International Energy Agency

Implementing Agreement on Process Integration

Annex I (Survey and Strategy)

Supported by

Canada, Denmark, Finland, Portugal, Sweden, Switzerland and UK

A Process Integration PRIMER

by

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3rd and Final version, 9 April 2002

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SUMMARY

This Primer can be regarded as a stand-alone document intended to convey the most basic aspects of Process Integration methods. It also contains towards the end some details about the more recent and advanced elements of Process Integration methods. Since emphasis is on learning, the number of references is kept at a minimum, especially in the core chapters of this Primer.

The Primer is one of the products from the Process Integration Implementing Agreement within the International Energy Agency (IEA). The other main products from Annex I (Survey and Strategy) of this IEA Agreement are:

- "IEA Implementing Agreement on Process Integration: Annex I End-User Survey", T.J. Pears, EA Technology, Capenhurst, UK, November 1997.

- "A Worldwide Catalogue on Process Integration", T. Gundersen, Telemark Institute of Technology, Porsgrunn, Norway, December 1997.

The Primer contains some general material on Process Integration (chapters 1, 2 and 3) in addition to the more tutorial chapters (5, 6, 7 and 8), and one chapter (9) on more advanced aspects of Process Integration. In the first part, there is some information about the IEA project, and the Primer attempts to put Process Integration into a broader perspective.

1. BACKGROUND

The Implementing Agreement on Process Integration within the ***International Energy Agency*** (IEA) was formally started in September 1995, motivated by the recognition that Process Integration was not used to its full potential in industry, with significant differences across geographical regions and industrial branches.

This Primer is based on the idea that there was a need for a document similar to the Pinch Technology Primer that was prepared by Linnhoff March and published by EPRI about ten years ago (1991). Since Process Integration has been expanded considerably and goes far beyond basic Pinch Technology, there is a need for a new Primer that describes the more recent developments. For completeness reasons, however, it was decided to include also the more basic and established parts of Process Integration methods.

Emphasis in the presentation of the material will be on what can be done, rather than how it is done. The Primer is written for ***practitioners*** in the process industries, and those that are interested in more details will have to consult some of the literature that is referenced here, both the journal papers and the growing number of text books available.

Finally, this Primer is expected to be one of many ways to ***disseminate*** knowledge about Process Integration into operating companies, engineering and contracting companies, consultants and even software vendors. Information about the IEA Agreement on Process Integration can be found on the ***Web*** site http://www.tev.ntnu.no/iea/pi/.

2. INTRODUCTION

The structure of the Primer has been designed to hopefully allow practitioners in industry to get a smooth introduction to Process Integration methods, with its powerful concepts, representations and graphical diagrams. Gradually, the Primer will provide more detailed information about application areas and relevant methods. For details about Process Integration Technologies, the reader will have to refer to some of the recommended literature given as references in this Primer. A number of good text books and overview articles also exist that will be referred to.

2.1 Definition of Process Integration

Process Integration is a fairly new term that emerged in the 80's and has been extensively used in the 90's to describe certain Systems oriented activities related primarily to Process Design. It has incorrectly been interpreted as Heat Integration by a lot of people, probably caused by the fact that Heat Recovery studies inspired by the Pinch Concept initiated the field and are still core elements of Process Integration. It appears to be a rather dynamic field, with new methods and application areas emerging constantly. The ***definition*** used in this context is the one used by the IEA since 1993:

> *"Systematic and General Methods for Designing Integrated Production Systems, ranging from Individual Processes to Total Sites, with special emphasis on the Efficient Use of Energy and reducing Environmental Effects".*

This definition brings Process Integration very close to Process Synthesis, which is another Systems oriented technology. Process Integration and Synthesis belong to Process Systems Engineering (see figure 2.1), which is Systems Engineering principles applied to Processes.

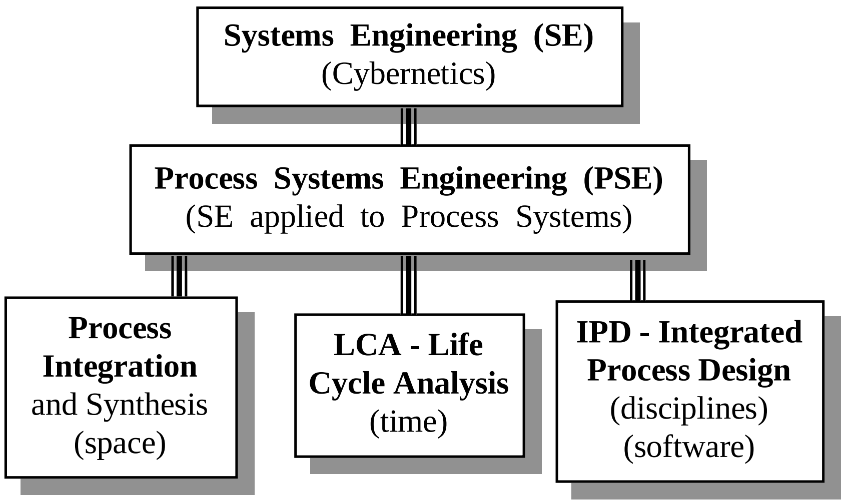


Fig. 2.1 Process Integration among similar Terms

While Process Integration and Synthesis are Systems approaches in ***space*** (the whole plant, the entire site, and sometimes even the whole region including domestic needs), Life Cycle Analysis is a Systems oriented methodology in ***time***, and Integrated Process Design is a Systems view across scientific ***disciplines*** and ***software*** systems.

When using the term Process Integration, we both refer to certain industrial tasks and to classes of methods to address these tasks. In this Primer, we will concentrate on methods that have been developed specifically to address these tasks, however, there are also a number of methods of a more general nature that can be used to a larger or smaller extent to solve Process Integration problems. These methods will also be briefly described.

2.2 Current Status of Process Integration

Process Integration is a strongly growing field of Process Engineering. It is now standard ***curriculum*** for process engineers in both Chemical and Mechanical Engineering at most universities around the world, either as a separate topic or as part of a Process Design or Synthesis course. At UMIST (Manchester, UK) there is a separate Department of Process Integration. ***Research*** at UMIST has for 15 years been supported by a large number of industrial companies through a Consortium that was established in 1984. As part of the IEA project on Process Integration, we have identified about 35 other universities around the world involved in research in this field.

While Heat Recovery was the initial focus of Process Integration, the ***scope*** has been expanded considerably during the late 80's and the 90's to cover several aspects of Process Design. A key feature of this ***expansion*** has been the use of basic concepts from heat recovery in other areas through the use of analogies. This has, for example, made it possible to use heat recovery techniques to study Mass Transfer processes in general and Water Management in particular. Unfortunately, the last attempt to review the entire field (Gundersen and Naess, 1988) is way out of date. The growth of Process Integration during the last 10 years has made it almost intractable to produce an updated review.

Appropriate tools, such as user-friendly and reliable ***software***, are keys to industrial use, and there are now around 50 computer programs available to assist the engineer in one or more areas of Process Integration. This software covers a wide range of problem areas, and the quality of the software is ranging from high standard commercial products being used routinely in industry, to prototype software from universities that were developed primarily to assist research. Some of this software may even be available free of charge.

There is an increasing ***international co-operation*** on Process Integration. The IEA project has already been mentioned, and it is at present supported by seven countries (Canada, Denmark, Finland, Sweden, Switzerland, Portugal and UK). Within the Nordic Energy Research Programme, Process Integration has been one of seven activities since 1995. The European Commission is also funding research in the area of Process Integration. Typically, these projects have a broad international representation.

While the field of Process Integration in the past has been allocated one or two single sessions on large international meetings, the trend today is to have separate ***conferences*** focusing specifically on Process Integration. Examples include PRES'98 (Prague), PI'99 (Copenhagen), PRES'99 (Budapest), and continuing with PRES'2000 (Prague).

In conclusion, Process Integration has evolved from a Heat Recovery methodology in the 80's to become what a number of leading industrial companies in the 90's regarded as a ***Major Strategic Design and Planning Technology***. With this technology, it is possible to significantly reduce the operating cost of existing plants, while new processes often can be designed with reductions in both investment cost and operating cost.

2.3 From History to the Future

Process Design has evolved through distinct "generations". Originally (first generation), inventions that were based on ***experiments*** in the laboratory by the chemists, were tested in pilot plants before plant construction. The second generation of Process Design was based on the concept of ***Unit Operations***, which founded Chemical Engineering as a discipline. Unit Operations acted as building blocks for the engineer in the design process. The third generation considered ***integration*** between these units; for example heat recovery between related process streams to save energy.

A strong trend today (fourth generation) is to move away from Unit Operations and focus on ***Phenomena***. Processes based on the Unit Operations concept tend to have many process units with significant and complex piping arrangements between the units. By allowing more than one phenomena (reaction, heat transfer, mass transfer, etc.) to take place within the same piece of equipment, significant savings have been observed both in investment cost and in operating cost (energy and raw materials). Most of the industrial applications of this idea have been based on trial and error. Research is progressing, however, trying to develop systematic methods in this area to replace trial and error. No doubt, this will affect the discipline of Process Integration, since we no longer look at integration between units only, but also at integration within units.

2.4 Process Integration and the "Pinch" Concept

The single most important concept and the one that originally gave birth to the field of Process Integration is the ***Heat Recovery Pinch***, discovered independently by Hohmann (71), Umeda et al. (78-79) and Linnhoff et al. (78-79). It was Linnhoff's group at UMIST in Manchester, however, that developed this concept into an industrial technology in the 80's. The concept has later been expanded into new areas by using various analogies.

The most obvious analogy is between heat transfer and mass transfer. In heat transfer, heat is transferred with temperature difference as the driving force. Similarly, in mass transfer, mass (or certain components) is transferred with concentration difference as the driving force. The corresponding ***Mass Pinch***, developed by El-Halwagi and Manousiouthakis (89-90), has a number of industrial applications whenever process streams are exchanging mass in a number of mass transfer units, such as absorbers, extractors, etc.

One specific application of the Mass Pinch is in the area of Wastewater Minimization, where optimal use of water and wastewater is achieved through reuse, regeneration and possibly recycling. The corresponding ***Water Pinch***, developed by Wang and Smith (94), can also be applied for design of Distributed Effluent Treatment processes.

The most recent extension is the ***Hydrogen Pinch*** technology, developed by Towler and Alves (96-99). Oil refineries experience these days an increasing need for hydrogen to meet new product specifications (for example on diesel and gasoline). The Hydrogen Pinch method is a tool to optimize the hydrogen distribution system and to evaluate the scope for introducing purification units (such as PSA, membranes and cryogenic units).

In summary, the Pinch concept is a Systems tool since it provides critical information on a total plant or even site level. The concept is also (as shown above) generally applicable in other areas than heat recovery. Actually, whenever an amount (heat or mass) has a quality (temperature or concentration), the concept of ***Composite Curves*** provides a Systems view of the problem concerning efficient recovery (or re-use) of resources. The ***"Pinch"*** then shows the location on these curves where there is an accumulated deficit of an amount above a certain quality. Figure 2.2 shows Composite Curves for a Heat Recovery problem (left) and a Wastewater Minimization application (right).

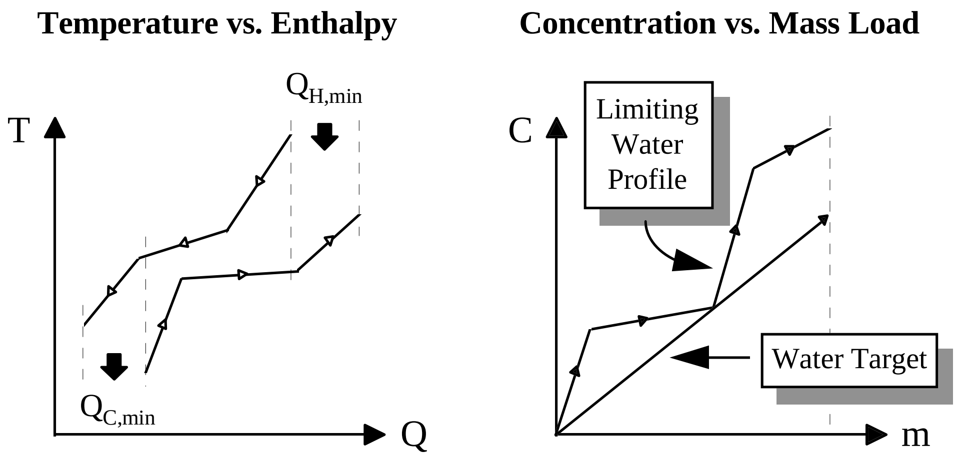


Fig. 2.2 The General Concept of Composite Curves applied to Heat and Mass Transfer

2.5 Performance Targets before Design

In the previous section, it was concluded that the Composite Curves represent a concept that is general and fundamental in Process Engineering. Another important concept from "early days" Pinch Technology is the idea of establishing objective performance ***targets before*** going into the ***design*** phase. Examples of such targets in the area of heat recovery are figures for minimum energy consumption, fewest number of heat transfer equipment, minimum total heat transfer area, and minimum total annual cost. While some of these targets are based on thermodynamics (such as energy), others are based on heuristic rules (such as the fewest number of heat exchangers). Finally, some targets are actually only <u>estimates</u> of the best performance (such as heat transfer area and total annual cost).

Examples of other targets in Process Integration include minimum wastewater, minimum shaftwork in low temperature processes, minimum emissions, maximum power production for total sites, etc. All these and previously mentioned targets have two important features:

1) Any design can be objectively compared with the "best possible".
2) The way some targets are calculated also provides guidelines for design.

In his text book on Chemical Process Design, Smith (1995) puts strong emphasis on the idea of establishing Performance Targets prior to Design.

2.6 Schools of Methods in Process Integration

The three major features of Process Integration methods are the use ***heuristics*** (insight), about design and economy, the use of ***thermodynamics*** and the use of ***optimization*** techniques. There is significant overlap between the various methods and the trend today is strongly towards methods using all three features mentioned above. The large number of structural alternatives in Process Design (and Integration) is significantly reduced by the use of insight, heuristics and thermodynamics, and it then becomes feasible to address the remaining problem and its multiple economic trade-offs with optimization techniques.

Despite the merging trend mentioned above, it is still valid to say that ***Pinch Analysis*** and ***Exergy Analysis*** are methods with a particular focus on Thermodynamics. ***Hierarchical Analysis*** and ***Knowledge Based Systems*** are rule-based approaches with the ability to handle qualitative (or fuzzy) knowledge. Finally, ***Optimization*** techniques can be divided into deterministic (Mathematical Programming) and non-deterministic methods (stochastic search methods such as Simulated Annealing and Genetic Algorithms).

One possible ***classification*** of Process Integration methods is to use the two-dimensional (automatic vs. interactive and quantitative vs. qualitative) representation in figure 2.3.

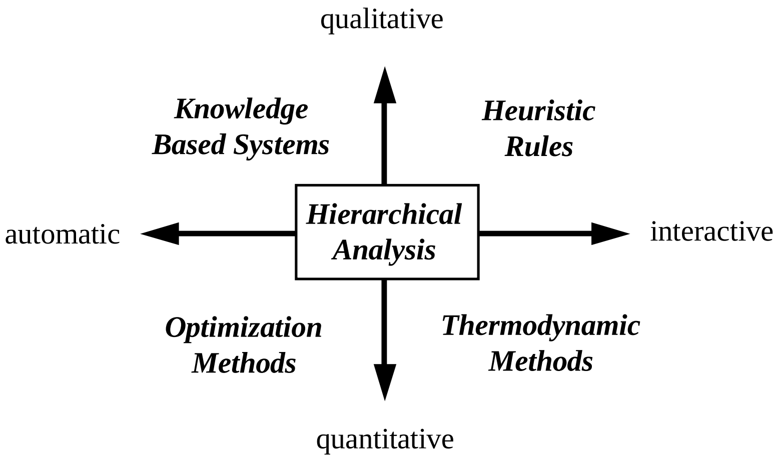


Fig. 2.3 One possible Classification of Process Integration Methods

Hierarchical Analysis is placed in the middle of the figure to indicate that all sensible design methods are (or should be) based on this idea in order to make the complete design problem tractable by systematic methods.

3. PROCESS INTEGRATION APPLICATION AREAS

This chapter will indicate major as well as minor industrial problem areas (or tasks) where Process Integration methods can be used. It is, however, of limited value to describe these tasks in detail without referring to Process Integration methods, representations, concepts and graphical diagrams. Therefore, this chapter has been made very brief, while chapter 5 and the following chapters go into more detail about the various tasks and in parallel present the most important features of the applicable Process Integration methods.

3.1 Classification of Industrial Tasks

Process Integration was defined in section 2.1 to be systematic and general methods for designing integrated production systems. Within the IEA project, special emphasis is given to efficient use of Energy and reducing Environmental impact. The methods do, however, go beyond these objectives. While reducing energy consumption through heat recovery normally increases investment cost, Process Integration methods also enable industries to reduce equipment cost for a specified level of energy usage.

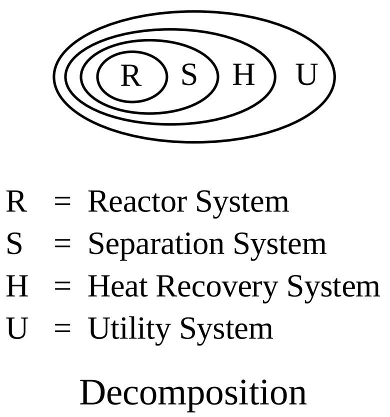
Process Integration also affects raw material utilization, since it has been shown that improved heat recovery will allow increased recycling in the process. In this way, raw material is used to maximize yield of the desired product and not being lost in less valuable outlet streams. Also, increased recycling may allow reduced reactor conversion (per pass), which may improve reactor selectivity and reduce byproduct formation.

The following list of keywords and activities indicate typical ***application areas*** of Process Integration for a large number of industrial branches:

- Planning, Design and Operation of Processes and Utility Systems
- Short Term (Scheduling) and Long Term Planning (including Strategic Planning)
- New Designs and various Retrofit Projects
- Improving Efficiency (Energy and Raw Material) and Productivity (Debottlenecking)
- Continuous, Semi-Continuous and Batch Processes
- All aspects of Processes, such as Reactors, Separators and Heat Exchanger Networks
- Integration between the Process and the Utility System
- Integration between Processes w.r.t. Material Streams and Energy Streams
- Integration between Industrial Sites, Power Stations and District Heating/Cooling
- Operability Issues (Flexibility, Controllability and Switchability)
- Waste and Wastewater Minimization
- Various aspects of Emissions Reduction

3.2 Some Useful Representations

A number of symbols and representations have been used to convey areas of application for Process Integration. The ***Onion Diagram*** in figure 3.1 indicates the hierarchy of most processes, with the reactor as the core of the system. Once the Reactor System has been designed, the compositions of the outlet streams from the reactors are known, and one can proceed to the design of the Separation System. The next level is the design of the Heat Recovery System, and finally the Utility System is addressed.



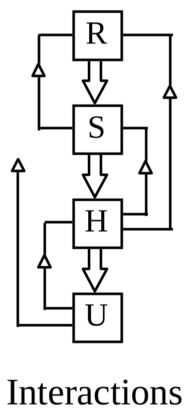


Fig. 3.1 The Onion Diagram of Process Design also applicable to Process Integration

While there is a logical information flow from the core of the onion towards the outer layers (suggesting a sequential and decomposed approach), there are important interactions that require iteration towards the center of the onion, alternatively a simultaneous approach is required. In addition, there are actually more layers to the onion. One example is total sites, where several processes interact with each other, and where there normally is a central utility system for heating, cooling, power, etc. Going even further, one may also want to include the community surrounding the plant.

Another frequently used representation, especially within Pinch Technology, is the ***Rubic Cube*** in figure 3.2. It indicates the start of Pinch Technology, focusing on Heat Exchanger Networks with minimum Energy consumption for Grassroots Designs. During the 80's and the 90's, Pinch Technology has expanded in all three dimensions of the cube to cover almost complete Process Design.

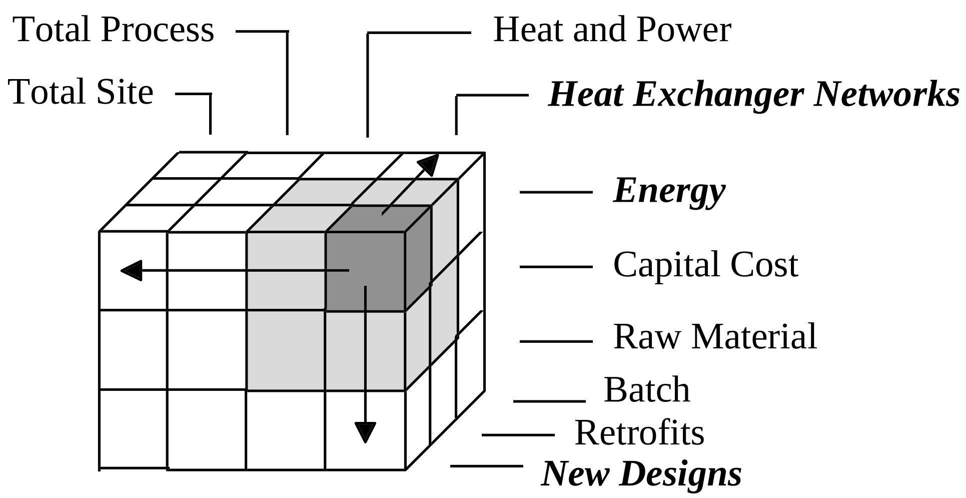


Fig. 3.2 The Rubic Cube indicating the development of Pinch Technology

Finally, Douglas (1988) in his text book on conceptual design of chemical processes is using a decomposition referred to as Hierarchical Levels in the decision making, where the reactor system is placed at the top level. This decomposition is very similar to the Onion Diagram in figure 3.1.

4. SEQUENCE OF PRESENTATION

The material will to some extent be presented in the sequence that the methods were developed historically. Hopefully, this will give the reader a smooth introduction to Process Integration by starting with the basic concepts and then subsequently move into more advanced applications.

The Primer will first discuss energy considerations in new design of continuous processes (chapter 5), followed by various retrofit situations (chapter 6). In chapter 7 there is a similar discussion of batch processes. In chapter 8 there is a brief introduction to representations and models used in Mathematical Programming approaches. Finally, the Primer will describe at some level of detail the more recent and advanced Process Integration applications in chapter 9. Since chapter 5 contains representations and concepts referred to in all the later chapters, it is assumed that the reader gets acquainted with this material before digging into special chapters of interest.

5. BASIC CONCEPTS FOR HEAT RECOVERY IN NEW DESIGN OF CONTINUOUS PROCESSES

It feels natural to start with the single most important industrial application area for Process Integration. The development that followed the discovery of the ***Heat Recovery Pinch*** has been unique in Process Design when it comes to real life applications in industry based on results from academic research.

One of the important advantages of basic Pinch Analysis is that a number of concepts, representations and graphical diagrams have been developed that both are excellent learning aids but also provide the engineer with powerful tools for industrial applications. In order to introduce and explain some of these concepts, representations and graphical diagrams, a simple example (figure 5.1) will be used throughout most of the presentation.

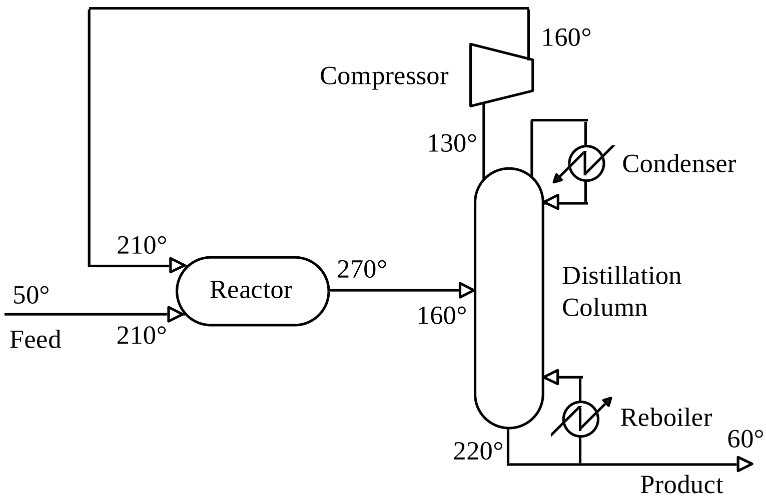


Fig. 5.1 A Simple Example of a Process with Reaction, Separation and Heat Exchange

The process example in figure 5.1 will also be used to illustrate the ***four phases*** of Pinch Analysis in the design of heat recovery systems for both new and existing processes:

1) Data Extraction, which involves collecting data for the process and the utility system.
2) Targeting, which establishes figures for best performance in various respects.
3) Design, where an initial Heat Exchanger Network is established.
4) Optimization, where the initial design is simplified and improved economically.

First, however, it is important to make sure that a proper problem definition has been established. This also includes relevant cost data and economic criteria. While this chapter looks at new designs (grassroots), chapter 6 will discuss the modification of existing plants (retrofits).

5.1 Data Extraction (Phase 1)

The most time consuming and often most critical phase is the identification of the need for heating, cooling, boiling and condensation in the process. This task is more art than science, and if not carried out properly, the final design will not be the best possible. It is quite easy to accept too many features of the proposed flowsheet, which inevitably results in the situation where many good opportunities are excluded from the analysis.

Once the Data Extraction and corresponding Targeting (Phase 2) activities are completed, it is time to look back and question some of the decisions made for the Reactor and Separation Systems. The idea is then to identify process modifications that will increase the potential for heat recovery and/or allow the use of cheaper utilities.

In practice, there are a number of situations where heat integration is not desirable. Examples include long distances (costly piping), safety (heat exchange between hydrocarbon streams and oxygen rich streams), product purity (potential leakage in heat exchangers), operability (start-up and shut-down), controllability and flexibility. A reasonable strategy is, however, to start by including all process streams and keep the degrees of freedom open. Later, practical considerations can be used to exclude some of these streams and degrees of freedom, and the engineer will then at any time be able to establish the consequences with respect to energy consumption and total annual cost.

A central part of data extraction is the identification of heating and cooling requirements in the process. The necessary data for each process stream are the following:

m	=	mass flowrate (kg/s, tons/h, etc.)
Cp	=	specific heat capacity (kJ/kg°C)
Ts	=	supply temperature (°C)
Tt	=	target temperature (°C)
ΔHvap	=	heat of vaporization for streams with a phase change (kJ/kg)

Table 5.1 shows the data extracted for the simple example in figure 5.1, including data for available utilities, where ΔQ values are variables that will be optimized during targeting and design. In order to analyze area and investment cost for heat exchangers, heat transfer conditions must be established. This is typically done by assigning a film heat transfer coefficient (h) to each process stream. The total coefficient (U) for heat transfer between a hot stream (H_i) and a cold stream (C_j) is then estimated by the simple equation:

$$1/U = (1/h_{Hi}) + (1/h_{Cj})$$

Table 5.1 Stream and Utility Data for the Example in Figure 5.1

Stream	ID	Ts(°C)	Tt(°C)	mCp(kW/°C)	ΔQ(kW)	h(kW/m² °C)
Reactor Outlet	H1	270	160	18	1980	0.5
Product	H2	220	60	22	3520	0.5
Feed	C1	50	210	20	3200	0.5
Recycle	C2	160	210	50	2500	0.5
Reboiler	C3	220	220		2000	1.0
Condenser	H3	130	130		2000	1.0
High pressure steam	HP	250	250		(var.)	2.5
Medium pressure steam	MP	200	200		(var.)	2.5
Low pressure steam	LP	150	150		(var.)	2.5
Cooling water	CW	15	20		(var.)	1.0

5.2 Performance Targets (Phase 2)

As indicated in section 2.5, an important feature of Process Integration is the ability to identify Performance Targets before the design phase is started. For heat recovery systems with a specified value for the minimum allowable approach temperature (ΔT_{min}), targets can be established for Minimum Energy Consumption (external heating and cooling), Fewest Number of Units (process/process heat exchangers, heaters and coolers) and Minimum Total Heat Transfer Area. In addition, the corresponding calculations will also identify the ***Heat Recovery Pinch***, which acts as a bottleneck for heat recovery.

For new designs, it is possible to return to data extraction and ***modify the process*** in such a way that the impact of the heat recovery pinch is reduced or even eliminated. Then a new Pinch point will be identified, and the procedure can be repeated.

It is also possible to combine targets for energy, units and total heat transfer area into an estimate of the total annual cost. By repeating these calculations for different values of ΔT_{min}, it is possible to identify a good starting value for the level of heat recovery. This exercise of ***pre-optimization*** (Linnhoff and Ahmad, 1990) has been referred to as "Super-Targeting" (which also gave name to one of the commercial software packages available).

While initial methods used a global value for ΔT_{min}, later methods allowed individual stream contributions to the overall minimum approach temperature (ΔT_i) reflecting the heat transfer conditions for each process stream, as indicated by its film heat transfer coefficient (h_i). One model that has been used is $\Delta T_i = C / \mathrm{sqrt}(h_i)$, where C is a common adjustable factor, reflecting the chosen level of heat recovery (see Ahmad et al., 1990).

5.2.1 Minimum Energy Consumption

By adding enthalpy changes for the hot and cold process streams in table 5.1 separately and for each temperature interval in the process, the hot and cold ***Composite Curves*** in figure 5.2 can be established. The distillation column (H3, C3) is <u>not</u> included at this stage.

