match violates the CP rules, however, the temperature difference of the heat exchanger must be checked.

In this simplified version of the PDM, it is assumed that utility exchangers are placed last where necessary to obtain the target temperatures of the streams. With multiple utilities at different temperature levels, this is obviously not a good strategy as it will be shown in Chapter 2.5.

6.2 Developing an initial MER Design

Next, the actual use of the PDM will be demonstrated using the simple process example presented in Fig. 4.1 with stream data in Table 4.1. The results from the targeting stage and the subsequent process modification stage can be summarised as follows:

- Increasing the pressure of the distillation column would allow complete integration of the column condenser with cold streams above the background process Pinch saving 3 MW of external heating and cooling, however, this change will not be made here in order to keep the case study simple, and the stream data of Table 4.1 will be used

- Integration of the distillation column without process modifications will save 2,200 kW of external heating and cooling, and the grid diagram in Fig. 4.4 will be used to design the heat exchanger network
When the condenser and reboiler are included in the stream data (with the intention to partly or fully integrate the column), the following targeting results were obtained:

- Overall process Pinch (with background process and distillation column):
  \[ T_{Pinch} = 120^\circ C/100^\circ C \] for hot/cold process streams.
- Minimum external heating: \[ Q_{H,\text{min}} = 3,100 \text{ kW} \]
- Minimum external cooling: \[ Q_{C,\text{min}} = 3,300 \text{ kW} \]
- Minimum number of units that is compatible with Maximum Energy Recovery (MER): \[ U_{\text{min,MER}} = 5 + 4 = 9 \] (5 units above and 4 units below Pinch)
- Minimum number of units when relaxing Pinch decomposition: \[ U_{\text{min,global}} = 7 \]

The grid diagram in Fig. 6.1 is the same as in Fig. 4.4, however, with additional information about \( CP \) and \( \Delta H \) to make the design process easier. The heat exchangers have been numbered according to the sequence they were introduced in the network. Utility exchangers are marked H for heaters and C for coolers. Notice that since hot stream H1 is cooled to 130°C which is above the hot Pinch temperature (120°C), there is only one Pinch exchanger above Pinch (for hot stream H2). Below Pinch, both cold streams C1 and C2 require Pinch exchangers. In Fig. 6.1, all relevant duties (in kW) and temperatures are provided.

![Figure 6.1 Maximum Energy Recovery Network for the Illustrative Example](image-url)
The grid diagram with the MER design in Fig. 6.1 clearly illustrates the advantages of this representation compared to the traditional way to draw process flowsheet. The Pinch point location and the decomposition into two separate heat exchanger networks above and below Pinch can be easily seen. For each heat exchanger it is very easy to check the driving forces since the process streams are drawn in a counter-current way. This means that the hot inlet to a heat exchanger is drawn vertically above the cold outlet from the same unit, with easily available information about the hot end of the exchanger. Likewise, the hot outlet from the heat exchanger is drawn vertically above the cold inlet to the same unit, with easily available information about the cold end of the exchanger. Further, as will become evident in the retrofit design case described in Chapter 2.5, if the 4 temperatures for each exchanger is logically positioned relative to the vertical dashed line that marks the Pinch (above/below or right/left), then cross Pinch heat transfer is easily identified graphically.

When an MER design is established using the Pinch Design Method (PDM), one should always check the developed network versus the performance targets calculated at earlier stages of the Heat Integration project. From Fig. 6.1, the following can be established:

- External heating: $Q_H = 100 + 3,000 = 3,100 = Q_{H,\text{min}}$
- External cooling: $Q_C = 3,300 = Q_{C,\text{min}}$
- Number of units: $U = 5 + 2 + 1 = 8 < 9 = U_{\text{min,MER}}$

Since the PDM decomposes the design problem at the Pinch, the resulting network will always meet the energy targets, unless errors are made during design. Such errors can be use of hot utility below Pinch or use of cold utility above Pinch. For industrial size problems there will be cases where it is not straightforward to develop an MER design, especially if the Composite Curves have a parallel shape or if there are near-Pinches. Some of these problems will be addressed in Chapter 2.5. It could also be argued that the target values for minimum external heating and cooling are firmly based on thermodynamics and therefore represent rigorous targets.

