LENI Systems

Advanced energetics

Process integration techniques for improving the energy efficiency of industrial processes

Dr. François Marechal

Maître d'Enseignement et de Recherches
Industrial Energy Systems Laboratory
LENI-IGM-STI-EPFL
Ecole Polytechnique Fédérale de Lausanne
CH-1015 Lausanne
http://leni.epfl.ch

The following document includes parts of : Marechal F., Pinch Analysis, chapter 3.19.1.7, and 3.19.1.8 of UNESCO Encyclopedia of Life Support Systems, ed. Christos Frangopoulos and Georges Tsatsaronis, EOLSS Publishers Co Ltd.

ÉCOLE POLYTECHNIQUE FÉDÉRALE DE LAUSANNE

LENI Systems

Advanced Energetics

Objectives of the lecture and organisation

Dr. François Marechal

Maître d'Enseignement et de Recherches
Industrial Energy Systems Laboratory
Station 9
LENI-IGM-STI-EPFL
Ecole Polytechnique Fédérale de Lausanne
CH-1015 Lausanne
http ://leni.epfl.ch, francois.marechal@epfl.ch

Titre / Title	**Advanced energetics**

Enseignant(s) / Instructor(s)	**Maréchal François: GM**		Langue / Language	**EN**
Programme(s) Période(s)		Nombre d'heures / Number of hours	Spéc / filière /orient	Type
Génie mécanique (2009-2010, Master semestre 3)		**C: 2 H hebdo, Proj: 3 H hebdo**	***D***	**opt**
Génie mécanique (2009-2010, Master semestre 1)		**C: 2 H hebdo, Proj: 3 H hebdo**	***D***	**opt**
Mineur en Énergie (2009-2010, Semestre automne)		**C: 2 H hebdo, Proj: 3 H hebdo**		**opt**

Objectifs:

Maîtriser les approches modernes d'analyse et de synthèse energétiques des procédés industriels. Acquérir les bases méthodologiques pour identifier les opportunités de récupération d'énergie par échange de chaleur et d'amélioration de l'efficacité énergétique des procédés industriels par l'intégration optimale de technologies de conversion d'énergie.

Contenu:

Méthodologie de l'analyse énergétique d'un procédés industriels.

Principes de **l'analyse exergétique** pour les procédés de conversion d'énergie y compris les liés aux processus de **combustion.**

Principe de l'intégration énergétique des procédés par la **méthode du pincement** : définition des besoins de transfert de chaleur, courbes composées d'un procédés, récupération d'énergie par échange de chaleur, conception d'un réseau d'échangeurs, identification des opportunités d'amélioration du procédé.

Principe de l'intégration optimale des unités de **conversion d'énergie**: combustion, cogénération, cycles frigorifiques et pompes à chaleurs industrielles.

Evaluation thermo-economique des options d'efficacité énergétiques d'un procédé industriel.

Application à un exemple industriel.

Prérequis:

Thermodynamique

Forme d'enseignement:

cours et projet

Forme du contrôle:

examen oral

Bibliographie:

Process integration and process improvement, F. Maréchal (LENI-EPFL)

Objectives:

To master the modern methods for the energy analysis and synthesis of industrial processes. To acquire the methodological bases required to realize the energy efficiency audit of an industrial process and to identify energy savings options by process improvement, heat recovery and optimal integration of energy conversion technologies.

Content:

Methodology for the energy efficiency audit of an industrial process.

Principles of the **energy analysis** of industrial processes and energy conversion systems including **combustion** processes.

Principles of the process integration using the **pinch analysis** method : heat transfer requirement, minimum approach temperature concept, composite curves, maximum heat recovery, heat exchanger network design. Identification of the process efficiency improvement.

Principles of the optimal integration of the **energy conversion** systems : combustion, combined heat and power, refrigeration, industrial heat pumps.

Thermo-economic evaluation of energy savings options in an industrial process.

Application to one industrial process test case.

Required prior knowledge:

Thermodynamics

Type of teaching:

Teaching and project

Form of examination:

oral exam

Matière examinée / subjects examined	Session	Coefficient / Crédits ECTS	Forme de l'examen / Type of examination
Advanced energetics	**HIV**	**5**	**Oral**

1 Course Information and organisation

Semester	$Master 1 or 3$ (autumn)
Duration	14 x (3+2) hours
Web :	http ://moodle.epfl.ch/
Place	Monday 13h15-16h00, MEB331
Lecturer **Assistant**	Francois Marechal LENI-ISE-STI-EPFL STATION 9 Tel + 41 21 693 35 16 E-mail : francois.marechal@epfl.ch Bureau ME A 2 402 Helen Becker helen.becker@epfl.ch - ME A2 396 Leandro Salgueiroleandro.salgueiro@epfl.ch - ME A2 394
Avalability	Please contact us during the lecture

2 Courses organisation

2.1 Lecture notes and other documentation

Lecture notes can be downloaded from the moodle web site http ://moodle.epfl.ch.

Printed copies are available on request (1 week delivery), a sum of 25 CHF will be asked for these paper copies.

PDF of the slides will be available for download on the moodle platform and will not be distributed.

2.2 The moodle platform

The moodle platform http ://moodle.epfl.ch is used to distribute the documents and other materials required for the lecture. Each week the slides of the corresponding lectures will be available for download and a short summary of the lecture is given.

You have to consult the moodle platform in order to confirm the lectures that are given as well as the location.

2.3 Progress assessment

Each week a group of students will be asked to make the summary of the previous lesson. The summary prepared using the wiki prepared on the http ://moodle.epfl.ch platform. The text should be short and should summarise the major goals and achievement reached during the lecture. The summary is added to the previous one so that at the end you will have the presentation of the whole lectures with a progression and a summary as prepared by the students. Comments, additions or corrections are always possible. In addition, the summary is presented orally by the students at the beginning of each lesson and the next candidates are designated.

2.4 Exercises

Exercises will be proposed during the lectures and answers will be given during the lecture. In case of doubt or questions, students are asked to organise a meeting with one supervisor to receive a feedback on what has been done. Team work is encouraged. The exercises will not receive a grade : they are proposed only to help you pratice the theory and apply it.

2.5 Project

In the second part of the lecture (from week 4), you will be asked to realise a process integration study. This study will result in a report that will serve as a basis for the examination. The project is realised within a team, the results will be presented orally on week 14 and a report has to be produced.

Each member is suppose to understand what is written in the report. Regular reviewmeetings will be organised with the supervisor. The meeting is prepared by the group based on a slide presentation to allow the supervisor to assess the work that has been done and fix the next goals. The major results of the project will be presented at the end of the semester (week 14).

When possible, project are proposed and evaluated by an expert from the industry.

2.6 Feedback

Feedback will be given to students that request it. The feedback will be given if a meeting is organised and if the questions and other materials are prepared. It is recommended to organise group meetings in order to avoid repetitions.

2.7 Evaluation

The lecture is evaluated in an individual oral exam. The exam has 2 parts : a preparation phase during which the student is asked to prepare an answer to a specific question referring to the project report and to present the answer to the examiner and another part related to the theory developed during the course. It is expected that the full content of the lecture is covered by the exam.

The report is including into the final grade, the jury will have however the possibility of giving the priority to the oral exam when fixing the final grade.

3 Specific objectives

At the end of the course, the student will be able to
- Analyse a process as a system considering not only the process operations but also all the enabling processes and the energy conversion units.
- Analyse the energy efficiency of an industrial process by analysing process flowsheets and analysing the process unit operation and the way they are integrated.
- Analyse the exergy efficiency of process units and of industrial energy systems
- Identify possible energy savings by heat recovery and design the heat recovery exchanger network
- Analyse the integration of the energy conversion systems and design efficient integrated energy conversion systems
- Thermo-economically evaluate energy savings options.

4 Detailed specific goals of the course

The following tables should help you assessing your progress. By filling the bullets you should be able to follow your progress during the lecture.

4.1 Minimum Energy requirement

	Goals
ooooo	Compute the energy bill of an industrial system
ooooo	Explain what is the meaning of the $\Delta Tmin$ and what are the major parameters that define its value
ooooo	Define hot and cold streams for a process integration analysis
ooooo	Compute the energy balance of an industrial system including the application of First Law balances to verify the coherency of the data set.
ooooo	Explain the construction of the hot and cold composite curves
ooooo	Explain the construction and the use of the grand composite curve
ooooo	Compute the maximum energy recovery in an industrial process and compute its minimum energy requirement
ooooo	Explain the major assumptions and how these can be assessed or overcome
ooooo	Estimate the heat recovery heat exchanger network cost and compute the optimal $\Delta Tmin$ value of the plant.
ooooo	Identify and quantify pinch violations in a process
ooooo	Explain the More-in More-out principle
ooooo	Explain the plus-minus principle
ooooo	Identify ways to improve the minimum energy requirement from the analysis of the composite curves

4.2 Exergy analysis

	Goals
ooooo	Explain what is exergy and how exergy balances can be formulated on the basis of First and Second Laws of Thermodynamics
ooooo	Explain what is exergy analysis and explain the different types of exergy losses : thermal, chemical,..
ooooo	Characterize the exergy losses of a given process unit operation
ooooo	Realise the exergy analysis of an industrial system
ooooo	Explain the application of exergy analysis principles in the field of process integration

4.3 Heat exchanger network design

	Goals
ooooo	State the problem of the heat exchanger network design
ooooo	Explain the major rules of the pinch design method
ooooo	Apply the pinch design method to an example
ooooo	Simplify a heat exchanger network design by following loops and path
ooooo	Compare heat exchanger network design options
ooooo	Explain and apply the remaining problem analysis
ooooo	Explain the principles of the heat load distribution calculation
ooooo	Formulate the problem of the optimisation of the heat exchanger network design.

4.4 Energy conversion systems integration

	Goals
ooooo	Explain the use of the Grand composite curve to characterise the energy conversion system
ooooo	Explain the principles of the integration of energy conversion systems
ooooo	Explain the integration of combustion and the conditions under which the efficiency can be improved using air preheating
ooooo	Explain the principle of the integration of a combined heat and power production units
ooooo	Explain how to solve the problem of the optimal integration of a steam network
ooooo	Explain the conditions of the integration of heat pumping systems
ooooo	Explain the integration of refrigeration systems
ooooo	Explain the integration of utility systems with multiple utility streams
ooooo	Explain the use of mathematical programming models to compute the optimal integration of energy conversion systems
ooooo	Explain the calculation of integrated composite curves of a sub-system

Table follows on the next page

	Goals
ooooo	Explain how exergy analysis helps designing well integrated energy conversion systems

4.5 Energy efficiency audits in industrial processes

	Goals
ooooo	What is the importance of the industry in the energy consumption of a country, what are the trends, what is the role of the energy efficiency with respect to such situation
ooooo	What the energy intensity of a process
ooooo	How to compute the energy cost of a production
ooooo	How to estimate the investment and evaluate the profitability of a energy saving investment
ooooo	What are the major steps of an energy audit in the industry, who is involved, what are the rules, etc...
ooooo	What are the major process flow diagrams, when to use which one, process flowsheet simplification for heat requirement definition and process integration analysis
ooooo	Define an industrial system, what are the boundaries, etc...
ooooo	Explain when and how to apply process integration techniques to study the energy efficiency of an industrial process

4.6 Application

	Goals
ooooo	Explain the steps followed during the process integration project
ooooo	Explain the major assumptions and limit of the study
ooooo	Explain the calculation of the minimum energy requirement
ooooo	Explain the integration of the energy conversion system and in particular the integration of the cogeneration unit using a internal combustion engine.
ooooo	Thermodynamically evaluate the solutions proposed
ooooo	Present the results in a report
ooooo	Use computer tools to solve the problem

Process integration techniques for improving the energy efficiency of industrial processes

Energy savings through process integration

Dr. François Marechal

Maître d'Enseignement et de Recherches
Industrial Energy Systems Laboratory
LENI-IGM-STI-EPFL
Ecole Polytechnique Fédérale de Lausanne
CH-1015 Lausanne
http://leni.epfl.ch

The following document has been developed on the basis of : Marechal F., Pinch Analysis, chapter 3.19.1.7, UNESCO Encyclopedia of Life Support Systems, ed. Christos Frangopoulos and Georges Tsatsaronis, EOLSS Publishers Co Ltd.

Energy savings through process integration

François Marechal
Industrial Energy Systems Laboratory
Ecole Polytechnique Fédérale de Lausanne
CH-1015 Lausanne (Switzerland)

January 27, 2009

Contribution in cross cutting technology section 8.5. or 8.3

1 Process integration

The goal of process integration is to take advantage of synergies between the various energy conversion processes and mass exchanges that occur in an industrial system in order to increase its overall energy efficiency. The process integration relies on the definition of the system and therefore strongly depends on the plant location and the plant infrastructure. It is therefore difficult to extrapolate the energy efficiency increase observed in a given case since such solutions are most of the time system depend. The process integration is the energy dimension of the industrial ecology that typically concerns material flows. Process integration techniques may be applied to any industrial production site as large as a complete harbour or even to a urban area, taking advantage of district heating and cooling distribution network or other networks like electricity or gas grids.

An industrial process is analysed as a system (fig. 1) in which raw materials are converted into products and by-products. This transformation is realised by a succession of process unit operations that use a product support like water or solvents and in which energy is the transformation driver. In order to realise the transformation, the energy resources are first converted before being distributed and used in the process units. Applying the mass and energy balances to the system shows that what is not useful products, goods or services leaves the system as waste in solid, liquid and gaseous form or as waste heat radiated to the ambience.

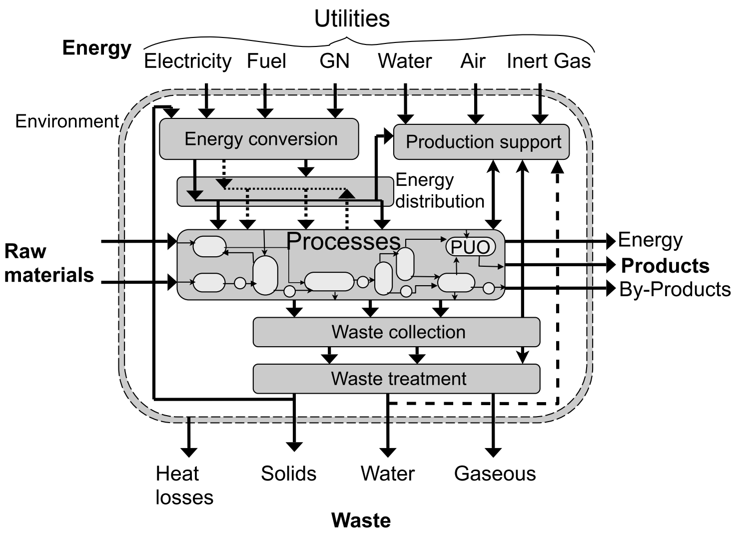


Figure 1: The systemic vision of an industrial process

By systematically analysing the materials and energy conversion processes in the different unit operations in the system and representing their possible interactions, the process integration aims at identifying synergies between the process unit operations by allowing :

Recycling of materials and energy : The goal is try to reuse materials for example by converting a waste into a product, to recycle waste streams or to convert the waste streams into useful energy. This is the typical approach of the industrial ecology and is a concept widely developed in the industrial chemicals production sites like refineries, petrochemical plants, etc...

Heat recovery : The heat recovery aim at recovering by heat exchange the heat of the hot streams (streams to be cooled down) to heat up cold streams (streams to be heated up).

Combined heat and power production : When the energy requirement of a process is in the form of heat, energy resources have to be converted into useful heat/cold to drive the processing transformations. Considering their temperature level, different types of conversion technologies can be used to produce the heating/cooling requirement. Combined heat and power (CHP) or in more general polygeneration production is typically used to convert fuels into useful heat and electricity as a by-product. Furthermore, considering the exergy content of the hot and cold streams, the CHP units such as steam cycles are also used to convert the available exergy in the heat exchange system into useful mechanical power.

Waste heat revalorisation : By energy balance of the system, the energy that is not embedded into the products leaves the system as waste heat. When its temperature level is sufficient and the heat support media is available (e.g. hot water), waste heat can be converted into mechanical power by thermodynamic cycles or can be revalorised by increasing its temperature level using heat pumping systems.

Although there exists a hierarchy between these different energy efficiency options, these have to be considered simultaneously within a systemic framework and applied considering the local conditions and the boundaries of the technically and economically accessible system.

The system boundaries definition is of major importance. Extending the system boundaries may change the impact of an energy saving action and lead to different solutions. For example, if the waste heat of a process can be reused to heat up other processes in the surroundings, the use of a heat pump to recover waste heat for the reference process will become counterproductive at the system level.

2 Heat recovery

Recovering heat from hot streams to preheat cold streams is constrained by the temperature levels of the heat required, by topology constraints and by heat exchangers investment. Due to the number of possible options, it is necessary to apply special techniques that allow to consider all streams simultaneously at the system level. Pinch analysis allows one to estimate the maximum heat recovery that can be realised in a system (without limitations on the number of streams and the size of the system) and without having to practically define the heat exchange interconnections. The Pinch analysis takes into account the temperature levels of the heat requirement and uses a parameter (so called $\Delta Tmin$) to represent the energy savings/capital trade-off that limits the heat exchanger cost by setting a minimum temperature difference between the hot and the cold streams considered for the heat recovery. The maximum heat recovery is obtained by calculating the integral of the heat available in the hot streams of the system and of the heat required by the cold streams as a function of the temperature. This draws two curves respectively the hot and cold composite curves that represent the system as one overall hot and overall cold streams between which counter current heat exchange will be used to recover heat. The maximum value is obtained by considering that the temperature difference between the two curves must always be higher than the $\Delta Tmin$ value. The point at which the curves are the closest is named the pinch point. It divides the system into two sub-systems above the pinch point, the process features a deficit of heat (heat sink) while below the process has a surplus of heat (heat source). This opens the doors to inter-processes integration since a process heat sink can profit from another process heat source provided that the temperature levels are compatible. Having defined the maximum heat recovery, the minimum energy to be supplied and removed from the process is obtained by energy balance while the pinch analysis defines their corresponding temperature levels. Comparing the minimum energy requirement with the present energy consumption defines the amount of energy that is directly transfered from the resources to the environment without having any real use in the process, as if this energy was bought to heat up the environment.