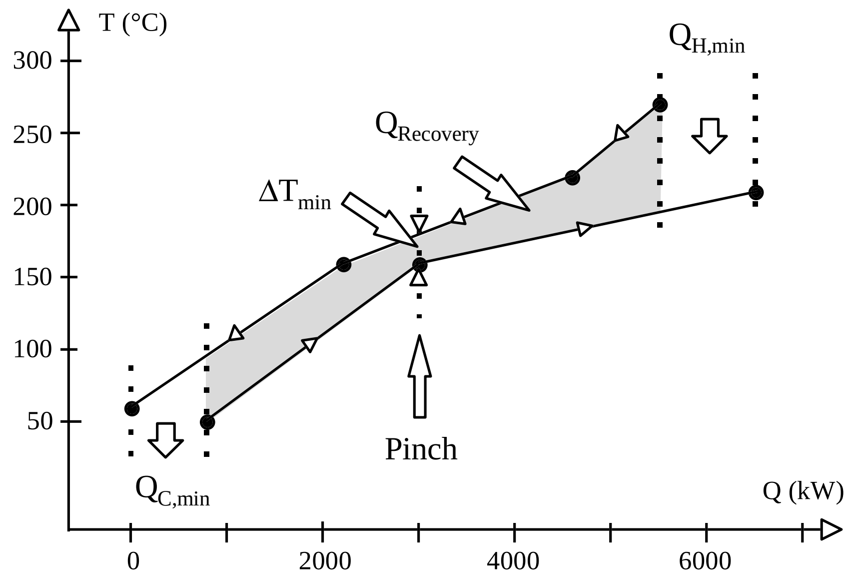


Fig. 5.2 Hot and Cold Composite Curves for part of the Example in Figure 5.1

Composite Curves provide valuable information about maximum heat recovery ($Q_{Recovery}$), minimum external heating ($Q_{H,min}$), minimum external cooling ($Q_{C,min}$) and location of the heat recovery Pinch for a given value of ΔT_{min}. As mentioned in section 2.4, Composite Curves can be applied and provide valuable information whenever an amount (such as heat) has a quality (such as temperature). The advantages of graphical representations (such as the one in figure 5.2) include a pedagogic aspect of understanding, they provide the engineer with an overview of the problem, they illustrate important economic trade-offs, and finally they represent information in a very concentrated form. The results (targets) that can be extracted from figure 5.2, where ΔT_{min} = 20°C, are the following:

Maximum Heat Recovery:	$Q_{Recovery}$	=	4700 kW
Minimum External Heating:	$Q_{H,min}$	=	1000 kW
Minimum External Cooling:	$Q_{C,min}$	=	800 kW
Pinch Point (caused by a cold stream):	$T_{Pinch,C}$	=	160 °C
Corresponding Pinch for hot streams:	$T_{Pinch,H}$	=	180 °C

As indicated in table 5.1, the values for mCp are assumed to be constant. This simplifies the calculations from numerical integration to a summation over intervals. When the value of Cp varies considerably with temperature, introducing stream segments can piece-wise linearize the temperature/enthalpy relation for the stream. The same applies for a stream that has a phase change.

Based on the Composite Curves in figure 5.2, a general strategy for ***Process Modifications*** can be established. In Pinch Analysis, this strategy has been referred to as the ***Plus/Minus*** principle (Linnhoff and Vredeveld, 1984), which means to increase ("plus") heat available above Pinch and/or heat demand below Pinch or to reduce ("minus") heat demand above Pinch and/or heat available below Pinch (see figure 5.3). Examples of such Process Modifications include changes in pressure for distillation columns and evaporators, changes in flowrates for some streams, and new target temperatures for streams when possible.

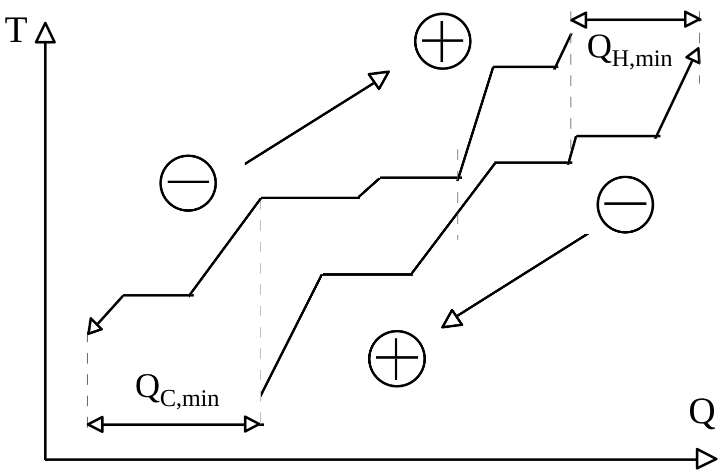


Fig. 5.3 The Plus/Minus principle applied to Composite Curves

While <u>graphical</u> diagrams such as the Composite Curves are excellent tools for learning the methods and understanding the overall energy situation, minimum energy consumption and the heat recovery pinch are more often obtained by <u>numerical</u> procedures. Typically, these are based on the ***Heat Cascade*** in figure 5.4. In the Heat Cascade, the supply and target temperatures of all process streams divide the temperature scale into Temperature Intervals, in the same way as the construction of the Composite Curves.

On the left side of the diagram in figure 5.4, hot streams supply heat into the various intervals according to a hot temperature scale. Similarly, on the right hand side of the diagram, cold streams extract heat from the various intervals according to a cold temperature scale. The difference between the hot and the cold temperature scale is the value of the minimum approach temperature, ΔT_{min}, thus the heat cascade ensures feasible heat transfer according to an economic criterion.

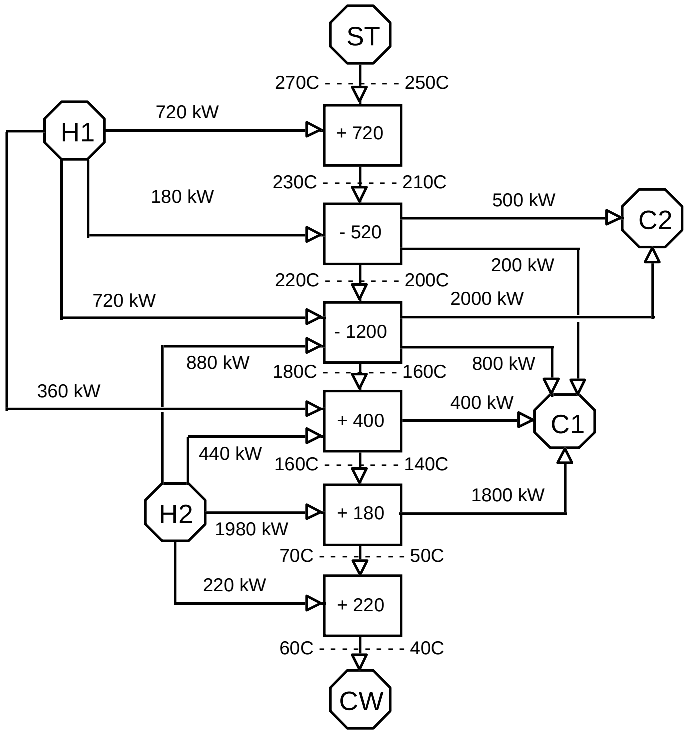


Fig. 5.4 The Heat Cascade for part of the Example in Figure 5.1

The objective is to allow heat surplus in one interval to cascade down to the next interval, in order to maximize heat recovery. In figure 5.4, a surplus of 720 kW in the first interval can be used to cover the deficit of 520 kW in the second interval. There is an accumulated heat surplus of 200 kW that can be cascaded further into the third interval. The heat deficit of 1200 kW in the third interval can then be covered by 200 kW of cascaded heat <u>and</u> the inevitable supply of 1000 kW from hot utility, such as steam (ST). The last three intervals in figure 5.4 all have heat surplus, and the total heat that must be removed from the cascade in this lower part by cold utility, such as cooling water (CW), is 800 kW.

Some ***important results*** can be extracted from the Heat Cascade in figure 5.4 and the subsequent discussion. First, we have identified the need of 1000 kW from hot utility (which, of course, is the same as indicated by the Composite Curves) and 800 kW removed by cold utility. By having this minimum exchange of heat between the process and the utility system, there is no heat flow between intervals 3 and 4. This is the ***Pinch Point*** (bottleneck for heat recovery). Above the Pinch temperature (180°C/160°C), we have a sub-system with heat deficit, and below Pinch there is a sub-system with heat surplus.

This ***decomposition*** effect is a very important property of the Process Pinch, and it has several important impacts on the design of energy efficient processes. In heat recovery, it becomes important not to transfer heat across pinch. Each single kW taken from the sub-system above Pinch (heat deficit) and transferred to the sub-system below pinch (heat surplus) will immediately require an extra 1 kW of both steam and cooling water. The easiest way to obey this rule is to design two separate heat exchanger networks, one above and one below the Process Pinch. Violations of the Pinch decomposition are also the key to identifying good retrofit projects, as will be discussed in chapter 6.

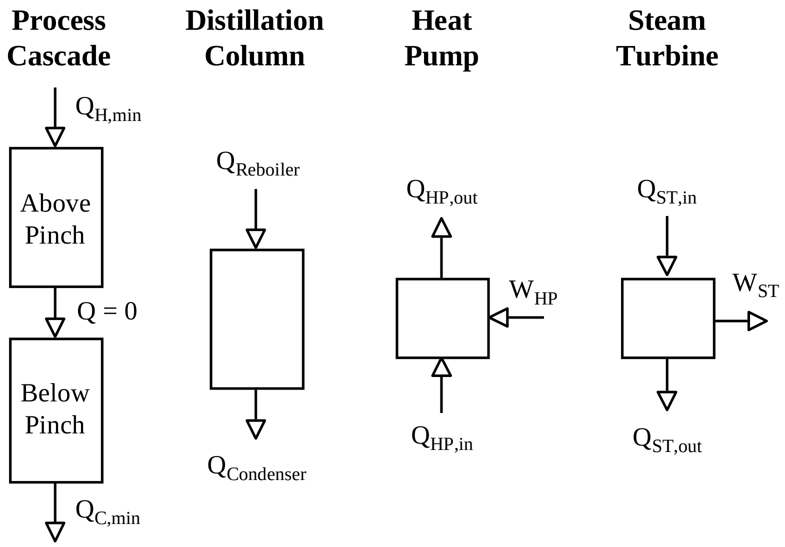


Fig. 5.5 Pinch Decomposition in a system with "Suppliers" and "Customers"

The decomposition property also has a major impact on the use of Heat Pumps, the use of back-pressure Steam Turbines and the integration of special equipment such as Distillation Columns and Evaporators. The ***general rule*** that can be formulated based on the decomposition principle is to try to match heat "suppliers" with heat "customers", otherwise heat integration does not serve any energy saving purpose, and will only introduce additional investment cost and less operable processes.

This scenario with suppliers and customers is indicated in figure 5.5 without any temperature details, however, for each item (process cascade, distillation column, heat pump and steam turbine), temperature is decreasing as we move from the upper to the lower part of the figure. Of course, the "supplier" must provide heat at a sufficiently high temperature to meet the needs of the "customer". The following ***explicit rules*** are derived from the decomposition principle using figure 5.5:

- A ***distillation column*** should only be integrated with the background process if:

  a) The reboiler temperature is lower than the Pinch temperature
  b) The condenser temperature is higher than the Pinch temperature

- A ***heat pump*** should only be integrated with the background process if it takes heat from below Pinch and lifts it above Pinch. Similarly, it should only be integrated with a distillation column if the column can not be integrated with the background process, since heat pumping is more expensive with respect to investment than direct process integration using heat exchangers.

- A ***steam turbine*** should only be integrated (i.e. back pressure or extraction turbine) with a process or distillation column if the outlet steam has a high enough condensing temperature (high enough pressure) to be used above the process Pinch or in a column reboiler. Otherwise a condensing turbine should be used.

While the Heat Cascade provides crucial insight about efficient use of energy through heat integration, it is also the basis for an important school of methods based on mathematical models. The heat cascade is a special case of the ***Transshipment Model*** which is frequently used in Operations Research and forms the basis for some of the optimization based methods such as ***Mathematical Programming***. This will be discussed in more detail in chapter 8.

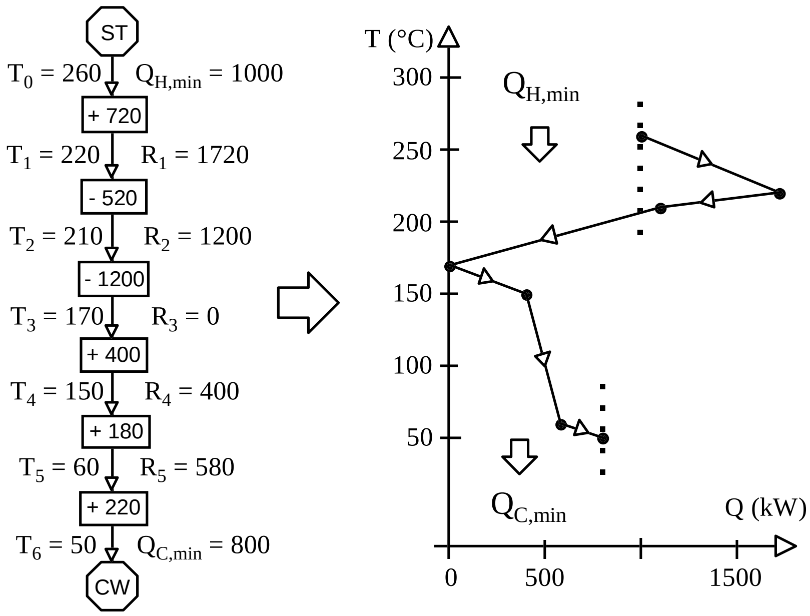


Fig. 5.6 The Grand Composite Curve for the Example in Figure 5.1 (except the Column)

The decomposition principle and the discussion following the scenario in figure 5.5 both have one major disadvantage. We need a better view of the amount of heat available at various temperatures from the "suppliers", and the corresponding need for heat at various temperatures among the "customers". This additional information about ***Load*** and ***Level*** is crucial for applying the explicit rules mentioned above. It does not help if the distillation column can provide heat above the Process Pinch, if the amount of heat needed at that specific temperature is very small.

Such information is implicitly available in the heat cascade, however, a much better overview of the situation is obtained if the Heat Cascade is transformed into another graphical diagram called the ***Grand Composite Curve*** (Linnhoff et al., 1982). This diagram, which also has been referred to as the Heat Surplus Diagram, is generated by plotting so-called modified interval temperatures against the corresponding flow of heat between intervals in the cascade. This is shown in figure 5.6 for the example process, after the addition of minimum hot and cold utility requirements. The modified temperatures are simply the average between the hot and the cold temperatures (+/- $\Delta T_{min}/2$), an adjustment that allows the drawing of hot and cold streams and utilities in the same temperature scale, while satisfying the need for minimum driving forces.

The Grand Composite Curve has a number of ***industrial applications***, mostly related to the utility system and heat and power considerations. Typically, the Grand Composite Curve can be used to qualitatively and to some extent quantitatively address the following tasks:

- Identify a near-optimal ***set of utility types*** (both load and level) to cover the need for external heating and cooling in the process. A Utility Grand Composite Curve (Hall, 1989) consisting of available utilities, such as for example various steam levels, flue gas from a furnace or gas turbine, hot oil circuits, cooling water, refrigeration, etc., can be combined in such a way that total utility cost is minimized.

- Identify potential for ***steam production*** below Pinch, if the process Pinch is at a sufficiently high temperature. This means that steam generation (typically LP steam) is acting as a cold utility.

- Identify potential for utilizing so-called "pockets" in the Grand Composite Curve for additional ***power production***. There is one such pocket above Pinch in figure 5.6. If the temperature difference had been sufficiently large between the part of the process where there is local heat surplus and the corresponding part where there is local heat deficit, there would have been some scope for producing steam that could have been used in a back pressure turbine. The turbine then borrows steam generated in the process and returns steam for heating at a lower level after power production.

- Identify scope for using ***heat pumps*** in the process to reduce both hot and cold utility consumption. Typically, this is the case where there is a distinct Pinch point, with flat profiles both immediately above and below the Pinch. In such cases, a significant amount of heat can be transferred from the heat surplus region below Pinch to the heat deficit region above Pinch, by using a heat pump with moderate temperature lift.

- Identify whether there is scope for ***integration*** of special equipment such as distillation columns or evaporators with the background process.

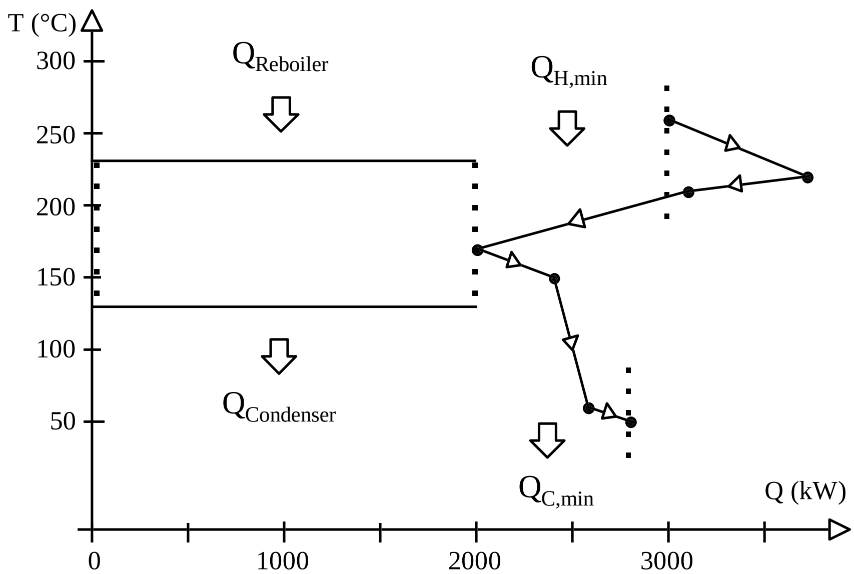


Fig. 5.7 Grand Composite Curve with Box Representation for Distillation Columns

Returning to the process example in figure 5.1, the Grand Composite Curve can be used to give a quick and simple answer about the scope for integrating the ***distillation column*** with the background process, or whether it should be operated with utilities (steam and cooling water). After heat integration with the process has been analyzed, the next step could be to evaluate the scope for heat pumping.

Figure 5.7 shows the Process Grand Composite Curve and the Temperature/Enthalpy Diagram for the distillation column in figure 5.1. Since the distillation column operates across the Pinch, there will be no energy savings from integration with the process. This also follows from the decomposition concept illustrated in figure 5.5. The graphical representation in figure 5.7 has also been referred to as the ***Andrecovich*** diagram.

Later extensions within Pinch Analysis include a refinement of the box representation, where a ***Column Grand Composite Curve*** (CGCC) shows the need for reboiling and condensation at various temperatures in the column. The CGCC is based on converged profiles from a rigorous column simulation, and can be used to identify the scope for distributed reboiling and condensing as well as feed pre-heating or pre-cooling, and finally changes in the reflux ratio for the column.

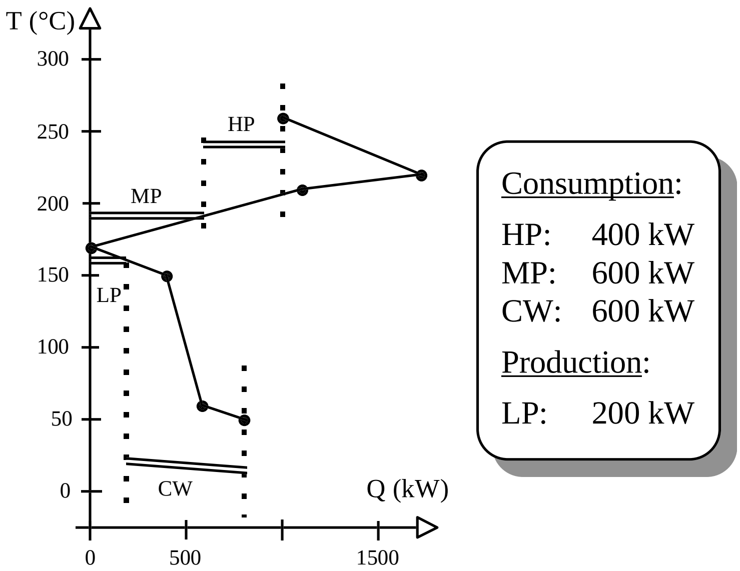


Fig. 5.8 Process and Utility Grand Composite Curves for the Example in Figure 5.1

Referring to the simple process example in figure 5.1 and the list of available utilities in table 5.1, the Grand Composite Curve can, as explained, be used to identify the set of utilities with minimum energy cost. Typically, this means to maximize the use of cheaper utilities in order to minimize the use of more expensive utilities. This is shown in figure 5.8, where the amount of MP steam is maximized and limited by the situation where the MP part of the Utility Grand Composite Curve touches the Process Grand Composite Curve. Similarly, a potential for LP steam production is identified below Pinch, and the amount is again limited by the point where the LP part of the Utility Grand Composite Curve touches the Process Grand Composite Curve.

If we assume that utility prices are 200 $/kWyr for HP steam, 170 $/kWyr for MP steam, 140 $/kWyr for LP steam and 20 $/kWyr for cooling water, the annual energy cost for the utility mix in figure 5.8 is 166,000 \$/yr. When using HP steam and cooling water only, the corresponding annual energy cost is 216,000 \$/yr, i.e. 30% higher.

The Grand Composite Curve enables the engineer to identify a set of utilities that gives minimum energy cost. As always, however, there is a ***trade-off*** between operating cost (energy) and investment cost (number of heat exchangers and their total heat transfer area). Thus, the following important factors need further investigation before accepting the set of utilities proposed in figure 5.8:

- Temperature driving forces will be reduced when introducing MP and LP steam, which means larger ***heat transfer area*** in some utility and process/process exchangers. As a result, there will be a significant increase in the investment cost.

- New ***Utility Pinch Points*** will be introduced when maximizing MP steam usage and LP steam production. This will result in tighter designs and more complex heat exchanger network structures.

- The ***decomposition*** feature of the Process Pinch also applies to Utility Pinches. This means for example that heat pumps can be used to transfer heat across (from below to above) all Pinch points in order to reduce total heating and cooling requirements (Process Pinch) or reduce the need for a more expensive utility (Utility Pinch).

- The number of ***heat transfer units*** will increase whenever new utilities are introduced and whenever Utility Pinches are created, which means increased investment cost.

- The ***complexity*** of the heat exchanger network (number of units, piping and stream splits) will increase with an increasing number of Pinch points included during design.

While significant savings in energy cost can be obtained by introducing intermediate (and thus cheaper) utilities, there will be a corresponding increase in investment cost (total heat transfer area and the number of units will increase). Minimum total annual cost is found by exploring these trade-offs.

The Grand Composite Curve (GCC) has the inherent limitation (which also in many respects is an advantage) that details about the individual streams are not shown. Thus, any conclusion about integration of distillation columns and heat pumps as well as steam generation, must be evaluated carefully by looking beyond the GCC and into the actual number of streams that would be involved. If a heat pump would have to extract (deliver) heat from (to) a large number of streams, it would not be economically interesting. The same applies if we end up with a large number of steam boilers.

5.2.2 Fewest Number of Units

A heuristic estimate for the minimum number of units is obtained by using Euler's Rule from Graph Theory as the basis:

$$U = N + L - S$$

where U is the number of units (process/process heat exchangers, heaters and coolers), N is the total number of process streams and utility types, L is the number of heat load loops in the network and S is the number of sub-systems in the network. Assuming there are no heat load loops (it will be shown later that loops can be removed) and no sub-networks (sets of hot and cold streams in perfect heat balance, which would be a coincidence), the following can be used as an estimate for the fewest number of units:

$$U_{min} = N + 0 - 1 = N - 1$$

In order to obtain Maximum Energy Recovery (MER) or minimum energy consumption, however, it was shown above that ***decomposition*** at the Process Pinch must be respected. This means that separate heat exchanger networks must be designed above and below Pinch, and the corresponding minimum total number of units is given by:

$$U_{min,MER} = (N - 1)_{above} + (N - 1)_{below}$$

In the case of multiple utilities, as indicated in figure 5.8, new Utility Pinch points will be introduced whenever a cheaper utility is maximized in order to minimize a more expensive utility. There are also cases with near-Pinches that could be included to make tight design situations easier. A more general equation for the fewest number of units is thus:

$$U_{min,MER} = \sum_{i=1}^{n_p+1} (N_i - 1)$$

where n_p is the number of Pinch points (Process and Utility Pinches), and N_i is the total number of process streams and utility types that are present between two neighboring Pinch points, alternatively above the highest Pinch and below the lowest Pinch.

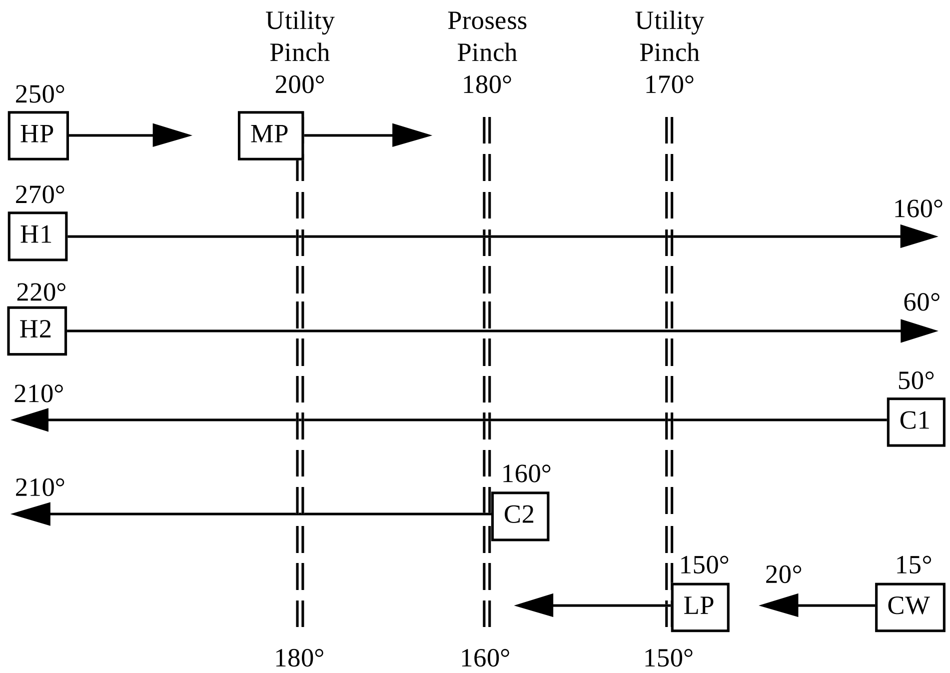


Fig. 5.9 Stream Grid for the Example in Figure 5.1 with Multiple Pinch Points

It is obvious from these equations that the target for minimum number of units depends on the number of utility types that are used and the number of Pinch points (process and utility pinches) where strict decomposition is implemented. For the simple example in figure 5.1 with 4 process streams (keeping the distillation column out of the discussion) and up to 4 utility types, the fewest number of units varies considerably. If we only use HP and CW and do not decompose at the process Pinch, the fewest number of units is 4, while it is 14 if we use all 4 utility types and decompose at all 3 Pinch points. Obviously, the economic trade-off between energy cost and equipment cost will have an optimum that is closer to 5 heat transfer units than 14.

The ***Stream Grid*** (Linnhoff and Flower, 1978a) shown in figure 5.9 is an important representation for the design of heat exchanger networks. It can also be used to assist in the application of the (N-1) rule to calculate the fewest number of heat exchangers for the various scenarios of multiple utilities and the existence of Process and Utility Pinch points.

5.2.3 Minimum Number of Shells

Refinements have been made in Pinch Analysis (Ahmad and Smith, 1989) to reflect the fact that very few industrial heat exchangers are pure counter-current. These refinements relate to both the number of heat exchange units (now counted as number of shells rather than heat exchangers) and to heat transfer area (see the discussion in section 5.2.4).

So far, these extensions only apply to Shell & Tube exchangers, where ***correction factors*** for heat transfer area (f_T) are used that depend on mCp values and temperatures for the streams. These factors represent deviations from pure counter-current heat exchange when using models and equations for 1-2 Shell & Tube exchangers. If the value of f_T falls under a minimum acceptable value, the number of shells must be increased by one, and the procedure is repeated. With these extended models, it is possible to obtain a target for the minimum number of shells rather than units. The next section on minimum heat transfer area also applies to shells in 1-2 configurations, with the addition of the f_T factor when calculating area.

5.2.4 Minimum Heat Transfer Area

Estimating the need for total heat transfer area in the network of heat exchangers <u>before</u> design is both the most time consuming (need software) and the most uncertain targeting activity. There are large uncertainties in heat transfer coefficients, and simplified assumptions are made about the network structure when calculating minimum total area.

In Pinch Analysis, a target for minimum area is obtained by applying and expanding the concept of counter current heat exchange between two streams to the situation with many hot and cold streams. The resulting heat flow model is the vertical one illustrated in figure 5.10. The idea of ***Vertical Heat Transfer*** between the Composite Curves is aiming at optimal use of the available driving forces in order to minimize total heat transfer area. Since, however, the general equation for heat transfer area is:

$$A = Q / (U \bullet \Delta T_{LM} \bullet f_T)$$

it is the <u>product</u> of heat transfer coefficient (U) and driving forces (ΔT_{LM}) that should be optimally distributed, not driving forces alone. This will be briefly discussed below.