The situation is quite different for the number of heat exchangers in the resulting network. The $(N-1)$ rule that forms the basis for the targeting formulas for minimum number of units
([4.1] and [4.2]) is a simplification of Euler’s Rule from Graph Theory \( U = N + L - S \). As explained in Section 4.2, in order to establish a formula that could be applied ahead of design, assumptions had to be made regarding loops \( (L = 0) \) and subnetworks \( (S = 1) \). Of course, this means that the target formulas are not rigorous, and there will be cases where the number of heat exchangers in the network can be both larger and smaller than the number of units obtained by the \((N-1)\) rule. The network in Fig. 6.1 has one unit less than the target value, and the reason is that there is a perfect match between the cooling requirements \((3,000 \text{ kW})\) in the condenser (CON) and the heating requirements \((3,000 \text{ kW})\) of cold stream C2 below Pinch, thus introducing subnetworks. This means that below Pinch, the number of units becomes \((N-2)\) rather than \((N-1)\). This perfect match saves one unit, and the two subnetworks are \([H2, C1, CW]\) and \([\text{CON, C2}]\).

6.3 A Strategy for Stream Splitting

In heat exchanger networks there are three different reasons why it is often beneficial and profitable to split process streams into two or more branches:

- Reduce energy consumption
- Reduce total heat transfer area
- Reduce the number of units

In this section, focus will be on the relation between stream splitting and energy consumption (external heating and cooling). In Chapter 2.5, the effect of splitting streams on heat transfer area will be discussed. The last of these three situations will not be discussed in detail, but refers to cases where there is a single or a few streams of one type (hot or cold) and many streams of the opposite type (cold or hot). The best process example is in oil refining, where the crude oil is preheated by a considerable number of hot streams coming from the main fractionator (distillation column) and elsewhere in the plant. In order to fully utilize these hot streams with much smaller \( CP \) values, cyclic matching have been used, resulting in a large number of heat exchangers. When the crude is split into several branches, each of the hot streams can be utilized in a single match, and the number of heat exchangers can be considerably reduced. This is in fact what has happened over the years in these so-called crude preheat trains.
In the application of the Pinch Design Method, situations are commonly encountered where stream splitting is an absolute requirement in order to design heat exchanger networks that achieve minimum external heating and cooling. Fig. 6.2 illustrates two such cases where stream splitting is required to develop an MER design. In Fig. 6.2a, there are three hot streams above Pinch, while there are only two cold streams available to bring these hot streams down to Pinch temperature. In this case, the topology rule ([6.2]) is broken, and a cold stream has to be split in order to have three cold streams (including branches). Considering the CP values in this case, however, it is not possible to split any of the cold streams into two branches that both have CP values large enough to bring a hot stream to Pinch temperature. This means that the CP rule ([6.2]) cannot be satisfied for all the Pinch Exchangers, and one of the hot streams will have to be split. The result then is a return to the original problem where the number of hot streams is larger than the number of cold streams, thus violating the topology rule, and further stream splitting is required.

![Figure 6.2 Stream Splitting Situations (a) above Pinch and (b) below Pinch](image)

In Fig. 6.2b, hot stream H1 cannot be used in a Pinch Exchanger with any of the cold streams, since the CP value of H1 is too small. Splitting one of the cold streams to make it possible for H1 to participate in a Pinch Exchanger is not a good idea, since then the topology rule would be broken (3 cold stream branches and only 2 hot streams). Since, however, hot stream H2 has a CP value that is larger than the sum of the CP values for the two cold streams, a split of H2
results in two branches that can bring the two cold streams to Pinch temperature and the problem is solved.