In addition, when experienced engineers are reading composite curves, they are able to identify pertinent process modifications that modify the temperature conditions of the requirement to move the heat exchange requirement from one sub-system to another, namely to change hot streams from heat source to heat sink sub-systems or the reverse for cold streams. This is obtained by changing the operating conditions of the process unit operations or by changing the type of technology that realises the operation.

2.1 Defining the hot and the cold streams

The definition of the maximum heat recovery strongly relies on the quality of the definition of the temperature levels and the corresponding heat loads of the hot and cold streams. The process integration analysis therefore requires a careful definition of the system boundaries and a systematic analysis of the process unit operations and their heat transfer requirement. Not only the amount of heat required is important but also their temperature level. Here are some rules of thumb when analysing processes for energy efficiency :

- Heat up streams at the lowest possible temperature;
- Cool down streams with the highest possible temperature;
- Streams leaving the system must be as close as possible of their equilibrium with the environment

3 Cogeneration and Combined heat and power

When the heat recovery is realised, engineers will use the pinch analysis to analyse the best way to supply or remove the heat to or from the process. In addition, based on the exergy analysis, one could try to convert the exergy lost in the heat exchanges into mechanical power or electricity. It is worth to mention that the evaluation of the integration of cogeneration units should be made after having identified the maximum energy recovery. If not, the heat recovery will not have the same value since the coproduction of electricity reduces the price of the heat supplied and the return on investment will impose to maximise the total time of operation at full cogeneration load.

Cogeneration concerns the combined production of heat and electricity as by-product when converting energy resources. After cogeneration of electricity or mechanical power, heat is sup-

Type	Typical size $[MW_e]$	η_e [%]	η_{th} [%]	Investment $[\$/kWe]$	Energy savings [%]
Gas turbine	[1:50]	[30:40]	[55:45]	[2500:1000]	[29%:36%]
Gas turbine combined cycle	[10:100]	50	35		[41%]
Engines	[0.1:10]	[30:50]	[55:40]	[4000:800]	[29%:43%]
Steam Turbine	[0.1:15.0]	[20:25]	[74:69]	[2500-450]	[26%:30%]
SOFC Fuel cell	0.1	40	40	[NA]	[33%]
Hybrid Gas turbine fuel cell	0.2	70	15	[NA]	[50%]

Table 1: Examples of cogeneration units to supply heat to processes

plied to the process by heat transfer. It must therefore have a temperature level compatible with the heat requirement of the process as defined by the pinch analysis. Only the heat of the cogeneration that is compatible with the heat sink above the pinch point will create a cogeneration effect.

The different type of cogeneration units are given on table 1. As for the specific investment, the electrical (η_e) and thermal(η_{th}) efficiencies depends on the size of the unit. The energy saving related to the integration of a cogeneration unit depends on way the co-procduced electricity is accounted. One may distinguish the avoided electricity import that substitute a centralised electricity production facility and that is accounted with the electricity mix efficiency (η_{grid}), from the electricity export that could be considered as an additional electricity production unit. In addition, one may also consider the fuel substitution effect. When the overall heat available from the cogeneration unit is used in the process, the energy saving (eq. 1) is calculated by comparing the resource consumed in the cogeneration unit with the one that would have been consumed for the separate production of heat in a boiler and of electricity from the grid.

$$\text{Energy savings } [\%] = 1 - \frac{1}{\frac{\eta_e}{\eta_{grid}} + \frac{\eta_{th}}{\eta_b}} \tag{1}$$

with
- η_e The electrical efficiency of the cogeneration unit
- η_{th} The thermal efficiency of the cogeneration unit
- η_{grid} The efficiency of the grid electricity production [38%]
- η_b The efficiency of the boiler replaced by the cogeneration unit [90%]

In the following we consider always the substitution of existing production facilities with electrcity mix corresponding to an efficiency (η_{grid}) of 38%.

Rankine cycle is another type of cogeneration unit. Rankine cycles require one heat source and one heat sink. Rankine cycles will be used above the pinch point using the process as the heat sink. In order to recover the exergy available in the heat exchange, Rankine cycles will also use the process as a heat source together with the resource conversion units. Typical applications of Rankine cycles is the steam network of industrial processes where high pressure steam is produced in boilers or in the process, is expanded in steam turbines to produce mechanical power before being condensed to supply heat to the process. The Rankine cycles are also used to valorise waste heat below the pinch point. In this case, the heat excess of the process is converted into mechanical power and the environment is used as a heat sink. When operated under low temperature conditions (typically below 100řC), the fluid used in the cycle will be an organic fluid such as the one used in refrigeration or heat pumping. One will try to use as much as possible process streams to create such CHP effects.

It is important to mention the importance of the pinch point which defines the process as a heat sink or a heat source. The cogeneration unit will produce its benefit only when it is located entirely above or below the pinch point. The heat that crosses the pinch point has no useful cogeneration effect.

4 Heat pumping

In industrial processes, heat pumps are used to upgrade the temperature level of a heat source. Referring to the pinch analysis, heat pumping will be profitable only if it allows to transfer heat from below to above the pinch, i.e. if it transforms excess heat from the system heat source into useful heat for the system heat sink. All the other situation do not lead to an overall saving : when operated only above the system pinch point, the heat pump is an expensive electrical heater for the system while when operated only below the pinch point the heat pump is an electrical heater that heats up the environment. The integration of heat pump has therefore to be studied with care.

Heat pumping systems may also use the environment as a heat source and the process as a heat sink.

The different type of heat pumps are :

Mechanical vapor recompression (MVR) : A vapour stream of the process in the heat source is compressed before being condensed to supply heat to the heat sink. MVR are typically used in evaporation, distillation or drying processes.

Mechanical Heat pumps : A fluid (typically a refrigerant) is evaporated using heat of the process heat source and compressed before being condensed to supply heat to the process heat sink.

Absorption heat pumps : Instead of using mechanical power, the absorption heat pumps are tri-therm systems. High temperature heat is used as a driver to raise the temperature of a fluid that is at a lower temperature in the heat source. The heat is send back to the process at a medium temperature in the process heat sink.

Heat transformers : Heat transformers use the same principle as the absorption heat pump but the driver in this case is at a medium temperature (evaporation from the process heat source) and send the heat back at a higher temperature (in the heat sink) while the remaining heat is sent back at the lowest temperature (typically to the environment).

Especially in mechanical driven heat pumps, the temperature level of the heat delivered and the temperature lift will define the heat pump performance. The coefficient of performance (COP) of a heat represents the amplifying effect of the heat pump, it is used to compute the energy saving of the heat pump. Considering constant temperature for the heat source ($\bar{T}_{source}$) and for the heat sink ($\bar{T}_{sink}$), the COP of the heat pump is given by eq. 2 where η_{COP} is the efficiency of the heat pump wrt the theoretical COP. Typical values for η_{COP} are around 50%.

$$COP = \frac{\dot{Q}_{th}}{\dot{E}_{hp}} = \eta_{COP} \frac{\bar{T}_{sink}(K)}{\bar{T}_{sink}(K) - \bar{T}_{source}(K)} \tag{2}$$

In integrated systems, the flow of the heat pump is limited by the heat available and required by the process and so is the heat recovery. On the composite curves, the flow limitation corresponds to the activation of a new pinch point in the system and therefore introduces the possibility of cascading heat pumps.

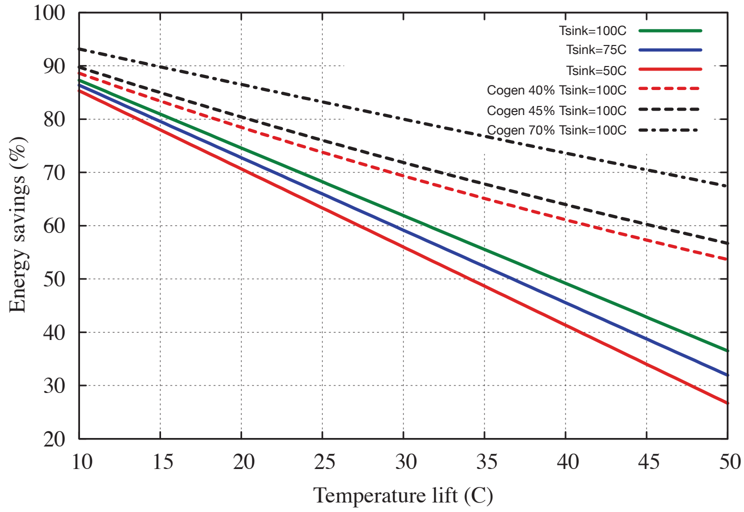


Figure 2: Energy savings by the integration of heat pumps ($\eta_{grid} = 38\%$, $\eta_b = 90\%$. $\eta_{COP} = 50\%$) and of cogeneration unit and heat pumps ($\eta_e = 40\%$, 45% and 70% with $\eta_e + \eta_{th} = 85\%$)

Heat pumps are mainly used in processes with low to medium temperature (up to 120 řC), especially in processes with evaporation systems like in the food industry or in the drying processes.

The energy saving of a heat pump is computed by eq. 3. It depends mainly on the efficiency of the grid electricity production and of the heat pumping temperatures, as shown on figure 2.

$$\text{Savings} = 1 - \frac{\eta_b(\bar{T}_{sink}(K) - \bar{T}_{source}(K))}{\eta_{grid}\eta_{COP}\bar{T}_{sink}(K)} \tag{3}$$

The the temperature level of the cogenerated heat permits, the integration of heat pumps together with a cogeneration unit is an attractive option. In this case, the electricity required by the heat pump is produced by the cogeneration unit that supplies an additional amount of heat to the process. In such a case, the energy saving is calculated by eq. 4.

$$\text{Savings} = 1 - \frac{\eta_b(\bar{T}_{sink}(K) - \bar{T}_{source}(K))}{\eta_e\eta_{COP}\bar{T}_{sink}(K) + \eta_{th}(\bar{T}_{sink}(K) - \bar{T}_{source}(K))} \tag{4}$$

Important rules of the heat pump integration

- Heat pump have to be used to raise the temperature of waste heat (i.e. below the system pinch point) to make it available in a heat sink (above the pinch point).
- The heat load of a heat pump is defined by the system heat cascade and not by the heat available in a given stream
- The environment can be used as a heat source for a heat pump
- The environment defines a pinch point, therefore refrigeration cycles may be considered as heat pumps.

5 Recycling of materials and energy conversion

When analysing process integration, it is worth to consider the value of the waste streams leaving the system as well as the one of the production support streams. This implies process improvement and modifications. The keywords will be : reduce-recycle-reuse :

Reduce : Try to minimise the waste streams production at the processing units

Recycle : Recycling requires the application of mass integration techniques to identify the possible recycling of streams in other process operation and typically requires new processing steps in order to upgrade the quality of the streams using for example membranes or distillation units.

Reuse : When the waste stream has to be treated, it can be converted into new product or converted into useful energy. Typical examples are solvent treatment by incineration or biomethanation in waste water treatement plant.

The different options described above have an impact on the energy process integration and have therefore to be considered as part of the process integration study. The waste management is usually related to an extension of the system boundary.

6 Examples

The following section presents some results obtained when applying process integration concepts. The results that are presented corresponds to energy consumption targets and not necessarily to heat recovery projects realised. Each process is obviously specific and can not be extrapolated easily, however the reported energy savings show the potential of process integration.

6.1 Pulp and paper

The table 2 shows the energy savings in a typical kraft pulp and paper processes ([2]). Results are considered considering the fossil fuel of the present process as a reference. "Process modifications" column refers to the best solution from the process integration perspective allowing for process modifications. "Heat exchangers" column refers to a situation where the process units are not modified and only the heat exchange system can be modified. For the resource consumption, we distinguish between the renewable resources used in the process (i.e the barks of the wood used to produce the paper) and the fossil fuel. The electricity is the net import of electricity. The primary energy includes the biomass and the electricity converted with a grid efficiency of 38%. The "Primary energy cogen" refers to the thermal processes only, the cogenerated energy is deduced from the fossil fuel and the biomass consumed using the grid efficiency for modeling the substituted electricity. In the Kraft process studied, the energy savings is of the order of 50 to 69%. One can see that the fuel consumption is decreasing by 71% while the cogeneration is increased by a factor 10. It should also be mentioned that in the reported study an in depth analysis of the process unit operations leads to a slight increase of the fossil fuel that is used to produce an additional amount of electricity by cogeneration (doubling of the cogenerated electricity). Kraft processes realise also the mass integration by recycling water (production support) and chemicals and converting waste (lignin) into useful process heat. In the sulfite pulp and paper process ([7]), steam import is converted in the equivalent fossil fuel consumption with a boiler efficiency of 85%. In the sulfite process, sulfite is used as chemicals for disolving the lignin. The sulfite is produced by oxidation of sulphur that provides heat to the process.

Kratf process	Present situation	Process modification	Heat exchangers
Fossil fuel	100.0	28.5(71.47%)	2.3 (97.74%)
Biomass	269.2	269.2	269.2
Electricity	353.1	289.8 (17.92%)	322.9 (8.56%)
Primary energy	1029.2	791.2 (23.12%))	851.9 (17.22%)
Cogeneration	5.6	68.9 (x12.3)	35.9 (x6.4)
Primary energy cogen	354.3	116.4 (67.16%)	177.1 (50.03%)

Sulfite process	Present situation	Process integration
Fossil Fuel	100	13 (85%)
Biomass	11	11
Electricity	48	36 (25%)
Primary energy	237	110 (53.4%)
Cogeneration	7	19 (x2.7)
Primary energy cogen	104	-23 (120%)

Table 2: Energy savings in the pulp and paper industry (data in [kJ/t dry pulp])

Black liquor, composed of lignin and chemicals, is converted into useful heat by combustion. The process under study realises mass integration by producing ethanol and lignosulfonate as by-products (biorefinery concepts). Paper is not produced in this site. Due to the high level of cogeneration obtained when recycling the chemicals in the process, the primary energy consumed in the process is negative. This is explained by the electricity to primary energy conversion factor ($\eta_{grid} = 38\%$).

6.2 Food industry

Due to the limited temperature level of the process unit operations, the food industry is a perfect candidate for the integration of heat pumping and cogeneration (see table 3). Heat pumping may be realised not only between process streams but also by using the heat of the environment. A special case is the use of the refrigeration system as a heat source for the process. In the chocolate process test case ([6]), the process integration solution concerns a boiler to supply the heat to the process and reports only the heat recovery potential, while the heat pump and cogeneration solution combines a heat pump that recovers the heat of the refrigeration systems and a cogeneration with a gas engine.

In the dairy process ([1]), heat from the evaporation process is recovered using a mechanical vapour recompression device. The recompressed flow is computed to reach the optimum integration allowing for the best integration of a gas engine as a cogeneration unit, the heat recovery from the refrigeration system and from the compressed air production. In the food processing processes, the heat/cold storage tanks are required to make the heat recovery possible between different period of operation. These tanks are also used to realise the mass integration and recycle water in the process. The reported case corresponds also to 30% water savings.

6.3 Chemical processes

For the chemical industry (table 4), 2 different processes are reported, the ethylen ([8], [4]) and the ammonia process ([3]). Ethylen and ammonia processes cover the whole range of temperature. A pinch point at very high temperature is observed in both processes which means that the process becomes a heat source at very high temperature and that actions to reduce the high

Table 3: Food processing energy savings through process integration (Primary energy concerns only the thermal aspects of the process, The reference is fossil fuel consumed in the process)

Dairy	Present situation	Heat pump + cogen.
Fossil Fuel	100	21 (79%)
Electricity	15	10 (36%)
Primary energy	140	47 (67%)

Chocolate	Present situation	Process integration	Heat pump + cogen.
Fossil Fuel	100	53 (46%)	30 (70%)
Electricity	9	3 (66%)	0 (100%)
Primary energy	123	61 (50%)	30 (75%]

temperature heat requirement together with the optimal integration of the combined heat and power production at lower temperature are critical to increase the efficiency of the process.

In the ethylen process, the efficiency increase is obtained by 3 different process integration actions :

- Fuel reduction by oxygen enrichment and air and gas preheating;
- Optimal integration of the steam cycle for combined heat and power production;
- Optimisation of the refrigeration cycle.The mechanical power requirement of the refrigeration system is reduced by 38 % by recovering the pressure and the cold effects available in the process streams and by optimising the integration of the refrigeration cycle.

In the ammonia process, the savings are related to

- the reduction of the high temperature requirement using a pre-reformer and additional feed preheating,
- the integration of the combined heat and power production in the steam cycle
- a better heat integration of the ammonia synthesis reactor by changing the reactor design.

The use of a high temperature cogeneration unit (a gas turbine is used to produce hot air for the combustion) is given as a second option. If the solutions with and without gas turbine appear to be similar for the substituted electricity accounted with a grid efficiency of 38%, the solutions differs when the grid efficiency is considered to be at the level of a new NGCC plant with an efficiency of 58%. In this case, the high temperature cogeneration appears to be less attractive than the conventional combustion solution.

6.4 Industrial chemical site

The next example (table 5) is based on a vinyl chloride production site ([5]). The results shows the interest of the site wide integration with combined heat and power realised by the steam network only or by the integrating in addition a gas turbine. The study concerned 5 processes to be integrated in a site that already contained 4 other processes of similar sizes. The purpose of

Table 4: Examples of energy savings in chemical processes (the reference is the fossil fuel consumption in the process, numbers is brackets represents the energy saving percentage)

Ethylen	Present situation	Process integration
Fossil Fuel	100	55 (45%)
Electricity	18	7 (59%)
Primary energy	147	74 (49%)

Ammonia	Present situation	Process integration	High temp. cogen.
Fossil Fuel	100	38 (62%)	77 (33%)
Electricity	0	13	-3
Primary energy (η_{grid} =38%)	100	71 (29%)	70 (30%)
Primary energy (η_{grid} =58%)	100	60 (40%)	73 (27%)

the integration was the mass integration between products and by products. The energy savings comes from the site wide integration that reduces the energy requirement by 56% in comparison with a not integrated site (obviously, in the present situation a certain level of integration is already reached). The site pinch point was used to identify some process modifications that allow heat recovery between production units in the chemical site. The heat recovery is realised using the steam network as a heat transfer media and reduces the site energy requirement by 30 %. The process modifications are only valid for the site integration and have no interest if it concerns the same process but not integrated.