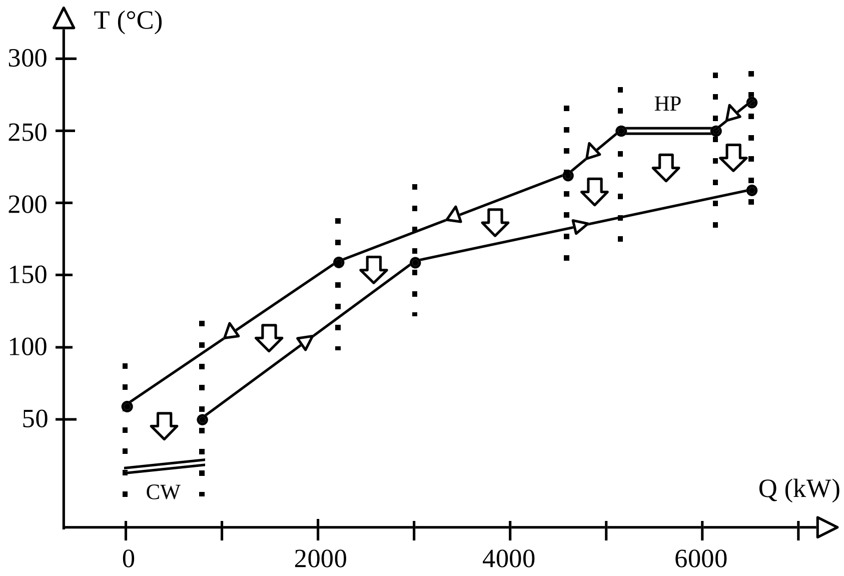


Fig. 5.10 Vertical Heat Transfer for the Example in Figure 5.1

In order to achieve vertical heat transfer in a heat exchanger network, however, all heat exchangers in the same ***Enthalpy Interval*** (marked by the dotted lines in figure 5.10) must have exactly the same temperature profiles. This can only be achieved by considerable splitting and mixing of streams and a large number of small heat exchangers (must apply the N-1 rule to each enthalpy interval). The corresponding network is therefore referred to as the "Spaghetti Design", and serves exclusively as a calculation model for total heat transfer area (Townsend and Linnhoff, 1984).

In figure 5.10, hot and cold utilities are included and the result is often referred to as the ***Balanced Composite Curves***. In this case only HP steam and cooling water are used. The actual calculation of minimum area, based on the concept of vertical heat transfer, is done with the so-called Bath formulae (after the place where the equation was presented):

$$A_{min} = \Sigma_j \left(1 / \Delta T_{LM,\, j} \right) \Sigma_i \left(q_i\right) / \left(h_i\right)$$

where q_i is the change in enthalpy and h_i is the film heat transfer coefficient for stream (i) in enthalpy interval (j). By applying this equation to the example problem we get the following results:

With MP/LP: $A_{min} = 775\ m^2$ Without MP/LP: $A_{min} = 632\ m^2$

Thus, while the introduction of MP and LP steam reduces total energy cost with about 30%, there is an increase in the target for heat transfer area of about 23%. In addition, as discussed in the previous section, there will be an increase in the number of units, and the network structure will be more complex.

As mentioned above, there are a number of uncertainties related to these target values for minimum area. In addition to the fact that heat transfer coefficients are uncertain by nature, the vertical model and the Bath equation have two severe limitations:

- To achieve minimum area, a large number of heat exchangers, splitters and mixers are required. Due to economy of scale effects, cost optimal heat exchanger networks will have close to the fewest number of units rather than close to minimum area. The so-called ***Spaghetti Design*** should only be regarded as a model for calculating A_{min}.

- The strict vertical model will only result in minimum area if all film heat transfer coefficients for the hot streams area equal (h_H), and that all cold stream film heat transfer coefficients are equal (h_C). With significant differences in these coefficients, streams with low film heat transfer coefficients should be matched and allowed more driving forces at the expense of matches between streams with large film heat transfer coefficients that will be assigned less driving forces. As a result, there may be considerable non-vertical (***Criss-Cross***) heat transfer.

These limitations are important, however, the main use of the target for minimum area is to be able to estimate total annual cost ahead of design for various values of ΔT_{min}, in order to identify a good starting point for the design exercise.

5.2.5 Total Annual Cost

By ***combining*** targets for minimum ***energy*** consumption, fewest number of ***units*** or ***shells*** and minimum heat transfer ***area***, as well as cost data for utilities, cost equations for heat exchangers, some economic factors such as payback time or interest rate, and the number of operating hours per year, it is possible to obtain figures for Total Annual Cost. There are also uncertainties in these estimates, for example related to the fact that we only have figures for total area, and not how this area is distributed among the heat exchangers. With an economy of scale type cost equation, such a distribution is important for the final result.

Experience from industrial projects have shown, however, that some of the uncertainties and assumptions in the calculation of area and total annual cost tend to cancel, and that the estimated total cost often is within a few percent from the total cost of the final heat exchanger network (using the same cost and economic data).

As mentioned above, the main purpose of estimating Total Annual Cost (TAC) is to identify a good ***starting point*** for network design. This is done by calculating the different targets and the resulting total annual cost for various values of ΔT_{min}. By selecting a value for ΔT_{min} where TAC has a minimum, the initial heat exchanger network (see next section) will have a <u>structure</u> that is compatible with the final optimal network.

In the case of multiple utilities, a similar economic trade-off should be explored in the targeting phase. Methods have been developed within Pinch Analysis that can be used to identify near-optimal amounts of the various utilities (Parker, 1989, and Hall et al., 1992). It should also be mentioned that utility selection and process modifications interact and must be considered simultaneously.

5.3 Network Design (Phase 3)

This section will be presented in much less detail than the previous section where a number of concepts, representations and graphical diagrams were introduced that are of a general nature with several different applications in Process Integration.

Design of Heat Exchanger Networks in various industries is primarily carried out using the now classical ***Pinch Design Method*** (Linnhoff and Hindmarsh, 1983). While the original method focused on minimum energy consumption and the fewest number of units, later graphical and numerical additions made it possible also to consider heat transfer area and total annual cost during design. Both the original features and the later extensions have been implemented in current state of the art commercial software packages for Heat Exchanger Network Design.

The basic Pinch Design Method respects the decomposition at Process and Utility Pinch points and provides a ***strategy*** and matching ***rules*** that enable the engineer to obtain an initial network, which achieves the minimum energy target. The ***Stream Grid*** presented in section 5.2.2 is very useful in the design phase and acts as a drawing board, where the engineer places one match at a time using these matching rules. The Pinch Design Method also indicates situations where stream splitting is required to reach the minimum energy target. Stream splitting is also important in area considerations and the optimal use of temperature driving forces.

The ***design strategy*** mentioned above is simply to start design at the Pinch, where driving forces are limited and the critical matches for maximum heat recovery must be selected. The matching rules simply ensure sufficient driving forces, and they attempt to minimize the number of units. The design then gradually moves away from the pinch, making sure that hot streams are utilized above Pinch (limited resource), and vice versa for cold streams below Pinch (limited resource).

The ***matching rules*** for Pinch exchangers (those situated immediately above or below Pinch) can be expressed mathematically by (where H_i and C_j are potential streams to be matched in a heat exchanger):

Above Pinch	Below Pinch
$mCp_{Cj} \geq mCp_{Hi}$	$mCp_{Hi} \geq mCp_{Cj}$
$n_C \geq n_H$	$n_H \geq n_C$

Making sure that every unit fully satisfies the enthalpy change of either the hot or the cold stream (the "tick-off" rule) minimizes the number of units. If the inequalities above are not satisfied for a complete set of Pinch exchangers, stream splitting has to be considered in order to reach Maximum Energy Recovery (MER). It is always possible by stream splitting to satisfy all the inequalities, since total mCp for cold streams are larger than total mCp for hot streams above Pinch, and vice versa below Pinch.

Later extensions enable the engineer to also consider investment cost during design, in particular the effect of each match on total heat transfer area. The ***Driving Force Plot*** (Linnhoff and Vredeveld, 1984) makes is possible to evaluate graphically whether a suggested match is using reasonable driving forces compared with what is available in that temperature region of the process.

The ***Remaining Problem Analysis*** (Ahmad, 1985) is more quantitative tool, that provides figures for energy (E), number of units (U), heat transfer area (A) and total annual cost (TAC), if a suggested match is accepted. Adding actual figures for partial designs under development to target values for the remaining problem provides accumulated figures for TAC.

Figure 5.11 shows an initial heat exchanger network for the process example in figure 5.1, when the distillation column is not integrated with the rest of the process. The targeting section (and figure 5.7) concluded that (in this particular case) integrating the column with the rest of the process would not result in any energy savings, since the column operates across the process Pinch. Also notice that only HP steam and cooling water is used for external heating and cooling. The network in figure 5.11 has been established using the Pinch Design Method and is drawn using the ***Stream Grid***.

The initial heat exchanger network in figure 5.11 reaches the targets values for energy consumption (1000 kW of heating and 800 kW of cooling) and minimum number of units. For the case with two utilities, four process streams and strict decomposition at the process Pinch, the minimum number of units is (5-1) above Pinch and (4-1) below Pinch, in total 7 heat exchangers including heaters and coolers.

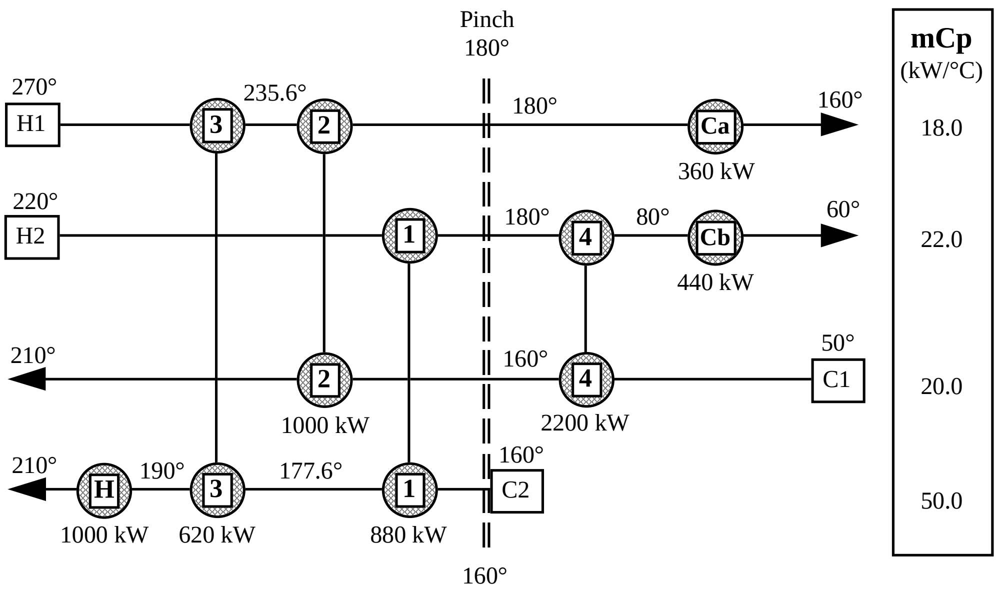


Fig. 5.11 Initial Heat Exchanger Network for the Example in Figure 5.1

It is important to notice that several initial networks may be generated. The Pinch Design Method provides rules for matching streams that eliminate certain configurations but still open up for alternatives. The larger the industrial problem is, the more alternatives exist, and the engineer is free to make choices based on practical considerations such as safety, operability, controllability, etc.

In the small process example of figure 5.1, only one significant alternative exists above Pinch. By splitting stream C2, it is possible to reduce the number of heat exchangers by one, as shown in figure 5.12. While this example illustrates the existence of subnetworks above Pinch, stream splitting is more often used to be able to reach minimum energy consumption. The best example is crude preheat trains in oil refineries, where there is one large cold stream (the crude oil) and many hot streams (intermediate products and pumparounds from the distillation tower), and the crude is typically splitted in two, three or four branches before and after the desalter.

Splitting of streams is also introduced to save total heat transfer area (better utilization of the available temperature driving forces), and in some rare cases splitting is also used to reduce the number of units, as indicated in figure 5.12. Here, the heating needed by stream C2 matches exactly the cooling required for streams H1 and H2 above Pinch.

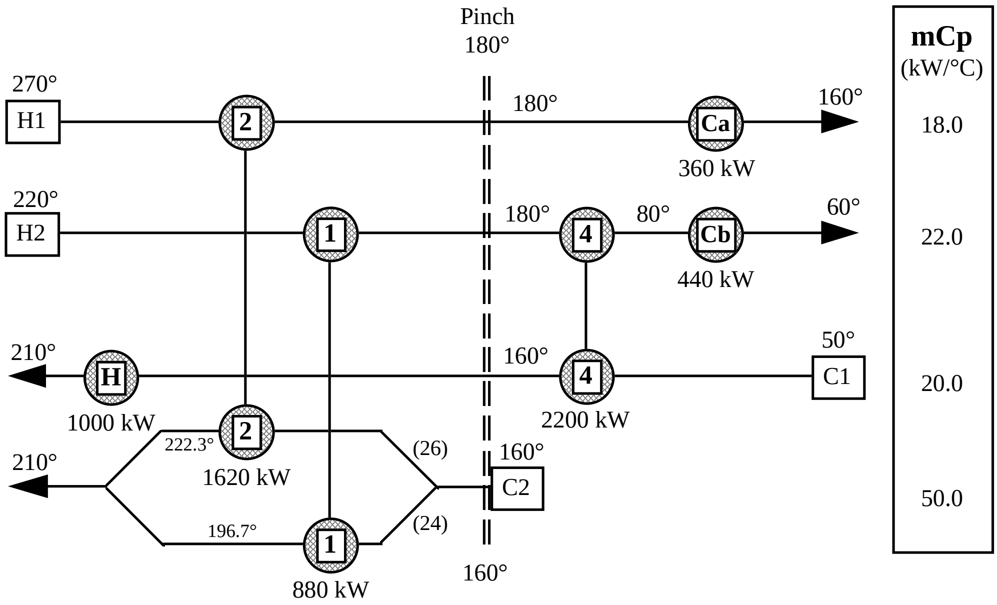


Fig. 5.12 An alternative initial Heat Exchanger Network for the Example in Figure 5.1

<u>Below</u> Pinch there is also a design alternative, since the cooler Ca could have been replaced by a match between H1 and C1, with a corresponding increase in the duty of cooler Cb and decrease in the duty for exchanger 4.

5.4 Network Optimization (Phase 4)

Heat Exchanger Networks for maximum energy recovery established by the Pinch Design Method, should only be regarded as initial designs. The strict decomposition at the Pinch normally results in networks with stream splits and a few rather small units. As mentioned above, the basic Pinch Design Method focuses on minimum energy consumption while using the fewest number of units. Even though extensions such as the Driving Force Plot and the Remaining Problem Analysis help the engineer to also minimize total heat transfer area, ***Total Annual Cost*** is not necessarily at its minimum, and some final <u>optimization</u> is required. With a good initial value for ΔT_{min}, only minor network changes (described as design evolutions by Linnhoff and Hindmarsh, 1983) are required in most cases. The matches of the initial network depend on the Pinch location, and since the Pinch point depends on the value of ΔT_{min}, this becomes a key parameter in Pinch based methods.

The ***Degrees of Freedom*** available for network optimization are the following:

- Since the initial network is produced by respecting strict Pinch decomposition, there will be more than the minimum number of units. Thus, there are ***Heat Load Loops*** in the network, where 2 or 4 or 6, etc. heat exchangers (including heaters and coolers) have duties than can be modified in a systematic way (see figure 5.13) without changing the stream target temperatures. Such loops can be used to remove small units in the network (<u>discrete</u> optimization) or simply to obtain an area distribution in the network with lower total annual cost (<u>continuous</u> optimization).

- There will also be ***Heat Load Paths*** from a hot utility exchanger through some of the process/process exchangers to a cold utility exchanger. These paths can be used to restore unacceptable temperature driving forces in some units after manipulation of heat load loops. Since increasing the duties of utility exchangers will affect the energy/area trade-off, this procedure has similarities to shifting the Composite Curves for the overall problem. A heat load path, however, affects only a limited number of units. In some cases, such heat load paths can also be used to remove small units.

- Flowrates of the individual branches of a ***Stream Split*** can be varied in order to reduce total heat transfer area (or actually investment cost) of the heat exchangers involved. This is a local optimization affecting a limited number of units, but interactions exist between this optimization and the manipulation of heat load loops and paths.

Figure 5.13 shows a Heat Load Loop in the initial heat exchanger network from figure 5.11, involving all four process/process exchangers. Another heat load loop exists between the two process/process exchangers 2 and 4 and the two coolers Ca and Cb. In figure 5.13, it is possible to remove exchanger 1 by selecting X = 880 kW or to remove exchanger 3 by selecting X = - 620 kW.

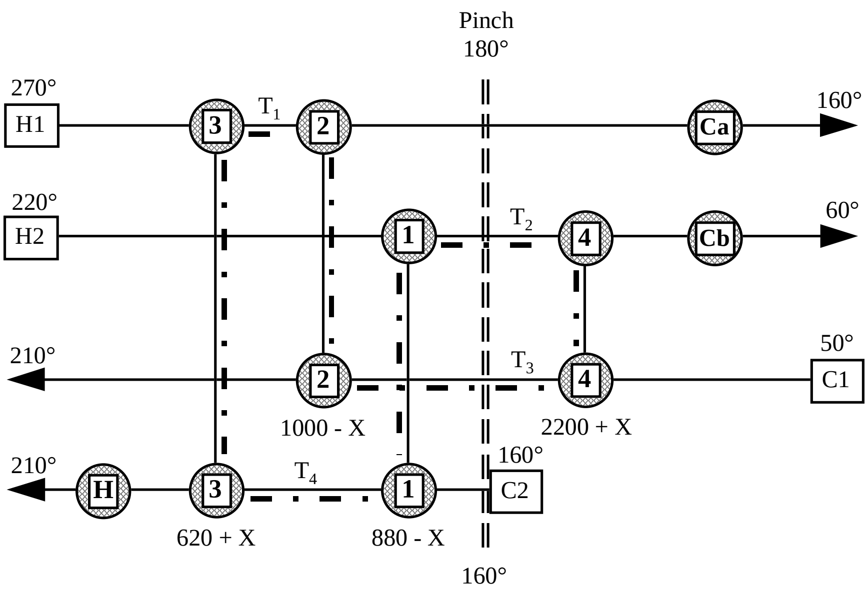


Fig. 5.13 Heat Load Loop in the initial Heat Exchanger Network

With economy of scale type cost equations, the obvious strategy is to remove small units, since these are expensive in terms of cost for the given heat recovered ($/kW). These effects are illustrated by the following cost equation:

$$C_{hex} = a + b \bullet (A)^{c}$$

where (a) is the fixed charge term, (b) is the cost factor for heat transfer area (A), and (c) is the area exponent which typically is less than 1.0 ("economy of scale").

A Heat Load Path from the heater (H) to one of the coolers (Cb) is illustrated in figure 5.14. This network is the result after removing heat exchanger (1) from the network in figure 5.13. As shown in figure 5.14, heat exchanger (2) has infeasible driving forces in both the hot (186.7ºC is less than 210ºC) and the cold end (180°C is less than 204°C) of the unit. When increasing external heating and cooling by Y kW, the temperature of stream C1 between exchangers (2) and (4) will be reduced, and similarly the temperature of stream H1 between exchangers (3) and (2) will be increased. The network eventually becomes feasible, however, in this case it requires Y = 880 kW to satisfy the requirement of ΔT_{min} = 20°C in the cold end of heat exchanger (2).

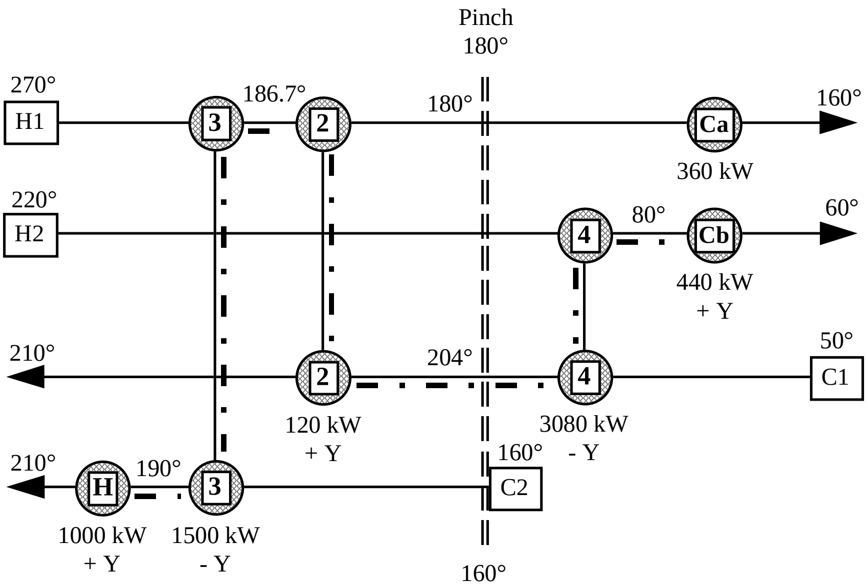


Fig. 5.14 Heat Load Path in a Heat Exchanger Network

The Heat Load Path in figure 5.14 could alternatively be used to remove heat exchanger (2) with infeasible driving forces, rather than to restore feasible heat transfer in that unit. This is achieved by selecting Y = - 120 kW. Exchanger (4) would then have a temperature difference of only 10°C in the hot end, a situation that cannot be resolved. This value is considerably less than the assumed value for ΔT_{min} (20°C), but it may still be economic. In this case, the manipulation of the heat load path results in the removal of a very small unit and at the same time a reduction in external heating and cooling, as shown in figure 5.15. In this network, the number of units is actually reduced to the absolute minimum.

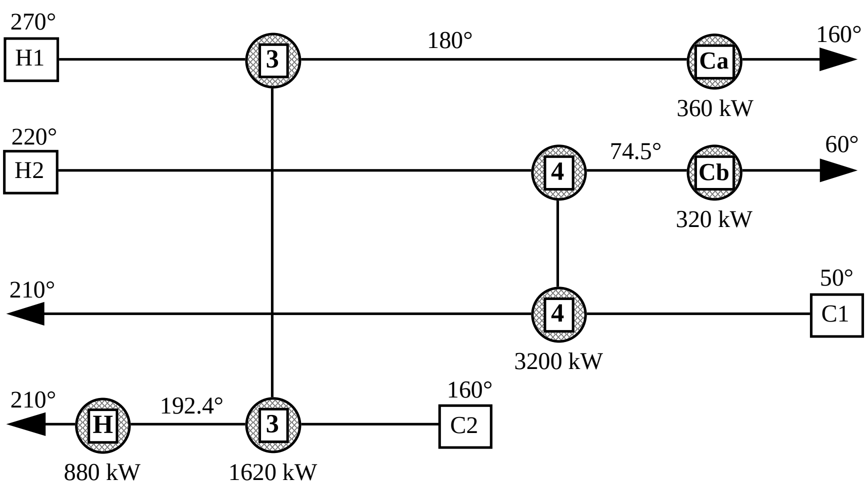


Fig. 5.15 Heat Exchanger Network for the simple Example with only 5 Units

In the general case, when a heat exchanger with duty X kW is removed from a network by breaking a heat load loop, it requires that Y kW is added to the external heating and cooling consumption through a heat load path in the network. The following is always valid:

$$0 \leq Y \leq X$$

In our example, Y = X = 880 kW, however, in many cases Y can be considerably less than X. The rare situation where Y is 0 kW only happens in cases where hot and cold streams or stream branches for the potential problem exchangers have equal mCp values.

While the Stream Grid is an excellent representation during heat exchanger network design and optimization, the flowsheet representation gives the engineer an idea about the piping that is involved in recovering the 4820 kW of heat indicated in figure 5.15. In this design, 1620 kW is recovered in exchanger 3 between streams H1 and C2, while 3200 kW is recovered in exchanger 4 between streams H2 and C1.

The resulting process flowsheet for the simple network solution in figure 5.15, is shown in figure 5.16, where the heat exchanger network is included among the other process units. As discussed earlier, the distillation column is not integrated with the rest of the process, and runs with steam (HP) and cooling water (CW).

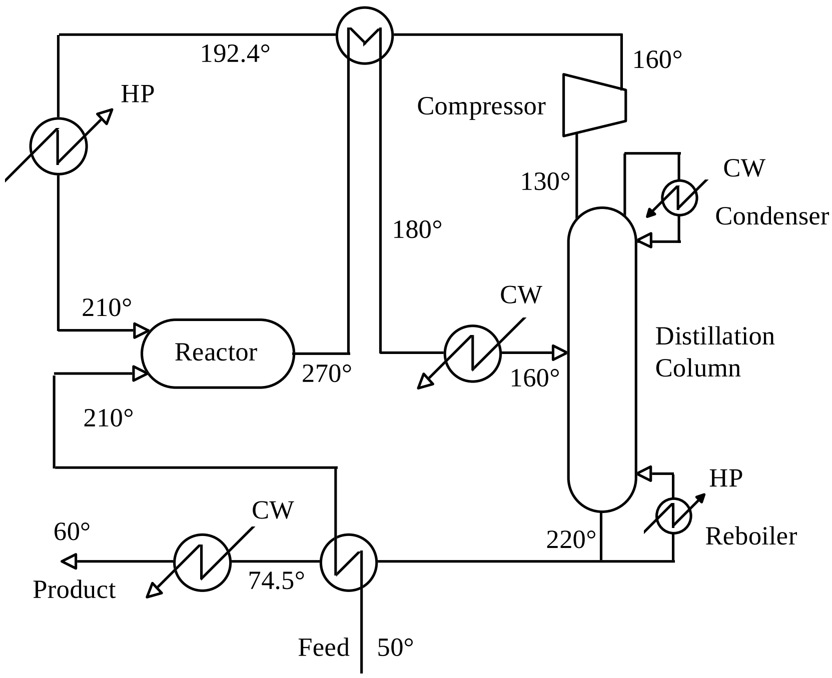


Fig. 5.16 Process Example with one possible Heat Recovery solution (from Fig. 5.15)

For comparison, the corresponding MER heat exchanger network from figure 5.11 is included in the process flowsheet in figure 5.17. This solution requires two more heat exchangers and considerably more piping, however, its total heat transfer area is considerably less than the solution in figure 5.16. This is not obvious, but can be explained by the fact that the MER design in figure 5.17 uses more energy (1000 kW of heating versus 880 kW in figure 5.16), more heat exchangers (7 units versus 5 units in figure 5.16), and finally that all exchangers have temperature driving forces that are at least 20°C. Figure 5.16 has one heat exchanger with a large duty (3200 kW) and only 10°C temperature difference in the hot end. In these arguments, the distillation column and its use of HP steam and cooling water is not included, since the column is operated in the same way in figures 5.16 and 5.17.

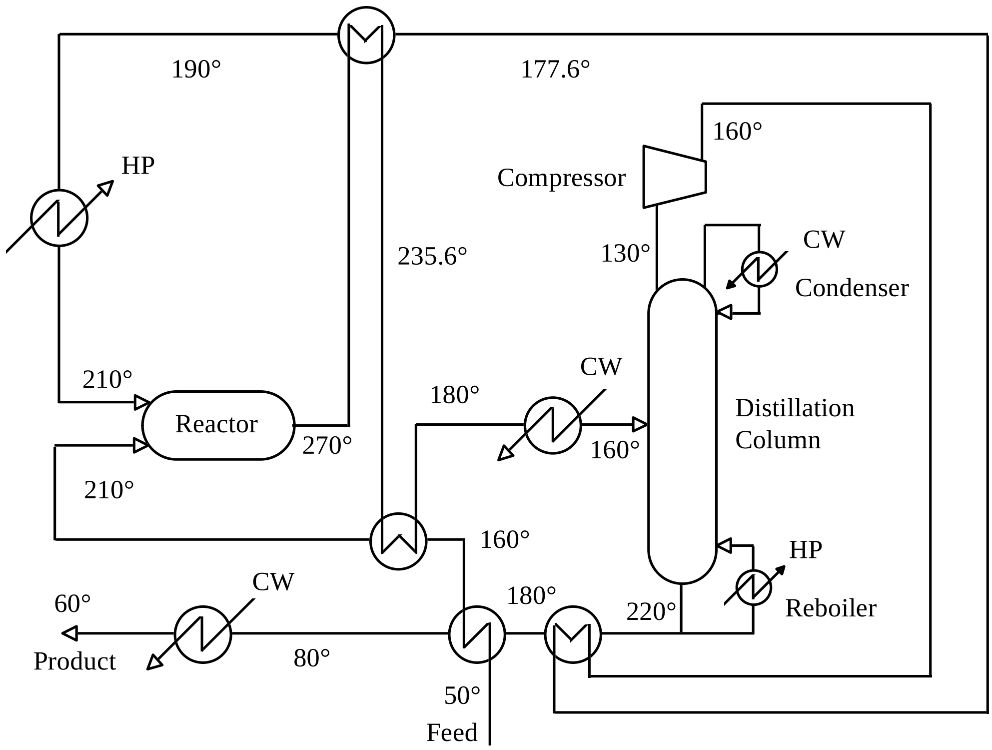


Fig. 5.17 Process Example with a maximum Heat Recovery solution (from Fig. 5.11)

In summary, this section has indicated how network optimization can be carried out as a design evolution, without large modifications to the basic network structure. This method requires a good initial design, as the ones that can be established by the Pinch Design Method. In practice, cost information is required to actually optimize the network, but the basic strategy outlined here is still valid:

a) Identify a good starting value for ΔT_{min} by pre-optimization based on individual targets for Energy, Area and Units (also referred to as SuperTargeting).

b) Design an MER network using the Pinch Design Method (section 5.3).

c) Remove the smallest unit by breaking a Heat Load Loop.

d) Restore driving forces by manipulating a Heat Load Path.