At this stage, one may question whether it is always possible to split a number of hot and cold streams in such a way that all Pinch exchangers satisfy the $CP$ rule while at the same time not violating the topology rule. The key to the answer to this question is an understanding of the concept of the heat recovery Pinch. The definition of the Pinch is that temperature driving forces are at its minimum, and in the immediate vicinity of the Pinch (just above and just below), the Composite Curves are opening up when moving towards higher and lower temperatures. This means that relations exist [6.3] between the $CP$ value for the sum of the hot streams and the $CP$ value for the sum of the cold streams that are present close to the Pinch. Equality in [6.3] only occurs when the Composite Curves are parallel, and the Pinch point expands into a Pinch region.

Above Pinch: $\sum_j CP_{Cj} \geq \sum_i CP_{Hi}$  
Below Pinch: $\sum_i CP_{Hi} \geq \sum_j CP_{Cj}$  \[6.3\]

Above the Pinch, the total $CP$ “resource” for the cold streams is large enough to allow the $CP$ rule to be satisfied for all Pinch Exchangers bringing hot streams down to Pinch temperature. Similarly, below the Pinch, the total $CP$ “resource” for the hot streams is large enough to allow the $CP$ rule to be satisfied for all Pinch Exchangers bringing cold streams up to Pinch temperature. The reason why it seemed difficult to find a good splitting scheme for the case in Fig. 6.2b (above Pinch) is that the total $CP$ resource for the cold streams (190 kW/°C) is only slightly larger than the total $CP$ value for the hot streams (180 kW/°C). As mentioned in Section 6.2, network design is more complicated when the Composite Curves have a near parallel shape.

7 Design Evolution

While Pinch decomposition guarantees the development of heat exchanger networks with minimum external heating and cooling (relative to a specified value of $\Delta T_{\text{min}}$), it also tends to produce networks where some of the heat exchangers have a rather small duty. Since heat exchangers have an economy-of-scale type cost law as indicated in [7.1], such small units will be quite expensive in capital cost for the limited amount of heat transferred or recovered.
In [7.1], \( a \) is the so-called fixed charge term, \( b \) is a cost coefficient, \( A \) is heat transfer area (for some exchanger types, volume could be used instead of area), and \( c \) is the exponent in this power law type cost equation. The economy-of-scale effect results from the fact that \( c \) is typically less than one, which means that the cost/area ratio reduces with increasing area. The fixed charge term is another economic argument for removing small heat exchangers.

The Pinch decomposition also tends to produce networks with a considerable number of stream splits (although this was not the case for the small process example in Fig. 4.1, with corresponding heat exchanger network in Fig. 6.1). However, the MER design should only be considered as an initial design that can be improved both with respect to economy and network complexity. This section on design evolution will investigate the removal of small units by accepting a modest increase in external heating and cooling (energy relaxation).

While the removal of small units has the potential to reduce total annual cost for the network, it may have the additional benefit of removing some of the stream splits. This is the case if the removed heat exchanger is located on a branch of a splitted stream. The effect will be reduced network complexity and cost and in most cases improved process operability.

### 7.1 A Three-way Trade-off in Network Design

Similar to any design activity, heat exchanger network design involves a number of economic trade-offs. The three most important cost elements are external heating and cooling (Energy), total heat transfer Area, and the number of heat exchangers (Units). This three-way trade-off between Energy, Area and Units is to a large extent affected by the selected value of \( \Delta T_{\text{min}} \), and the optimum trade-off is when the total annual cost of the network is minimised. Other elements that should be considered during optimisation are piping, control valves for stream splitting, network complexity, issues related to operability (including start-up and shut-down) and controllability, however, these aspects are beyond the scope for this chapter.