Table 5: Industrial site integration results, the reference is the fossil fuel used in the site, reported electricity concerns only cogenerated electricity

Vinyl Chloride Site	Present situation	Site integration	Gas turbine
Fossil Fuel	100	57 (43%)	88 (12%)
Electricity	-4	-13 (x2.5)	-38 (x9)
Primary energy	91	24.2 (74%)	-12 (113 %)

7 Conclusions

Process integration techniques are used in order to reduce the energy consumption in a given process or preferably in a given system. When waste heat is available from a process, the heat can be used for other purposes like heating. Extending the system boundaries from the process system to its neighbor process or built area will allow to recover the waste energy but also change the pinch point location and therefore the decisions for energy efficiency improvement. The process integration perspective allows for a systematic analysis of the synergies in a system and therefore to target the energy efficiency improvement with a holistic and systemic perspective. The reported energy savings of more than 30 % show that considerable amount of energy can be saved by the proper integration of process heat sources and sinks with the appropriate integration of heat pumping and polygeneration technologies. Process integration also shows the importance of extending as much as possible the system boundaries which in turns requires new contracting mechanism (like energy services or waste management) to realise a large scale process integration.

7.1 Implementation Policies

There are some conditions

References

[1] Helen Becker. Intégration eau-énergie dans une laiterie. Master's thesis, Ecole Polytechnique Federale de Lausanne, 2007.

[2] D. Brown, F. Marechal, and J. Paris. A dual representation for targeting process retrofit, application to a pulp and paper process. *Applied Thermal Engineering*, Applied Thermal engineering(25):1067–1082, 2005.

[3] Boris Kalitventzeff, Francois Marechal, and Herve Closon. Better solutions for process sustainability through better insight in process energy integration. *Applied Thermal Engineering*, 21(13-14):1349–1368, 2001.

[4] F. Marechal, H. Closon, B. Kalitventzeff, and S. Pierucci. A tool for optimal synthesis of industrial refrigeration systems : application to an olefins plants. *AIChE annual meeting, New-Orleans, March 2002*, (Paper 89g), March 2002.

[5] F. Marechal and B. Kalitventzeff. Energy integration of industrial sites: tools, methodology and application. *Applied Thermal Engineering*, 18(11):921–933, 1998.

[6] Damien Muller. *Web-Based Tools for Energy Management in Large Companies Applied to Food Industry*. PhD thesis, ECOLE POLYTECHNIQUE FEDERALE DE LAUSANNE, THESE NËŽ3785 (2007), 2007.

[7] Zoe Perin-Levasseur and Francois Marechal. Annual report 2007, efficient energy conversion in the pulp and paper industry. Annual report, Ecole Polytechnique Fédérale de Lausanne, 2007.

[8] D. Tsahalis and et al. An expert system of energy savings technologies for the process industry. Technical report, EU Project no JOE3950020, program NNE-JOULE C final report, 1998.

Process integration techniques for improving the energy efficiency of industrial processes

Part I : Pinch analysis

Dr. François Marechal

Maître d'Enseignement et de Recherches
Industrial Energy Systems Laboratory
LENI-IGM-STI-EPFL
Ecole Polytechnique Fédérale de Lausanne
CH-1015 Lausanne
http://leni.epfl.ch

The following document has been developed on the basis of : Marechal F., Pinch Analysis, chapter 3.19.1.7, UNESCO Encyclopedia of Life Support Systems, ed. Christos Frangopoulos and Georges Tsatsaronis, EOLSS Publishers Co Ltd.

Contents

1 Introduction **4**

2 Energy-Capital trade-off for heat recovery by a heat exchanger. **4**

3 Defining the minimum energy requirement of a process **7**
3.1 The composite curves . . . 7
3.2 The pinch point . . . 9
3.3 The heat cascade . . . 10
3.4 The problem table method . . . 11
3.5 The grand composite curve . . . 11
3.6 Example . . . 13

4 Consequences of the pinch point location **14**
4.1 Heat sink and heat source . . . 14
4.2 The more in, the more out . . . 14
4.3 Penalising heat exchangers . . . 15

5 Utility integration **16**

6 Targeting the Investment **18**
6.1 The minimum number of connections target . . . 18
6.2 Total Area Target . . . 19
6.3 Capital cost estimation . . . 20
6.4 Optimal ΔT_{min} value . . . 21
6.5 Physical meaning of the ΔT_{min} . . . 22

7 Summary of the targeting method **23**

8 Heat Exchanger Network (HEN) design **24**
8.1 Representing a heat exchanger network . . . 24
8.2 The HEN design target . . . 25

9 The Pinch Design Method **26**
9.1 Feasibility rules . . . 26
9.2 Heuristic rules . . . 28

10 Mathematical programming approach **32**
10.1 Heat load distribution . . . 32

11 Optimising the heat exchanger network design **33**
11.1 Loops and Path for reducing the number of heat exchangers . . . 33
11.2 Using Mixed Integer Non Linear Programming methods . . . 36

12 Final remarks concerning the heat exchanger network design **37**

Glossary

Nomenclature

Capital letters correspond to total quantities while lower-case letters correspond to specific values.

A	$[m^2]$	Heat exchanger area
$\dot{Q}$	[kW]	Heat load or heat exchanged
$\dot{Q}^+$	[kW]	Hot utility requirement
$\dot{Q}^-$	[kW]	Cold utility requirement
T	[K]	Temperature
T^*	[K]	Corrected temperature
c_p	[kJ/(kg K)]	Specific heat at constant pressure
$\dot{M}$	[kg/s]	Mass flowrate
ΔT_{min}	[K]	Minimum approach temperature in a heat exchanger
c	[MU/unit]	Specific cost in (MU) monetary units of one unit per mass or energy.

Sub and superscripts

$^+$	entering the system
$^-$	leaving the system
$_h$	hot stream
$_c$	cold stream

Abbreviations

MER	Minimum Energy Requirement
HEN	Heat exchanger network
MI(N)LP	Mixed Integer (Non) Linear Programming
EMAT	Exchanger Minimum Approach Temperature
HRAT	Heat Recovery Approach Temperature

Summary

The **pinch analysis** is a technique that has been developed to identify the possible energy recovery by counter-current heat exchange between the hot streams to be cooled down and the cold streams to be heated up in a system. The pinch analysis is based on the definition of the **minimum approach temperature** (ΔT_{min}) that represents the energy-capital trade-off between the energy savings obtained by heat exchange and the required heat exchangers investment. For a given system, a pinch analysis is made in three steps : 1) the definition hot and cold streams, 2) the calculation of the minimum energy requirement (**targeting step**) and 3) the design of the heat exchanger network (**synthesis step**). The first step relies on the definition of the process unit operations and their required thermodynamic operating conditions. The second step of the analysis is made by computing the **hot and cold composite curves** of the process and identifying the **pinch point** location. The hot and cold composite curves represent respectively as a function of the temperature, the heat load available for heat exchange in the hot streams of the process and the heat required by the cold streams. The pinch point is identified by computing the **heat cascade** of the process that represents the maximum heat recovery between the hot and the cold streams, considering the minimum approach temperature constraint. The identification of the pinch point allows one to compute the **minimum energy requirement** of the process prior to any heat exchangers network rearrangement. The method

allows one to target the possible energy savings and to identify what goes wrong with the present heat exchange system in the process. Based on the pinch point location, **heat exchanger networks** that realise the targeted heat recovery can then be synthesised using either heuristic and feasibility rules or applying mathematical programming methods. The pinch analysis method has been applied in various industrial sectors where heat transfer plays an important role. By its holistic nature, pinch analysis allows one to solve large scale problems like industrial sites or eco-industrial clusters.

1 Introduction

Pinch analysis is a method that aims at identifying the heat recovery opportunities by heat exchange in complex thermal processes. Based on the pioneering work of Umeda, the pinch analysis has been mainly developed in the early 70's by Linnhoff and co-workers who developed a graphical method to calculate the minimum energy requirement of a process and design the heat recovery exchanger network and by Grossmann and co-workers who developed a mathematical programming framework for the design of heat exchanger networks. The graphical tools and the mathematical methods have converged to propose nowadays tools and methods that help in the identification of energy recovery by heat exchange and energy savings in site wide complex systems. The pinch analysis has been first developed for studying energy savings in the chemical process industry. Since then, the pinch analysis has been applied in the other industrial sectors where thermal operations occur like food, cement, pulp and paper, metallurgy, power plants, urban systems, etc.

The power of the pinch analysis stands mainly in its ability to offer a holistic analysis of the possible heat exchanges in a large and integrated system.

2 Energy-Capital trade-off for heat recovery by a heat exchanger.

When analysing a process system, the basic goal of the pinch analysis is to identify the possible heat recovery between the streams to be cooled down (so-called **hot streams**) and the streams to be heated up (so-called the **cold streams**) by using counter-current heat exchangers.

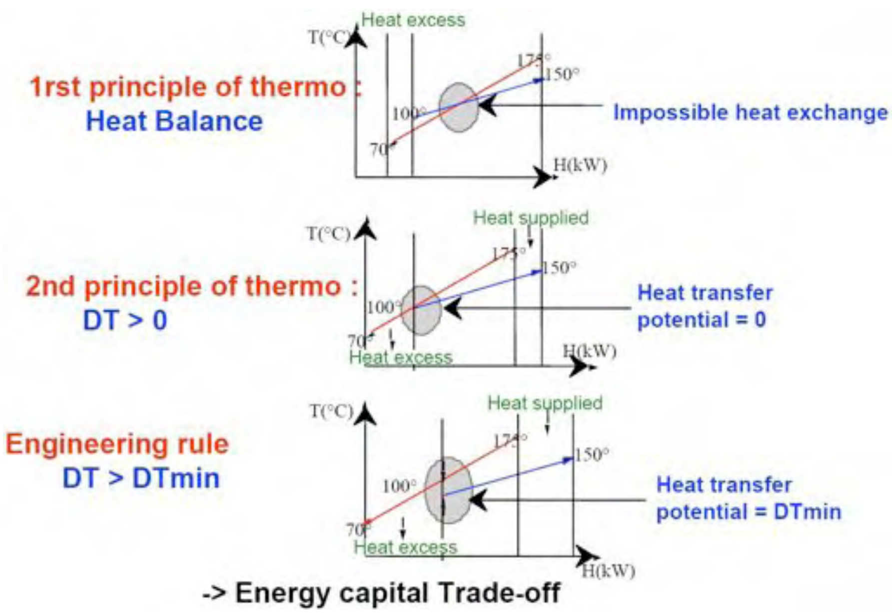


Figure 1: Needing of ΔT_{min}

Figure 1 represents a heat exchanger between a cold stream to heat up (from 100°Cto 150°C)

and a hot stream to cool down (from 175°Cto 70°C). Different situations may occur. In the first situation, we consider only the heat balance representing the first principle of thermodynamics. The cold stream heat requirement is balanced by the heat available in the hot stream, leading to a heat requirement equal to zero. Unfortunately, in this situation, the second law of the thermodynamic says that the exchange is impossible.
It becomes possible only if the temperature difference between the hot and the cold stream is greater than zero. As enthalpy has no fixed origin, the Cp curve of the streams can be shifted horizontally, until the temperature difference is greater or equal to the chosen ΔT_{min}. When the value is zero, the exchange is always thermodynamically feasible but, because of a zero heat transfer potential, the heat exchange will require an infinite area. This is not acceptable for the engineer who will try to obtain an acceptable area.

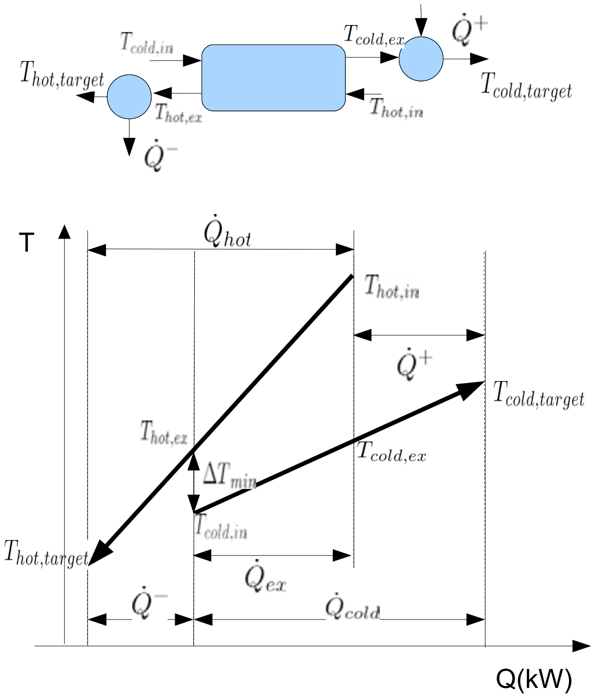


Figure 2: Heat exchanger example

$$OC^{ref} = (c^{-}\dot{Q}^{-}_{cold} + c^{+}\dot{Q}^{+}_{hot}) \cdot time_{year} \tag{1}$$

Let us consider the case of one cold stream to be heated up from an inital temperature ($T_{cold,in}$) to a target temperature ($T_{cold,target}$) and one hot stream to be cooled down from $T_{hot,in}$ to $T_{hot,target}$ (Figure 2). Assuming a constant specific heat capacity ($c_{p_{cold}}$), the heat required by the cold stream is calculated by $\dot{Q}^{-}_{cold} = \dot{M}_{cold}\, c_{p_{cold}}(T_{cold,target} - T_{cold,in})$ and the heat available in the hot stream is $\dot{Q}^{+}_{hot} = \dot{M}_{hot}\, c_{p_{hot}}(T_{hot,in} - T_{hot,target})$. Without heat recovery, the annual cost of the energy requirement (OC^{ref}) is computed by eq. 1 considering c^{+}, the cost of the hot utility used to supply the heat to the cold stream, c^-, the cost of the cold utility used to cool down the hot stream and $time_{year}$, the annual operating time of the process.

A counter-current heat exchanger may be used to recover the heat from the hot stream in order to preheat the cold stream (Figure 2). The energy savings correspond to the heat load exchanged in the heat exchanger ($\dot{Q}_{ex}$), it is obtained at the expense of an investment (I_{ex}) that is a function of the heat exchanger area (A_{ex}) . The heat exchange area is computed from eq. 2.

$$\begin{aligned}
\dot{Q}_{ex} &= \dot{M}_{hot}\, c_{p_{hot}}(T_{hot,in} - T_{hot,ex}) = \dot{M}_{cold}\, c_{p_{cold}}(T_{cold,ex} - T_{cold,in}) \\
\dot{Q}_{ex} &= U_{ex} A_{ex} \Delta T_{lm} \\
\Delta T_{lm} &= \frac{(T_{hot,in} - T_{cold,ex}) - (T_{hot,ex} - T_{cold,in})}{ln\left(\dfrac{(T_{hot,in} - T_{cold,ex})}{(T_{hot,ex} - T_{cold,in})}\right)} \\
\frac{1}{U_{ex}} &= \frac{1}{\alpha_{cold}} + \frac{e}{\lambda} + \frac{1}{\alpha_{hot}}
\end{aligned} \tag{2}$$

With:

U_{ex}	$[kW/m^2/K]$	the overall heat transfer coefficient of the heat exchanger;
α_{cold}	$[kW/m^2/K]$	the convective heat transfer coefficient of the cold stream;
α_{hot}	$[kW/m^2/K]$	the convective heat transfer coefficient of the hot stream;
λ	$[kW/m/K]$	the thermal conductivity of the tubes;
e	[m]	the thickness of the tubes.

The installed cost of the heat exchanger may for example be estimated by a power law relation : $I_{ex} = a_{ex}(A_{ex})^{b_{ex}}$. In order to compare the annual investment cost IC_{ex} with the energy savings, an annualised value of the investment is obtained by considering the annualisation interest rate (i) and the expected lifetime of the equipment (ny_{ex}). The possible heat exchange is limited by the **approach temperature** between the hot and the cold stream in the heat exchanger. When the approach temperature is small, the energy savings are high but the investment required is also high, when the approach temperature is bigger, the investment decreases while the operating costs are increasing. The **minimum approach temperature (ΔT_{min})** is the smallest temperature difference between the hot and the cold streams in the heat exchanger. The minimum approach temperature can be used as a parameter to determine the optimal size of the heat exchanger. The calculation of the optimal value of the ΔT_{min} is shown on figure 3. The trade-off curve is obtained by combining the annual operating cost $OC_{ex}(\Delta T_{min})$ considered over a yearly operating time of $time_{year}$ (eq 3) and the annualised investment $IC_{ex}(\Delta T_{min})$ computed by eq. 4.