One of the major limitations in chapter 5 is the assumption of a global value of ΔT_{min} for all process streams and heat exchangers. In industrial applications, differences in heat transfer coefficients must be accounted for in Targeting, Design and Optimization. Another limitation is the fact that sequential procedures as the one outlined here have problems handling complicated multiple trade-offs and so-called topology traps as explained by Gundersen et al., 1990 and 1991.

6. BASIC CONCEPTS FOR HEAT RECOVERY IN RETROFIT DESIGN OF CONTINUOUS PROCESSES

While the majority of early days methods developed within Process Integration were related to the design of new plants, most of the projects in industry are trying to make the most out of existing facilities. Typically, these ***projects*** are related to improved operation, removal of plant bottlenecks, improved efficiency with respect to energy and raw material utilization, and the introduction of new technology into an existing process.

Many terms are used for plant modifications, such as retrofit, revamp and debottlenecking. In this section, the term ***retrofit*** is used for projects trying to reduce energy consumption in the most economic way. Typical economic parameters or constraints are maximum allowed values for Payback Time and Investment Cost. The objective of a retrofit project is then to save as much energy as possible while satisfying these economic constraints.

The economy of most energy saving projects (cost of new equipment versus reductions in operating cost) is not good enough to include the losses in production if the plant has to be stopped for a period of time while the modifications are installed. Thus, the timing of retrofit projects into regular plant maintenance periods is extremely important. Further, the best retrofit projects are the ones that combine pure energy saving features with more general plant modifications.

6.1 Some Useful Representations

Grassroots Pinch Analysis can and has been used to a large extent in industry to establish the potential for energy savings in existing plants. When comparing the current energy consumption with grassroots targets, however, the identified ***potentials*** tend to be rather optimistic. In the process industries there is no "second hand" market, thus one of the prime objectives in retrofit projects is to try to improve the utilization of already invested and installed equipment. There will be discrepancies in the existing design that cannot be completely removed, only improved by smaller or larger process modifications. As a result, the optimal heat exchanger network after retrofit is likely to be quite different from the optimal grassroots design.

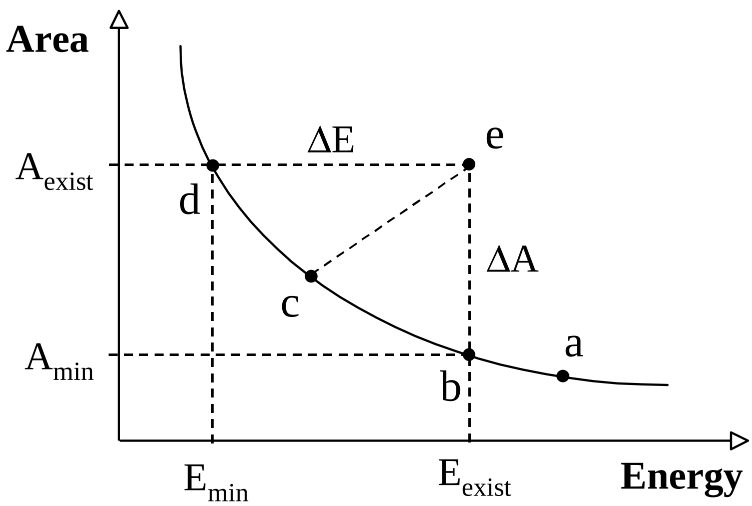


Fig. 6.1 Area-Energy Plot for Heat Exchanger Networks

In an existing plant, the heat recovery system can suffer from two types of errors, as illustrated in figure 6.1. Each point on the curved line indicates the minimum amount of heat transfer area that is required to have a certain energy consumption (or level of heat recovery). Similarly, the curve also indicates minimum energy consumption for a given total heat transfer area. The points (a) to (e) represent different design solutions that will be discussed in the following.

The curved line in figure 6.1, also referred to as the ***Area-Energy Plot***, is constructed by calculating minimum ***target values*** for energy and area as indicated in section 5.2 for different values of the minimum allowable approach temperature, ΔT_{min}. With small values of ΔT_{min}, the minimum area target is large, while the energy target is low, and opposite for large values of ΔT_{min}.

Assume that design (c) is the optimal grassroots heat exchanger network, with an optimal trade-off between operating cost and investment cost for the current energy and area prices. Network (a) has been correctly designed in the sense that it uses minimum area to achieve a certain level of heat recovery. Most likely, this design has been established by the Pinch Design Method. The ***trade-off*** in this design is wrong, however, as it uses more energy than would have been optimal with the "current" prices. In a retrofit project, it will be very hard and costly to improve this network. Moving along the curve from (a) to (c) would mean that a number of heat exchangers would have to be taken out of the network. What would be done in practice is to keep most of the existing exchangers and invest in some new ones. The corresponding retrofit project would move along a curve above the minimum target line, and this curve would be steeper than the target line.

Next, consider design (e), which is located far above the target line. If this had been a suggested new design, both investment cost (area) and operating cost (energy) could have been reduced as indicated by ΔA and ΔE. If this is an ***existing*** network, however, it is not economically tractable to try to reach design (c), since that would involve throwing away a large number of invested heat exchangers. Again, the retrofit project would follow a curve to the left, but in this case it would be flat in the beginning, since the existing network has major errors that can be corrected by moderate investments, such as re-piping and the addition of strategically placed new heat exchangers. After correcting the most obvious errors in the existing design, the cost of recovering additional heat will gradually become more costly. This means that the retrofit curve would become steeper, and payback time therefore increases with the amount of energy saved.

While network (a) is a "good" design (unfortunately with a wrong trade-off), network (e) is a "poor" design, since it uses much more energy than what could have been achieved with the amount of invested heat transfer area. The errors in design (e) are important in retrofit projects and will be discussed in detail in this section.

It should also be mentioned that the minimum area figures used to establish the target curve in figure 6.1 actually require a large number of heat exchangers, splitters and mixers (referred to as the Spaghetti Design in section 5.2.4). Thus, one would never design on the target line, but some small distance above. Figure 6.1 is a quantitative tool to identify the potential for improved heat recovery, and at the same time a qualitative picture of the situation indicating how costly the corresponding retrofit projects will be. What are needed next are some guidelines on how to actually modify the network.

The reason why an existing design, such as network (e) in figure 6.1, is using more than the minimum amount of energy (both heating and cooling), is the fact that ***heat*** is being ***transferred across*** the heat recovery ***Pinch***. Such heat transfer can take three different forms, as indicated in figure 6.2.

The heat recovery Pinch divides the process into a heat deficit part above Pinch and a heat surplus part below Pinch. Of course, it would not make sense to transfer heat from the deficit part to the surplus part. Nevertheless, when heat exchanger networks are designed without the knowledge about the heat recovery Pinch, such heat transfer is often inevitable. This is why ***large potentials*** for energy savings have been identified in existing plants, and the more complex these processes are, the more likely it is that considerable cross Pinch heat transfer takes place. Typical examples are petrochemical plants and oil refineries, however, significant potentials have also been identified in other industries as well.

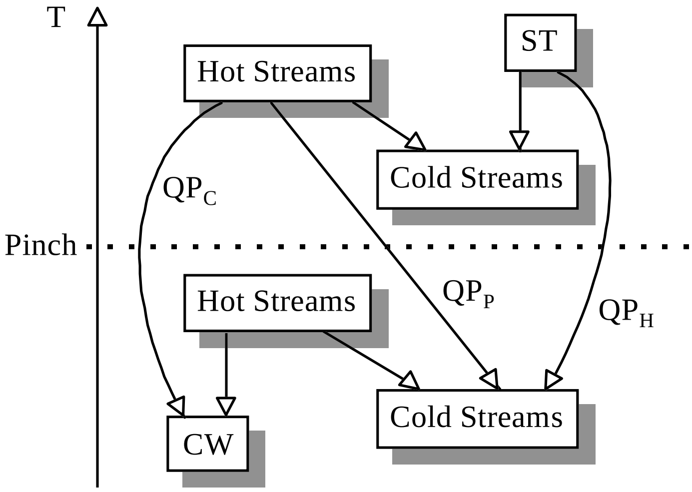


Fig. 6.2 The Penalty Heat Flow Diagram for Heat Exchanger Networks

There may also be ***practical*** reasons for such heat transfer across the Pinch. One of the major limitations of the Composite Curves and the corresponding Pinch Analysis, is that hot and cold streams are regarded to be heating and cooling resources that can be used without limitation. In practice, however, there will be match combinations among hot and cold streams that one would avoid. Examples include safety considerations, geographical distance, start-up considerations, ensuring product purity, etc.

In many of these cases, heat transfer across the Pinch is inevitable, however, there are some degrees of freedom in how this heat transfer takes place. The ***Penalty Heat Flow Diagram*** (Linnhoff and O'Young, 1987) in figure 6.2 shows that heat can be transferred across the heat recovery Pinch in the following ***three*** ways:

- Heat transfer from a hot stream above Pinch to a cold stream below Pinch: QP_P
- Heating a cold stream below Pinch with hot utility, such as steam: QP_H
- Cooling a hot stream above Pinch with cold utility, such as cooling water: QP_C

The total ***Energy Penalty*** for heat transfer across the Pinch is then the sum of these individual heat flow components:

$$QP = QP_P + QP_H + QP_C$$

This penalty is then the difference between the current energy consumption and the minimum energy consumption for a given value of ΔT_{min}:

$$Q_{H,exist} = Q_{H,min} + QP$$
$$Q_{C,exist} = Q_{C,min} + QP$$

The three components of penalty heat flow (QP_P, QP_H and QP_C) can be considered as ***variables*** that can be used to take advantage of the situation when practical constraints result in an energy penalty. When trying to minimize the cost penalty of such constraints, the three variables provide ***two degrees of freedom***. This is obvious from figure 6.2, since there are two heat load loops that can be manipulated. The following advantages can be taken from a constrained situation:

- QP_P means heat transfer at larger driving forces, thus heat transfer area is reduced.
- QP_C can be realized as steam production, if the Pinch temperature is high enough.
- QP_H means that a cheaper hot utility with lower temperature can be used.

Since the energy target depends on the chosen ***value of ΔT_{min}***, the corresponding potential for reduced energy consumption is larger for a smaller value of ΔT_{min}. The corresponding retrofit project will, however, also be more complex and costly. While targeting methods exist for the retrofit case that can identify a proper value for ΔT_{min} (will be described later), it is common practice in industry to use a larger value for ΔT_{min} in a retrofit situation than the corresponding and optimal value of ΔT_{min} in a grassroots case.

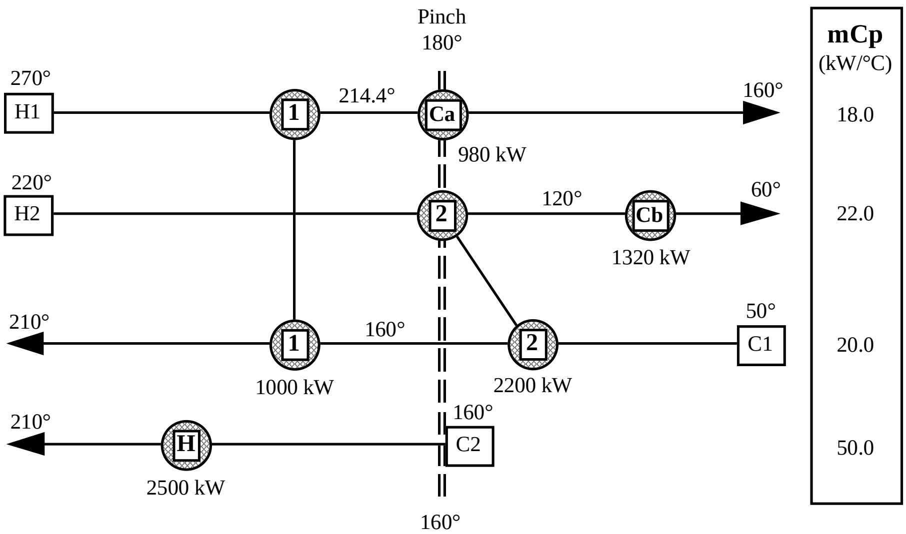


Fig. 6.3 Cross Pinch Heat Transfer in a Stream Grid

Having explained the features of an existing design that is responsible for a larger than minimum energy consumption, the next logical step is to look at the actual heat exchanger network to identify which process/process heat exchangers, external heaters and coolers that are actually transferring heat across pinch. The ***Stream Grid*** is an excellent tool for this purpose. In figure 6.3, an existing heat exchanger network is drawn in a stream grid in such a way that the relative position (in temperature) to the Pinch is indicated for all units.

6.2 A Preliminary Retrofit Discussion

It is now easy to identify which heat exchangers that transfer heat across the Pinch, and what amount of heat that is transferred across Pinch in each of these units. The sum of all these cross Pinch occurrences should add up to the total energy penalty. It should be noted, however, that there sometimes are cases where a heat exchanger operating with small temperature driving forces is transferring heat from below to above Pinch. These heat flows must then be subtracted when calculating the total energy penalty.

The heat exchanger network in figure 6.3 uses 2500 kW of hot utility and 2300 kW of cold utility. The corresponding minimum target values for ΔT_{min} = 20°C are $Q_{H,min}$ = 1000 kW and $Q_{C,min}$ = 800 kW (see the small example used for illustration in section 5.2). The total energy penalty for this existing design is thus:

$$QP = Q_{H,exist} - Q_{H,min} = Q_{C,exist} - Q_{C,min} = 1500 \text{ kW}$$

For the network in figure 6.3, cross Pinch heat transfer can be identified in heat exchanger (2) and cooler (Ca). The actual amount of heat transfer across the Pinch in these units can be calculated as follows:

Exchanger (2): $QP_P = 22 \cdot (220 - 180) = 880$ kW
Cooler (Ca): $QP_C = 18 \cdot (214.4 - 180) = 620$ kW

In this case, there is no external heating below Pinch, and the total energy penalty can be calculated from the occurrences of cross Pinch heat transfer:

$$QP = QP_P + QP_H + QP_C = 880 + 620 + 0 = 1500 \text{ kW}$$

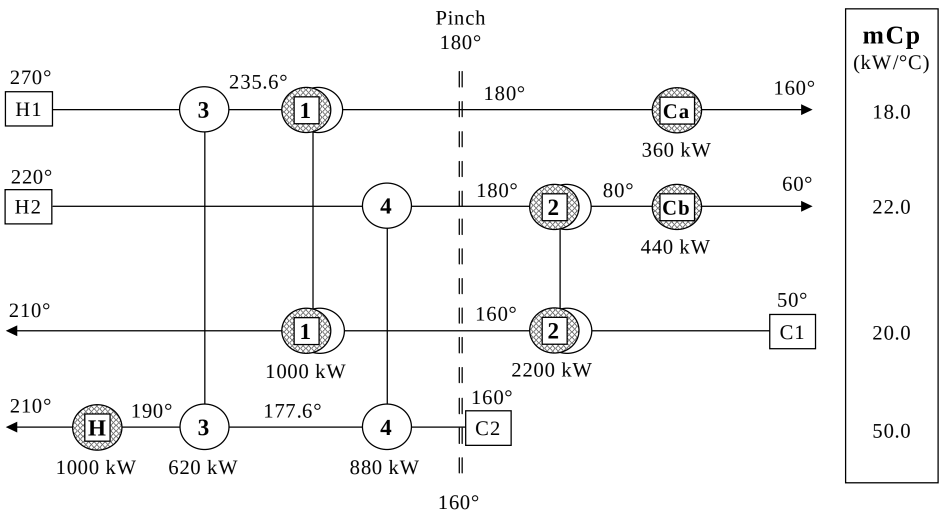


Fig. 6.4 A Retrofit Solution reaching Minimum Energy Consumption

Obvious retrofit projects should involve attempts to reduce heat transfer across the Pinch. Hot stream H2 is a heating resource above Pinch that could be used to heat up cold stream C2 and thus reduce the use of steam in the heater (H). Hot stream H1 is also a heating resource above Pinch, where some heat in the existing design is lost to cooling water.

Trying to realize the total potential for energy savings (1500 kW), would involve two new heat exchangers (3 and 4) and additional area in the existing ones (1 and 2), due to reduced driving forces. The corresponding heat exchanger network shown in figure 6.4 is actually identical to the initial MER design for the grassroots case shown in figure 5.11. Without actually performing cost calculations, it is obvious that the retrofitted network in figure 6.4 will be very expensive. It is almost an entirely new heat exchanger network.

An alternative solution would be to try to recover some of the heat that is lost from hot stream H1 into cooling water, by adding a new unit between hot stream H1 and cold stream C2. In this case, the existing heat exchangers are not modified (no additional area is needed), and the simple question is whether the saving of 620 kW of steam and cooling water will justify the investment in a new heat exchanger (3). The corresponding network is shown in figure 6.5.

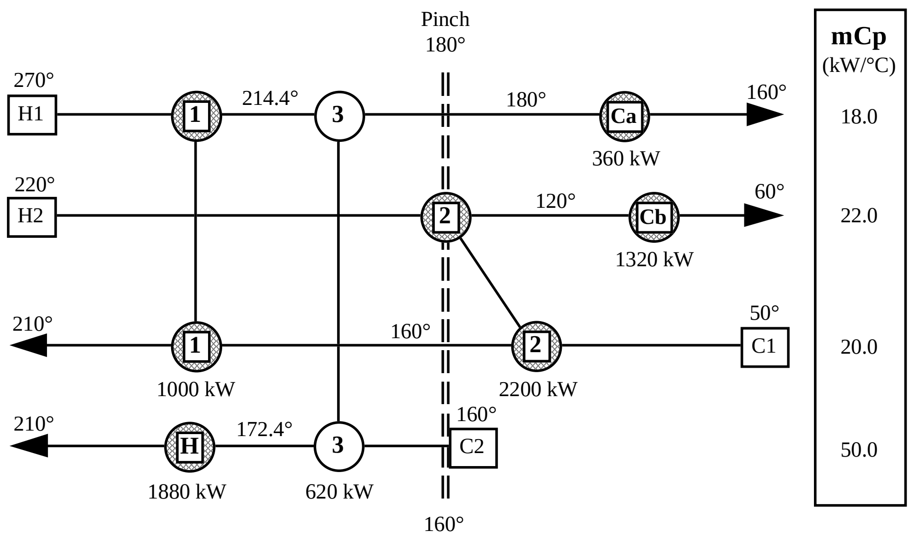


Fig. 6.5 A cheaper Retrofit Solution recovering part of the Potential Energy Savings

Having shown some of the useful representations and indicated a possible "thinking" in retrofit situations based on cross Pinch heat transfer, the remaining part of chapter 6 will be devoted to a presentation of the methods that can be used for heat exchanger network retrofit. Similar to the grassroots case, there are ***four*** distinct ***phases*** also for retrofit design:

1) Data Extraction 2) Targeting 3) Design 4) Optimization

There will, however, be significant differences in all of these phases when compared to the grassroots situation. These differences and the new objectives will be highlighted in the description of each of these phases.

6.3 Data Extraction (Phase 1)

While there are a number of similarities between data extraction in the retrofit situation and the grassroots case as described in section 5.1, there are also significant differences that will be highlighted here. In both cases, data extraction is a time consuming and critical activity for the outcome of a Process Integration project.

Typically, for a new design there will be material and energy balances available either manually derived or based on a simulation model. A rigorous simulation model has the distinct advantage that stream data can be extracted directly and even automatically with today's software. Unfortunately, such models may not always be available for an existing plant. In general, the following are possible sources for data that are needed in a retrofit heat recovery project, and often these sources have to be used in combination:

- Measurements (that are often not complete and not reliable)
- Design data (that are often outdated after plant modifications)
- Simulation models (that may not always reflect true plant behavior)

As a result, data reconciliation is important in retrofit projects. If measurements indicate that heat extracted from a hot stream in an existing heat exchanger does not match the heat absorbed by the cold stream in the same exchanger, it is necessary to analyze the situation. Stream data must be modified in such a way that heat balance is obtained; otherwise the heat recovery project will produce unrealistic results. It is important to notice that data accuracy is most important in the near Pinch region of the plant. Thus, it is common practice to try to establish a first draft of the Composite Curves, and then try to improve the accuracy only for process streams in the near Pinch region.

Another typical retrofit issue is related to which streams to include in the analysis. There may be a number of practical considerations suggesting that certain streams should not be included, since heat integration of these streams could cause operational problems. It is, however, good practice to start by including all streams that need heating or cooling, and then later exclude these streams one by one from the analysis. In this way, the engineer will know the loss in heat recovery potential from excluding certain streams.

In retrofit projects it is not necessary to iterate between data extraction and targeting, since the basic process (reactors, separators) is given and cannot easily be modified for improved heat recovery potential as the case is for grassroots projects. It would also be expensive to modify these process units, and would seldom "pay off" in pure energy based projects.

6.4 Retrofit Targets (Phase 2)

Targeting in the Retrofit situation is far more difficult than for Grassroots design. This is so because a number of different changes can be made to the heat exchanger network in order to reduce energy consumption. Typically, these modifications include:

- Addition of a new heat exchanger
- Additional area to an existing unit (for example a new shell)
- Change internals in heat exchangers
- Modify piping on one side of the exchanger
- Modify piping on both sides of the exchanger
- Moving a heat exchanger to a new location

Most of these retrofit actions will change the operating conditions for many of the heat exchangers, and a rigorous rating exercise is required to evaluate whether an existing unit will be able to operate in the new situation. The cost function for the retrofit project will exhibit a discontinuity whenever a heat exchanger switches from being large enough to become too small for the new operation.

In other words, the targeting of capital investment (new heat transfer area and new units) is much more difficult than in the grassroots case. Energy consumption, on the other hand, is much easier to predict, however, knowing the savings in energy cost is of limited value if it is not correctly linked with its corresponding investment cost. That is the true challenge in retrofit targeting.

6.4.1 Different ΔT Representations

The ultimate goal of the targeting exercise is to establish a good starting value for the level of heat recovery. In grassroots heat exchanger network design (Chapter 5), the parameter ΔT_{min} (minimum approach temperature) was used to represent this level of heat recovery. In most industrial processes, it does not make sense to require that all heat exchangers (and thus all process streams and utilities) obey the same minimum value for driving forces, since streams (and utilities) in general have very different heat transfer coefficients. Quite often, the difference in film heat transfer coefficients can be two orders of magnitude. Thus, some heat exchangers require large ΔT-values in order to avoid excessive heat transfer area, while other units manage well with much smaller ΔT-values.

Since this document is a Primer with focus on the key concepts in Process Integration, we did not discuss the details about heat transfer conditions and driving forces in Chapter 5. When considering the retrofit case, however, there are many reasons why we need to reconsider this question. Without going into too many details, we should at least acknowledge the need for two different approach temperatures:

HRAT = Heat Recovery Approach Temperature
EMAT = Exchanger Minimum Approach Temperature

While HRAT, as the name indicates, is a key parameter for the level of heat recovery (it is simply defined as the smallest vertical distance between the Composite Curves), EMAT is the minimum allowable temperature difference for the individual heat exchangers. In order to reach a certain level of heat recovery, (as specified by HRAT), the following inequality must be satisfied:

$$0 \leq \text{EMAT} \leq \text{HRAT}$$

As mentioned in section 5.2, it is also possible to assign individual contributions to the minimum driving forces for each stream and utility. Typically, these ΔT contributions should reflect heat transfer conditions, but they can also be used to represent the need for expensive material of construction, expensive heat exchanger types, etc. In this case, EMAT becomes stream dependent, and the following must be satisfied for a match between hot stream/utility (i) and cold stream/utility (j):

$$\Delta T_{i,j} \geq \text{EMAT (i,j)} = \Delta T_i + \Delta T_j$$

where ΔT_i and ΔT_j are the individual stream contributions. To illustrate how these different ΔT-values apply in retrofit situations, consider a typical oil refinery with a crude preheat train that warms up the crude from ambient temperature to the inlet of the furnace just before the crude fractionation tower. This is a complex heat exchanger network with many units, a large number of stream splits and considerable heat recovery from various hot streams in the refinery. In retrofit projects for such plants, it is common to design for a level of heat recovery that corresponds to HRAT = 30°C, however, the actual energy consumption in many such crude preheat trains corresponds to a value of HRAT well above 50°C. At the same time, there will be some heat exchangers typically where ΔT in one end of the units is in the range between 10 and 15°C. Thus, EMAT and the individual contributions ΔT_i and ΔT_j are considerably less than HRAT.

6.4.2 A Simple Energy Target

An obvious way to establish a target for energy savings in a retrofit project is to calculate the minimum external heating requirements for different values of HRAT (previously referred to as ΔT_{min}). One of these values of HRAT (typically a large one) corresponds to the current energy consumption, and the targeting exercise then becomes the identification of a new value of HRAT that is less than the "existing" value of HRAT:

$$HRAT_{new} \leq HRAT_{existing}$$

By plotting minimum energy consumption (or minimum energy cost in the case of multiple utilities) as a function of HRAT, it is possible to identify potential starting values of HRAT for the retrofit project.

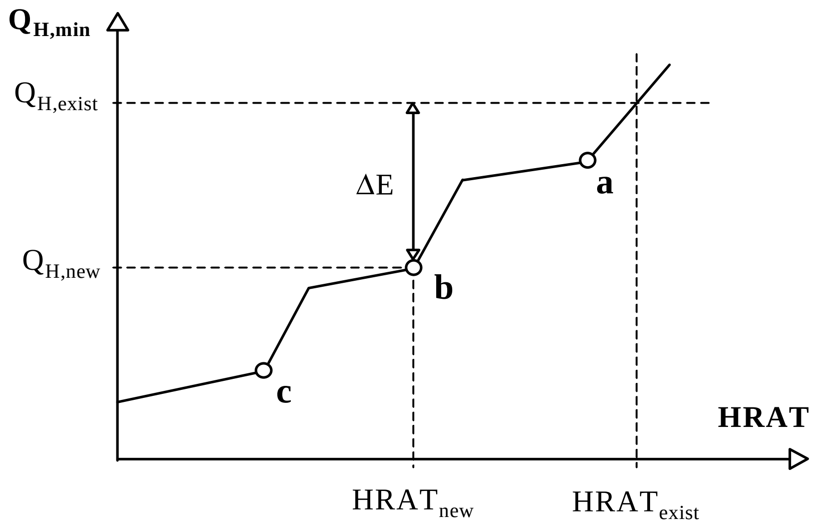


Fig. 6.6 A typical Energy Target Plot for the Retrofit Case

Consider the ***Energy Target Plot*** in figure 6.6, where the change in the slope illustrates the typical situation that minimum energy consumption does <u>not</u> always increase directly proportional with minimum driving forces. As a result, there are certain levels of heat recovery (represented by HRAT) that are more likely to be good retrofit candidates than others. Consider point (a) in figure 6.6. When trying to move towards larger energy savings, the change in $Q_{H,min}$ is relatively small, while the reduction in HRAT is considerable. Normally, this means large investments with moderate savings.

Figure 6.6 also indicates how the targeted savings in energy consumption (ΔE) can be read from the diagram for different values of HRAT. Also, by looking at figure 6.6, one may conclude that point (a) seems to save too little energy, while point (c) involves too large investments. Qualitatively, it may look as if point (b) provides a good trade-off between investments and savings in the retrofit case; thus $HRAT_{new}$ is a good starting value for the retrofit project. Cross Pinch Analysis (section 6.1 and figure 6.3) will then be performed, where the existing heat exchanger network is drawn in a stream grid with a Pinch point according to $\Delta T_{min} = HRAT_{new}$.

6.4.3 Targets for Area and Investment Cost

As stated in the beginning of section 6.4, targeting for heat transfer area and investment cost is far more complicated and uncertain in the retrofit situation than in the grassroots case. The identification of "promising" starting points in figure 6.6 may work in some cases, however, there is a need to quantify not only the energy saving part, but also the investment in new equipment and changes in piping.

Within Pinch Analysis, a Retrofit Targeting procedure has been proposed that is based on the concept of ***Area Efficiency*** (Tjoe and Linnhoff, 1986). This parameter can be easily obtained from the existing design and can be mathematically formulated as:

$$\alpha = [A_{min}] / [A_{exist}]$$

where α = Area Efficiency
A_{min} = Minimum area for the current level of heat recovery ($HRAT_{exist}$)
A_{exist} = Total heat transfer area in the existing network

A conservative assumption is that any new invested heat transfer area will at least have the same utilization level (area efficiency) as the installed area. This assumption (also referred to as the ***"constant α"*** approach) proved to work nicely for oil refineries and crude preheat trains, where area efficiency in existing plants was quite high (above 80%), while it did not work equally well in other industries. In processes with less heat integration, the constant α assumption can be too conservative. Attempts have been made to overcome this problem; one is the so-called ***"incremental α"*** approach (Silangwa, 1986), which means that area efficiency will change (improve) during the retrofit project.