Design evolution is a simplified form of network optimisation, since the basic structure of the network is maintained, while smaller changes are made to the network by removing small heat exchangers. As in any optimisation, design evolution also needs degrees of freedom. In
heat exchanger networks, the degrees of freedom that can be used to optimise the three-way trade-off mentioned above belong to one of three categories:

- Heat Load Loops
- Heat Load Paths
- Stream Splits

Heat Load Loops in the network is a result of Pinch decomposition and represent a relation between an even number of heat exchangers (2, 4, 6, etc.) in the sense that their duties can be modified without changing the enthalpy balance of the affected streams. The number of independent loops is equal to the difference between the actual number of heat exchangers in the network ($U$) and the minimum number of units without Pinch decomposition ($U_{\text{min}}$). Heat Load Paths in the network represent a relation between an odd number of units (3, 5, 7, etc.) in the sense that their duties can be modified similar to the heat load loops. A major difference is that a path starts with a hot utility exchanger and ends with a cold utility exchanger. While loops are primarily used to remove small units from the network, paths are primarily used to restore violations of $\Delta T_{\text{min}}$ in the network that result from manipulating loops. Stream Splits in the network allow mass flowrates in the stream branches to be varied in order to minimise heat transfer area and thereby minimise cost.

7.2 Using Loops and Paths to evolve Network Design

In Fig. 6.1 there is a very small heater ($H_a$) with a duty of only 100 kW, while the smallest process-to-process heat exchanger is 900 kW. The number of units in this MER design is 8, while the minimum number of units without Pinch decomposition as calculated in Section 4.2 is 7. Using Euler’s Rule from Graph Theory, the number of independent Loops can be found by [7.2], when the number of subnetworks for the overall problem is assumed to be $S=1$.

$$U = N + L - S \quad \Rightarrow \quad L = U - (N - S) = U - (N - 1) = U - U_{\text{min}}$$  \[7.2\]

The number of independent loops in the MER design in Fig. 6.1 is then $8 - 7 = 1$. This single loop (here referred to as L1) can be found by inspection to be:

(L1): $H1 \rightarrow (III) \rightarrow C1 \rightarrow (V) \rightarrow H2 \rightarrow (I) \rightarrow C2 \rightarrow (II) \rightarrow H1$
A heat load loop represents a degree of freedom in the sense that the duties of the heat exchangers that are involved in the loop can be varied within limits. Another characteristic of such a loop is that the process streams involved in the loop flow through two of the units in the loop. This means that if the duty of one of these units is increased, then the duty of the other unit can (and has to) be reduced by the same amount, so that the enthalpy change for the stream remains unchanged, and the stream will reach its target temperature. It should be emphasized that utility exchangers can also be part of such heat load loops, as long as these units operate with the same type of utility. This means that HP steam must be distinguished from LP steam, and cooling water must be distinguished from refrigeration, etc.

For loop L1, assume that the duty of the participating units is changed by $x$ kW. In order to satisfy the enthalpy change for the participating streams (H1, H2, C1 and C2), the change in duty for these heat exchangers must follow the scheme given in [7.3].

$$
Q_{III} = 2,400 + x, \quad Q_{V} = 3,000 - x, \quad Q_{I} = 900 + x, \quad Q_{II} = 6,600 - x
$$

Since heat exchangers with negative duties are infeasible, the optimisation variable $x$ that represents the degree of freedom for this loop is bounded by: $-900 \leq x \leq 3,000$. While in principle one could optimise the network (i.e. minimise its total cost) by varying $x$ in this region, the main objective of design evolution (which distinguishes it from optimisation) is to reduce cost and complexity in the MER design by removing small heat exchangers. For this particular case, unit I with the smallest duty in the loop can be removed if $x = -900$ kW.