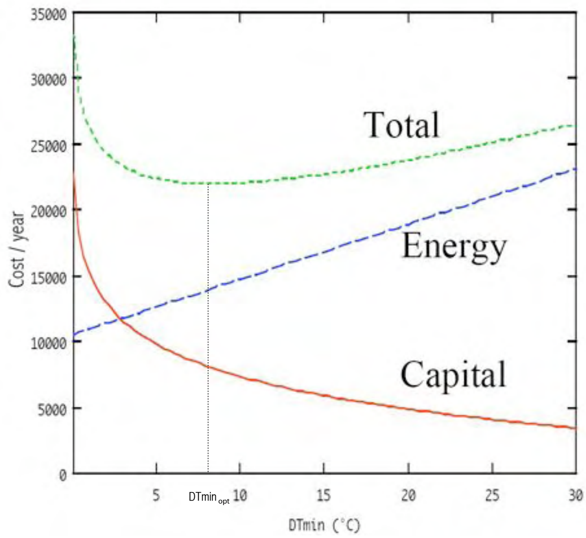


Figure 3: Energy Capital trade-off

$$OC_{ex}(\Delta T_{min}) = (c^-(\dot{Q}^-_{cold} - \dot{Q}_{ex}(\Delta T_{min})) + c^+(\dot{Q}^+_{hot} - \dot{Q}_{ex}(\Delta T_{min}))) \cdot time_{year} \quad (3)$$

The hot and cold streams heat load being constant, the energy saving of the hot utility is identical to the energy saving of the cold utility.

$$IC_{ex}(\Delta T_{min}) = (\frac{i(1+i)^{ny_{ex}}}{(1+i)^{ny_{ex}} - 1}) a_{ex}(A_{ex}(\Delta T_{min}))^{b_{ex}} \quad (4)$$

where

$$A_{ex}(\Delta T_{min}) = \frac{(1-\kappa)}{\dot{M}_{hot} c_{p_{hot}} U_{ex}} (ln((T_{hot,in} - T_{cold,in})(1-\kappa) + \kappa \Delta T_{min}) - ln(\Delta T_{min}))$$

$$\kappa = \frac{\dot{M}_{hot} c_{p_{hot}}}{\dot{M}_{cold} c_{p_{cold}}}$$

$$\dot{Q}_{ex}(\Delta T_{min}) = \dot{M}_{hot} c_{p_{hot}} (T_{hot,in} - (T_{cold,in} + \Delta T_{min})) \quad (5)$$

3 Defining the minimum energy requirement of a process

3.1 The composite curves

When seeking to identify the possible heat recovery by heat exchange in a process, one has first to define the hot and cold streams. For this purpose, the process will be defined using the **energy flow diagram**. This diagram is obtained by considering the production process as a succession of unit operations that transform given inlet streams with specified thermodynamic states (temperature , pressure, composition and flowrate) conditions into resulting streams defined by their

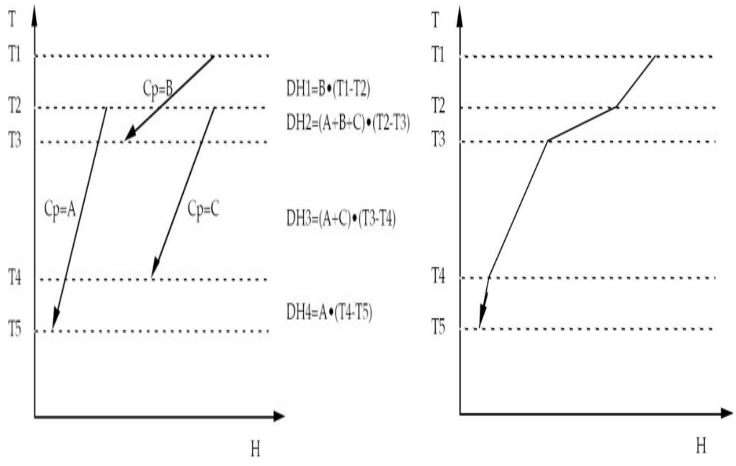


Figure 4: Hot composite curve construction

corresponding thermodynamic states. The hot and cold streams of the system will be defined by the necessary change of enthalpy (eventually pressure) between an **initial state** (outlet conditions of one unit operation or inlet state in the system) and a **target state** (required inlet conditions at the entry of the next operation or at the outlet of the system). From this definition, the process heat transfer requirement will be defined as a list of hot and cold streams.

Considering that all the exchange between the hot and the cold streams may be realised, the heat of the hot streams available for heat exchange in the process will be drawn as a function of the temperature. The obtained enthalpy-temperature profile will constitute the **hot composite curve**. The same approach is used for establishing the **cold composite curve** that represents as a function of the temperature the heat required from a heat exchange by the cold streams of the process.

The construction mechanism of the hot composite curve is illustrated for three hot streams in figure 4 varying from T_1 to T_5. For each elementary temperature interval in the temperature range, the cumulated heat load is calculated as a sum of the contributions of each of the streams present. When using constant c_p, the curve can be calculated by dividing the temperature range in successive linear segments defined by the extreme temperatures of the streams. For this construction, we assume that the streams have constant c_p. Fluid phase changing streams or highly non linear c_p are represented into successive segments with constant c_p. The heat load required is computed by $\sum_{h \in \{\text{Hot streams in interval k}\}} \dot{M}_h c_{p_{h,k}} (T_k - T_{k-1})$.

For the whole system, the composite curves can be seen as hot and cold streams that could exchange heat using counter-current heat exchangers (figure 5). Heat recovery between hot and cold composite curves is feasible when the hot composite is above the cold composite. Assuming a ΔT_{min} value, the cold composite may be shifted horizontally until the smallest vertical distance between the two composite reaches the ΔT_{min} value. Like in the two streams example we can then read on the figure the minimum hot ($\dot{Q}^+$) and cold ($\dot{Q}^-$) energy requirement.

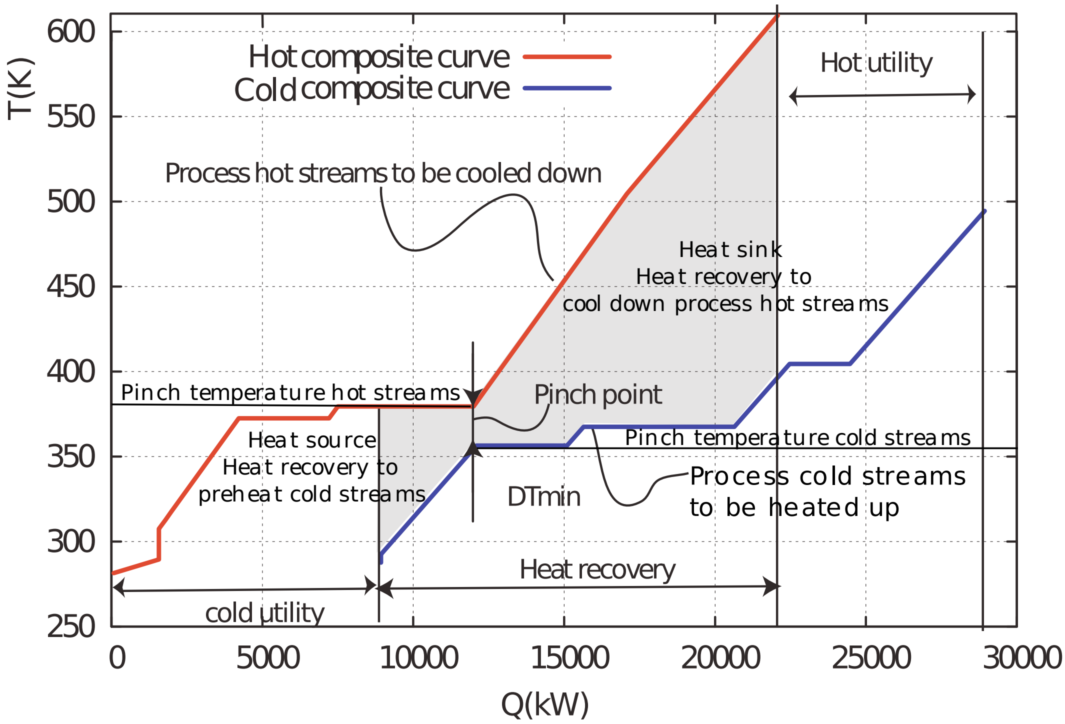


Figure 5: Hot and cold composite curves of a process

3.2 The pinch point

The point where the approach temperature between the two curves is equal to the chosen ΔT_{min} value is called the **pinch point**. Usually, the pinch point does not appear at an extreme temperature like in the two streams exchange. Its position localises the temperature of the process where the heat transfer is the most difficult and the temperature differences (the driving force) in the heat exchangers will be the smallest. Away from this point, the exchanges will be easier with higher approach temperatures. The pinch point identifies the bottleneck of the process in terms of heat recovery potential. The analysis of the streams in the vicinity of the pinch point will be of great help to further improve the energy efficiency of the process by changing the operating conditions of the unit operations concerned in order to create new energy recovery opportunities.

3.2.1 Graphical localization of the pinch point

The graphical localization of the pinch point is done by shifting horizontally the cold composite curve, until it is at any point DTmin under the hot composite curve. The figure 5 shows the graphical localization of the pinch point. Like in the two streams example, we can read on the graphic the minimum hot ($\dot{Q}_{hmin}$) and cold ($\dot{Q}_{cmin}$) utility requirements. These represent the minimum energy requirements (MER).

3.3 The heat cascade

Mathematically, the minimum energy requirement is computed by solving the heat cascade (8). This model is based on the definition of the **corrected temperatures** that are obtained by reducing the initial and target temperatures of the hot streams by $\frac{\Delta T_{min}}{2}$ and increasing the temperatures of the cold streams by $\frac{\Delta T_{min}}{2}$ (eqs. 6 and 7)

$$T_h^* = T_h - \frac{\Delta T_{min}}{2} \quad \forall h \in \{hot\, streams\} \tag{6}$$

$$T_c^* = T_c + \frac{\Delta T_{min}}{2} \quad \forall c \in \{cold\, streams\} \tag{7}$$

When a pinch occurs between the hot stream h and the cold stream c, the approach temperature between the two streams is equal to ΔT_{min}. When $T_h - T_c = \Delta T_{min}$, then $T_h^* - T_c^* = 0$ which corresponds to an intersection between the two curves.

The corrected temperatures define an ordered list of $n_r + 1$ increasing temperatures. A temperature interval r is defined by two successive temperatures : from T_r^* to T_{r+1}^*. Considering R_r, the heat cascaded from the system at a temperature higher than T_r, the energy balance may be written for each temperature interval. The heat cascade model (eq. 8) is a one degree of freedom linear programming problem that computes the minimum energy required $\dot{Q}^+ = R_{n_r+1}$ to balance the needs of the cold streams when recovering the maximum energy from the hot streams by counter-current heat exchange and cascading the heat excess to the lower temperatures.

$$\min_{R_r} \dot{Q}^+ = R_{n_r+1} \tag{8}$$

subject to heat balance of the temperature intervals :

$$\begin{aligned} R_r = & R_{r+1} \\ & + \sum_{h_r \in \{\text{hot streams in interval } r\}} \dot{M}_{h_r} c_{p_{h_r}} (T_{r+1}^* - T_r^*) \\ & - \sum_{c_r \in \{\text{cold streams in interval } r\}} \dot{M}_{c_r} c_{p_{c_r}} (T_{r+1}^* - T_r^*) \quad \forall r = 1, ..., n_r \end{aligned} \tag{9}$$

and the heat cascade feasibility

$$R_r \geq 0 \quad \forall r = 1, ..., n_r + 1 \tag{10}$$

With this definition, the value of the heat cascaded from the highest temperature (R_{n_r+1}) represents the *minimum energy requirement* (**MER**) of the process ($\dot{Q}^+$). It is assumed to be supplied to the process with a hot utility stream with a temperature higher than $T_{n_r+1}^* + \frac{\Delta T_{min}}{2}$. By heat balance, $\dot{Q}^- = R_1$ represents the heat to be removed from the process by a cold utility with an expected temperature lower than $T_1^* - \frac{\Delta T_{min}}{2}$.

The corrected temperature $T_{r_{min}}^*$ corresponding to the inequality constraint $R_{r_{min}} = 0$ is the pinch point temperature, it corresponds to a real temperature of $T_{r_{min}}^* + \frac{\Delta T_{min}}{2}$ for the hot streams and $T_{r_{min}}^* - \frac{\Delta T_{min}}{2}$ for the cold streams. When $r_{min} = 1$ or $r_{min} = n_r + 1$ the problem is said to be a *threshold* problem with respectively no cold utility or hot utility and without pinch point.

3.4 The problem table method

The problem table method is an algorithm proposed by Linnhoff to solve the heat cascade problem and to localize pinch point.

1. Define the hot and the cold streams
2. Divide the enthalpy-temperature profile into linear segments
3. Compute the ordered list of corrected temperatures
4. Compute the temperature difference $\Delta T_r = T^*_{r+1} - T^*_r$
5. For each temperature interval r compute :
   $(\sum_{h_r} \dot{M}_{h_r} c_{p_{h_r}} - \sum_{c_r} \dot{M}_{c_r} c_{p_{c_r}})\Delta T_r$
   $h_r \in \{\text{hot stream segments in interval } r\}$
   $c_r \in \{\text{cold stream segments in interval } r\}$.
6. Assume $R_{n_r+1} = 0$
7. From eq. 9, compute R_r sucessively for $r = n_r, ..., 1$
8. Compute $R_{min} = min_{r=1,...,n_r+1}(R_r)$
9. Set $R_{n_r+1} = -R_{min}$
10. Recompute $R_r = R_r + R_{n_r+1}$ for $r = 1, ..., n_r$ using eq. 9

An alternative set of equations (eq. 11) may be used to compute the heat cascade. This formulation has the advantage of involving only one R_r per equation, each of the equations being related to the initial temperature of one of the streams.

$$\dot{Q}^+ = \max_s(0, R_s), s \in \{\text{hot and cold stream segments}\} \tag{11}$$

$$\begin{aligned}\text{with } R_s = &\sum_c \dot{M}_c c_{p_c}(max(T^*_s, T^*_{c,target}) - max(T^*_s, T^*_{c,in})) \\ &- \sum_h \dot{M}_h c_{p_h}(max(T^*_s, T^*_{h,in}) - max(T^*_s, T^*_{h,target}))\end{aligned}$$

$h \in \{\text{hot stream segments}\} \quad c \in \{\text{cold stream segments}\}$

3.5 The grand composite curve

The **grand composite curve** (Figure 6) is the plot of the heat cascaded as a function of the corrected temperatures $((R_r, T^*_r), \forall r = 1, ..., n_r + 1)$. Since heat can be supplied and removed from the process at any temperature provided that the heat cascade remains feasible (i.e. $R_r \geq 0$), the grand composite defines the enthalpy-temperature profile of the energy requirement of the process. In the grand composite curve, the pinch point is located at the intersection between the curve and the temperature axis, identifying the activation of the $R_r \geq 0$ constraint.

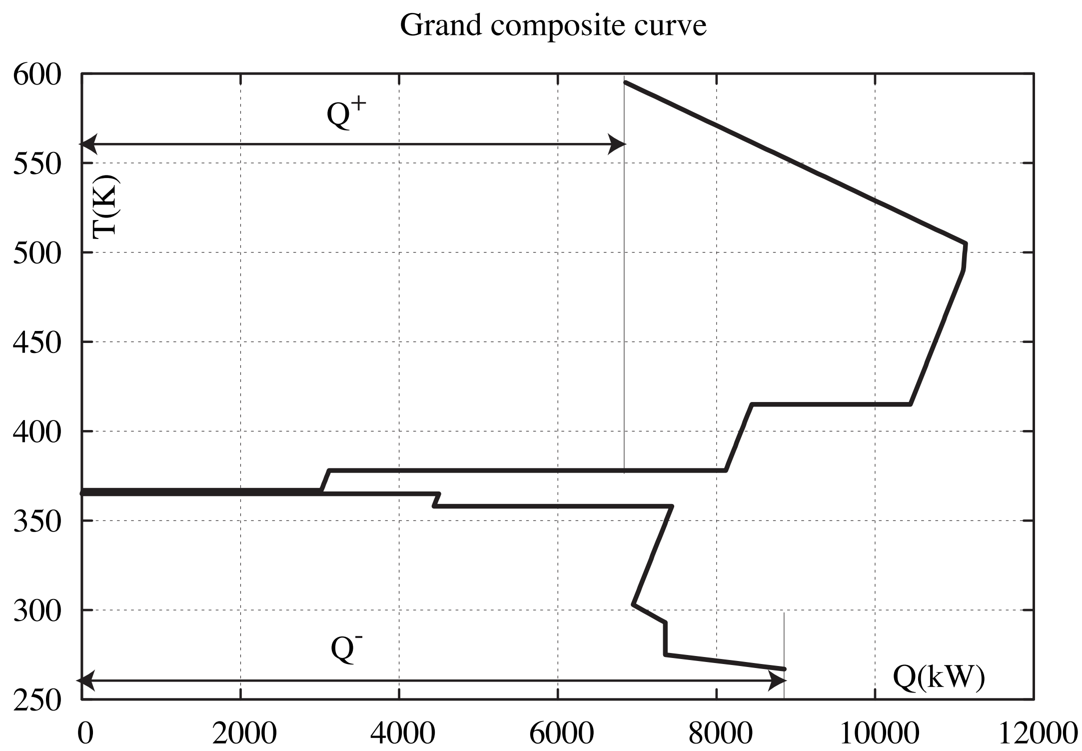


Figure 6: Grand composite curves of the process

3.6 Example

To illustrate the problem table method and the grand composite curve construction, we will solve the following example:

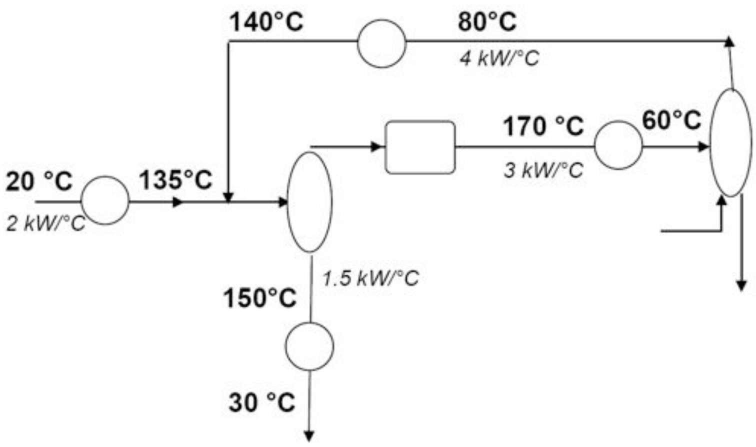


Figure 7: Example for the problem table method

$°N$	T_{in} [°C]	T_{out} [°C]	$\dot{M}c_p$ [kW/K]	$\dot{Q}$ [kW]
1	20	135	2	230
2	170	60	3	-330
3	80	140	4	240
4	150	30	1.5	-180