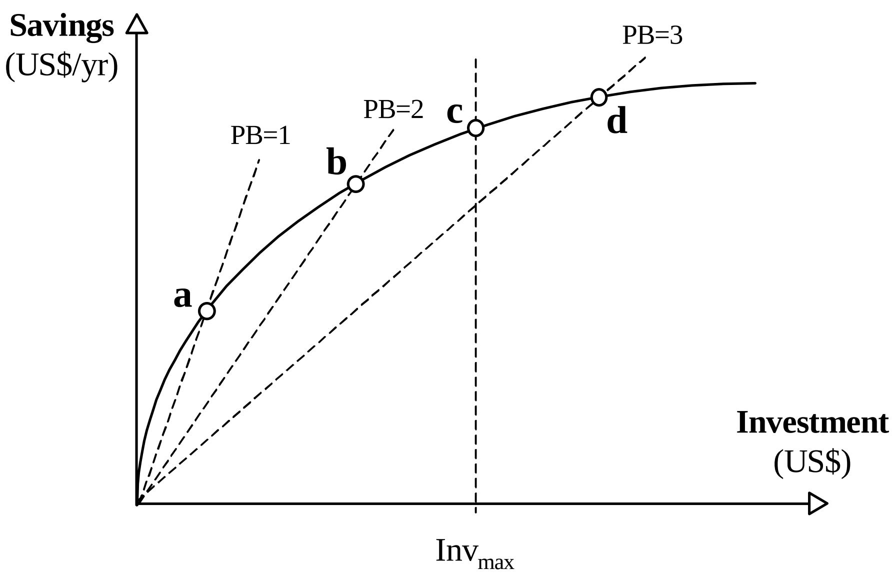


Fig. 6.7 A typical Retrofit Target Plot for Savings and Investment Cost

Irrespective of the actual approach chosen, using some kind of assumptions about area efficiency for new heat exchangers or additional shells, it becomes possible to estimate the need for new area (and thus investment cost) when targeting for different values of HRAT. By combining target values for area and energy for different values of HRAT (starting with the existing HRAT and then decreasing this value gradually), the ***Retrofit Target Curve*** in figure 6.7 can be obtained.

The curved line in figure 6.7 shows corresponding values for annual savings in energy cost and the total investment for new equipment (including installation). Each point on the curve represents a certain level of heat recovery (HRAT), starting in the origin of the diagram, where there are no investments made and no savings earned. Thus, the origin represents the situation before the retrofit project is started, and moving along the curve to the right means heading for smaller HRAT values and higher levels of heat recovery. As discussed in section 6.1, the most obvious errors in the existing network can be corrected first, often with small or moderate investments. Thus, the target line is initially steep, but then becomes more flat as we move towards higher levels of heat recovery.

Payback Time is simply defined as the Investment Cost divided with the Annual Savings in Operating Cost (energy). Considering the nature of the target curve in figure 6.7, it is obvious that Payback Time increases as we move along the curve towards larger energy savings. The dashed lines in the diagram illustrates typically Payback Times (PB) of one, two and three years. It is also quite common that management has set an upper limit on the investment that will be put into a certain energy saving project (indicated as Inv_{max} in figure 6.7). There will also be constraints on the Payback Time, and depending on whether maximum Payback Time for this particular example is set to two or three years, the retrofit targeting exercise will identify points (b) if minimum payback is two years or point (c) limited by maximum investment, if maximum payback is three years.

Points (b) and (c) in figure 6.7 correspond to different values of HRAT, which means that a target for the level of heat recovery has been identified. This target is an improvement compared with the more simplified discussion in section 6.4.2, since investment cost has been included and quantified, even though there are large uncertainties in these numbers. Again, once the new value for HRAT has been identified, the next stage is a cross Pinch analysis, as described in section 6.1.

6.5 Retrofit Design (Phase 3)

The Cross Pinch Analysis mentioned in the previous section is a good starting point for the design exercise. The first methods suggested to remove heat exchangers that transferred heat across the Pinch and to try to reuse these units in new locations. Since, however, heat exchangers in most cases are tailor made for a certain application (flowrates and types of streams) it is not easy and quite expensive to follow this approach.

6.5.1 Temperature “Shifting” of Heat Exchangers

Instead, Tjoe and Linnhoff, 1986, suggested to “shift” heat exchangers away from a cross Pinch situation. This shifting involves changes in operating conditions for the unit in such a way that hot stream temperatures are reduced and/or cold stream temperatures increased. The result is that cross Pinch heat transfer in that particular unit is reduced and possibly eliminated. Heating resources are released above Pinch and or cooling resources are released below Pinch. Consider the existing heat exchanger network in figure 6.3 that was used in the preliminary retrofit discussion of this chapter.

The cooler Ca and heat exchanger (2) are transferring a total of 1500 kW across Pinch, which is why external heating (2500 kW) and cooling (2300 kW) requirements are larger than the established minimum figures (1000 kW of heating and 800 kW of cooling). The shifting procedure means that the inlet temperature of stream H1 to the cooler Ca should be reduced from 214.4°C to at least 180°C (Pinch temperature for hot streams). This will release a heating resource from stream H1 above Pinch equal to 620 kW, and the duty of cooler Ca is reduced from 980 kW to 360 kW.

Similarly, the inlet temperature of stream H2 to heat exchanger (2) should be reduced from 220°C to 180°C. Assuming that the duty of this unit remains unchanged at 2200 kW (should always be questioned during network optimization), the duty of cooler Cb will be reduced by 880 kW to 440 kW. Figure 6.8 shows the incomplete network after these shifting operations. As indicated by the rectangles, there are two heating resources that have been released and not yet utilized above Pinch.

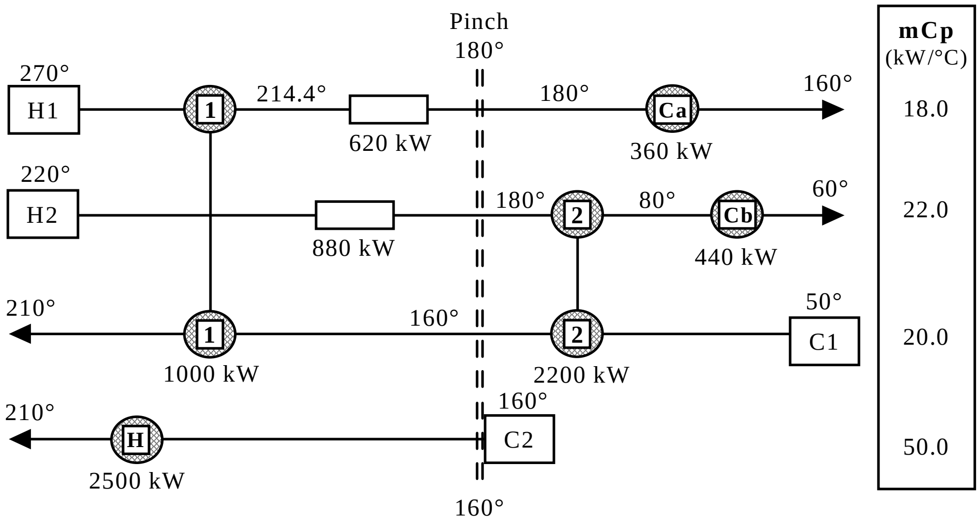


Fig. 6.8 Incomplete Heat Exchanger Network after "Shifting"

6.5.2 Introduction of New Heat Exchangers

The next obvious question is how to utilize these new heating resources above Pinch. Since cold stream C1 already is fully covered through heat recovery from hot stream H1, the obvious option is to try to use heat from hot streams H1 and H2 to partially heat cold stream C2 in order to reduce steam consumption in the heater.

Following the basic philosophy of the Pinch Design Method, cold stream C2 cannot fully utilize the two new heating resources (would involve taking both streams H1 and H2 down to Pinch temperature) unless stream C2 is split into two branches. Since mCp for stream C2 (50 kW/°C) is larger than the sum of mCp (18+22 kW/°C) for streams H1 and H2, this is a feasible option. Alternatively, the heating resource related to hot stream H1 could be shifted to the beginning (hottest part) of the stream. This option has already been shown in figure 6.4, however, as indicated in the same figure, heat exchanger (1) has considerably reduced driving forces and additional area is inevitable.

Figure 6.9 shows the initial retrofitted heat exchanger network when the stream split option is chosen. In this case, the operating conditions (duty and temperatures) for heat exchanger (1) is unchanged, and no additional area is needed. Heat exchanger (2) has, however, reduced driving forces with the same duty, and additional area is needed as indicated. A comparison with the alternative retrofit design in figure 6.4 will be made before going into the optimization stage.

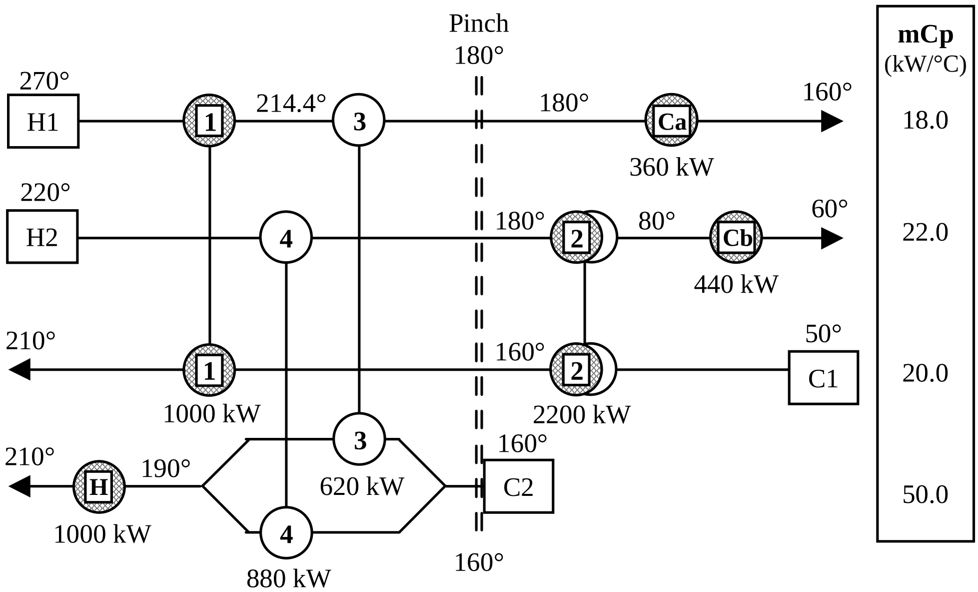


Fig. 6.9 Complete Retrofitted Heat Exchanger Network with Stream Split

In order to compare the two alternative initial retrofit heat exchanger networks in figure 6.4 and figure 6.9, details about heat transfer conditions and cost equations are needed. In this particular case, we do have information about heat transfer coefficients for streams and utilities (table 5.1), however, for the purpose of this Primer we do not want to go into detailed cost calculations. Instead, comparison between the two alternatives will be made on the basis of a simple ***UA analysis***.

It is easy from the heat transfer equation to calculate UA-values for the heat exchangers before and after retrofit modifications. If we assume that the units are pure counter current, UA-values can be obtained from:

$$UA = Q / \Delta T_{LM}$$

Table 6.1 shows UA-values for existing and new heat exchangers before and after retrofit modifications for the two alternative designs A (figure 6.4) and B (figure 6.9). Utility exchangers are not included, since the duty of these units are reduced in such a way that no additional area is needed (actually, these units will not be fully utilized after the retrofit modifications). Isothermal mixing is assumed for stream C2 after the split.

As indicated in table 6.1, the UA analysis does not give any strong preference for design A (figure 6.4) or design B (figure 6.9). The difference in total UA needed in the retrofitted networks is not significant, and a stronger argument for choosing design B is probably that the number of modifications is less, since there is no change needed for heat exchanger (1) in this case, however, there is a stream split introduced.

Table 6.1 UA-values (kW/°C) for two alternative Retrofit Designs

Heat Exchanger	Existing Design	Retrofit Networks Design A	Design B
1	17.49	44.12	17.49
2	33.91	89.20	89.20
3	0	9.07	27.99
4	0	29.52	35.68
Total	51.40	171.91	170.36

While both design A and design B fully recover the energy saving potential of 1500 kW, in most cases one can only justify economically to realize some fraction of this potential. The figures for UA listed in table 6.1 indicate that heat transfer area must be more than tripled in order to reduce energy consumption to its minimum for HRAT = 20°C. Thus, more recent retrofit methods use a "greedy" approach trying to identify the most economic retrofit projects with the fewest number of topological changes.

6.5.3 Matrix Methods for Retrofit Design

Some interesting matrix based methods have also been proposed for heat exchanger network retrofit situations. Shokoya, 1992, focused on heat transfer area in a method where targeting and design are closely linked. The so-called ***Area Matrix*** method is an adoption of the vertical heat transfer model (see the Area Targeting method in section 5.2.4). For various levels of heat recovery, the best vertical match area contribution is found using Linear Programming (LP). The result is a significantly improved retrofit area targeting method when compared with the constant α or incremental α methods mentioned in section 6.4.3. While the Area Matrix method primarily is a targeting procedure, the results from the LP optimization can also be used for retrofit design.

Another matrix based method for retrofit design is the ***Cost Matrix*** method developed by Carlsson et al., 1993. The method is based on the experience from a number of retrofit projects that other costs such as pumping and piping may have a larger influence on the optimal design than the number of units and heat transfer area. A Cost Matrix for possible matches is established, where the cost for each match is estimated taking into account parameters such as physical distance between process streams, material requirements, type of heat exchangers, auxiliary equipment (such as valves), heat transfer coefficients, space requirements, pumping cost, maintenance cost and fouling. The method uses the greedy approach due to its sequential nature, and there is no targeting involved.

6.5.4 More Recent Retrofit Methods

A number of more recent methods for retrofit heat exchanger networks using optimization (Mathematical Programming) to a large extent have been developed (e.g. Asante and Zhu, 1996, Briones and Kokossis, 1996). Due to the rather complex nature of these methods, however, they are only briefly mentioned in chapter 9 and omitted here. The complexity of these methods also means that software is an absolute requirement. Typically, these methods acknowledge the fact that only a few carefully selected modifications will be economically worthwhile, and the approach is to identify these retrofit actions.

6.6 Network Optimization (Phase 4)

As mentioned in the previous section, optimization is used in some of the more recent retrofit methods for network design, and the distinct classification into targeting, design and optimization has been reduced and almost eliminated. This section, however, will discuss how initial retrofit designs developed using the methods described in sections 6.5.1 and 6.5.2 can be improved economically and simplified with respect to network structure, using the same optimization philosophy as in the grassroots case.

Degrees of freedom in the form of ***heat load loops*** and ***paths*** as well as ***stream splits*** can be used to improve the initial retrofit design. One important new aspect in the retrofit case is to maximize the utilization of existing heat exchangers. After the shifting of cross Pinch heat exchangers and the introduction of new units, some of the existing heat exchangers may have a reduced duty and therefore no longer require all the area installed. In such cases it may be worthwhile to shift duty in heat load loops and/or paths until the existing units are better utilized.

Similar to grassroots situations, retrofit network optimization is a combination of discrete and continuous adjustments. The discrete part takes care of the removal of small new heat exchangers or small area additions to existing units, while the continuous part takes care of the trade-off between investment cost and obtained energy savings. The continuous part also includes, as mentioned above, the maximum utilization (if possible) of existing units.

Figure 6.5 shows an alternative retrofit heat exchanger network for the example problem, where only one topological modification is suggested. The introduction of the new heat exchanger (3) between H1 and C2 recovers heat that is lost to cooling water above Pinch in the existing design. In this retrofit alternative, the existing heat exchangers (1) and (2) are not changed, and the optimization simply becomes a one-dimensional search to identify the largest duty for heat exchanger (3) that satisfies constraints such as maximum Payback Time and maximum Investment Cost. Figure 6.10 illustrates that the optimization problem is reduced to the issue of finding the best duty for the new heat exchanger (variable y may take positive and negative values), and how this affects the heat load path from the heater through the new unit to the cooler Ca. Temperatures T_1 and T_2 depends on the value of y.

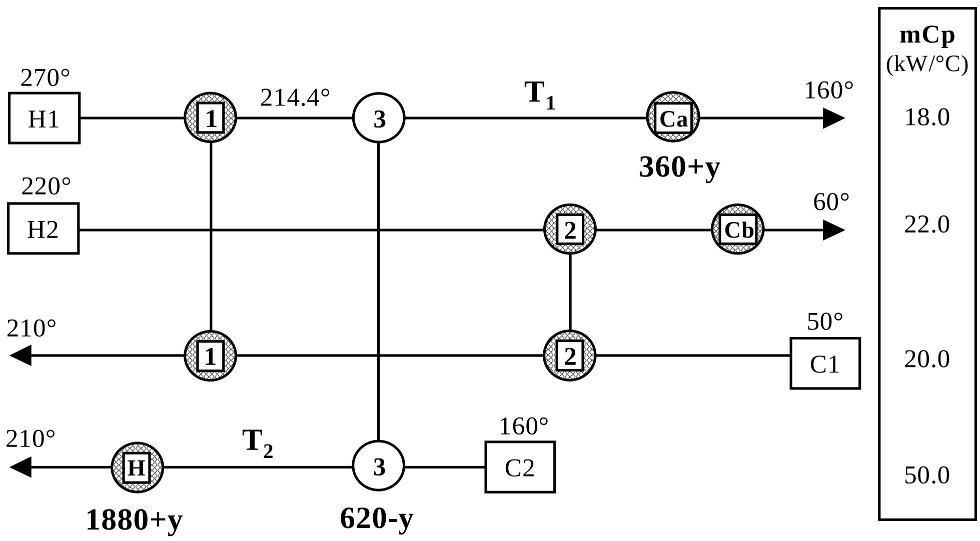


Fig. 6.10 A limited Retrofit Network Optimization Problem

7. BASIC CONCEPTS FOR HEAT RECOVERY IN BATCH PROCESSES

Batch processes have several advantages compared with continuous processes. While continuous processes are "tailor-made" to serve basically <u>one</u> major purpose for 15-20 years, batch processes typically consist of more ***general-purpose*** equipment, which makes such plants far more flexible. This becomes increasingly important in a world of high value products with shorter lifetime.

By its very nature, however, batch processes cannot reach the same degree of utilization of the equipment as continuous processes. Typically, time is lost in a number of operations such as feeding, unloading, cleaning, etc. ***Time analysis*** of the operation of such plants (also called scheduling) is therefore extremely important. The time aspect is also a key factor when studying heat recovery of such processes.

7.1 Introduction

In order to maximize the use of invested equipment, ***Gantt Diagrams*** are heavily used in batch industries. In these diagrams, both the duration of each operation and the relative sequence of operations depending on each other, can be plotted as function of time. Since efficient schedules are important, there is a very strong link between design, scheduling (short term) and planning (long term) of such processes.

This section will primarily concentrate on the use of Pinch Analysis based methods for the identification of ***Heat Recovery*** opportunities. While energy cost is small compared with the cost of equipment and raw material for a majority of batch processes, it has been shown in some cases that the energy system may be responsible for bottlenecks in the plants. One should also mention the more general spin-off from Systems oriented studies such as Pinch Analysis, that more insight about the various aspects of the process is obtained. Also, there are quite a few batch industries where energy cost is considerable, and where improved energy efficiency may give the company a competitive edge.

7.2 Heat Recovery and Design Phases

There are several differences between batch and continuous processes when studying heat recovery. The most important difference is the time aspect in batch processes where a certain amount of heat is available between specified temperatures, and also between ***specified times***. Thermal energy in such plants has two qualities, temperature and time.

While most plants are operated with repeated cycles, some of the material in this section relates to a single batch only, however, the effect of ***cyclic operation*** will be addressed. When plants are operated in a cyclic manner, the question about "before" or "after" becomes more vague than the situation when considering a single batch.

It should also be mentioned that the majority of the literature covering design, scheduling, planning and heat integration of batch processes assumes that these plants are operated with a rigid schedule. Some plants do not operate according to a predefined schedule, and quite often the schedule is somewhat uncertain due to upsets in plant operation. Delays may then destroy the nice schedule that was planned.

Heat exchange between process streams in batch processes can take place in two different ways:

- ***Direct*** using a Heat Exchanger if the streams exist in the same time period
- ***Indirect*** using a Heat Storage System if the streams do not co-exist in time

When heat storage is expensive, an alternative solution would be to consider rescheduling in such a way that direct heat exchange becomes possible.

When using Pinch Analysis for heat recovery in batch processes, the ***Design Stages*** are similar to the case with continuous processes, however, the actual content of each phase can be quite different. The following four phases should be carried out in sequence:

1) Data Extraction (collecting data for the process and the utility system)
2) Targeting (identify maximum heat recovery using direct/indirect heat transfer)
3) Design (establish an initial network of heat exchangers and heat storages)
4) Optimization (improve the initial network using degrees of freedom)

First, however, it is important to make sure that a proper problem definition has been established. This also includes relevant cost data and economic criteria. It is also important to identify the actual schedule, and to what extent the schedule is fixed.

7.3 Data Extraction (Phase 1)

In section 5.1 it was mentioned that the most time consuming and often most critical phase of heat recovery projects, is the collection and evaluation of process data to be used in the analysis and design. In batch processes, data extraction is even more complicated. In particular, there are two new problems to deal with in batch processes. The first is the obvious ***time*** aspect, and the second is the fact that several ***types*** of streams exist.

A central part of data extraction is the identification of heating and cooling requirements in the process. The necessary data for each process stream are the following:

m	=	mass flowrate (kg/s, tons/h, etc.)
Cp	=	specific heat capacity (kJ/kg°C)
T_s , T_t	=	supply and target temperature (°C)
t_s , t_f	=	start and finish time (s, h, etc.)
ΔHvap	=	heat of vaporization for streams with a phase change (kJ/kg)

In order to analyze heat transfer area and investment cost for heat exchangers or heat storage systems, film heat transfer coefficients (h) are needed. In continuous processes, the heat flow (often referred to as heat duty) of a hot stream can be calculated by (index C for continuous):

$$Q_C = m \bullet Cp \bullet (T_s - T_t) \quad \text{(kW)}$$

In batch processes, the total amount of heat is a more representative property, and can be calculated for a hot stream by (index B for batch):

$$Q_B = m \bullet Cp \bullet (T_s - T_t) \bullet (t_f - t_s) \quad \text{(kWh)}$$

Based on the start and finish times of each stream exchanging heat, a Gantt Diagram that only relates to heat recovery can be constructed.

During data extraction the following ***four types*** of streams can be encountered:

1) Streams with fixed or constant T_s, T_t, t_s, t_f, and Q. This corresponds to the situation in continuous processes apart from the time aspect.

2) Streams with a gradual change of Q with time, even though temperature is constant. This can be the case when a volatile component is vaporizing from a batch reactor.

3) Streams with a gradual change in temperature with time, even though Q is constant. This can be the case when a liquid tank is heated with electric coils.

4) Streams with a gradual change in both Q and temperature with time. This can be the case when a batch reactor is heated or cooled with steam or cooling water. During heat exchange, the temperature driving forces are reduced, resulting in a reduced Q.

7.4 Energy Targeting (Phase 2)

A number of different methods exist for identifying minimum energy consumption in batch processes. Many of these methods are inspired by Pinch Analysis originally developed for continuous processes. Various assumptions are used, and as a result, different target values are obtained providing the engineer with valuable information. Thus, it is difficult to claim that any of these methods are better than the others, they just serve different purposes. The most important among these assumptions are:

- Whether Heat Storage is used or not.
- Type of Heat Storage used and the degree of Temperature Loss.
- Selected ΔT_{min} value for the Heat Storage
- Single Batch or Cyclic Operation

Methods for Energy Targeting in batch processes can be classified in two major groups:

I. Methods where Temperature is considered the Primary Constraint, while Time is considered to be a secondary concern, in some cases not included at all:

TAM : Time Average Model
TSM : Time Slice Model
CA : Cascade Analysis
BUC : Batch Utility Curves

II. Methods where Time is considered the Primary Constraint, while Temperature is considered to be a secondary concern:

TPA : Time Pinch Analysis

The Time Average Model (Linnhoff et al., 1988), for example, neglects time completely and assumes that heating or cooling of a stream takes place in the entire batch period. An "average heat duty" obtained from the real heat duty multiplied with the actual time span for the stream and divided with the total batch period is used in TAM. This may seem to give a target of limited value, but is actually the minimum energy consumption for cyclic batches and unlimited ideal heat storage. The obtained number can then be used as a rigorous lower bound on energy consumption.

Time Pinch Analysis (Wang and Smith, 1995), on the other hand, neglects temperature feasibility, which again may seem to give a target of limited value. This method does, however, identify the ***Time Pinch*** of the processes, and also suggests what is possible through process modifications, including the use of heat pumps to overcome negative temperature driving forces.

While TAM and TPA are the extremes, this section attempts to indicate that each method will result in different target values for energy consumption, and this will be illustrated by a small and simple example. Table 7.1 shows the process streams for the example. No utilities are included, however, they are assumed to be available in sufficient amounts and temperature levels. In the targeting exercise, it is assumed that ΔT_{min} = 10°C for both direct (heat exchangers) and indirect heat transfer (heat storage systems). In practice, this is not very realistic, but serves the purpose of illustrating the various targeting methods.

7.4.1 Time Average Model (TAM)

This method has already been discussed above, and the energy targets obtained serve the only purpose to indicate the absolute maximum heat recovery that can be obtained for a cyclic process, when an unlimited number of ideal heat storages (no temperature loss) are allowed. As such, it provides a rigorous lower bound on energy consumption, and indicates the potential for rescheduling or heat storage systems.

The main problem with the TAM method, apart from its extremely optimistic assumptions, is that one cannot distinguish between direct and indirect heat transfer in the target values, which is important information when estimating total cost.

Table 7.1 Stream Data for a small Example

Stream Name	T_s (°C)	T_t (°C)	mCp (kW/°C)	t_s (h)	t_f (h)	Q_C (kW)	Q_B (kWh)
H1	170	60	4.0	0.25	1.00	440	330
H2	150	30	3.0	0.30	0.80	360	180
C1	20	135	10.0	0.50	0.70	1150	230
C2	80	140	8.0	0.00	0.50	480	240

7.4.2 Time Slice Model (TSM)

The next logical step is to split the batch period into smaller ***Time Intervals*** (time slices), and consider heat recovery separately in each of these intervals. By increasing the number of intervals, the target values for minimum external heating and cooling gradually changes from the TAM targets to new values for heat recovery that is feasible without heat storage.

An obvious way to identify these time intervals is to use the start and finish times for all process streams involved in heating or cooling. The Gantt Diagram in figure 7.1 illustrates this for the small example in table 7.1.

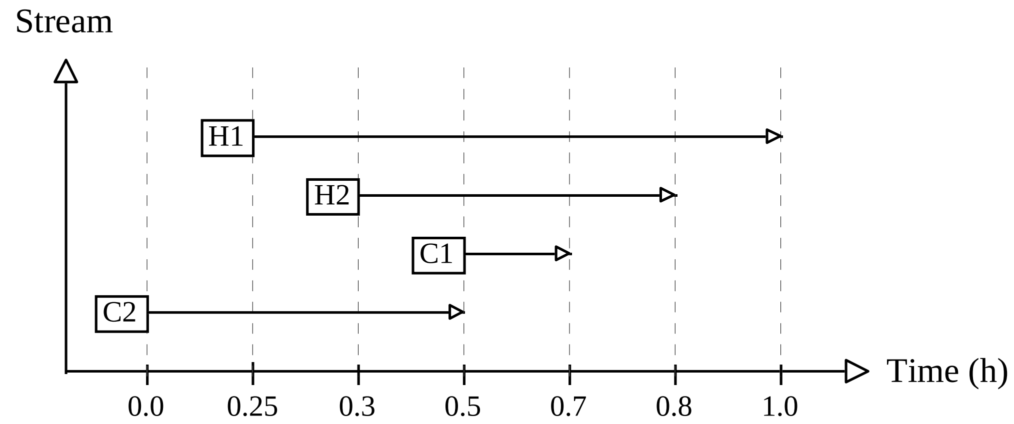


Fig. 7.1 Gantt Diagram for Hot and Cold Streams in the small Example

The difference between the TAM and TSM targets indicates the potential for reducing energy consumption through process modifications such as rescheduling and/or the use of heat storage, <u>after</u> direct heat exchange has been maximized.

7.4.3 Cascade Analysis (CA)

A more systematic way to consider the time aspect is to use a Two-Dimensional Heat Cascade (Kemp and Deakin, 1989), where heat can be transferred to lower temperature (direct heat exchange) or a later time interval (indirect heat exchange using storage). Again, the obvious way to establish this cascade is to use the supply and target temperatures as well as the start and finish times for all process streams. In some cases, this may give a large number of intervals, however, when the method is implemented into computer software, this should not be a major obstacle.

Temp. Interv.	Time Intervals						
	0-0.25	0.25-0.3	0.3-0.5	0.5-0.7	0.7-0.8	0.8-1.0	Sum
165 – 145	0	+ 4	+ 16	+ 16	+ 8	+ 16	+ 60
145 – 140	- 10	- 1	- 1	+ 7	+ 3.5	+ 4	+ 2.5
140 – 85	- 110	- 11	- 11	- 33	+ 38.5	+ 44	- 82.5
85 – 55	0	+ 6	+ 42	- 18	+ 21	+ 24	+ 75
55 – 25	0	0	+ 18	- 42	+ 9	0	- 15

Fig. 7.2 Two-Dimensional Heat Cascade for the small Example

Figure 7.2 shows the two-dimensional heat cascade for the small example problem, where the 4 process streams result in 5 temperature intervals and 6 time intervals. The last column takes the heat balance for each temperature interval over all time intervals, thus it shows the overall heat cascade used in the TAM method.

Notice that modified temperatures have been used (average between hot and cold interval temperatures) in the same way as for continuous processes. The dotted lines indicate the individual Pinch points for each time interval. The numbers in the boxes (intervals) are heat balance figures given in kWh, where a positive sign means heat surplus.

Different ***strategies*** can be used in the Cascade Analysis (CA). When transferring heat within time intervals only, the TSM targets will be obtained, and no heat storage is needed. If heat primarily is transferred within the same temperature interval to later time intervals, heat degradation is minimized, and it may be possible to use a cheaper hot utility at lower temperature. This strategy will, however, result in a considerable amount of heat being stored for later use. Alternatively, one could transfer heat from below Pinch in one time interval to the region above Pinch in a later time interval.