The resulting network with updated duties and temperatures are shown in Fig. 7.1. Changing duties for the exchangers in loop L1 will actually violate Pinch decomposition that was the basis for the MER design in Fig. 6.1. It should thus not come as a surprise that there is a penalty involved in removing one of the heat exchangers in the network. As indicated in Fig. 7.1, the temperature driving forces ($\Delta T_{min} = 20^\circ C$) are violated in heat exchanger V since the hot inlet temperature is $130^\circ C$, while the cold outlet is $118^\circ C$, i.e. the temperature difference in the hot end of this unit is reduced to $12^\circ C$. One option is to accept this situation, which means that the penalty for removing unit I is an increase in heat transfer area for unit V. Alternatively, one could require that the driving forces in the network should be restored to $\Delta T_{min}$, which means that the penalty will be increased energy consumption. In an industrial
setting, the penalty will be split into some additional heat transfer area and some additional external heating and cooling, dictated by the trade-off between capital cost and operating cost. For illustration purposes in this case, a heat load path from hot utility to cold utility will be used to restore the temperature driving forces. The selected path must affect the heat exchanger with too small $\Delta T$ in such a way that the driving forces will be restored.

![Evolved Network for the Illustrative Example](image)

**Figure 7.1** Evolved Network for the Illustrative Example

For the network in Fig. 7.1, there is only one heat load path:

(P1): $ST \rightarrow (H_a) \rightarrow C1 \rightarrow (V) \rightarrow H2 \rightarrow (C) \rightarrow CW$

It should be obvious that more steam and cooling water must be used to restore the driving forces. This is in line with the Composite Curves in Fig. 4.2, where increased external heating and cooling will move the two curves apart and increase the temperature driving forces. The change in duty for the units in the heat load path should follow the scheme given in [7.4].

$$Q_{H_a} = 100 + y \quad Q_v = 3,900 - y \quad Q_c = 3,300 + y \quad [7.4]$$

The hot inlet temperature to heat exchanger V will remain at 130°C, which means that the cold outlet temperature must be reduced to 110°C to restore $\Delta T_{min}$ for this unit. This
observation will determine the minimum value of the energy penalty $y$. By following cold stream C1 through heat exchanger V, the relation in [7.5] can be established.

$$40 + \frac{(3,900 - y)}{50} = 110 \quad \Rightarrow \quad y = 400 \text{ kW}$$  \[7.5\]

This means that the energy penalty for removing a heat exchanger with duty $x = 900 \text{ kW}$ is considerably less, i.e. $y = 400 \text{ kW}$. The general relation between $x$ and $y$ is given by [7.6].

$$0 < y \leq x$$ \[7.6\]

The fact that $y < x$ for the example above can be explained as follows: When unit V (a Pinch exchanger) is increased in duty from 3,000 to 3,900 kW, both the hot inlet and the cold outlet temperatures are changed (increased). This means that the duty of this heat exchanger does not have to be reduced to its original value of 3,000 kW in order to restore the driving forces. The resulting network (Design A) has the same structure as the design in Fig. 7.1 and will not be shown here. The units in the heat load path ([7.4]) will have their duties changed by 400 kW, with corresponding changes of the internal network temperatures for streams H2 and C1.

While the initial MER network typically is designed under stringent conditions of Pinch decomposition and $\Delta T \geq \Delta T_{min}$ for all heat exchangers, the design evolution stage allows for relaxation and creative solutions. One should also keep in mind the distinction between hard and soft stream data discussed in Section 3.2. In the MER network in Fig. 6.1, there is a very small steam heater with a duty of only 100 kW in order to increase the temperature of cold stream C1 from 148°C to 150°C. If there is some “softness” in the target temperature for C1, then this heater should not be installed, and one would accept 148°C as the modified target temperature for C1. If the original target temperature of 150°C is a hard specification, then this temperature can still be reached without installing such a small heater. The clue to this rather creative solution is to use the heat load path P1 in the “opposite” direction (see [7.7]) to remove the small heater Hₐ rather than to use heat load loop L1 to remove heat exchanger I.

$$Q_{Hₐ} = 100 - y \quad , \quad Q_V = 3,000 + y \quad , \quad Q_C = 3,300 - y$$ \[7.7\]
By choosing \( y = 100 \text{ kW} \), the resulting network uses 100 kW less steam and cooling water and it has one unit less than the MER design. Of course, this does not come without a penalty, and in the final network for this alternative (Design B) shown in Fig. 7.2, heat exchanger V has a temperature difference of 18°C in the hot end, which is a very small violation of \( \Delta T_{\text{min}} \).