Table 1: Table of exchanges

Choosing a $\Delta T_{min}/2$ of 5 [°C], we obtain the following results (figures 8 and 9):

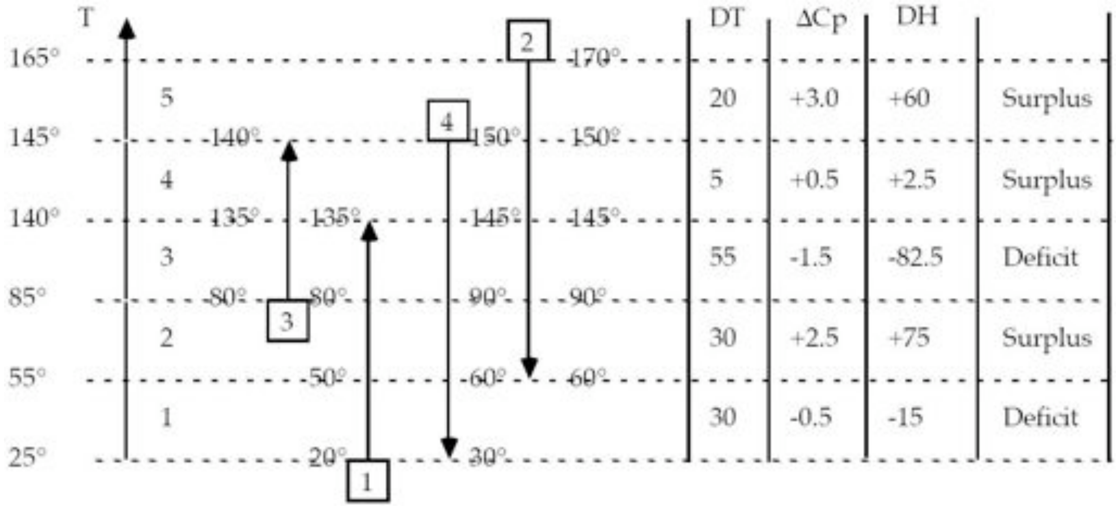


Figure 8: Details on the calculation

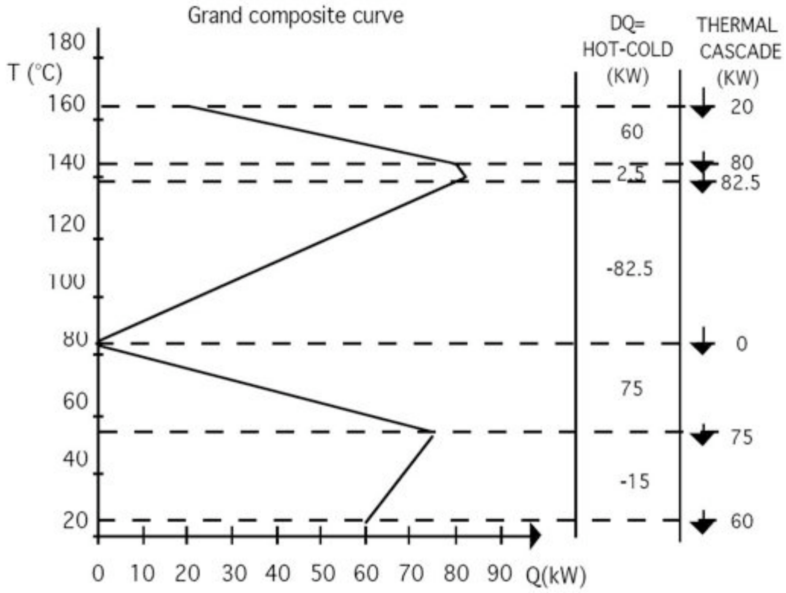


Figure 9: Grand composite curve and the thermal cascade

The pinch point temperature is 85 [°C]: 80 [°C] for the cold streams and 90 [°C] for the hot ones.

The minimum energy requirements are:
Hot utility: 20 [kW]
Cold utility: 60 [kW]

4 Consequences of the pinch point location

4.1 Heat sink and heat source

From the constraints of eq. 11, when $R_r = 0$, the heat of the hot streams above the corrected temperature T_r^* has imperatively to be used to preheat the cold streams above the same corrected temperature, the energy balance being obtained by importing the MER in the system. Under a given ΔT_{min} assumption, the pinch point divides therefore the system into two independent sub-systems: a **heat sink** above the pinch temperature and a **heat source** below (figure 6) between which no heat exchange will occur.

4.2 The more in, the more out

One important principle results from this analysis: by comparing the MER obtained for one given ΔT_{min} and the present energy consumption, we are able to quantify the possible energy savings by heat recovery with a holistic vision of heat exchange in the system. In addition, considering that the heat load of the hot and cold streams is constant, each unit of energy supplied to the system by a hot utility in addition to the MER will, by energy balance, correspond to the same additional heat load to be evacuated by the cold utility. This is known as *"the more in - the more out"* principle. This means also that any hot utility saving will result in the same energy saving for the cold utility.

> *With respect to the MER, any additional unit of heat that is added to the system by a hot utility is in reality bought to heat up the cold utility, i.e. "to heat up the environment".*

This principle is also important for checking the calculations since the difference between the hot and the cold utility MER should be equal to the difference between the present hot and cold utility consumption (before pinch analysis).

4.3 Penalising heat exchangers

For the pinch point location, important heat exchange system diagnosis rules will allow the identification of the penalising heat exchangers that are responsible of the energy penalty.

4.3.1 Exchangers using hot utility below the pinch point

Below the pinch point, the process is a heat source : i.e. the heat available in the hot streams is greater than the heat of the cold streams and the corrected temperatures of the hot streams are always greater or equal to the one of the cold streams. The system being a heat source there is no need to supply from a hot utility in this sub-system. If a hot utility is used below the pinch point, it will be added to the hot streams and therefore it will increase the cold utility requirement. This is an application of *"the more in - the more out"* principle : the hot utility used below the pinch point is purchased for heating the environment.

4.3.2 Exchangers using cold utility above the pinch point

Similarly, above the pinch point the system has a heat deficit. The heat from the hot streams should be used to heat up the cold streams. If a cold utility is usedt to cool down the hot streams, less heat will be available for the cold streams and therefore additional heat will be required to balance the heat requirement of the cold streams. From *"the more in - the more-out"* principle, the cold utility used above the pinch point removes the additional heat supplied by the hot utility.

4.3.3 Exchangers that do exchange heat across the pinch point

As the two sub-systems defined by the pinch point are independent, heat can not be cascaded across the pinch point. When a hot stream above the pinch point exchanges heat with a cold stream below the pinch point, its heat load is not anymore available for the cold streams above the pinch point. The missing heat load above the pinch point will have to be compensated by an additional heat supplied above the pinch point by a hot utility, and, by balance, the heat transferred to the cold stream through the pinch point will not allow the cold streams to cool down the hot streams leading to the corresponding increase of the cold utility. By applying *"the more in - the more out"* principle, the extra cold utility requirement will be equal to the extra hot utility requirement both equal to the heat load transferred across the pinch point by the penalising heat exchanger.

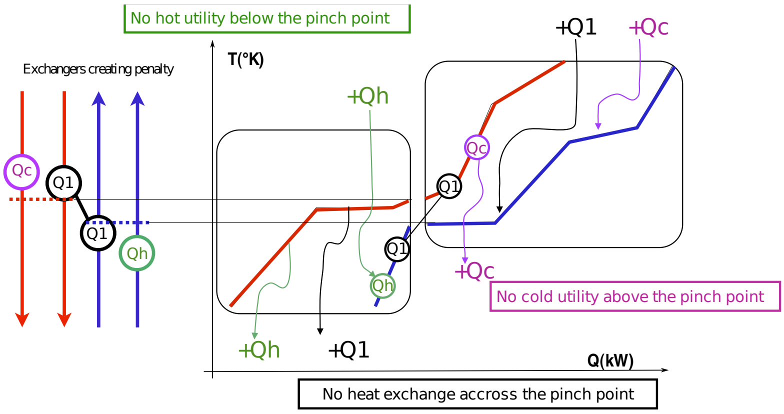


Figure 10: penalising heat exchangers

5 Utility integration

The calculation of the MER assumes that the hot and cold utilities will have the appropriate temperatures to supply the energy to the process. In reality, the energy requirement is satisfied by converting energy resources into useful energy for the heat exchange requirements of the process. The streams resulting from the conversion of the resources will therefore be added to the hot and cold streams of the system and their enthalpy-temperature profiles have to be considered for the calculation of the heat cascade. In contrast to the process streams that have constant temperature and flowrates, the utility system streams have more degrees of freedom. Their temperatures, pressures and flowrates will be chosen in order to minimise the cost of satisfying the process energy requirement.

The grand composite curve (Figure 6) gives information on the temperatures and flows conditions to be met by the utility system. Considering the heat cascade formulation, the grand composite curve represents the value of the pinch constraints. Any combination of utility streams that will maintain the $R_r \geq 0$ constraints will be feasible. The selection will however seek the utility cost minimisation. The latter being essentially proportional to the flowrate as defined by eq. 12.

$$CU = time_{year} \sum_{u=1}^{n_u} \dot{M}_u cu_u \tag{12}$$

With

cu_u	$[MU/kg]$	the cost of the utility stream u
$\dot{M}_u$	$[kg/s]$	the flowrate of utility u

Considering the hot utility requirement above the pinch point, the grand composite curve presents the process as being a cold stream to be heated up by the hot utility. A cold stream should have an increasing enthalpy content with temperature. Therefore the so-called self sufficient zones as identified on figure 11 will be ignored. In such zones, the heat delivered by the hot streams balances the requirements of the cold ones. The goal of the hot utility integration will be to define a hot stream (or a set of hot streams) whose enthalpy-corrected temperature diagram (hot composite curve) will always be above the grand composite curve. Similarly, below the pinch,

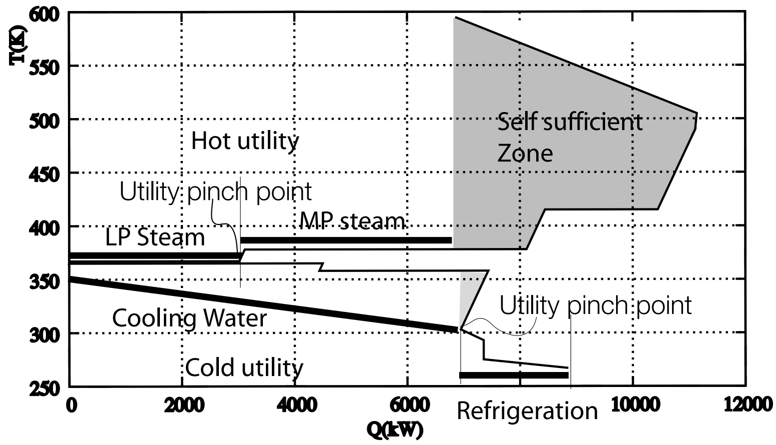


Figure 11: grand composite curve for the definition of the utility system

the grand composite curve of the process will appear as a hot stream to be cooled down by the cold utility. At this stage, it will be necessary to distinguish between the cold utility requirement above the ambient temperature that may be realised by heat exchange with the ambient streams (air or cooling water) from the requirement below the ambient temperature that will require refrigeration. The example in figure 11 shows the integration of a hot utility system made of two steam flows one at low pressure (LP Steam) and one at medium pressure (MP Steam). The flow of the cheapest stream (LP Steam) has been maximised by activating a heat cascade constraint that defines a **utility pinch point**. Below the pinch point, the flow of the cooling water has been minimised by increasing its outlet temperature, Below the ambient temperature, the refrigeration requirement is satisfied by a fluid that evaporates with a constant temperature. The operating cost of this utility stream is defined by the mechanical power consumption of the refrigeration compressor. To be rigourous, the heat load of the hot stream of the refrigeration cycle should be integrated with an additional flow of the cooling water.

The selection of the utility streams is of major importance since it defines the complete list of streams to be considered when designing the heat exchanger network. In practice, thermochemical models of the energy conversion units will be used to determine the operating temperatures and compositions allowing the definition of the hot and cold streams of the utility sub-system. The flowrates will be determined by optimisation in order to minimise the cost of the energy delivered. The constraints of the heat cascade will be considered in the problem and the solution will be characterised by a list of pinch points. One of these being the process pinch point, representing the maximum energy recovery between the process streams , the others corresponding to the maximum use of the cheapest utility. If, in the simplest cases, the calculation of the utility streams may be done graphically, it is more convenient to use optimisation techniques to solve the problem, especially when cycles like steam network integration is considered. This will be discussed in more detail in the next chapter. In conclusion :

The grand composite curve defines the enthalpy temperature profile to which the utility system has to be integrated. Above the pinch point, it defines a cold stream to be heated up by a hot utility, below the pinch point and above the ambient temperature the process is a hot stream to be cooled down by a cold utility. Below the ambient temperature, the grand composite defines a hot stream to be cooled down by a refrigeration system.

6 Targeting the Investment

As in the case of a single heat exchanger, the ΔT_{min} assumption has to be tested against the energy-capital trade-off. In order to confirm the validity of the ΔT_{min} value, it is necessary to estimate the investment of the heat exchanger network (HEN) that will realise the targeted heat recovery.

The installed cost of the heat exchanger network I_{HEN} (eq. 20) is obtained considering the area of the heat exchangers and summing up their installed cost. This calculation obviously depends on the value of the ΔT_{min}. However as at this stage, we do not know the heat exchanger network structure but only the expected heat recovery. There is a need to estimate the HEN cost without having to design it. The HEN cost will be estimated by calculating the minimum number of heat exchangers required to realise the MER and estimating the heat exchangers area.

6.1 The minimum number of connections target

Considering the list of hot and cold streams and assuming that each time a heat exchanger is placed it will satisfy the heat load of one of the two streams involved, the Euler graph theory defines the minimum number of connections by eq. 13.

$$N_{ex} = N_s + N_{u\_s} + L - S \tag{13}$$

With:

N_{ex}	the number of connections
N_s	the number of process streams involved in the heat recovery
N_{u_s}	the number of utility streams used to close the energy balance of the system
L	the number of loops
S	the number of independent systems, i.e. systems where the heat load of one connection completes the simultaneously the requirements of the two streams involved

Usually $S = 1$ because the overall heat balance is verified, and $L = 0$ as the goal is to calculate the minimum number of connections. Considering that the pinch point separates the system into two separated subsystems and that the change in the slope of the composite curves will not allow heat exchanges across the pinch point (except if the streams have equal c_p), the formula (eq. 13) has to be applied separately above and below the pinch point. As a consequence, the minimum number of connections in the MER conditions is given by eq. 14, where the N_p streams that cross the pinch will be accounted twice.

$$N_{ex_{min,mer}} = (N_{s\_a} - 1) + (N_{s\_b} - 1) = (N_s + N_{u\_s} - 1) + (N_p - 1) \tag{14}$$

With:

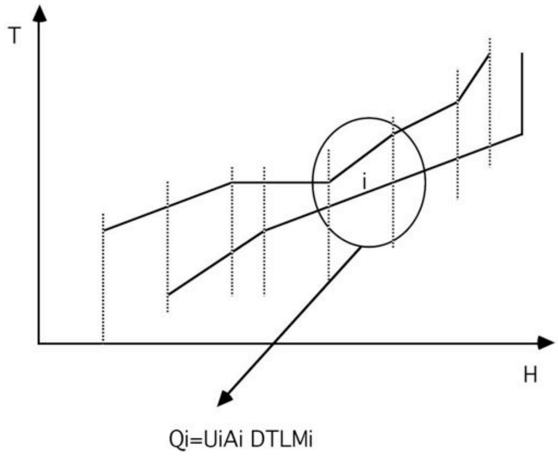


Figure 12: Cumulated area calculation

$N_{ex_{min,mer}}$	the minimum number of exchangers in the MER conditions
N_{s_a}	the number of streams above the pinch point including the utility streams.
N_{s_b}	the number of streams below the pinch point including the utility streams.

6.2 Total Area Target

Prior to any design, it is possible to obtain an estimation of the cumulated area required to achieve the MER.

The analogy with the two streams exchange gives an easy way to calculate the total area corresponding to an optimal utilisation of the heat transfer potential between the two composite curves. If the hot and cold composite curves are considered as hot and cold streams whose enthalpy-temperature profiles are defined by linear segments and assuming a global heat transfer coefficient, the composite curve of figure 12, can be divided into successive vertical section of constant c_p for which the area may be computed by the ΔT_{lm} formulation (eq. 2). When the overall heat transfer coefficient is considered as identical for all the streams, the heat exchange area is calculated by eq. 15. The experience shows that MER heat exchanger networks obtained with the method described in ¤9 usually feature a cumulated are ($A_{tot,mer}$) near the minimum calculated value.

$$A_{tot,mer} = \sum_{i=1}^{nb_{vert\,section}} A_i = \sum_{i=1}^{nb_{vert\,section}} \frac{\dot{Q}_i}{U_i \Delta T_{lm\,i}} \tag{15}$$

This approach can be adapted to account for different heat transfer convective coefficients. When the hot stream h exchanges heat with the cold stream c in the vertical section i, the area of the vertical heat exchange section i ($A_{h,c,i}$) is computed by eq. 16, neglecting the conduction resistance. This is feasible because the temperatures considered in the calculation are those of the composite curves.

$$A_{h,c,i} = \frac{\dot{Q}_{h,c,i}}{U_{h,c,i}\Delta T_{lm i}} = \left(\frac{1}{\alpha_{i,h}} + \frac{1}{\alpha_{i,c}}\right)\frac{\dot{Q}_{h,c}}{\Delta T_{lm i}} \tag{16}$$

Considering all the streams j and their respective heat transfer film coefficient $\alpha_{i,j}$ in the vertical heat exchange section i, and considering that the heat load of the vertical section ($\dot{Q}_i$) equals $\sum_{h=1}^{n_{hot\_streams}} \dot{Q}_{h,i}$ and also $\sum_{c=1}^{n_{cold\_streams}} \dot{Q}_{c,i}$, the total area of the vertical section i (A_i) is obtained by summing the contributions of the streams to the vertical exchange i (eq. 17), the total area being then obtained by summing up the vertical sections.

$$A_{tot,mer} = \sum_{i=1}^{nb_{vert\,sections}} A_i = \sum_{i=1}^{nb_{vert\,sections}} \left(\frac{\dot{Q}_i}{\Delta T_{lm i}}\left(\sum_{j=1}^{n_{streams_i}} \frac{1}{\alpha_{i,j}}\right)\right) \tag{17}$$

6.3 Capital cost estimation

Knowing the minimum number of exchangers target $N_{ex_{min,mer}}$ and the overall minimum cumulated area $A_{tot,mer}$, it is possible to estimate the heat exchanger network investment if the repartition of the area between the heat exchangers is known. Assuming that the total area will be distributed equally among the heat exchangers, we compute a mean area for the heat exchangers by eq. 18.

$$A_{mean}(\Delta T_{min}) = \frac{A_{tot,mer}(\Delta T_{min})}{N_{ex_{min,mer}}(\Delta T_{min})} \tag{18}$$

$$A_{mean}(\Delta T_{min}) = \frac{\sum_{i=1}^{nb_{vert\,sections}(\Delta T_{min})} \frac{\dot{Q}_i(\Delta T_{min})}{\Delta T_{lm i}(\Delta T_{min})}\left(\sum_{j=1}^{N_{streams_i}} \frac{1}{\alpha_{i,j}}\right)}{(N_s + N_{s_u}(\Delta T_{min}) - 1) + (N_p(\Delta T_{min}) - 1)} \tag{19}$$

Using the investment cost estimation for the heat exchangers (eq. 4), the capital cost of the heat exchanger network may be estimated by eq. 20.