For small problems, as the one illustrated here, it is possible to carry out the analysis and obtain the targets by studying the two-dimensional heat cascade in figure 7.2. For larger problems, however, this becomes quite a task. For this reason, graphical diagrams have been developed, such as the ***Heat Recovery Plot***. Figure 7.3 shows such a diagram for the two neighboring time intervals 3 (0.3 - 0.5 h) and 4 (0.5 - 0.7 h). Time interval 3 has a net heat surplus of 64 kWh, while time interval 4 has a net heat deficit of 70 kWh. If heat from time interval 3 is available at a high enough temperature, it should be possible to transfer 64 kWh indirectly from time interval 3 to time interval 4 by using heat storage. Figure 7.3 shows that this indeed is possible, at least in theory (or more correctly: from a thermodynamic point of view).

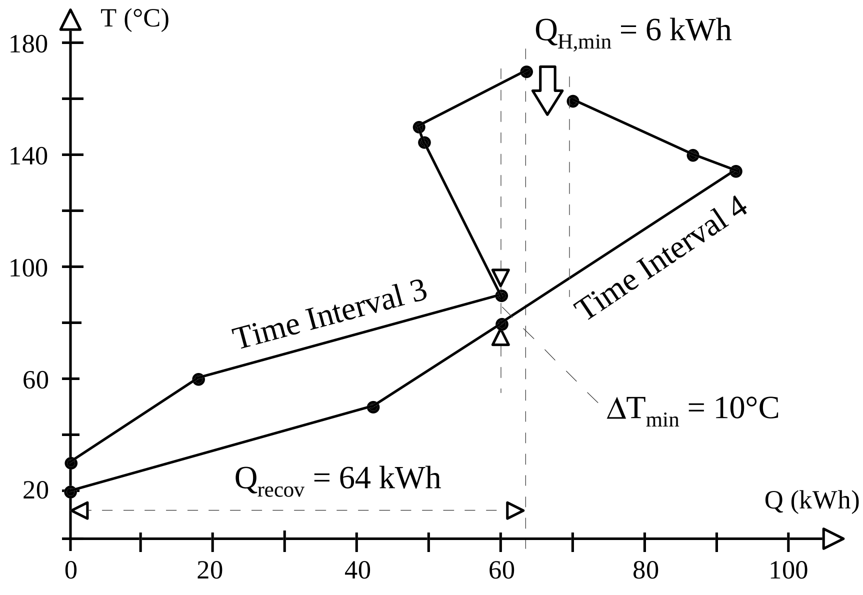


Fig. 7.3 Heat Recovery Plot for Time Intervals 3 and 4 for the Example

The Heat Recovery Plot is similar to the ***Grand Composite Curve***, in the sense that heat cascade figures are plotted versus temperature. One time interval is plotted as a heat source, and the other interval as a heat sink. In figure 7.3, the scope identified for heat recovery by indirect exchange through a heat storage system, is rather optimistic. If only latent (constant temperature) heat storages are considered, and indirect heat transfer from a hot stream through storage to a cold stream is allowed at ΔT_{min} = 10°C (the same as for direct heat exchange), an ***infinite*** number of storages would result as we approach the heat recovery pinch for heat transfer between time interval 3 and 4. The situation would be somewhat easier if sensible heat storage systems are assumed.

Nevertheless, the Heat Recovery Plot can be used to find a rigorous upper bound for indirect heat transfer, after direct heat transfer is maximized in each time interval. It is also possible to plot more than one time interval as heat source and/or heat sink. The reason for plotting two neighboring time intervals is to reduce time for storage, which affects heat loss in the storage system.

The Heat Recovery Plot can also be used to find the maximum amount of indirect heat transfer through storage when only a limited and specified number of storages is allowed. This is illustrated in figure 7.4, where only one latent heat storage system is allowed, and where the total value for ΔT_{min} is equally distributed among heat transfer between the hot stream(s) in time interval 3 and the heat storage, and heat transfer between the heat storage and the cold stream(s) in time interval 4. Notice that one of the curves has been inverted in order to identify the temperature where the heat transferred to the storage from hot streams equals the heat transferred from the storage to the cold streams.

As shown in figure 7.4, the amount of heat recovered is reduced from 64 kWh for ideal and unlimited storage to 43.2 kWh when only one heat storage is allowed. With larger values for ΔT_{min}, the amount of heat recovered would be reduced even further. If two or more storages are allowed, the graphical procedure becomes more complicated.

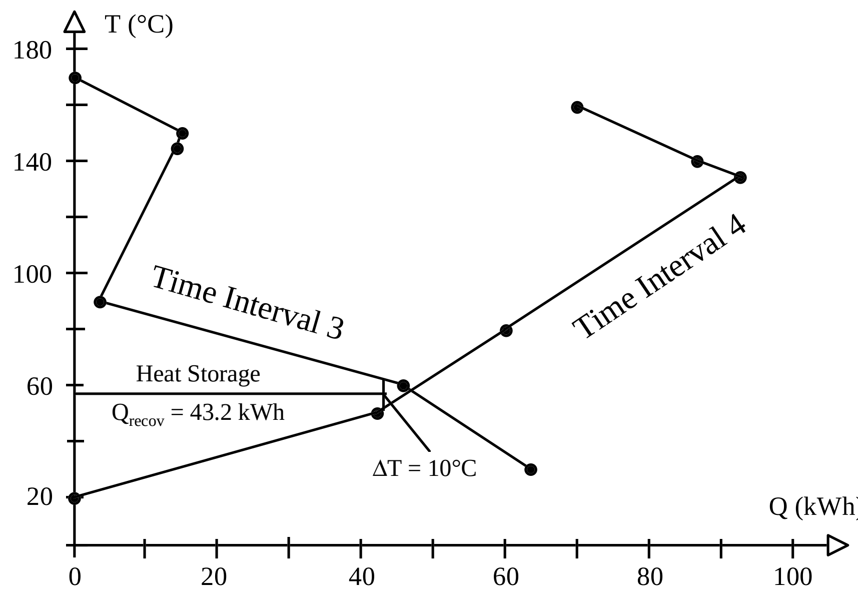


Fig. 7.4 Heat Recovery Plot for the Example with only one Heat Storage

7.4.4 Batch Utility Curves (BUC)

Another graphical diagram has been developed for the identification of appropriate temperature levels for heat storages and external utilities. The Batch Utility Curves (Gremouti, 1991) are based on the Grand Composite Curve for each time interval. First, these Grand Composite Curves are separated into two different curves at the individual Pinch temperature for each time interval. The part below Pinch is a Heat Source, while the part above Pinch is a Heat Sink. Then all existing "pockets" are removed from the curves, and finally the remaining segments are added into Total Heat Source and Sink Profiles.

These curves are similar to the Total Site Source and Sink Profiles for studies of total sites where continuous processes exchange heat indirectly through the steam system. In batch processes, these so-called Batch Utility Curves (see figure 7.5) are used to consider indirect heat exchange through heat storage systems.

Since the Hot and Cold Utility Curves are monotonic with temperature, the amount of heat that can be sent to storage from the Heat Source Profile (Cold Utility Curve) is increasing as we move to lower temperatures. The need for such heat in the Heat Sink Profile (Hot Utility Curve) is, however, decreasing towards lower temperatures.

By considering the two curves it is quite easy to identify the temperature level for a single heat storage system. Once the number of heat storages is allowed to be more than one, however, the picture gets more complicated. In addition, the load and level for the various utilities must also be considered simultaneously with the selection of temperature levels and amounts of heat for the storage systems.

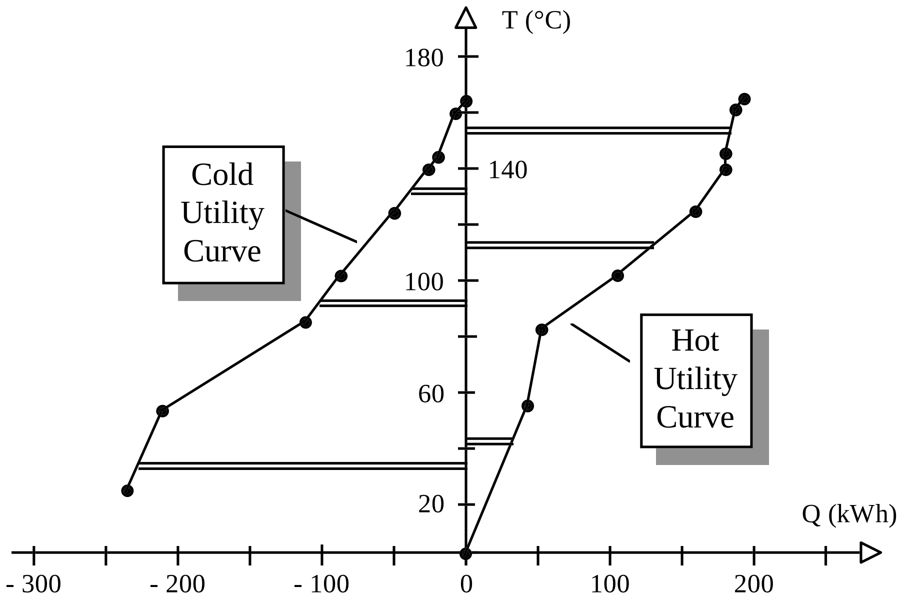


Fig. 7.5 Batch Utility Curves for the Example

As always with such graphical diagrams, a major limitation is the fact that the identities of the individual streams are lost. For the Heat Recovery Plot in figures 7.3 and 7.4, we need to go behind the curves to identify the actual hot and cold streams that should be linked with the heat storage system. In the Batch Utility Curves in figure 7.5, the situation is even worse, since we also have lost the identities of the various time intervals.

This is a major criticism and disadvantage of almost all graphical diagrams that have been introduced within Pinch Analysis. Without information about the number of streams involved as well as their actual identity, it is almost impossible to make any reasonable estimation of the investment cost involved.

7.4.5 Time Pinch Analysis (TPA)

This method will only be briefly described, since neglecting temperature feasibility appears to be a rather drastic assumption. By plotting heat duties as function of time, two new diagrams are produced that are similar to the Composite Curves and Grand Composite Curve for continuous processes. It should also be mentioned that the approach taken in this method minimizes degradation of heat to lower temperatures.

The TPA Composite Curves can be used to identify bottlenecks for heat recovery in batch processes, referred to as the ***Time Pinch***. The TPA Grand Composite Curve can be used to distinguish indirect heat exchange (storage) from direct heat exchange.

7.4.6 Summary of Methods for Batch Targets

As mentioned earlier, rescheduling is an alternative solution to the use of heat storage, which tends to be quite expensive. Using heat storage, however, adds flexibility to the process, and heat recovery becomes less dependent on a fixed schedule. Tools that can be used to consider rescheduling include the Gantt Diagram, the Two-Dimensional Heat Cascade and Time Pinch Analysis.

To illustrate the different targets that can be calculated with the various methods discussed in this section, some scenarios are listed in table 7.2. For reference and comparison, total heating needed by the cold streams is 470 kWh, while total cooling needed by the hot streams is 510 kWh.

Table 7.2 Various Energy Targets for the small Example (all numbers in kWh)

Method	Assumption	$Q_{H,min}$	$Q_{C,min}$	$Q_{storage}$
TAM	Ideal Storage, Cyclic Operation	20	60	n.a.
TSM	Maximum Direct, No Storage	198	238	0
CA-1	Maximum Direct, No Storage	198	238	0
CA-2	Single Batch, Ideal Storage	134	174	64
CA-3	Cyclic Operation, Ideal Storage	20	60	178
CA-4	Single Batch, Only One Storage	174	214	24

The actual numbers for TAM can be obtained by considering the last column of figure 7.2, where the net heat deficit above Pinch (dotted line) is 20 kWh and the net heat surplus below Pinch is 60 kWh. The minimum external heating requirements ($Q_{H,min}$) for TSM and CA-1 are obtained by adding the minimum heating requirements for each time interval in figure 7.2:

$$120 + 8 + 0 + 70 + 0 + 0 \quad = \quad 198 \text{ kWh}$$

7.5 Network Design (Phase 3)

Since most of the targeting methods mentioned in the previous section (except Time Pinch Analysis) consider temperature as the primary variable (or constraint), the ***Pinch Design Method*** developed for continuous processes can, with some adjustments, be applied for the synthesis of networks of heat exchangers and heat storages for batch processes.

Use of the Pinch Design Method is, however, considerable more complicated for batch processes. The most important ***new features*** that must be dealt with are:

- The ***Time*** aspect normally means that separate initial networks must be developed for each time interval. The overall network can then be found by combining the features of the individual networks.

- The various time intervals normally have different ***Pinch*** points. This affects the individual networks, since the matches depend on the Pinch location.

- At all times, the use of ***Heat Storage*** must be considered as an alternative to the use of external heating and cooling utilities.

- While both targeting and design methods for heat recovery systems in batch processes use amount of heat (in kWh) rather than heat flow or heat duty (in kW), it is necessary to consider the actual time of operation for the individual heat exchangers, to be able to calculate the required heat transfer ***Area*** in each unit.

Returning to the small example, the Gantt Diagram in figure 7.1 shows that three of the six time intervals only involve one type of streams (hot or cold). In such cases, the design simply means adding utility exchangers. In time interval (1), only cold stream C2 is present, and a heater is needed. In time interval (5), hot streams H1 and H2 are present, and two coolers are needed. In time interval (6), only hot stream H1 is present, and a cooler is needed. For the other time intervals (2, 3 and 4) both hot and cold streams are present, and the Pinch Design Method can be used to design the corresponding heat exchanger networks.

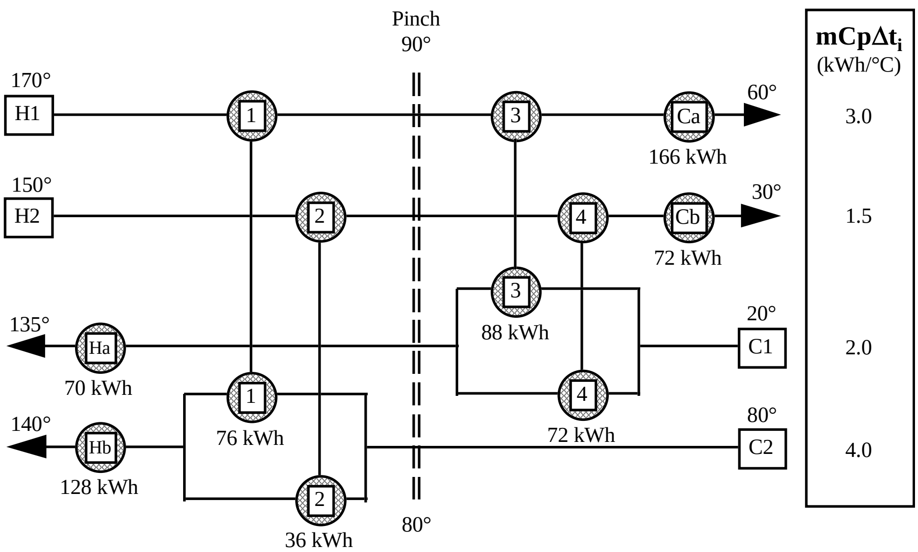


Fig. 7.6 Initial Heat Exchanger Networks for the small Example without Heat Storage

By combining the networks for the six individual time intervals, an initial heat exchanger network without heat storage can be established, as shown in figure 7.6. The utility consumption in this network corresponds to the target value for maximum direct heat exchange and no heat storage. The Pinch point refers to the overall bottleneck for heat recovery identified in the last column ("Sum") of figure 7.2.

It is important to notice that temperatures and heat transfer areas in the network cannot be obtained directly from figure 7.6. This can be explained by the fact that heat transfer area is related to the heat duty of a match, while figure 7.6 shows the total amount of heat that is transferred from a hot stream to a cold stream across all time intervals. The stream temperatures and heat transfer areas must be calculated for each time interval, and the actual heat exchanger area is then equal to the largest required area for these time intervals.

The use of ***Heat Storage*** systems does not introduce any major new aspects in the design exercise. The storages are simply represented as hot or cold streams in the various time intervals.

It should also be mentioned that a so-called ***Combinatorial Method*** (Stoltze et al., 1995, and Mikkelsen, 1998) has been developed for the design of heat recovery networks, where all heat exchange takes place through the use of heat storage. The objective of this method is to find the number of storages needed to reach the TAM targets for minimum energy consumption. At the same time, optimal temperatures are identified for each heat storage. The method is limited to storages with constant temperature and cyclic batch processes.

7.6 Network Optimization (Phase 4)

The conventional methods for optimization and simplification of heat exchanger networks for continuous processes described in section 5.4 can also be used without adjustments for batch processes. As a brief reminder, the degrees of freedom that can be utilized include Heat Load ***Loops*** and Heat Load ***Paths***, as well as Stream ***Splits***. Since the initial design is based on a combination of networks for the individual time intervals, there is normally significant scope for cost reductions and network simplifications.

Considering the initial network in figure 7.6, heat exchanger (2) only recovers 36 kWh and can be removed by manipulating the heat load loop that contains all four process/process exchangers. This will automatically remove the split on cold stream C2. It is also possible to remove the split on cold stream C1. The resulting heat exchanger network is shown in figure 7.7. There is a small penalty in energy consumption (36.8 kWh) that must be compared with the savings in investment cost.

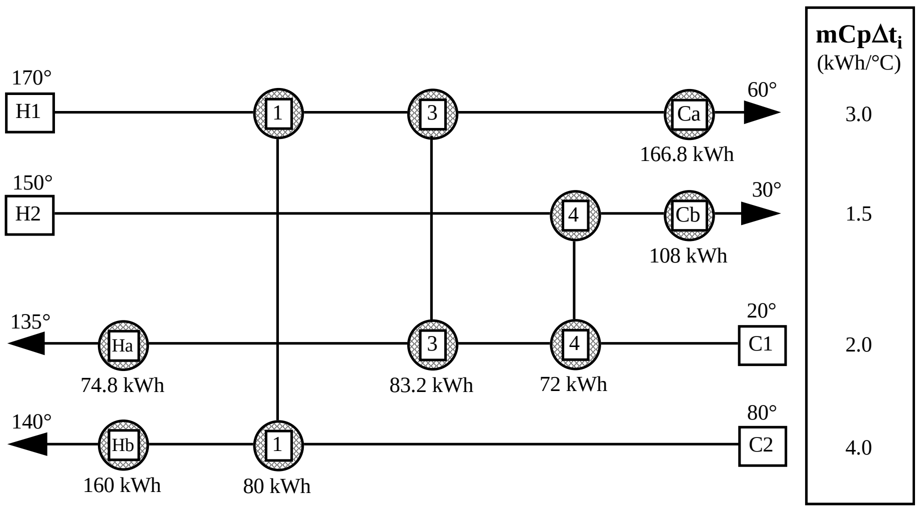


Fig. 7.7 Simplified Heat Exchanger Network for the small Example without Heat Storage

8. BASIC CONCEPTS FOR USING MATHEMATICAL PROGRAMMING IN PROCESS INTEGRATION

Mathematical Programming is a class of methods for solving constrained Optimization Problems. Since both continuous and discrete (or binary) variables can be used in the corresponding Mathematical Programming models, these methods are perfectly suited for typical design tasks encountered in Process Synthesis and Process Integration. The binary variables (0 or 1) can be used to model discrete decisions such as selection of equipment, the interconnection and sequencing of equipment and other non-continuous choices made by the designer. The continuous variables can be used to model process stream properties such as flowrate, temperature and pressure, as well as operating conditions and geometrical sizes such as area and volume of process equipment.

While Mathematical Programming belongs to the so-called deterministic optimization methods, there have also been applications of stochastic or non-deterministic methods such as Simulated Annealing and Genetic Algorithms in Process Integration. These methods have been used to overcome numerical problems such as local optima caused by non-convexities in the models, discontinuities and combinatorial explosion. It is, however, beyond the scope of this Primer to describe these methods. Use of Simulated Annealing in Process Integration has been described by Dolan et al., 1989, Pohlig et al., 1991, and by Sandvig Nielsen, 1995. Apparently, there has been less use of Genetic Algorithms in Process Integration, but one application has been described by Lewin, 1998a,b.

8.1 Motivation

Even though Pinch Analysis is used routinely in a large number of industrial companies, indicating that the technology is realistic and applicable, there are a number of limitations inherent in Pinch based methods. These limitations are actually the major motivation for using Mathematical Programming in Process Integration.

One of the most important and basic concepts in Pinch Analysis is the Composite Curves that in a single diagram gives the engineer a “bird’s eyes view” of the opportunities for heat recovery in the total process. The diagram provides details about the bottleneck(s) for heat recovery (Process and Utility Pinches), the minimum external heating and cooling requirements as a function of the specification of minimum allowed driving forces in the heat exchangers, and an indication of the total need for heat transfer area.

It is, however, an underlying assumption that all hot and cold streams are resources that can be used without limitation for heat recovery purposes. Once the streams have been merged into the hot or cold composite curve, their identity is lost. If pairs of hot and cold streams are not allowed to exchange heat (for reasons such as safety, operability, piping difficulties, contamination prevention, etc.) it becomes extremely difficult to evaluate the effect on heat recovery of such forbidden matches from the composite curves. Using Mathematical Programming, however, it is extremely simple to formulate such situations, and the corresponding solution phase actually becomes easier.

In Pinch Analysis, there is also a combination of Heuristic Rules and Thermodynamics. Of course, Thermodynamic Methods do not fail, but the Heuristic Rules are by nature only approximations that have a limited validity. One example in Process Integration is the so-called (N-1) rule for the fewest number of units (heat exchangers). Since this formula only counts the number of process streams and utility types, with no reference to temperature, it sometimes fails to properly identify the correct minimum number of units. The calculation of minimum total heat transfer area is based on the assumption (or Heuristic Rule) that vertical heat transfer minimizes total heat transfer area. As discussed in section 5.2.4, this assumption is not valid when there are significant differences in the film heat transfer coefficients for the streams.

In the design phase, Pinch based methods fail to properly address the multiple trade-offs involved due to the sequential nature of these methods. The Pinch Design Method is also quite time-consuming, and even though the matching rules are simple, it often becomes a major effort to develop a valid initial design. The strict Pinch decomposition has also been shown to be counter-productive, since the subsequent design evolution is trapped into the structure of the initial decomposed design (Sagli et al., 1990).

In summary, there are limitations in many phases of Pinch Analysis, such as the problem definition phase (hard to handle forbidden matches), the targeting phase (approximations and heuristic rules that fail), as well as the design and optimization phase (multiple trade-offs, topology traps, etc.). In theory, Mathematical Programming overcomes all these limitations, however, some of the corresponding models are extremely difficult to solve. Finally, it should be mentioned that Mathematical Programming provides a framework for Automatic Design, which means that time (which is a limiting factor in many engineering projects) can be saved and used for more high level decisions.

8.2 A Brief History

A number of attempts were made in the 60's and the 70's to apply Linear Programming (LP) in the form of assignment models to solve the matching problem of heat exchanger networks. These methods were, however, not able to produce results that could be used in industry, and therefore no references are included in this Primer.

A major step forward and a breakthrough in the use of optimization methods such as Mathematical Programming in Process Integration came in the early 80's when results from Pinch Analysis were included in the mathematical models. Cerda and Westerberg, 1983, and Cerda et al., 1983, used Transportation models to calculate minimum energy consumption in cases with restricted matching between process streams. At the same time, Papoulias and Grossmann, 1983, used the more compact and efficient Transshipment model for the same purpose. Models were also developed for minimum number of units, and three years later, Floudas et al., 1986, showed that heat exchanger networks could be generated and optimized using Mathematical Programming based on the matches identified during the targeting exercise for minimum energy and fewest number of units.

In the late 80's, completely simultaneous Mathematical Programming models were developed for automatic network design by Yee and Grossmann, 1990, and by Ciric and Floudas, 1991, however, these methods were extremely hard to solve, and the 90's were used to try to overcome these inherent numerical difficulties. In a Sequential Framework proposed by Gundersen et al., (96-99), the engineer is kept inside the decision loop and the design task divided into smaller models. At the same time, there has been increased use of Mathematical Programming within Pinch based methods during the 90's. More details about these topics are given later in this chapter and also to some extent in chapter 9.

8.3 Classes of Mathematical Programming Models

Generally, a Mathematical Programming model consists of an objective function (typically some economic criteria) and a set of equality constraints as well as inequality constraints. The general form is indicated below:

$$\begin{aligned} &\min f(x,y) \\ &\text{subject to} \\ &\quad g(x,y) \le 0 \\ &\quad h(x,y) = 0 \\ &\text{where} \\ &\quad x \in R^n \\ &\quad y \in [0,1]^m \end{aligned}$$

It should be noticed that the variables x and y in general are vectors of variables, and that the constraints g and h similarly are vectors of functions. The objective function (f) is assumed to be a scalar.

If there are no binary variables (dim(y) = 0), and all functions f, g and h are linear, we have the simplest class of problems, the ***Linear Programming (LP)*** models. Using the Simplex algorithm, for example, LP models with hundreds of thousands variables and constraints can be solved in reasonable times with today's computer resources. If there are no binary variables (dim(y) = 0), and at least one of the functions f, g and h are non-linear, we have a ***Non-Linear Programming (NLP)*** problem. These are generally much harder to solve, especially if the non-linearities are non-convex, because a <u>local</u> optimum may be found.

If there <u>are</u> binary variables in the model (dim(y) > 0), and all functions f, g and h are linear, we have a ***Mixed Integer Linear Programming (MILP)*** problem. These can be solved to global optimality provided the number of binary variables does not cause a combinatorial explosion. Finally, if there <u>are</u> binary variables in the model (dim(y) > 0), and at least one of the functions f, g and h are non-linear, we have the hardest class of problems; ***Mixed Integer Non-Linear Programming (MINLP)*** models. Unfortunately, most real design problems <u>are</u> of the MINLP type with significant problems related to computer time (combinatorial explosion) and local optima (non-convex nature).

Fortunately, in most Process Design, Synthesis and Integration applications, the binary variables do not occur in the <u>equality</u> constraints (h(x) = 0) and they appear <u>linearly</u> in the objective function and the inequality constraints.

8.4 Rigorous Targets for Heat Integration

As mentioned in section 8.1, one motivation for the use of Mathematical Programming in Process Integration is that rigorous targets can be obtained for energy requirement, number of units and heat transfer area, even in cases where there are restrictions on the matches.

Before indicating how these targets can be obtained, we have to introduce some of the basic concepts used to establish the relevant models. As indicated in the beginning of this chapter, the breakthrough in the use of Mathematical Programming in Process Integration came when results from Pinch Analysis were included in the models. The most important concept in this respect was the use of the heat cascade (as shown in figures 5.4 and 5.6 of section 5.2.1) to formulate Transshipment Models. In these models, the Temperature Intervals act as Warehouses (or "intermediate storages") between the Sources/Producers (the hot process streams and utilities) and the Sinks/Consumers (the cold process streams and utilities).

8.4.1 Transshipment Model

Figure 8.1 indicates possible heat flows to and from temperature interval (k) in a general heat cascade. A number of variables and sets must be defined, in order to develop the actual Mathematical Programming models for the targeting phase.

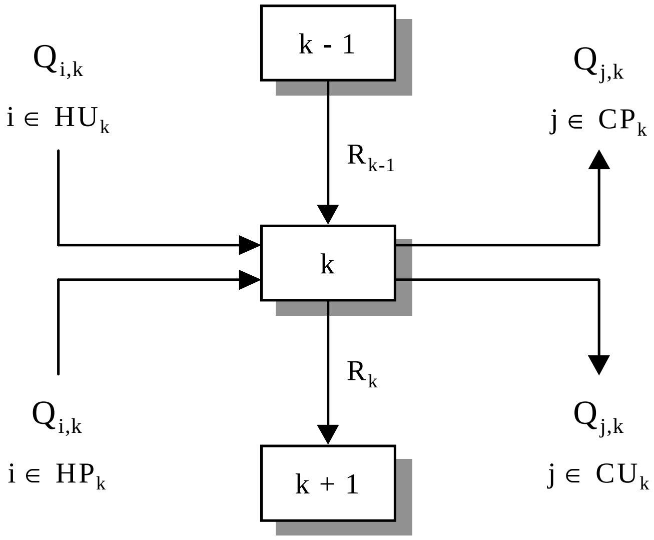


Fig. 8.1 Heat Flows related to Temperature Interval (k) in a Heat Cascade

The following definitions of variables and sets apply to figure 8.1 as well as the targeting models that will be developed in this chapter:

i	:	Index for hot process streams and utilities
j	:	Index for cold process streams and utilities
k	:	Index for temperature intervals (k = 1 is the hottest and k = K is the coldest)
TI	:	Set of all temperature intervals (k = 1,K)
HP	:	Set of all hot process streams
HP_k	:	Set of hot process streams that provide heat to interval k
HP_k'	:	Set of hot process streams that provide heat to interval k or any interval higher up in the heat cascade (higher temperatures)
HU HU_k HU_k'	:	Similar sets for hot utilities
CP	:	Set of all cold process streams
CP_k	:	Set of cold process streams that require heat from interval k
CU	:	
CU_k	:	Similar sets for cold utilities

R_k : Accumulated heat surplus (residual heat) from interval k (in kW)
$Q_{i,k}$: Heat delivered from hot process stream or utility i to interval k (in kW)
$Q_{j,k}$: Heat delivered from interval k to cold process stream or utility j (in kW)

c_i : Price for hot utility i (for example in US$/kWyr)
c_j : Price for cold utility j (for example in US$/kWyr)

8.4.2 Models for Minimum Energy Consumption (or Cost)

A simple model for minimum energy consumption or minimum energy cost in the case of multiple utilities can now be developed based on figure 8.1 and simple heat balances for each temperature interval. Model (P1) shown below only contains continuous variables and constants in linear relations; thus it is a ***Linear Programming*** model.

$$\min \left[\sum_{k \in TI} \left(\sum_{i \in HU_k} c_i \cdot Q_{i,k} + \sum_{j \in CU_k} c_j \cdot Q_{j,k} \right) \right]$$

subject to for all $k \in TI$: **(P1)**

$$R_k - R_{k-1} + \sum_{j \in CU_k} Q_{j,k} - \sum_{i \in HU_k} Q_{i,k} = \sum_{i \in HP_k} Q_{i,k} - \sum_{j \in CP_k} Q_{j,k}$$

where

$$R_k \geq 0 \quad \text{for} \quad k = 1, 2, \ldots, K\text{-}1$$

$$R_0 = R_K = 0$$

Notice that the model only contains one equation (equality constraint) for each temperature interval (a simple heat balance). These equations are written in the model in such a way that the unknowns (variables) are placed on the left side and the known variables on the right hand side of the equality sign.