![Diagram of the Alternative Network (Design B) for the Illustrative Example](image)

**Figure 7.2  An Alternative Network (Design B) for the Illustrative Example**

Finally, when removing both heat exchangers I and heater \( H_u \) from the initial MER design, it is not possible to restore \( \Delta T_{\text{min}} \) in the network, since there is no heat load path. The only \( \Delta T_{\text{min}} \) violation in this case is again heat exchanger V that has a temperature difference of 10°C in the hot end. This design is referred to as Design C and is shown together with the other network alternatives in Table 7.1. While focus in Table 7.1 is on \( \Delta T_{\text{min}} \) violation in heat exchanger V, it should not be forgotten that the driving forces for some of the other heat exchangers in the network are also affected, but these effects are marginal for this particular example, and there are no other units that violate the \( \Delta T_{\text{min}} \) specification.

In summary, even for very small processes (as in Fig. 4.1), there are a large number of different heat exchanger networks that can be designed, with differences both in network
topology and network parameters (stream temperatures and heat transfer area for the heat exchangers).

### Table 7.1 Alternative Networks for the Simple Process in Fig. 4.1

<table>
<thead>
<tr>
<th>Design</th>
<th>$Q_H$ (kW)</th>
<th>$Q_C$ (kW)</th>
<th>Units</th>
<th>$\Delta T_V$ (°C)</th>
</tr>
</thead>
<tbody>
<tr>
<td>MER</td>
<td>3,100</td>
<td>3,300</td>
<td>8</td>
<td>20</td>
</tr>
<tr>
<td>A</td>
<td>3,500</td>
<td>3,700</td>
<td>7</td>
<td>20</td>
</tr>
<tr>
<td>B</td>
<td>3,000</td>
<td>3,200</td>
<td>7</td>
<td>18</td>
</tr>
<tr>
<td>C</td>
<td>3,000</td>
<td>3,200</td>
<td>6</td>
<td>10</td>
</tr>
</tbody>
</table>

### 8 Concluding Remarks

In this chapter, the main concepts, representations, procedures and graphical diagrams of basic Pinch Analysis for Heat Integration have been defined and illustrated with application to a small and simple process example. The most important topics covered in this chapter are:

- Data Extraction and Process Modifications
- Plus/Minus Principle, Appropriate Placement and Correct Integration
- Performance Targets for Heating, Cooling and Number of Units
- Heat Recovery Pinch and Pinch Decomposition
- Composite Curves, Heat Cascade and Grand Composite Curve
- Grid Diagram and Pinch Design Method with Matching Rules
- Design Evolution with Heat Load Loops and Paths
- Differences between Grassroots and Retrofit Design

The treatment of the topics in this chapter has been kept on a fairly simple level to convey the most important concepts and provide insight and understanding that will hopefully make the reading of the other chapters in the Handbook easier. For more reading material on these topics, readers are referred to textbooks published over the last 15 years by Shenoy (1995), Smith (2005), Kemp (2007) and Klemeš et al. (2010). Each book provides the state-of-the-art until the year when it was published.
For a more detailed and advanced approach to Heat Integration with focus on Heat Exchanger Network Design for both Grassroots cases and Retrofit projects, readers are referred to Chapter 2.5. A number of issues will be discussed and described in Chapter 2.5 that will add industrial realism to the material. Examples of topics that will be treated are Targets for Heat Transfer Area and Total Annual Cost, Multiple Utilities, Forbidden Matches, different Heat Exchanger Types and a brief introduction to the use of Optimisation in Heat Integration. The different stages in Retrofit Design of heat recovery systems will also be described.

9 Sources of further Information

The primary objective of this chapter has been to introduce the basics of Pinch Analysis as a foundation and core technology in the discipline of Process Integration. For further and more detailed information, the reader is advised to consult Chapter 2.5 on Grassroots and Retrofit Analysis and Design as well as all the other chapters of this Handbook.

References