Considering the property of the powered function with an exponent lower than 1, this function overestimates the real investment of a heat exchanger network with the same total area and the same number of exchangers.

$$I^*_{HEN}(\Delta T_{min}) = \sum_{ex=1}^{N_{ex}} I(A_{ex}(\Delta T_{min})) \approx N_{ex_{min,mer}} a_{ex} \left(\frac{A_{mean}(\Delta T_{min})}{N_{ex_{min,mer}}(\Delta T_{min})}\right)^{b_{ex}} \tag{20}$$

6.3.1 Overestimation illustration

In the case of two exchangers in figure 13, if the total area is At, each heat exchanger will feature an area of A1 = At/2 for the cost estimation and the cost estimation will be C* = 2C1. On the other hand if the exchangers feature respectively A2 and A3 so as A2 + A3 = At and A2 ≠ A3, the real cost C = C2 + C3 is smaller than C*. C* is the maximum value of CI. The minimum bound of the cost correspond to A1 or A2 is negligible with regards to the other.

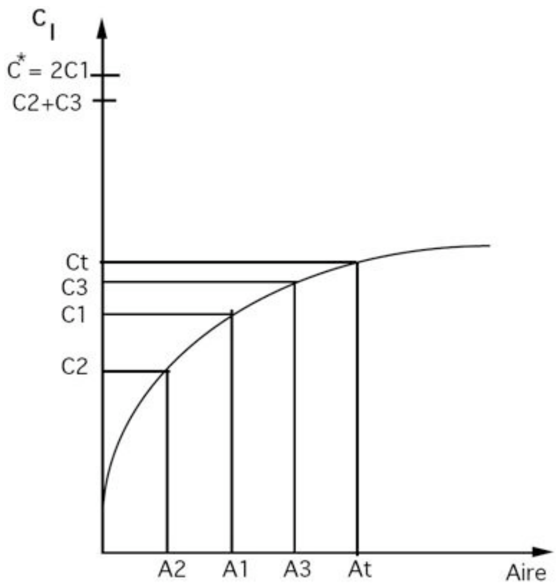


Figure 13: Capital cost overestimation C*

6.4 Optimal ΔT_{min} value

Having an estimation of the investment and the energy cost, it is possible to calculate the evolution of the total cost ($Ct(\Delta T_{min})$ defined by eq. 21 as a function of the ΔT_{min} value. Graphically, this corresponds to moving horizontally the cold composite curve while keeping the hot composite curve unchanged (figure 14).

$$Ct(\Delta T_{min}) = (\frac{i(1+i)^{ny_{ex}}}{1+i)^{(ny_{ex}} - 1})I^*_{HEN}(\Delta T_{min}) + \left(c^+\dot{Q}^+(\Delta T_{min}) + c^-\dot{Q}^-(\Delta T_{min})\right) time_{year} \tag{21}$$

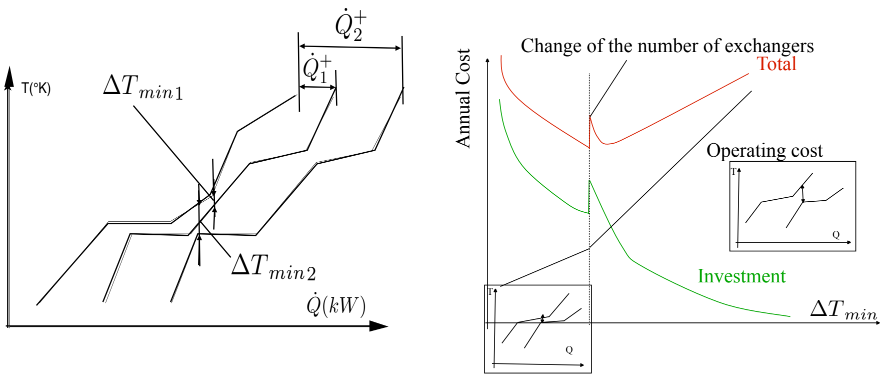


Figure 14: Total Cost ($Ct(\Delta T_{min})$), annual operating cost ($CU(\Delta T_{min})$) and annual investment cost ($C_I(\Delta T_{min})$) as a function of ΔT_{min}

The most important consideration is that discontinuities in the curve can occur due to pinch point changes. Considering that the pinch point location defines the list of the streams in the different sub-systems that will be used for the heat exchanger network design, the identification of the optimum ΔT_{min} value is therefore necessary before starting any heat exchanger network design.

As energy and investment costs, as well as investment strategies vary from one plant to another, the ΔT_{min} value will be specific to the plant under study and the results can not be translated to similar plants with other production or in other locations without a ΔT_{min} study. The important points are:

- Pinch point changes are related to the fact that infeasible exchanges becomes feasible and vice-versa. The examination of these exchangers allows to judge if the new value can be accepted or not.
- The optimal value does not have to be determined with precision. Any value between two pinch point changes that frame the optimum value might be adopted since non linear optimization of the heat exchanger network will be performed once the HEN structure is determined.
- A wrong ΔT_{min} value might lead to sub optimal solutions in the heat exchanger network synthesis because the streams concerned with the pinch point will change from one pinch point to another.

6.5 Physical meaning of the ΔT_{min}

The ΔT_{min} value is an "experience value" chosen by the engineer. It is directly linked to the area of the heat exchanger through the ΔT_{lm} but also through the heat transfer coefficient.

At constant area and fixed inlet temperatures, the increase of the heat transfer coefficient decreases the approach temperature in the heat exchanger. When neglecting the thermal conductivity of the heat exchanger, the overall heat transfer coefficient is the sum of stream dependent contributions (eq. 2). Similarly, the ΔT_{min} in an exchanger is defined as a sum of stream dependent contributions $\Delta T_{min}/2_h$ and $\Delta T_{min}/2_c$, which will be respectively proportional to $\frac{1}{\alpha_h}$ or $\frac{1}{\alpha_c}$. With this definition the corrected temperatures will be calculated by eqs. 22 and 23.

$$T_h^* = T_h - \Delta T_{min}/2_h \quad \forall h \in \{hot\,streams\} \tag{22}$$

$$T_c^* = T_c + \Delta T_{min}/2_c \quad \forall c \in \{cold\,streams\} \tag{23}$$

With this definition, consider two hot streams at the pinch point. The first is a gas stream with a $\Delta T_{min}/2 = 10$°C. The second is a liquid stream with a $\Delta T_{min}/2 = 5$°C. Suppose they exchange with a cold stream with $\Delta T_{min}/2 = 5$°C. At the pinch point the corrected temperatures are equal. But when calculating the real temperature values, the ΔT_{min} will be $5 + 10 = 15$°Cfor the heat exchange with the gas stream and $5 + 5 = 10$°Cfor the exchange with the liquid stream.

Heuristic rule (eq. 24) can be used to estimate the value of $\Delta T_{min}/2$ of one stream when knowing its convective heat transfer coefficient.

$$\Delta T_{min}/2_j = K \frac{1}{(\alpha_j)^{b_{ex}}} \tag{24}$$

Type	Heat transfer coefficient $W/m^2/C$	$\Delta T_{min}/2$
Gas stream	60	15
Liquid stream	560	5
Condensing stream	1600	3
Vaporizing stream	3600	2

Table 2: Typical values for the $\Delta T_{min}/2$ as a function of the heat transfer film coefficient

Varying the K coefficient will allow one to compute the Energy-Capital trade-off. Values in table 2 refer to the calculation of the typical heat transfer coefficients.

7 Summary of the targeting method

The former section defines a method for targeting the MER of a process, identifying the possible heat recovery between the hot and cold streams based on an optimal value of the ΔT_{min} and defining the temperature conditions and the flows in the utility system. By comparing with the present energy consumption in the process, this method gives a holistic view of the possible energy savings by heat exchange and allows one to identify the faulty heat exchangers in the system. The method is summarised as follows :

Step 1: Composite curves and pinch point location Assuming $\Delta T_{min}/2_j$ values for the process streams, the calculation of the composite curves defines the pinch point location, the maximum heat recovery between the hot and the cold streams of the process and its Minimum Energy Requirements (MER).

Step 2: Optimization of the $\Delta T_{min}/2$ contributions An estimation of the capital cost based on heat transfer film coefficient is used to analyse the influence of the $\Delta T_{min}/2$ values.

Step 3: Utility and thermodynamic cycles selection The resulting grand composite curve of the process is used to define the utility streams to be considered to supply the MER. The flowrates of the utility streams are defined in order to maximise the use of the cheapest utility streams by activating utility pinch point.

Step 4 : Update the value of the $\Delta T_{min}/2$ considering the utility streams An analysis of the influence of the $\Delta T_{min}/2$ values should be done to confirm its optimal value. In this analysis, it should be reminded that the temperature conditions of the utility streams in fact depend on the values of the $\Delta T_{min}/2$ which makes the problem more complex.

At the end of the targeting phase, we obtain the following results.

1. The maximum energy recovery by heat exchange between hot and cold streams and the corresponding possible energy saving.
2. The identification of the pinch point and the definition of the heat source and sink.
3. The optimal flows of the utility streams.
4. The pinch point location with the penalising heat exchangers :
   (a) Heat exchangers that exchange heat across the pinch
   (b) Hot utility streams that heat up streams below the pinch
   (c) Cold utility streams that cools down hot streams above the pinch
5. The complete list of streams to be considered in the heat exchanger network design phase.

8 Heat Exchanger Network (HEN) design

Starting from the definition of the complete list of streams to be considered, the heat exchanger network design task can be completed by answering three questions:

"Which hot and cold streams should exchange heat ?"
"What is the heat load corresponding to this exchange?"
"How are the heat exchangers interconnected?"
(What are input and output temperatures? Is there a split? Etc...)

Such a problem is by nature combinatorial with a huge number of possible matches. Two types of approaches may be used to solve the heat exchanger network design problem.

1. The pinch design method is a sequential method mainly based on heuristic and feasibility rules deduced from the properties of the pinch point definition.
2. The mathematical programming methods state the HEN design problem as a Mixed Integer (Non)Linear programming problem, solving the combinatorial nature of the problem using specific optimisation algorithms.

Both approaches have advantages and disadvantages, it is therefore advantageous to combine them.

The heat exchanger network (HEN) obtained has to be "optimal". For the engineers, this optimality is not only the minimum total cost of the system. It has also to fulfil a set of criteria specific to the plant under study. These include the flexibility, reliability, safety, layout, starting, operability, retrofit (i.e. existing units) and technological constraints. The HEN design method should therefore be flexible enough to incorporate such criteria.

8.1 Representing a heat exchanger network

The grid representation is a convenient graphical representation of a HEN structure. Both vertical and horizontal representations may be used. In the vertical grid representation, the streams of matter participating in the heat exchanger network are represented as vertical lines

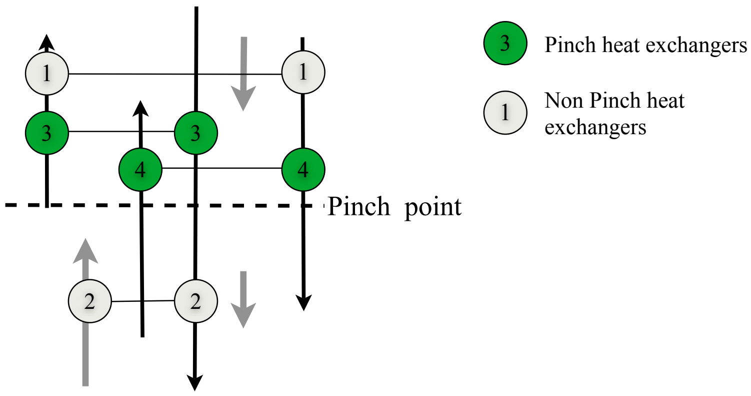


Figure 15: HEN grid representation

from lower to upper temperatures (figure 15). Hot streams go top-down while the cold stream go bottom-up. In this representation, a heat exchanger is a horizontal line linking to vertical lines (streams). The counter-current heat exchanger is clearly identified, the approach temperature and the ΔT_{lm} are easy to calculate even by hand.

In the following, we will name "*pinch stream*" a stream whose temperatures follow eq. 25 :

$$T^*_{low} \leq T_{pinch} \leq T^*_{high} \tag{25}$$

With T^*_{high} the highest temperature of the stream (hot end for a cold stream, inlet condition for a hot stream) in the corrected temperature domain, T^*_{low} is the lowest temperature of the stream in the corrected temperature domain and T_{pinch} is the pinch point temperature.

A *pinch heat exchanger* is an exchanger that involves **two** pinch streams **at the pinch**. The exchangers 3 and 4 of figure 15 are pinch heat exchangers, while exchangers 1 and 2 are not.

8.2 The HEN design target

Let us first consider a problem where hot and cold utility streams have respectively the highest and the lowest temperatures of the system. From the heat cascade, we learn that the pinch point distributes the streams in two different sub-systems between which no heat can be exchanged. For these sub-systems, the following target for the heat exchanger network design task can be stated:

- *Above the pinch point*
  The system **above the pinch point** is a heat sink, therefore the heat of the hot streams has to be recovered to heat up the cold streams above the pinch point. **The hot streams are therefore the Key streams**
  The **hot streams** have to be cooled down to their target temperature only by heat exchange with the cold streams. In particular, the pinch hot streams have to be cooled down to their pinch temperature (pinch temperature corrected by $\Delta T_{min}/2$) by heat exchange with the pinch cold streams.
  As the pinch point defines a place where the approach temperature between the streams is known, the HEN design start from the pinch point tempertaure and goes to the higher temperatures.

- *Below the pinch point but above the ambient temperature*
  The system **below the pinch point and above the ambient temperature** is a heat source, therefore the heat requirement of the cold streams has to be recovered from the hot streams to heat the cold streams below the pinch temperature. **The cold streams are therefore the Key streams**
  The **cold streams** have to be heated up to their target temperature only by heat exchange with the hot streams. In particular, the pinch cold streams have to be heated up to their pinch temperature by heat exchange with the pinch hot streams.
  In this sub-systems, the energy balance of the **hot streams** is realised by the use of ambient cold utility.
  Below the pinch point, the HEN design starts from the pinch point and goes to the lower temperatures.

- *Below the ambient temperature*
  The system below the ambient temperature is also a heat source, the **key streams** are therefore the **cold streams** that have to be heated up to their target temperature using the hot streams in the corresponding sub-system.

9 The Pinch Design Method

The pinch design method is a heuristic method for design heat exchanger network. It is based on a set of feasibility and heuristic rules deduced from the analysis of the pinch point definition.

9.1 Feasibility rules

The feasibility rules concern only pinch point exchangers. They are obtained from the definition of the pinch point, knowing the value of of the approach temperature at the pinch point. The rules will be presented for the sub-system above the pinch point where the key streams are the hot streams, by analogy the rules will be translated for the streams below the pinch point.

9.1.1 Number of streams rule

Above the pinch point, the hot pinch streams have to be cooled down to the pinch temperature only with the help of cold pinch streams. Therefore, the number of cold pinch streams above the pinch point $N_{pinch,c}$ must be greater or equal to the number of hot streams $N_{pinch,h}$. When

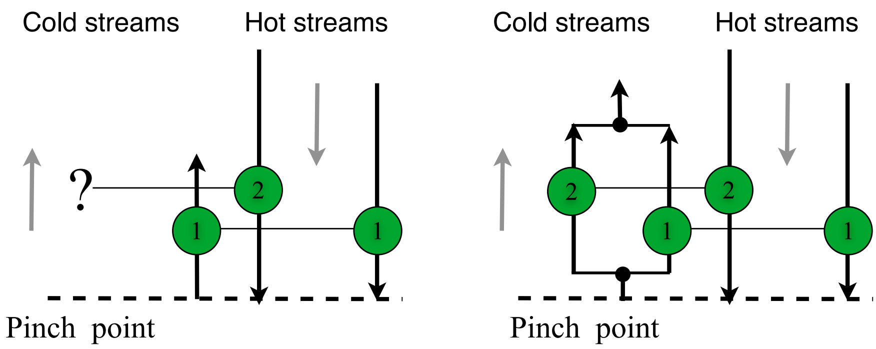


Figure 16: Number of streams rules above the pinch point and the cold stream splitting

$N_{pinch,c} < N_{pinch,h}$, cold streams have to be split as shown on figure 16. In this case, the heat exchange is said to be performed in parallel.