Consider the heat cascade in figure 5.4 for the small example used for illustration purposes in chapter 5. Stream and utility data are given in table 5.1. The same example will be used in this chapter, however, the distillation column will not be included, and we will assume only two utility levels, high pressure steam and cooling water. The simplified stream and utility data are given in table 8.1. For simplicity reasons we will not discuss heat transfer area, and the film heat transfer coefficients have not been included.

As a convenience for the reader, the heat cascade in figure 5.4 is repeated in figure 8.2. Before developing the Mathematical Programming model for minimum utility cost, the heat cascade can, however, be simplified. If the only purpose of the model is to find minimum figures for external heating and cooling as well as the Pinch point, there is no need to include both supply and target temperatures of all the streams. This was done in chapter 5, where the heat cascade also was used to establish the Grand Composite Curve. Actually, only supply temperatures should be used to establish the temperature intervals, since these are the potential Pinch candidates. A target temperature can never become a Pinch point since the total mCp for the corresponding composite curve (hot or cold) will be reduced at such points. Since the size of the mathematical models strongly depends on the number of temperature intervals in the heat cascade, there is a need to keep the number of intervals at a minimum. This is indeed achieved by using supply temperatures only. The corresponding reduced heat cascade (and transshipment model) for the same example is shown in figure 8.3.

Table 8.1 Stream and Utility Data for the Example in Figure 5.1

Stream	ID	Ts(°C)	Tt(°C)	mCp(kW/°C)	ΔQ(kW)
Reactor Outlet	H1	270	160	18	1980
Product	H2	220	60	22	3520
Feed	C1	50	210	20	3200
Recycle	C2	160	210	50	2500
High pressure steam	HP	250	250	(var.)	
Cooling water	CW	15	20	(var.)	

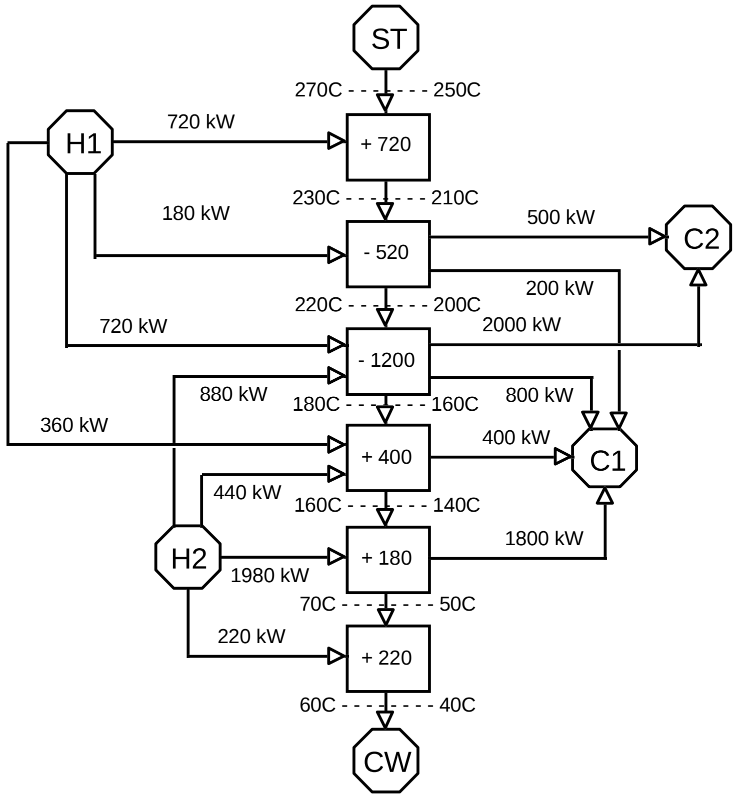


Fig. 8.2 The Heat Cascade for part of the Example in Chapter 5

We assume the same minimum approach temperature (ΔT_{min} = 20°C) as in chapter 5 and only use supply temperatures 270°C (H1), 220°C (H2), 160°C (C2) and 50°C (C1). The resulting heat cascade (figure 8.3) has only four temperature intervals compared with the six intervals in the original heat cascade in figure 8.2. The lowest interval is introduced since hot stream H2 has a lower temperature (60°C) than the hot side temperature that corresponds to the lowest cold interval temperature (50°C + ΔT_{min} = 70°C).

Since only one hot and one cold utility is considered here, minimizing utility cost is the same as minimizing utility consumption. Further, since external heating and external cooling are linked ($Q_H - Q_C$ = constant), the objective function can be simplified to minimizing the use of high pressure steam. Cooling water consumption will then also be minimized. Also notice that no temperature intervals are included for the utilities. This is the case in this example, since HP steam has a higher temperature than any of the cold process streams and cooling water has a lower temperature than any of the hot process streams. One should also point out that there is a small difference in the modeling of utilities with constant as opposed to variable temperatures.

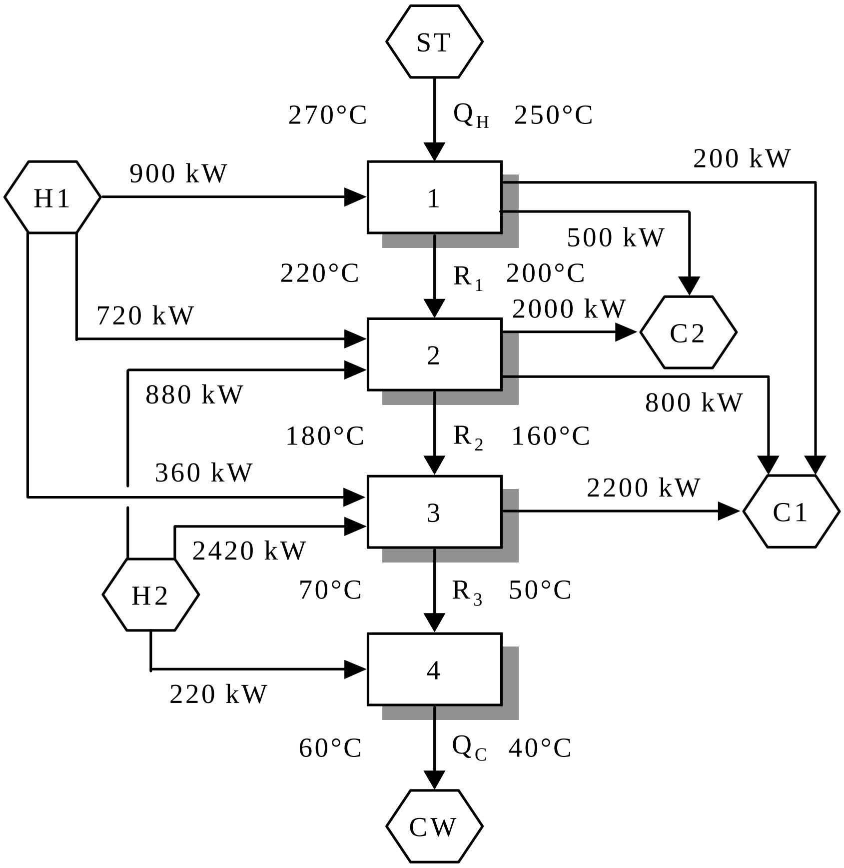


Fig. 8.3 Reduced Heat Cascade for the Example

Using the numbers from figure 8.3, the LP transshipment model (P1) for minimum utility consumption becomes:

$$\min \; Q_H$$

subject to:

$$R_1 - R_0 - Q_H = Q_{H1,1} - Q_{C1,1} - Q_{C2,1}$$
$$R_2 - R_1 = Q_{H1,2} + Q_{H2,2} - Q_{C1,2} - Q_{C2,2}$$
$$R_3 - R_2 = Q_{H1,3} + Q_{H2,3} - Q_{C1,3}$$
$$R_4 - R_3 + Q_C = Q_{H2,4}$$

where

$$R_1, R_2, R_3, Q_H, Q_C \geq 0 \quad \text{and} \quad R_0 = R_4 = 0$$
$$Q_{H1,1} = 900,\ Q_{C1,1} = 200,\ Q_{C2,1} = 500, \text{ etc.}$$

By introducing the known values for the variables on the right hand side of the equations, it is straightforward to solve this small optimization problem without using the Simplex algorithm. The model then becomes:

$$\min \; Q_H$$

subject to:

$$
\begin{aligned}
R_1 - Q_H &= 900 - 200 - 500 = 200 \\
R_2 - R_1 &= 720 + 880 - 800 - 2000 = -1200 \\
R_3 - R_2 &= 360 + 2420 - 2200 = 580 \\
-R_3 + Q_C &= 220
\end{aligned}
$$

The equations in the model can be rewritten in order to make the solution obvious:

$$
\begin{aligned}
Q_H &= R_1 - 200 \\
R_1 &= R_2 + 1200 \\
R_2 &= R_3 - 580 \\
R_3 &= Q_C - 220
\end{aligned}
$$

Since minimizing utility consumption (Q_H and Q_C) is the same as minimizing the residuals (R_1, R_2 and R_3) while the residuals are non-negative, the solution is obtained when one of the residuals becomes zero. The obvious solution is thus that $R_2 = 0$, which subsequently leads to the following results:

$$R_1 = 1200 \;,\; R_3 = 580 \;,\; Q_H = 1000 \;,\; Q_C = 800$$

Notice that compared with figure 5.6, we have lost two points on the Grand Composite Curve (heat flows of 1720 kW at 230°C/210°C and 400 kW at 160°C/140°C). This is the result from using supply temperatures only (no target temperatures) to define the temperature intervals of the heat cascade.

In summary, the Linear Programming model (P1) has resulted in the identification of a Pinch point ($R_2 = 0$) at 180°C/160°C, and minimum utility consumption of respectively $Q_{H,min}$ = 1000 kW and $Q_{C,min}$ = 800 kW. The calculations are (of course) similar to the ones used in chapter 5; so model (P1) does not represent anything new.

The true advantage of using Mathematical Programming for targeting minimum energy consumption is related to the situation where there are ***practical constraints*** on the matches between hot and cold streams. These restrictions can take the form of forbidden, required or restricted matches, however, in this Primer we will only look at situations with forbidden matches.

Assume that for some reason, hot stream H1 and cold stream C1 are not allowed to exchange heat with each other. This is a so-called ***forbidden match***. Hot stream H1 is still allowed to exchange heat with cold stream C2, in the same way as cold stream C1 can exchange heat with hot stream H2.

The problem with the heat cascades in figure 8.2 and 8.3 is exactly the same as with the Composite Curves in Pinch Analysis. As soon as a hot stream (say H1) has delivered its heat to the cascade through some temperature intervals, we do not have the tools to follow the flow of heat from this stream through the cascade and thus identify where the heat is used. To solve this problem, the heat cascade (and the transshipment model) must be expanded by introducing new residuals between the intervals that relate to a particular hot process stream or utility. Now, the purpose of the sets $HP_k^{'}$ and $HU_k^{'}$ defined in section 8.4.1, is related to the fact that once a hot process stream or utility has introduced heat into the cascade, it can be used in that particular or any other lower interval in the cascade. The reason why there are no similar sets defined for cold process streams or utilities is the fact that we find it easier (more obvious) to think that heat is flowing down through the cascade towards lower temperatures than to think that cold is flowing up through the cascade towards higher temperatures.

A general temperature interval in an ***expanded heat cascade*** is shown in figure 8.4, while the expanded heat cascade for the example is shown in figure 8.5.

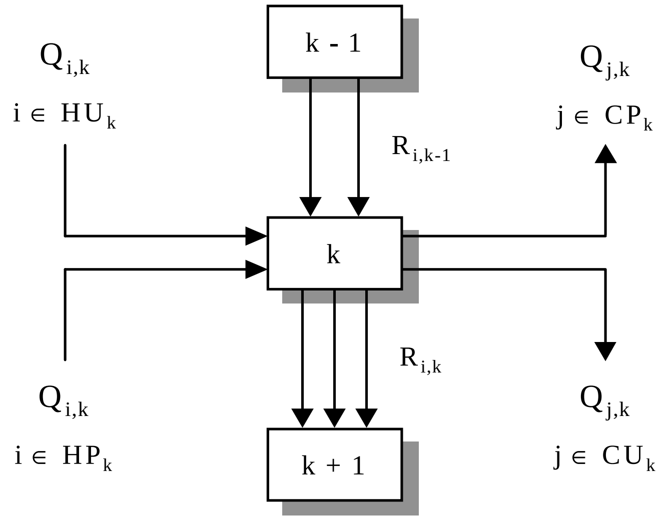


Fig. 8.4 Heat Flows related to Interval (k) in an Expanded Heat Cascade

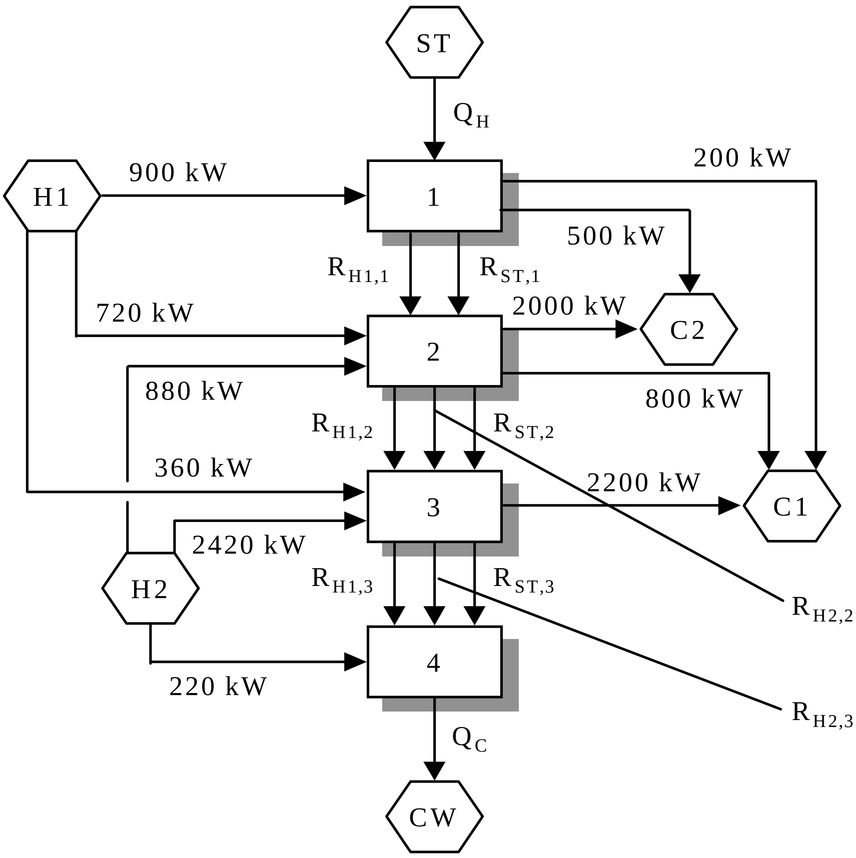


Fig. 8.5 Expanded Heat Cascade for the Example

Before developing the corresponding expanded LP Transshipment Model for targeting energy consumption in the constrained situation, the set of forbidden matches must be defined and we must also introduce new variables for heat exchange:

P : [(i,j) | $i \in HP,HU$ and $j \in CP,CU$ where a match between i and j is forbidden]

$Q_{i,j,k}$: Heat transferred from hot stream or utility i to cold stream or utility j in temperature interval k (in kW)

While the original LP model only has one total heat balance for each temperature interval, the expanded model must balance the heat for each process stream and utility in each interval. The expanded model (P2) is shown below:

$$\min \left[\sum_{k \in TI} \left(\sum_{i \in HU_k} c_i \cdot Q_{i,k} + \sum_{j \in CU_k} c_j \cdot Q_{j,k} \right) \right]$$

subject to for all $k \in TI$: **(P2)**

$$R_{i,k} - R_{i,k-1} + \sum_{j \in CP_k, CU_k} Q_{i,j,k} = Q_{i,k} \qquad i \in HP_k'$$

$$R_{i,k} - R_{i,k-1} + \sum_{j \in CP_k} Q_{i,j,k} - Q_{i,k} = 0 \qquad i \in HU_k'$$

$$\sum_{i \in HP_k', HU_k'} Q_{i,j,k} = Q_{j,k} \qquad j \in CP_k$$

$$\sum_{i \in HP_k'} Q_{i,j,k} - Q_{j,k} = 0 \qquad j \in CU_k$$

where

$$R_{i,0} = R_{i,K} = 0 \text{ and } R_{i,k} \geq 0 \text{ for } i \in HP_k', HU_k' \text{ and } k = 1, 2, \ldots, K-1$$

$$Q_{i,j,k} \geq 0 \text{ for } i \in HP_k', HU_k' \text{ and } j \in CP_k, CU_k$$

$$Q_{i,j,k} = 0 \text{ for } (i,j) \in P$$

Notice that (similar to model P1) unknown variables are placed on the left side, while known variables are placed on the right hand side of the equality sign. Also notice that the size of model (P2) is much larger than model (P1) counted in the number of variables and constraints (equations). Finally, notice that model (P2) is still linear, and that no binary variables had to be introduced in order to model forbidden matches.

Some of the high level modeling languages available to set up and solve Mathematical Programming problems (such as GAMS) allows the engineer to formulate the model using indexed equations as they occur in model (P2). For educational purposes, however, model (P2) will be written in full detail below for the example.

Again, we assume only one hot (ST) and one cold (CW) utility. First, the model is written using symbols for all variables (known and unknown), then the model is simplified using all available knowledge. Translating model (P2) using figure 8.5 results in the following optimization model:

$$\min \; Q_H$$

subject to:

(k=1):

$$R_{H1,1} - R_{H1,0} + Q_{H1,C1,1} + Q_{H1,C2,1} = Q_{H1,1}$$
$$R_{ST,1} - R_{ST,0} + Q_{ST,C1,1} + Q_{ST,C2,1} - Q_H = 0$$
$$Q_{H1,C1,1} + Q_{ST,C1,1} = Q_{C1,1}$$
$$Q_{H1,C2,1} + Q_{ST,C2,1} = Q_{C2,1}$$

(k=2):

$$R_{H1,2} - R_{H1,1} + Q_{H1,C1,2} + Q_{H1,C2,2} = Q_{H1,2}$$
$$R_{H2,2} - R_{H2,1} + Q_{H2,C1,2} + Q_{H2,C2,2} = Q_{H2,2}$$
$$R_{ST,2} - R_{ST,1} + Q_{ST,C1,2} + Q_{ST,C2,2} = 0$$
$$Q_{H1,C1,2} + Q_{H2,C1,2} + Q_{ST,C1,2} = Q_{C1,2}$$
$$Q_{H1,C2,2} + Q_{H2,C2,2} + Q_{ST,C2,2} = Q_{C2,2}$$

(k=3):

$$R_{H1,3} - R_{H1,2} + Q_{H1,C1,3} = Q_{H1,3}$$
$$R_{H2,3} - R_{H2,2} + Q_{H2,C1,3} = Q_{H2,3}$$
$$R_{ST,3} - R_{ST,2} + Q_{ST,C1,3} = 0$$
$$Q_{H1,C1,3} + Q_{H2,C1,3} + Q_{ST,C1,3} = Q_{C1,3}$$

(k=4):

$$R_{H1,4} - R_{H1,3} + Q_{H1,CW,4} = 0$$
$$R_{H2,4} - R_{H2,3} + Q_{H2,CW,4} = Q_{H2,4}$$
$$R_{ST,4} - R_{ST,3} = 0$$
$$Q_{H1,CW,4} + Q_{H2,CW,4} - Q_C = 0$$

where

$$R_{i,k} \geq 0 \quad \text{for} \quad i \in [H1, H2, ST] \quad \text{and} \quad k = 1, 2, 3$$
$$R_{i,0} = R_{i,4} = 0 \quad \text{for} \quad i \in [H1, H2, ST]$$
$$Q_{H1,C1,1} = Q_{H1,C1,2} = Q_{H1,C1,3} = 0$$

Notice again that all variables on the right hand side have known values. By introducing all known variables and omitting those with zero value, the model above becomes:

$$\min \; Q_H$$

subject to:

(k=1):

$$R_{H1,1} + Q_{H1,C2,1} = 900$$
$$R_{ST,1} + Q_{ST,C1,1} + Q_{ST,C2,1} - Q_H = 0$$
$$Q_{ST,C1,1} = 200$$
$$Q_{H1,C2,1} + Q_{ST,C2,1} = 500$$

(k=2):

$$R_{H1,2} - R_{H1,1} + Q_{H1,C2,2} = 720$$
$$R_{H2,2} + Q_{H2,C1,2} + Q_{H2,C2,2} = 880$$
$$R_{ST,2} - R_{ST,1} + Q_{ST,C1,2} + Q_{ST,C2,2} = 0$$
$$Q_{H2,C1,2} + Q_{ST,C1,2} = 800$$
$$Q_{H1,C2,2} + Q_{H2,C2,2} + Q_{ST,C2,2} = 2000$$

(k=3):

$$R_{H1,3} - R_{H1,2} = 360$$
$$R_{H2,3} - R_{H2,2} + Q_{H2,C1,3} = 2420$$
$$R_{ST,3} - R_{ST,2} + Q_{ST,C1,3} = 0$$
$$Q_{H2,C1,3} + Q_{ST,C1,3} = 2200$$

(k=4):

$$- R_{H1,3} + Q_{H1,CW,4} = 0$$
$$- R_{H2,3} + Q_{H2,CW,4} = 220$$
$$- R_{ST,3} = 0$$
$$Q_{H1,CW,4} + Q_{H2,CW,4} - Q_C = 0$$

Notice that some of the equations in the model above only contain a single variable, thus some of the variables can be obtained immediately. Introducing these known values into some of the other equations may reveal some additional occurrences where variables can be calculated before the optimization algorithm is started.

Here, we shall use another approach. Because this is a rather small problem it should be possible to solve it by using insight and logic. The forbidden match is between streams H1 and C1, thus in order to reduce energy consumption one should try to utilize H1 to heat up C2 as much as possible. Considering the heat cascade in figure 8.5, it is obvious that H1 can deliver 500 kW to C2 in interval 1. In the next interval, H1 can deliver its 720 kW as well as the 400 kW that was not utilized in interval 1, thus a total of 1120 kW can be transferred to C2 in interval 2. Stream H2 is delivering 880 kW to interval 2, where stream C1 needs 800 kW and stream C2 needs 880 kW after having received 1120 kW from stream H1. Whatever option we choose for H2 in interval 2, all its heat will be recovered. Moving down to interval 3, H2 can deliver 2200 kW to stream C1, thus 220 kW is not utilized and will end up in cooling water.

Total cooling water consumption can now be calculated by adding heat from the hot streams H1 and H2 that is not utilized to heat up the cold streams. Hot stream H1 delivers 360 kW to interval 3, while hot stream H2 delivers 220 kW to interval 3 and 220 kW to interval 4, in total 800 kW is not utilized and will end up in cooling water. Now, this is the same cooling water target as was obtained in the unconstrained case, thus in this particular case there is no energy penalty from forbidding a match between streams H1 and C1.

If, on the other hand, H1 was not allowed to exchange heat with C2, using the same logic will reveal an energy penalty of 620 kW. This penalty of course applies to <u>both</u> external heating (steam) and external cooling (cooling water), since the enthalpy changes of the process streams have not been changed. The worst penalty on energy consumption would result if H2 and C1 were not allowed to exchange heat. The penalty in that case would have been 1840 kW. When forbidden matches result in an energy penalty, there is some flexibility with respect to cross Pinch heat transfer, as indicated in figure 6.2.

In summary, we have demonstrated that Mathematical Programming can be used for targeting purposes to provide rigorous lower bounds for energy consumption even in the case of forbidden matches. Even though the corresponding expanded model (P2) is considerably larger than the basic unconstrained model (P1), the size of these models will never be a problem no matter how large the process is. Even for very large processes, model (P2) will never involve more than a few thousand variables and equations at worst.

8.4.3 A Model for Fewest Number of Units

An important spin-off from developing model (P2) is that it can be used with only minor modifications to find the fewest number of units, whether we want to decompose at the Pinch ($U_{min,MER}$) or not (U_{min}). It should also be noticed that this model should be applied after energy targeting by model (P1) or (P2), thus energy consumption figures for external heating and cooling are known.

First, the objective function of model (P2) must be replaced by one that counts the number of units (or matches). This is done by a simple summation over all the binary variables that are used to represent a potential match:

$y_{i,j}$: Binary variable that denotes whether there is ($y_{i,j}$ = 1) a match between hot process stream or utility i and cold process stream or utility j or not ($y_{i,j}$ = 0)

Secondly, there is a need to make a connection between the binary variables ($y_{i,j}$) for a potential match and the corresponding continuous variables for heat transferred between the same pair of streams ($Q_{i,j,k}$). The following logical constraint forces the binary variable ($y_{i,j}$) to become 1 as soon as any of the heat duties ($Q_{i,j,k}$) become positive (non-zero):

$$\sum_{k \in TI} Q_{i,j,k} - U_{i,j} \cdot y_{i,j} \leq 0 \quad \text{for} \quad i \in HP,HU \text{ and } j \in CP,CU$$

where $U_{i,j}$ can be any large number that is always larger than the sum over $Q_{i,j,k}$. For the purpose of the numerical solution of the model for minimum number of units, the value of $U_{i,j}$ should not be made larger than necessary, it introduces so-called "gap" in the model and makes the corresponding Mixed Integer Linear Programming (MILP) model hard to solve. One possible set of values for $U_{i,j}$ is to compare the maximum amount of heat that the two streams in question can exchange:

$$U_{i,j} = \min \left[\left(\sum_{k \in TI} Q_{i,k} \right), \left(\sum_{k \in TI} Q_{i,k} \right) \right]$$

It is also possible to use thermodynamic considerations such as feasible temperature driving forces between streams i and j to come up with even lower feasible values of $U_{i,j}$, however, this is beyond the scope of this Primer.

The Mathematical Programming model for minimum number of units (matches) has one important new feature compared with model (P2); it contains binary (or discrete) variables together with the continuous variables. Thus, it is a ***Mixed Integer Linear Programming (MILP)*** model, which is many orders of magnitude more complicated to solve than the corresponding LP model (P2). MILP models are solved using a Branch and Bound algorithm, where a sequence of LP models are solved in a binary search tree.

The basic optimization model (P3) for fewest number of units (or matches) without the obvious non-negativity constraints is shown below:

$$\min \quad \left[\sum_{i \in HP,HU} \; \sum_{j \in CP,CU} y_{i,j} \right]$$

subject to for all $k \in TI$: **(P3)**

$$R_{i,k} - R_{i,k-1} + \sum_{j \in CP_k,CU_k} Q_{i,j,k} = Q_{i,k} \qquad i \in HP_k'$$

$$R_{i,k} - R_{i,k-1} + \sum_{j \in CP_k} Q_{i,j,k} = Q_{i,k} \qquad i \in HU_k'$$

$$\sum_{i \in HP_k',HU_k'} Q_{i,j,k} = Q_{j,k} \qquad j \in CP_k$$

$$\sum_{i \in HP_k'} Q_{i,j,k} = Q_{j,k} \qquad j \in CU_k$$

and where:

$$\sum_{k \in TI} Q_{i,j,k} - U_{i,j} \cdot y_{i,j} \leq 0 \quad \text{for} \quad i \in HP,HU \text{ and } j \in CP,CU$$

Of course, forbidden matches (and any other practical constraints) can be incorporated in model (P3) in the same way as for model (P2). The main advantage of model (P3) is to provide a ***rigorous target*** for the fewest number of units as a replacement of the heuristic (N-1) rule that sometimes fails. The results from solving model (P3) also include duties for the matches, however, one should notice that there is normally more than one set of matches that result in the fewest number of units.

8.4.4 Mathematical Programming Models for Minimum Area

In section 5.2.4 the limitations of the targeting method for minimum area used in Pinch Analysis (the so-called Bath formula) were discussed. The basic assumption of vertical heat transfer is not always valid, and several methods have been proposed in the literature using Mathematical Programming to come up with a rigorous value for minimum total heat transfer area. Since, however, heat exchanger networks with close to minimum heat transfer area would be very expensive (requires a large number of heat exchangers and stream splitters/mixers), the corresponding methods are omitted here. After all, the main purpose of calculating minimum total area is to be able to estimate Total Annual Cost, as discussed in section 5.2.4.

8.5 Network Design using Mathematical Programming

Of course, it is especially in the Design Phase that the use of Mathematical Programming is attractive due to its ability to handle multiple trade-offs. That is the topic of this section, where some of the alternative approaches are briefly mentioned. A detailed description is beyond the scope of this Primer.

8.5.1 A Sequential Approach to Network Design

As mentioned in section 8.4.3, a set of matches and the corresponding duties are spin-offs from model (P3) for minimum number of units. This does not, however, mean that a heat exchanger network can be straightforward derived from the results of model (P3). First, one has to consider whether there is a one-to-one relationship between the matches in model (P3), where heat can be exchanged in one or more temperature intervals, and the resulting and implemented heat exchangers in the heat recovery network achieving the targets obtained from models (P2) and (P3). Fortunately, this one-to-one relationship has been proven by Floudas et al., 1986. Secondly, the network structure requires decisions about sequence of matchews, parallel vs. series heat exchange, flowrates for split-streams, etc.

In the same paper by Floudas et al., 1986, a ***Sequential Procedure*** was outlined where heat exchanger networks are synthesized in three steps (figure 8.7):

1) Identify minimum utility consumption ($Q_{H,min}$ and $Q_{C,min}$) and the corresponding Pinch point using a Linear Programming (LP) model such as (P1) or (P2).

2) Identify the fewest number of units and a corresponding set of matches (i,j) with heat duties ($Q_{i,j}$) for the decomposed case ($U_{min,MER}$) or the global case (U_{min}), using a Mixed Integer Linear Programming (MILP) model such as (P3).

3) Generate and optimize a heat exchanger network structure based on a stream superstructure and a Non-Linear Programming (NLP) model.

The main contribution in the paper by Floudas et al., 1986, was a ***stream superstructure*** that in a clever way allows the representation and inclusion of all possible network structures or topologies. Such a stream superstructure is shown in figure 8.6, where it is assumed that the MILP model (P3) has revealed that hot stream H1 should exchange heat with cold streams C1 and C2. With more matches, the complexity of the stream superstructure grows rapidly, however, the important point is to notice that all possible combinations of series and parallel heat exchange is included in the superstructure.

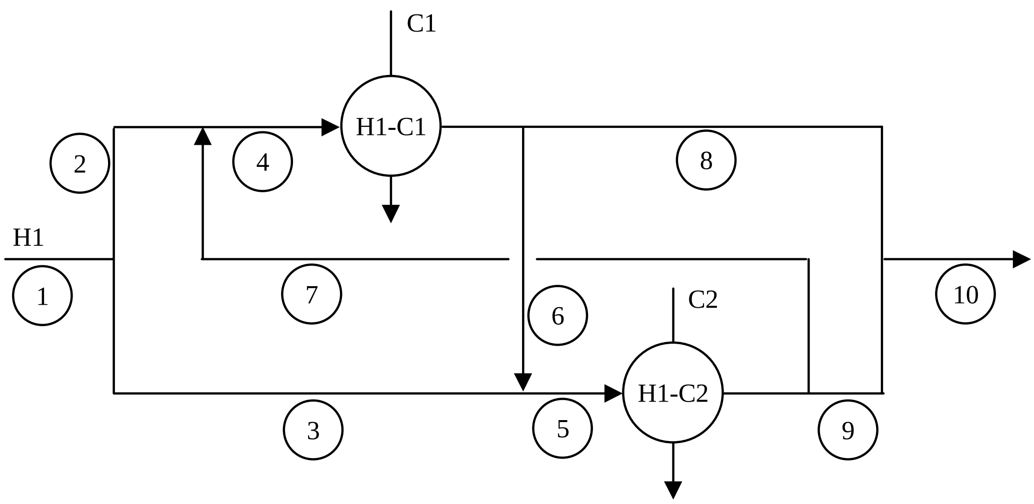


Fig. 8.6 Stream Superstructure for a hot stream (H1) involved in two Matches

Depending on the flowrates of the streams in the superstructure in figure 8.6, it is possible to have pure parallel, pure series as well as a combination of parallel/series configurations with bypass streams. Pure parallel heat exchange would be the case when streams 6 and 7 have zero flowrate, while streams 2 and 3 have non-zero flowrate. If streams 2, 6 and 9 have zero flowrate, there is a series heat exchange where stream H1 first exchanges heat with stream C2, and then exchanges heat with stream C1. The opposite sequence would be the result if streams 3, 7 and 8 have zero flowrate. The flowrates are the result of an optimization model based on the stream superstructure in figure 8.6. The objective function of this model is investment cost, which normally is a non-linear function of the area of the individual heat exchangers. These areas are also non-linear functions of the temperatures involved in the superstructure. Finally, the model consists of mass and/or enthalpy balances for the splitters, mixers and heat exchangers in the network. Since both flowrates and temperatures are unknown, these will result in non-linearities in the model since they occur in multiplications in the heat balances.

8.5.2 Alternative Design Approaches

As mentioned above, unfortunately there are multiple solutions to the MILP model for the fewest number of units. This is the main weakness of the three-stage procedure mentioned above and implemented in a software system from the 80's at Carnegie Mellon University called MAGNETS. The structure of this package is shown in figure 8.7

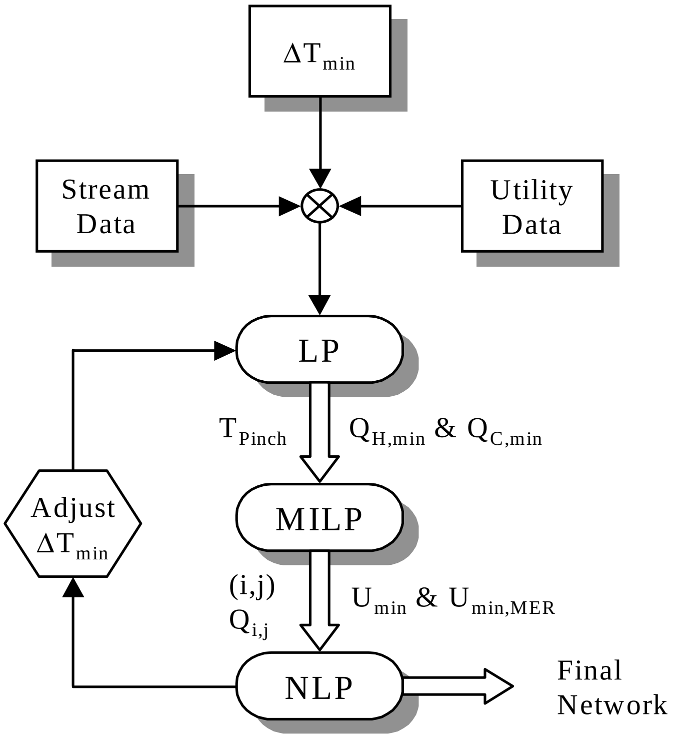


Fig. 8.7 Sequential Procedure for Heat Exchanger Networks

The problem with the ***multiple solutions*** from the MILP is that even though these solutions are equally good when it comes to the number of units, they may be drastically different when it comes to the total heat transfer area of the resulting heat exchanger network produced by the NLP. Without paying special attention, and since the "first" among the feasible sets of matches satisfying the fewest number of units is selected in a random way, the corresponding heat exchanger network may have large investment costs.

In order to overcome this problem, the obvious solution seemed to be to get rid of the decomposition indicated in figure 8.7. Both Yee and Grossmann, 1990, and Ciric and Floudas, 1991, proposed completely simultaneous ***MINLP*** models for heat exchanger network synthesis. In these models, the level of heat recovery, the number of matches, the size distribution among the matches as well as the network structure were all addressed and optimized as a single Mathematical Programming problem.

Initial results were indeed promising, and this seemed to be the way forward. These MINLP models provided a framework for automatic design, they were able to handle the multiple trade-offs in a superior way compared with Pinch Analysis, and they were also in an elegant way able to handle practical constraints in the design.

There are severe ***numerical problems*** related to these MINLP models. The non-linear relations occur in the models in a non-convex way, which means that solution algorithms have a tendency to be trapped in local optima rather than to identify the global optimum. Another problem is the combinatorial nature caused by the very large number of possible matches and network structures. After ten years of research trying to overcome these limitations, there is still little progress and industrial sized problems cannot be handled.

8.5.3 An Improved Sequential Approach

An alternative approach was initiated by Gundersen and Grossmann, 1990, where the match selection procedure (the MILP model) was improved by adding a penalty term for non-vertical heat transfer to the objective function. The corresponding ***Vertical MILP*** model was gradually improved during the 90's (the most recent improvements were described by Hashemi-Ahmady et al., 1999) and is now the core feature of a Sequential Framework for heat exchanger network synthesis (figure 8.8).

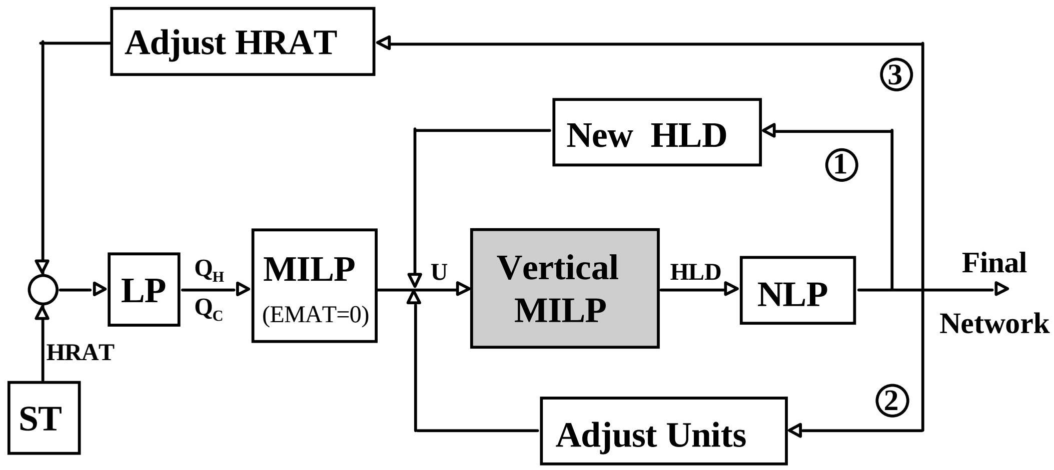


Fig. 8.8 Sequential Framework for Heat Exchanger Network Synthesis

The philosophy behind the Sequential Framework is first to get a good starting point for the level of heat recovery (represented by HRAT) by estimating and minimizing Total Annual Cost. These estimates are based on target values for energy, units and area (often referred to as SuperTargeting and denoted ST in figure 8.8). Further, the absolute fewest number of units is found (for this particular level of heat recovery) from an MILP model where EMAT (see section 6.4.1) is set to zero to maximize the matching options.

The engineer can then explore three possible iteration loops, however, it is not expected that any of these loops need more than two or three passes. The vertical MILP model produces a set of matches and corresponding heat duties (Heat Load Distributions – HLD) that is supplied to the NLP model using the same stream superstructure that was proposed by Floudas et al., 1986, for network generation and optimization. The engineer can then choose to run the vertical MILP once more to find the second (and even third) best HLD as judged by the vertical MILP to see if better networks (lower cost and/or simpler network structure) can be established. This is the innermost loop (1). Next, the engineer can repeat this after having increased the number of units by one, and thereafter adding yet another unit (loop 2). The final adjustment indicated as loop 3 is to make small changes to the level of heat recovery (HRAT).

Of course, the combinatorial issues and local optima problems mentioned for the MINLP models have not been removed by the Sequential Framework. At least, however, we do not have to cope with both problems at the same time since there is a separate MILP model for match selection and a separate NLP model for network generation. Also, it has been shown that it is much easier to find convex estimators for the NLP model in the Sequential Framework (with known heat duties) than it is for the simultaneous MINLP model.

8.5.4 Mathematical Programming in Hybrid Methods

Yet another approach has been taken at UMIST, where Mathematical Programming models and Thermodynamic Concepts and insights are combined (also referred to as Conceptual Programming) into what appears to be very powerful semi-automatic methods for both grassroots and retrofit heat exchanger network synthesis. These methods are described in some detail in chapter 9, but is generally beyond the scope of this Primer.

8.6 Summary

This chapter has briefly introduced some of the concepts, models, targeting methods and design approaches when Mathematical Programming is used to address and solve relevant issues in Process Integration. Stochastic Optimization methods were briefly mentioned in the beginning of this chapter but is beyond the scope for this Primer.

When using Mathematical Programming in Process Integration, there are <u>three</u> distinct activities that are of importance for the final result:

1) The structural alternatives must be represented in a ***Superstructure*** that is rich enough to contain all promising solutions, but not too wide to create prohibitive combinatorial problems.

2) The Superstructure must then be converted into a ***Mathematical Model*** that trades off complexity and rigor with speed in the numerical solution phase. This modeling activity is almost an art, and the same problem can be formulated quite differently.

3) The numerical solution phase where powerful ***Algorithms*** are needed to efficiently and reliably identify the “global” optimum for the proposed models.

Even though process engineers primarily should deal with the first two of these activities, history has shown that major contributions have also been made in the last activity.

9. ADVANCES IN PROCESS INTEGRATION

Process Integration has grown significantly during the 80's and the 90's from the early day's fairly simple concepts and methods that are well suited for a Primer such as this one. These days, Process Integration also consists of a set of rather complex technologies that are powerful in industrial applications, but are too elaborate to be easily conveyed in a simplified Primer. This chapter is meant to be a taste of what is available and under development, it is not the intention to go into any kind of detail, however, references will be made to relevant literature for those eager to get more involved.

It is always hard to draw the line between Process Synthesis and Process Integration, and it is always a danger that the material selected for this chapter reflects methods published in the Process Integration "community", while more Process Synthesis related material is not included. There have also been a large number of advancements in using Mathematical Programming for various tasks in design and operation of plants. Many of these relate to Process Integration, however, we have chosen to guide the reader to a comprehensive and recent overview article by Grossmann et al., 1999, for more details on these subjects.

9.1 Structuring the Material

Many of the new technologies that have been developed within Process Integration during the 90's have a wide scope, and it is therefore difficult to classify the material. We have, nevertheless, decided to try to follow the Onion Diagram (figure 3.1) in the presentation of the material in this chapter.

9.2 Reactor Systems

There has been a considerable activity on Reactor Systems in the research community, however, most of this work would be classified as Process Synthesis rather than Process Integration. The heat integration of reactors and appropriate "placement" (meaning operating conditions) for such equipment has been discussed by Glavic et al., 1988.

Much of the work on chemical reactors has applied Mathematical Programming to a large extent. Kokossis and Floudas, 1990, considered optimal design and operation of complex reactor networks. Lakshmanan and Biegler, 1996, discussed the synthesis of optimal chemical reactor networks with simultaneous mass integration. Finally in this brief section on reactors we would mention the work by Mehta and Kokossis, 1997, on the development of novel multiphase reactors using a systematic design framework.

9.3 Separation Systems

Most of the work in Process Integration on Separation Systems has been focusing on Distillation based systems, with some efforts also in ways to improve energy consumption in evaporation systems. These systems are both large energy consumers in process plants, and heat integration is important.

Figure 5.7 indicated how the potential integration of a single distillation column against a background process could be evaluated using a box representation of the column and the Grand Composite Curve for the background process. This box representation can also be used when there are more than one distillation column. The corresponding graphical representation has been referred to as the ***Andrecovich Diagram*** (Andrecovich and Westerberg, 1985).

These "boxes" were later refined by Dhole and Linnhoff, 1993, by using results from a converged rigorous simulation of the column to produce a ***Column Grand Composite Curve***. This new and powerful diagram can be used to set targets for the distillation column, and more precisely to investigate the opportunities for side reboilers and condensers, the use of feed preheating or cooling and the possible change of reflux ratio.

The synthesis of a distillation train to separate one or more multi-component mixtures into a set of single and multi-component products is a complex task. The sequence of the columns, the operating conditions (pressure, reflux ratio, etc.) and the possible heat integration of the columns must all be addressed in a simultaneous manner since these decisions have strong interactions. Apparently, this is still a research issue.

9.4 Heat Exchanger Networks

Even though the synthesis of Heat Exchanger Networks was the first task to be "solved" in early days Pinch Analysis and Process Integration, there are still unsolved issues. There is no doubt that this field has a lot of maturity, nevertheless, there have been a number of advancements during the last ten years that was not mentioned in chapters 5 and 6.

Heat recovery is achieved in heat exchangers, and in these units there is always a trade-off between heat transfer conditions and ***pressure drop***. This simple fact is important both for grassroots design and in retrofit projects. When improving heat recovery in existing plants, there is always a chance that additional heat exchangers will increase the pressure drop in such a way that pumps can no longer operate. Issues such as these, and the three-way trade-off between investment cost, thermal energy and mechanical energy was thoroughly discussed by Polley et al., 1990, in a paper on pressure drop considerations. Closely related to this topic is the question about internal ***heat transfer enhancement*** equipment in heat exchangers and the corresponding pressure drop in retrofit projects, as it is discussed by Nie and Zhu, 1999.

Continuing with retrofit considerations, it became obvious during the 90's that the Pinch point related to the Composite Curves was too strongly connected with the grassroots situation, and therefore too ideal to be applied in retrofits. Instead, newer methods for heat exchanger network retrofit relied on the so-called ***Network Pinch*** (see e.g. Asante and Zhu, 1996), which is a feature (bottleneck) of the existing network structure. While previous retrofit methods were too occupied trying to fix all "errors" in the current network, methods developed during the 90's (such as the work by Asante and Zhu, 1996) use a more "greedy" approach. The new objective was to try to come up with good retrofit projects that featured minimum number of topological changes to the existing design.

Perhaps the most important development during the 90's on heat exchanger network design is the combined use of conceptual targets from Pinch Analysis with optimization methods such as Mathematical Programming. Recently, a 3-paper series on the use of so-called ***Conceptual Programming*** for both grassroots and retrofit design was published by Briones and Kokossis, 1999. Even though the distinction between targeting, design and optimization becomes more vague when automatic methods such as Mathematical

Programming are used, Briones and Kokossis, 1999, identified so-called ***Hypertargets***, which can be explained as promising regions for the final design. In this approach, there is a development from very simple mathematical models in the early phase to more complex models in the final design stage.

Computer software is an absolute requirement in order to apply these new combined or hybrid methods for heat exchanger networks in grassroots and retrofit situations. Such software is not at present generally available, only as membership software for companies supporting the industrial Research Consortium at UMIST.

9.5 Exergy Considerations in Process Integration

Apparently, there has been a battle for quite some time between those who believe in Pinch Analysis and those in favor of using Exergy Analysis. Unfortunately (and as always), this is the result of academic attitudes where there is a strong need to promote ones own ideas, and to expand these ideas as broadly as possible. As always, it is not a question of what method to use, but rather how to get the maximum benefit from a combined use.

Actually, Exergy has been used within Pinch Analysis for quite some time, especially in areas such as Distillation and Refrigeration Cycle design. If temperature is replaced by the Carnot-factor:

$$\eta_C = (T - T_0) / T$$

on the y-axis in Temperature/Enthalpy diagrams such as the Composite Curves and the Grand Composite Curve, the area between the hot and cold Composite Curve as well as the area between the Process Grand Composite Curve and the Utility Grand Composite Curve is directly proportional to the Exergy losses involved. This fact was utilized by Linnhoff and Dhole, 1992, to produce a ***Shaftwork Target*** for subambient processes.

Unfortunately, while minimizing exergy losses sometimes also minimizes cost, there are a number of examples where the opposite is true. For a distillation column we can minimize exergy losses (or entropy production) by avoiding mixing at the feed tray, for example by making sure the feed location is right. We can also jointly minimize exergy losses and total cost for the column by some limited pre-fractionation and in some cases by limited use of distributed reboiling and/or condensing. If taken to the limit (approaching ideal or reversible distillation), however, there will be a cost increase.

Exergy losses are generally linked to non-reversible behavior in the processes. Reversible behavior is an ideal situation that can never be approached in practice. It is, however, possible to reduce these non-reversibilities for example by allowing chemical reactions to run slowly (means very large reactors) or to allow heat transfer to take place with small driving forces (means very large heat exchangers).

Another problem with Exergy Analysis is that it does not easily apply to networks or total flowsheets. Exergy Analysis tells us the exergy losses in the process equipment, but it is hard to distinguish between avoidable and inevitable exergy losses. Further, it is quite often the case that exergy losses in one unit are caused by decisions somewhere else in the process. The best example is heat exchanger network synthesis, where exergy losses are constant and independent of network structure as long as the stream data (mCp values and supply and target temperatures) are unchanged and the set of utilities (load and level) has been selected.

Here, decisions made at a higher level (when designing the reactor and separator system) are responsible for the exergy losses in the heat exchanger network, and apart from careful selection of utilities, there is nothing we can do to reduce the exergy losses. A fairly recent discussion on combined Pinch and Exergy Analysis for process modifications, is given by Feng and Zhu, 1997.

9.6 Advanced Methods for Utility Systems

The ***Utility Grand Composite Curve*** (Hall, 1989) has already been mentioned several times in this Primer. This TQ-diagram is constructed by merging the contributions from hot and cold utilities such as flue gas from a furnace or a gas turbine, various condensing steam levels, hot oil circuits, cooling water, air (for cooling), refrigeration cycles, etc. The "art" is to come up with a mix of utilities such that the Utility Grand Composite Curve as closely as possible follows the Process Grand Composite Curve.

As such, the Utility Grand Composite Curve is an important tool when designing Furnaces with air preheat and single, multiple and possibly cascaded refrigeration cycles. Another, more recent graphical representation that can be applied for Steam Turbine networks (Mavromatis and Kokossis, 1998a,b) is the ***Hardware Composites***.

For integration of heat pumps into an existing heat exchanger network, Wallin, 1996, presented ***New Grand Composite*** type curves that gives the engineer an indication of the level of modifications that are required in the existing network. The motivation behind these new curves is that the Grand Composite Curve is based on stream data and neglects the existing heat exchanger network. As such, the argument is similar to the need for a Network Pinch rather than a Process Pinch when retrofitting heat exchanger networks.

Application of Process Integration to Total Sites has been considered one of the major developments during the 90's. One of the first publications on this topic is the one by Dhole and Linnhoff, 1992, where ***Total Site Targets*** for Fuel, Co-generation, Emissions and Cooling are presented. One important part of this technology is the Site Source and Sink Profiles that are constructed from the Grand Composite Curves of the individual processes on the site. More recently, Kimura and Zhu, 1999, proposed to use the ***R-Curve*** concept for Total Site Analysis.

9.7 Analogies to the Heat Recovery Pinch

There was a brief description in section 2.4 on how the heat recovery Pinch concept has been applied through the use of analogies to other areas of process design. It should be emphasized that these analogies have produced powerful and advanced technologies that have already found significant industrial application. It is, however, beyond the scope of this Primer to provide a detailed description.

El-Halwagi, 1997, has actually written a book on Pollution Prevention through Process Integration. The ***Mass Pinch*** mentioned in section 2.4 plays an important role in this technology.

The ***Water Pinch*** technology developed by Smith and coworkers at UMIST has also been heavily applied for Wastewater Minimization and design of Distributed Effluent Treatment Systems. A more recent publication on this subject was given by Alva-Argaez et al., 1998. The Water Pinch technology has evolved similar to the Heat Recovery Pinch technology in the sense that after starting with basic concepts and powerful graphical diagrams, there has been a development towards increased use of Mathematical Programming. This is for example the case when handling multiple contaminants (Doyle and Smith, 1997).

Smith and Petela systematically addressed the more general ***Waste*** problem in a 5 paper series (1991-92), where the presentation was based on the onion diagram, and where the sources for various types of waste were discussed.

The last example on how analogies to the heat recovery Pinch can be used to develop new technologies is the so-called ***Hydrogen Pinch*** methodology. These methods have been developed primarily for oil refineries, but can also to some extent be applied for certain petrochemical plants. Hydrogen Pinch is a powerful technology for targeting and designing Hydrogen Management systems. Due to changes in the market as well as new tight environmental regulations for gasoline and diesel, most upgrading refineries will experience a shift from a hydrogen surplus situation (where hydrogen is sent to the fuel gas system) to a situation where hydrogen as a chemical becomes limiting.

In order to avoid the investment in a new Hydrogen plant (a steam reformer), there is a need for refineries to look into the optimal use, recovery and upgrading of hydrogen rich streams in the plant. While the first publications in this field (e.g. Towler et al., 1996) focused on Cost and Added Value (so-called ***Value Composites***) based on interactions with the refinery LP model for operational planning, a more fundamental approach was taken by Alves, 1999. In his thesis, two important new graphical diagrams were presented. The ***Purity Profile*** shows Hydrogen Sources and Sinks drawn as Composite Curves in a purity versus gas flowrate diagram. Based on the Purity Profile, a Grand Composite Curve type diagram referred to as the ***Hydrogen Surplus Diagram*** can be constructed.

The Hydrogen Surplus Diagram indicates the location (purity) of the Hydrogen Pinch, and the diagram can be used to investigate “appropriate placement” of hydrogen recovery processes such as Pressure Swing Absorption (PSA), membranes and cryogenic processes. Even though this technology has been developed very recently, there have already been a considerable number of applications in the refining industry. For many refineries it will be of critical importance how their hydrogen shortage problems are solved.

9.8 Component Considerations in Systems Technologies

Process Integration is, as discussed in section 2.1 a “Systems” technology. This means that focus is more on the total system (site, process, heat exchanger network, etc.) than on the individual components that the system consists of. However, one should never forget that there is an ***interaction*** between the System and its components. In Heat Integration, there has “always” been an underlying assumption that the basic building blocks are pure counter-current heat exchangers.

Of course, these units are, for various reasons, extremely rare in the process industries, and care must be taken when conclusions are drawn on the Systems level based on such simplified components. Within Pinch Analysis, this problem has to some extent been addressed by the development of targeting and design methods as well as software that considers 1-2 Shell & Tube exchangers rather than the pure counter-current exchangers. There are, however, a large number of different heat exchanger configurations applied in various industries.

There is, therefore, a need for interaction between activities on the Systems level and activities related to the design of Equipment. The interfacing of heat exchanger network synthesis and detailed heat exchanger design was discussed by Polley and Panjeh Shahi, 1991. An obvious way to utilize results from Process Integration is to try to come up with new equipment that makes it possible to realize some of the potential savings indicated by Process Integration. This is exactly the philosophy behind an article by Polley, 1993, where heat exchangers for the future are discussed.

Consider a distillation column, where both 2nd law considerations as well as more basic heat recovery philosophies indicate that using distributed reboiling and condensation could result in worthwhile savings. With today's technology (equipment) this is realized with considerable piping and external heat exchangers, which means both costly systems and practical limitations. Research is therefore conducted today to come up with new highly efficient heat transfer material that can be used <u>inside</u> a distillation column (between the trays).

To some extent, this philosophy is quite similar to how cryogenic heat exchangers ("cold boxes") were designed by companies such as Linde long before Pinch Analysis had been developed. Apparently, the first application of Composite Curves was related to the design of such multi-stream heat exchangers, where thermodynamic representations (such as TQ-diagrams) were used to identify where to supply and remove the individual streams.

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Text Book References

This section gives references (in chronological order) to text books in the area of Conceptual Process Design and Synthesis with varying focus on Process Integration methodologies.

Linnhoff B. et al. ***"A User Guide on Process Integration for the Efficient Use of Energy",*** *The Institution of Chemical Engineers, Rugby, UK, 1982.*

- Basic Pinch Technology, including targets and design for Heat Exchanger Networks, Appropriate Placement, Heat and Power Systems
- Heat Transfer Equipment
- Applications

Douglas J.M. ***"Conceptual Design of Chemical Processes",*** *McGraw-Hill, 1988.*

- Hierarchical Analysis, Design Decisions and Heuristics / Rules of Thumb
- Economic Evaluation and Quick Screening
- Recycle Structures, Reactor Systems
- Separation Systems
- Heat Exchanger Networks
- Applications

Smith R. ***"Chemical Process Design",*** *McGraw-Hill, 1995.*

- Hierarchical and Thermodynamic ("Pinch") Approach to Design
- Choice, Synthesis and Integration of Reactor and Separation System
- Targets, Design and Optimization of Heat Exchanger Network Systems
- Design of Utility Systems
- Waste Minimization and Effluent Treatment
- Safety and Health Considerations

Floudas C.A. ***"Nonlinear and Mixed-Integer Optimization: Fundamentals and Applications",*** *Oxford University Press, 1995.*

- Convex Analysis and Nonlinear Optimization (NLP)
- Mixed-Integer Optimization (MILP and MINLP)
- Applications in Process Synthesis and Integration, including Heat Exchanger Networks, Distillation-based Separation Systems, Reactor Networks and Reactor-Separator-Recycle Systems

Shenoy U.V. et al. ***"Heat Exchanger Network Synthesis",*** *Gulf Publ. Co., 1995.*

- Targeting, Design and Optimization of Heat Exchanger Networks
- Basic Pinch Design Method with Extensions, including Retrofit Situations
- Interfacing Network Synthesis with Detailed Exchanger Design
- Mathematical Programming Formulations for Heat Exchanger Networks
- Heat and Power Integration

Biegler L.T., Grossmann I.E. and Westerberg A.W. ***"Systematic Methods of Chemical Process Design",*** *Prentice-Hall, Upper Saddle River, New Jersey, 1997.*

- Preliminary Analysis and Evaluation of Processes
- Analysis with Rigorous Process Models
- Basic Concepts in Process Synthesis
- Optimization Approaches to Process Synthesis and Design
- Heat Exchanger Networks, Separation Systems, Reactor Networks
- Design and Scheduling of Multiproduct Batch Plants

El-Halwagi M.M. ***"Pollution Prevention through Process Integration - Systematic Design Tools",*** *Academic Press, San Diego, 1997.*

- Mass Integration Methodology and Tools for Pollution Prevention
- Integration of Pollution Prevention with other Process Objectives
- Graphical, Algebraic and Optimization Techniques for Allocation, Separation and Generation of Streams and Species
- Targeting Strategies for Pollution Prevention using Mass Integration
- Mass Exchange Networks, Reactive Separation, Heat Induced Separation, Membrane Separation, Benign Chemistry, and Synthesis of Environmentally acceptable Species
- Integration of Synthesis and Analysis
- CD-ROM with Software for Mass Exchange Networks and Mathematical Optimization