9.1.2 The c_p rule

In a pinch heat exchanger, the approach temperature at the pinch point is equal to the ΔT_{min} value (Figure 17). Above the pinch point, the ΔT_{min} occurs on the cold side of the exchange. An exchange is therefore feasible if the $\dot{M}c_p$ of the cold streams is higher than the one of the hot streams. In other words, the temperature difference at the hot side of exchange must be greater or equal to the ΔT_{min}. (eq. 26
Furthermore, when the heat exchanger is placed, the remaining streams should still satisfy the pinch point, which means that the cumulated $\dot{M}c_p$ of the remaining cold streams should be greater than the cumulated $\dot{M}c_p$ of the hot streams.

$$\begin{aligned} \dot{M}_{i_{hot}} c_{p_{i_{hot}}} &\leq \dot{M}_{j_{cold}} c_{p_{j_{cold}}} \\ \sum_{i=1, i \neq i_{hot}}^{n_{hot}} \dot{M}_i c_{p_i} &\leq \sum_{j=1, j \neq j_{cold}}^{n_{cold}} \dot{M}_j c_{p_j} \end{aligned} \tag{26}$$

When no possible matching verifies the c_p rule, the **hot stream** featuring the too big $\dot{M}c_p$ will be split. In this case, the number of streams rule has to be reexamined.

It is always possible to find a configuration which satisfies the former rules because at the pinch point, the inequality (27) is valid.

Above the pinch point:

$$\sum_{c=1}^{N_c} \dot{M}_c c_{p_c} - \sum_{h=1}^{N_h} \dot{M}_h c_{p_h} \geq 0 \tag{27}$$

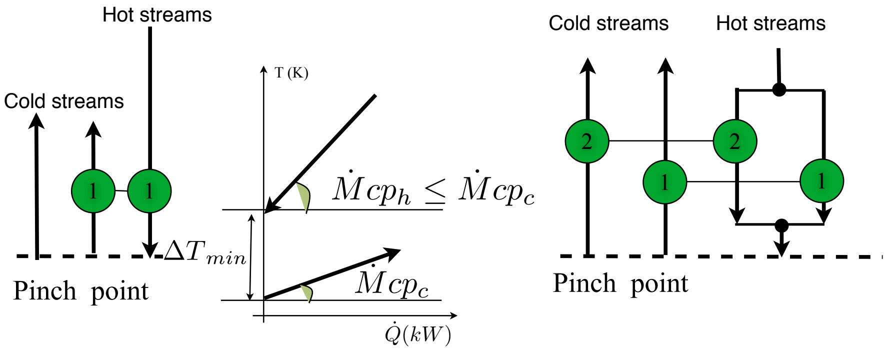


Figure 17: The c_p rule at the pinch point

9.2 Heuristic rules

9.2.1 Tick-off rule

Once a feasible heat exchanger is found, its heat load is defined such that it satisfies the heat requirement of one of the two streams involved. This rule results from the application of the minimum number of connections rule. This rule is known as being the "**tick-off**" rule.

9.2.2 Remaining problem analysis

Once a heat exchanger is placed, the streams involved in the heat exchanger are not anymore available for other heat exchanges. The list of hot and cold streams involved in the heat cascade is therefore updated and a new heat cascade is calculated. This is called the remaining problem analysis. A new minimum energy requirement ($\dot{Q}^+_{remaining}$) is defined and possibly a new pinch point is created. When $\dot{Q}^+_{remaining} = \dot{Q}^+$ the heat exchanger is well placed according to the MER target. In this case, the remaining problem heat cascade redefines the c_p rules for the next heat exchanger placement. Otherwise, the heat exchanger creates a penalty and one should decide to accept or not this penalty.

9.2.3 Driving force plot and splitting factors

When stream splitting is required for the heat exchanger placement, the splitting factor is chosen according to the streams that are connected. One may choose the $\dot{M}c_p$ ratio of the streams involved, another possibility is the calculation of the splitting factor to reach isothermal mixing. In this case, the splitting factor would be a function of the heat load and the $\dot{M}c_p$.

The driving force plot (18) is the plot of the temperature difference between the hot and the cold composite as a function of the temperature. The driving force plot is plotted as a function of the hot and the cold stream temperatures. Placing the temperatures of the heat exchanger against the driving force plot gives a good indication of the quality of the heat exchanger placement.

9.2.4 Other heuristics

- When trying to identify the streams to be connected, a lot of possibilities remain. It is recommended to start the design procedure by considering first the cold and hot streams with the highest $\dot{M}c_p$.

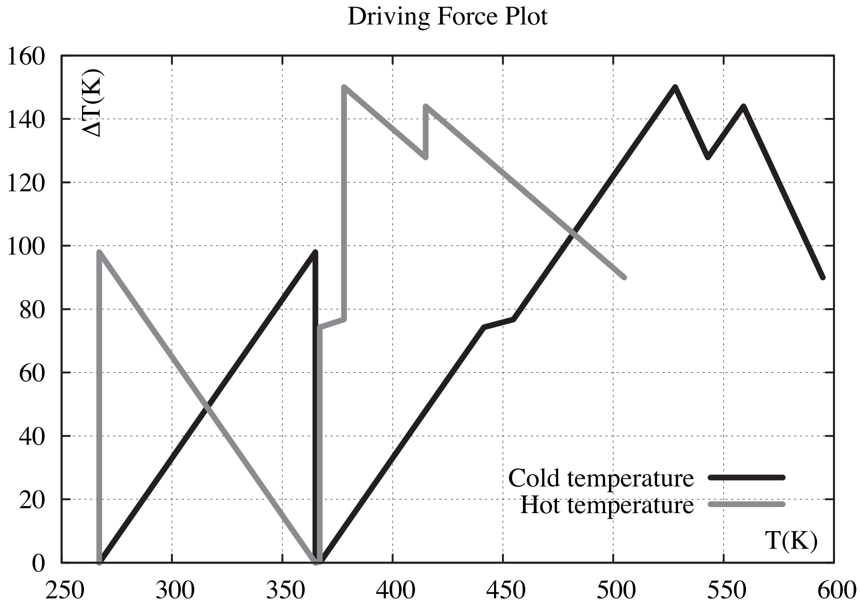


Figure 18: Driving Force plots

- The utility streams are usually at the end of the temperature range and have flowrates that can be manipulated, it is therefore recommended to place heat exchangers with the utility streams at the end of the heat exchange in order to control the target temperature of the associated process streams.

- For reason of flexibility, the utility streams are preferably split if they are matched with more than one process streams.

Away from the pinch point, the exchanges are usually easier than at the pinch point. But the remaining problem analysis shows that while the exchangers are added during the synthesis calculation, new pinch points can be created.

9.2.5 A synthesis method.

Using the feasibility and the heuristic rules, the pinch design method is summarised below. It is only a way to approach the problem, not a systematic method to solve it. The know-how and the experience of the engineer will be very useful to complete the synthesis task.

1. Compute the heat cascade and identify the pinch points

2. Analyze separately the sub systems defined by the pinch point(s)
   The key idea is to bring the key {k} streams in the sub-system to their target temperature or to the pinch point.
   Above a pinch point, the key streams are the hot streams
   Below a pinch point, the key streams are the cold streams.
   {k} refers to the key streams, {k-1} to the other ones.

3. Identify the pinch point streams.
4. If the streams number rule is not verified : i.e. $N_{\{k\}} \leq N_{\{k-1\}}$
split the {k-1} streams until the rule is verified and go back to step 2
5. Find a connection $(i_{\{k\}}, j_{\{k-1\}})$ verifying the c_p rules (eq. 28).

$$\begin{aligned} \dot{M}_{i_{\{k\}}} c_{p_{i_{\{k\}}}} &\leq \dot{M}_{j_{\{k-1\}}} c_{p_{j_{\{k-1\}}}} \\ \sum_{i=1, i \neq i_{\{k\}}}^{n_{\{k\}}} \dot{M}_i c_{p_i} &\leq \sum_{j=1, j \neq j_{\{k-1\}}}^{n_{\{k-1\}}} \dot{M}_j c_{p_j} \end{aligned} \tag{28}$$

If no connection is found and the problem is not terminated, split the {k} streams which do not verify the rule, until a connection is found, or suppress an already defined connection and go back to step 1.

6. For the feasible connection, apply the tick-off rule and calculate the exchange.
7. Perform the remaining problem analysis and accept or not the connection.
8. The remaining problem defines a new set of streams. With the new data restart the procedure (step 1) to find the next connection until all the streams requirements are satisfied.

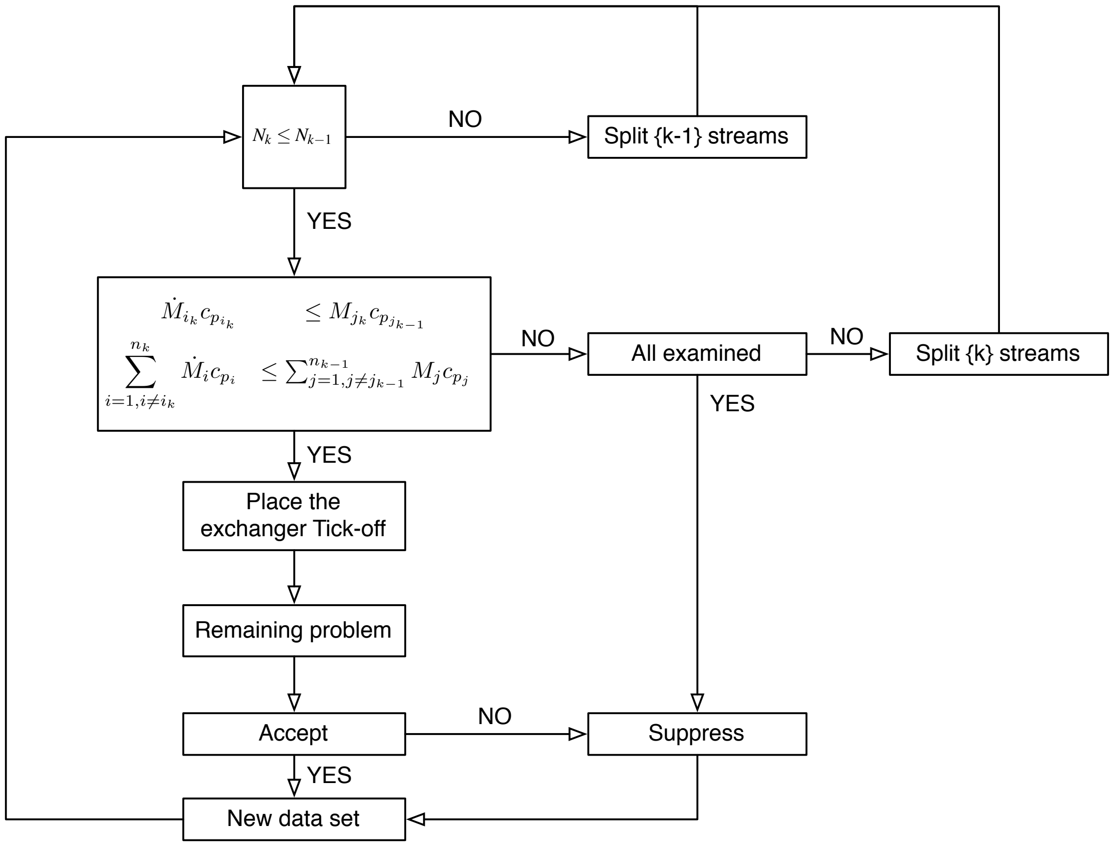


Figure 19: The synthesis method: a simple approach

10 Mathematical programming approach

The pinch design method is a sequential method, that does not guarantee that the sequence of decisions will lead to the best network neither to a network that will have the expected minimum number of heat exchangers. Instead of applying heuristic rules in order to identify the streams to be connected, the choice of the streams to be connected may be formulated as a mixed integer linear programming (MILP) problem : the **Heat Load Distribution** calculation.

10.1 Heat load distribution

The Heat load distribution answers two questions: the determination of the streams involved in a connection and the corresponding heat load with a holistic vision of the overall heat exchanger system. It solves therefore in part the problem of the sequential nature of the pinch design method the pinch method will then be applied on the results of the heat load distribution in order to determine the interconnections between the heat exchangers in the network.

For a sub-system defined by the pinch point, the heat load distribution problem minimises the number of connections in a network by computing the heat exchanged ($\dot{Q}_{hkc}$) between the hot stream h in the temperature interval k of the heat cascade and the cold stream c. The connection between a hot stream h and a cold stream c is represented by an integer variable y_{hc} : if the connection exists $y_{hc} = 1$, if not $y_{hc} = 0$. Assuming that each connection identified by y_{hc} may be realised by one heat exchanger, the solution of the MILP problem (29) gives the heat load ($\dot{Q}_{hc}$) exchanged between hot stream h and cold stream c that minimises the number of exchangers while satisfying the approach temperature (ΔT_{min}) constraint in each heat exchanger and the minimum energy requirement of the system.

$$\min_{y_{hc}\dot{Q}_{hkc}} N_{ex} = \sum_{c=1}^{nc_z}\{\sum_{h=1}^{nh_z} y_{hc}\} \tag{29}$$

Subject to:
Thermal balance of hot stream h in temperature interval k.

$$\sum_{c=1}^{nc_z} \dot{Q}_{hkc} = \dot{Q}_{hk} \qquad h = 1...nh_z \qquad k = k_z...kend_z \tag{30}$$

Thermal balance of the cold stream j.

$$\sum_{h=1}^{nh_z}\{\sum_{k=k_z}^{kend_z} \dot{Q}_{hkc}\} = \dot{Q}_{c,z} \qquad c = 1...nc_z \tag{31}$$

Existence of the connection ij.

$$\sum_{k=k_z}^{kend_z} \dot{Q}_{hkc} - y_{hc}\dot{Q}_{max} \leq 0 \qquad h = 1...nh_z \qquad c = 1...nc_z \tag{32}$$

Feasibility of the connection: the heat load demand of the cold stream c in the intervals k and above must at least be satisfied by the hot streams in the intervals from $kend_z$ to k.

$$\sum_{r=k}^{kend_z} \{\sum_{h=1}^{nh_z} \dot{Q}_{crh}\} - \sum_{r=k}^{kend_z} \dot{Q}_{ck} \geq 0 \qquad c = 1...nc_z$$
$$k = k_z + 1...kend_z \tag{33}$$

Each heat load must be positive.

$$\dot{Q}_{hkc} \geq 0 \qquad h = 1...nh_z$$
$$c = 1...nc_z$$
$$k = k_z...kend_z \tag{34}$$

$$y_{hc} \in \{0, 1\}$$

$$\sum_{k=k_z}^{kend_z} \dot{Q}_{hkc} = \dot{Q}_{hc} \qquad \forall h = 1, ..., nh_z \qquad \forall c = 1, ..., nc_z \tag{35}$$

Where:

- k_z and $kend_z$ are respectively the lower and upper intervals of the calculated sub-system z.
- nc_z, nh_z are respectively the total number of cold and hot streams in the sub-system z including the utility streams.
- $\dot{Q}_{hkc}$ is the heat load exchanged between the hot stream h in the temperature interval k and the cold stream c.
- y_{hc} is an integer variable representing the connection between hot stream i and cold stream c.
- $\dot{Q}_{ik}$ is the heat load of the stream i in interval k.
- $\dot{Q}_{h,z}$, $\dot{Q}_{c,z}$ are respectively the load of hot stream h and cold stream c in the calculated sub-system.
- $\dot{Q}_{max} = \max_{h=1,...,nh_z,c=1,...,nc_z}(\dot{Q}_{h,z}, \dot{Q}c, z)$ is the maximum heat that can be exchanged between stream i and j in the sub-system z.

11 Optimising the heat exchanger network design

11.1 Loops and Path for reducing the number of heat exchangers

Resulting from the application of the methods described above, it might happen that the number of exchangers exceeds the minimum number of connections. The analysis of the heat exchanger network structure may be applied in order to reduce the number of heat exchangers. In a heat exchanger network, a loop is defined as a path through the exchangers and the streams linking a heat exchanger to itself. In the grid representation, a loop follows verticals and horizontals to return back to the starting exchanger (fig. 20, left). An increase or decrease of the heat load of a heat exchanger in a loop is propagated along the loop. In order to maintain the overall energy balance, an increase induces a decrease of the next exchanger in the loop on the same stream. If the decrease corresponds to the smaller heat load of the heat exchangers in the loop, the corresponding heat exchanger is removed (e.g. $\Delta\dot{Q} = \dot{Q}_1$ on figure 20). When a heat exchanger is suppressed, one has to verify the ΔT_{min} constraint by computing the new temperatures. When this constraint is not fulfilled, e.g. when the loop is crossing the pinch point, one of three options may be used :

1. If the heat exchange remains feasible, i.e. the approach temperature of the heat exchangers is positive, one may accept the heat exchanger suppression by comparing the total cost of the new heat exchanger network with the previous one. If the heat exchanger annualised investment of the suppressed heat exchanger is higher than the increase of the cost of the remaining heat exchangers, then the new heat exchanger network structure is accepted. In order to have a certain flexibility in the design of the heat exchanger network, some authors have suggested to use two different minimum approach temperatures : the EMAT (Exchanger Minimum Approach Temperature) which is used for the heat exchanger network design and the HRAT (Heat Recovery Approach Temperature) that is used for computing the Minimum Energy Requirement and the utility streams flow prior the HEN design.

2. When the heat exchanger network is not feasible (i.e. when the approach temperatures in the HEN is lower then the EMAT), it is possible to re-establish the network feasibility by identifying in the HEN a path through the remaining heat exchangers in the loop and that goes from a hot utility to a cold utility (figure 20 right). Along that path, the exchanger heat loads successively increase and decrease when changing from hot to cold streams. The approach temperature is re-established by increasing the hot utility in order to reduce the heat load of the heat recovery exchangers. The heat load of the hot utility ($\dot{Q}_{EMAT}$) is computed to reach the EMAT value in the heat exchangers of the network, and new operating and investment costs are obtained and compared with the one of the initial network.

3. Do not suppress the heat exchanger and leave the HEN unchanged.

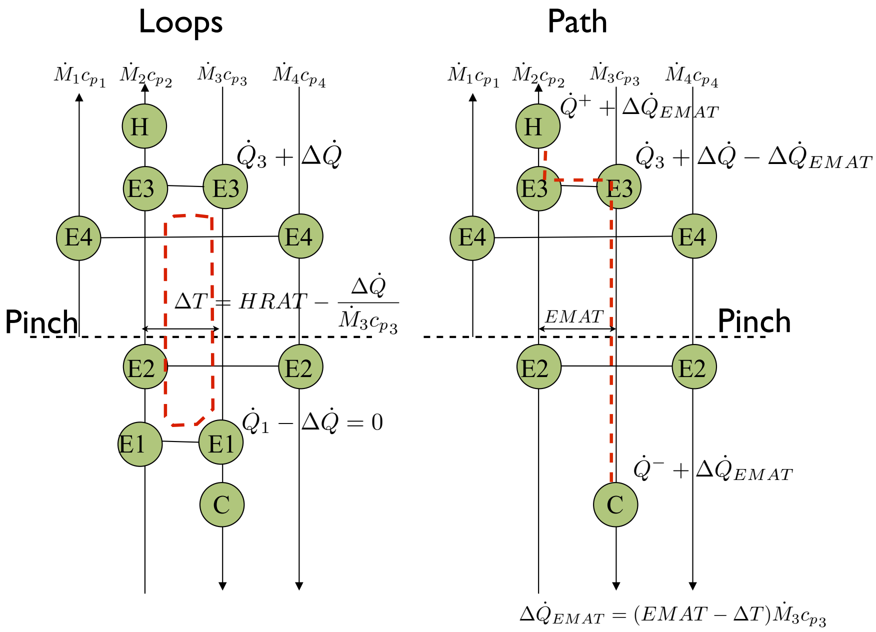


Figure 20: Loops and paths in a heat exchanger network

11.2 Using Mixed Integer Non Linear Programming methods

Once the heat exchanger network structure is known, the heat exchanger network is optimised using non linear programming (NLP) techniques. Because of the non linear nature of the ΔT_{lm}, solving non linear programming problem requires a careful formulation of the problem. The application of the previous steps allows one to obtain a good initial value for the variables. As the optimisation may be concerned with the existence of a heat exchanger, it is necessary to introduce integer variables that will define the existence or not of the heat exchangers and the flowrates in the network. The integer variables allow as well to avoid solutions with too small exchangers or flowrates. In this case the problem becomes a Mixed Integer Non Linear Programming (MINLP) problem (eq. 36). In this optimisation problem, both the area of the heat exchangers, the flows in the splitters and the utility flowrates will be optimised together in order to minimise the overall HEN cost, the HRAT and EMAT constraints being relaxed.

$$Minimise = (\sum_{u=1}^{N_u} c_u^+ \dot{M}_u) time_{year} + (\frac{i(1+i)^{ny_{ex}}}{1+i)^{(ny_{ex}}-1}) \sum_{ex=1}^{N_{ex}} (a_{ex}(A_{ex})^{b_{ex}} \tag{36}$$

Subject to heat exchanger models ($\forall ex = 1, ..., N_{ex}$)

$$\begin{aligned}
&\dot{Q}_{ex} = \dot{M}_{hot}\, c_{p_{hot}}(T_{hot,in} - T_{hot,ex}) = \dot{M}_{cold}\, c_{p_{cold}}(T_{cold,ex} - T_{cold,in}) \\
&\dot{Q}_{ex} = U_{ex} A_{ex} \frac{(T_{hot,in} - T_{cold,ex}) - (T_{hot,ex} - T_{cold,in})}{ln\left(\frac{(T_{hot,in} - T_{cold,ex})}{(T_{hot,ex} - T_{cold,in})}\right)} \qquad (37) \\
&\frac{1}{U_{ex}} = \frac{1}{\alpha_{cold}} + \frac{e}{\lambda} + \frac{1}{\alpha_{hot}}
\end{aligned} \tag{38}$$

Mixers and splitters models

$$H(T_i, \dot{M}_i) = 0 \tag{39}$$

Heat exchanger network specifications

$$T_i = T_{i,spec} \qquad \dot{M}_i = \dot{M}_{i,spec} \tag{40}$$

Inequality constraints

$$\begin{aligned}
\Delta T_{min} &\geq 0 \\
G(T_i, \dot{M}_i) &\geq 0 && (41) \\
y_i \dot{M}_{min} \leq \dot{M}_i &\leq y_i \dot{M}_{max} \quad i = 1, ..., N_{streams} && (42) \\
y_{ex} A_{min} \leq A_{ex} &\leq y_{ex} A_{max} \qquad ex = 1, ..., N_{ex} && (43) \\
y_i, y_{ex} &\in \{0, 1\} && (44)
\end{aligned}$$

with
- y_i an integer representing the existence of stream i
- y_{ex} an integer variable representing the existence of exchanger ex.
- N_u the number of utility streams
- N_{ex} the number of heat exchangers
- $N_{streams}$ the total number of streams including the utility streams

Nowadays, the availability of powerful methods for solving MINLP problems (e.g. outer approx-

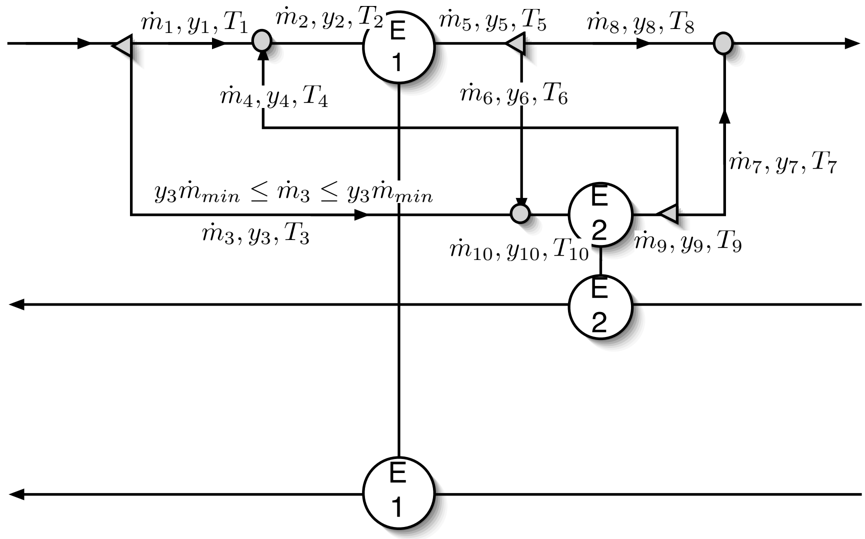


Figure 21: HEN superstructure for 1 cold stream and 2 hot stream

imation algorithm) allows one to avoid the use of manual/sequential procedure and solve the HEN design as a MINLP problem using a superstructure concept. The superstructure concept is based on the use of mixers and splitters to define systematically all the possible connections between the hot and the cold streams in the HEN. An example of such systematic representation is given in figure 21 for the case where one hot stream exchanges with two streams. In order to limit the combinatorial nature of the problem, it is recommended to generate the superstructure based on the results of the heat load distribution.

12 Final remarks concerning the heat exchanger network design

The HEN design problem is by essence a multi-modal problem, i.e. an optimisation for which several local optima exist. Furthermore, several criteria are not explicitly considered in the optimisation problem, it is therefore important to consider that several solutions exist and that a good HEN for one situation or location will not necessarily be an optimal system for another location or situation.

One has also to keep in mind that the ΔT_{min} constraint is just an assumption that allows one to solve the problem and that engineering minded analysis should be applied together with the pinch analysis solving procedure.

The pinch analysis seeks to identify the heat recovery opportunities between hot and cold streams in a system. The utility system integration define additional streams that have to be considered together with the process streams, not only because their interface with the process will be heat exchangers but also because these streams will be used to solve integration issues like start-up and shut-down procedures, flexibility and control issues. In most of the cases, the utility streams, like hot water or steam, will also play the role of heat transfer fluids between process streams allowing one to overcome restricted matches constraints and topology restriction problems.

Bibliography

[1] B. Linnhoff. Pinch analysis - a state-of-the-art overview. *Chemical Engineering Research and Design*, 71(A5):503–522, 1993. [Gives an overview of the different applications of pinch technology.]

[2] Robin Smith. *Chemical Process design and integration*. ISBN 0471486809. John Wiley & Sons, 2005. [This book presents the principles of process synthesis of process integration]

[3] L.T Biegler, I.E. Grossmann, and A.W. Westerberg. *Systematic Methods of Chemical Process Design*. Number ISBN 01334924223. Prentice Hall PTR, Marsh 1997. [This book presents computer aided process modelling and design methods focussing on the application of mathematical programming techniques for the optimisation of process system design.]

[4] M.M. El-Halwagi. *Process integration*. Number ISBN0123705320 in Process system engineering. Academic Press, 2006. [In addition to the pinch analysis, this books presents other applications of process integration techniques for materials integration]

[5] *Cost/capacity relationships part 2* Process engineering, April, 1998, pp 16

[6] B. Linnhoff, D.W. Townsend, P. Boland, G.F. Hewitt, B.E.A. Thomas, A.R. Guy, and R.H. Marsland. *A User Guide on Process Integration for the Efficient Use of Energy.* Number ISBN 0852953437. The Institution of Chemical Engineers, IChemE, Revised edition, 1994. [Pioneering user guide on process integration based on the application of pinch technology and pinch design method]

[7] R. Turton, W. Whiting, and J. Shaeiwitz. *Analysis, Synthesis and Design of chemical processes.* Number ISBN:130647926. Prentice Hall, NJ, 1998. [This book presents methods for process system analysis and design as well as cost estimation methods for process system design and optimisation.]

[8] Warren D. Seider, J. D. Seader, and Daniel R. Lewin. *Product and process design principles, Synthesis, Analysis and Evaluation.* Number ISBN 04712116631. Wiley, 2nd Edition, July 2003. [This book presents product and process design principles for reducing the environmental impact of the process design and including the application of new techniques : artificial intelligence, optimisation and computer modelling.]

[9] Luis Puigjaner, Georges Heyen (eds.) *Computer Aided Process and Product Engineering*, Number ISBN-10: 3-527-30804-0, Wiley-VCH, 1rst Edition, September 2006. [This book in 2 volumes presents the state of the art in the field of computer aided process and product engineering methods.]

LENI Systems

Process integration techniques for improving the energy efficiency of industrial processes

Example : calculation of a heat recovery heat exchanger network

Dr. François Marechal

Maître d'Enseignement et de Recherches
Industrial Energy Systems Laboratory
LENI-IGM-STI-EPFL
Ecole Polytechnique Fédérale de Lausanne
CH-1015 Lausanne
http://leni.epfl.ch

ÉCOLE POLYTECHNIQUE FÉDÉRALE DE LAUSANNE

Example : calculation of a heat recovery heat exchanger network

October 11, 2010

Abstract

This is a simple example presented to illustrate the application of the **problem table method** that allows one to calculate the minimum energy requirement of a process and the **pinch design method** that allows one to calculate one of the possible heat exchangers network that realizes the maximum heat recovery in the system. The heat exchanger network are further improvement by applying path following and loops breaking techniques.

1 Problem statement

Let us consider the example of the 4 streams : two hot and two cold streams as described in table 1.

	T_{in} [C]	T_{out} [C]	$\dot{M}cp$ [kW/C]	$\dot{Q}$ [kW]	α $[kW/C/m^2]$
A	20	130	-1.5	-165.0	0.5
B	80	140	-4.0	-240.0	0.5
C	160	60	+2.5	250.0	0.5
D	150	50	+2.0	200.0	0.5

Table 1: Streams definition

The following economical data will be used to calculate the optimal value of the ΔT_{min} and the economical performances of the heat recovery project.

Operating conditions

- Cooling water can be used to cool down process streams to 20°C
- Process operating time 2000h/year
- Process fluid is assimilated to liquid water with a heat tranfer coefficient of 500 $W/C/m^2$
- Investment equation: Purchased cost ref year (2000) $C_{p,ref} = 800(A)^{0.7}[CHF]$

Operating costs

- Natural gas: 0.05 CHF/kWh
- Water: 0.01 CHF/ m^3

- Electricity: 0.15 CHF/kWhe

Useful values

- Boiler efficiency based on lower heating value: 85%
- Interest rate: 8 %
- Expected life time of heat exchangers: 15 years
- Chemical engineering plant cost index (2000) : 400.
  Chemical engineering plant cost index (May 2010) : 510.
- Bare module factor: 4.74 for heat exchangers.

Assuming that the cold streams requirement are today satisfied by the hot utility, meaning that there is today no heat recovery realised for these streams. The present energy bill is therefore of $(240 + 165)[kW] * 2000[h/year] * 0.05[CHF/kWh]/0.85 = 47647[CHF/year]$ if we neglect the cooling requirement costs.

1.1 Minimum Energy Requirement

Before calculating the minimum energy requirement, one has to decide a value of the minimum approach temperature. We will calculate the ΔT_{min} value by considering the optimization for one of the possible heat exchangers in the system. This will be done in order to consider the same value for the ΔT_{min} of all the streams.

Solving the optimization problem for the heat exchange between streams B and C, we obtain the optimal value of $\Delta T_{min} = 9.5C$, as presented on figure 1. Therefore an overall value of 10 can be considered for the MER calculation.

As the heat transfer coefficient are similar for the hot and the cold streams, one can consider the same $\Delta T_{min}/2$ contribution for each the streams (table 2).

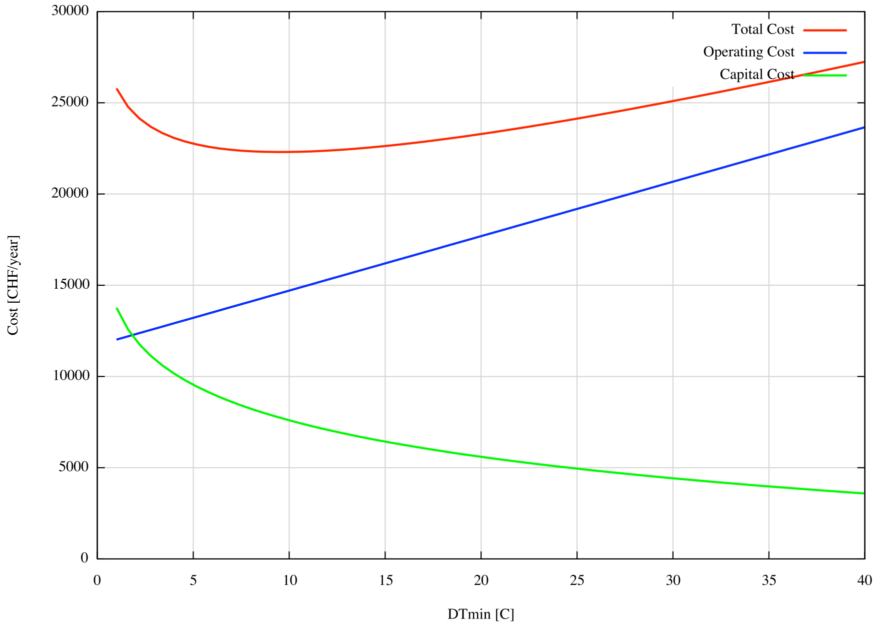


Figure 1: ΔT_{min} optimization results for the heat recovery between streams B and C

1. The ordered list of temperature is then set up (Column T^*, table3) and defines temperature intervals.

2. The temperature difference in each of the temperature interval is calculated by $\Delta T^*_r = T^*_{r+1} - T^*_r$ ((Column ΔT, table3).

3. Considering the corrected temperatures of each stream i, the $\dot{M}_i^+ cp_i$ contributions of each stream i in each temperature interval r is calculated. By convention $^+$ refers to positive for inputs in the heat exchangers. We consider therefore a positive sign for the hot streams and a negative sign for the cold streams.

4. The sum of the $\dot{M}cp$ in each temperature interval is calculated (Column $\sum \dot{M}cp$, table3)

5. The heat balance in each temperature interval is calculated by $\Delta\dot{Q}_r = \Delta T^*_r \cdot \sum_{i\in r} \dot{M}_i^+ cp_i$ (Column $\Delta\dot{Q}$, table3). A positive sign indicates that in the corresponding temperature interval the heat delivered by the hot streams exceeds the heat required by the cold streams.

6. The heat cascade is calculated assuming that the heat supplied from outside is equal to 0 : $\dot{R}^0_r = \dot{R}^0_{r+1} + \Delta\dot{Q}_r \quad \forall r = 1,..,n_r, \quad \dot{R}^0_{n_r+1} = 0$ (Column $\dot{R}^0_r$, table3).

7. Identify the smallest $\dot{R}^0_r$ (bold element in column $\dot{R}^0_r$, table3).

8. The Minimum Energy Requirement ($\dot{R}_{n_r+1}$)is calculated by $\dot{R}_{n_r+1} = -min(0, \dot{R}^0_r, r = 1, n_r + 1)$

9. Calculate the new heat cascade by $\dot{R}_r = \dot{R}_{r+1} + \Delta\dot{Q}_r \quad \forall r = 1,..,n_r, \quad \dot{R}_{n_r+1} = -min(0, \dot{R}^0_r)$. Note that this is also calculated by $\dot{R}_r = \dot{R}^0_r - min(0, \dot{R}^0_r)$.

The minimum heat requirement of the system is therefore $20[kW]$ while the heat to be evacuated from the system is $65[kW]$, calculated by the heat balance of the system. The minimum heat requirement represents the heat that has to be supplied to the system ($\dot{Q}^+$ in order to satisfy the heat requirement of the cold streams while maximizing the heat recovery from the hot streams and satisfying the constraint of the minimum temperature difference of $10[C]$ in the heat exchangers.

Supplying $20[kW]$ and removing $65[kW]$ closes the heat balance of the system. In order to check that no errors has been made in the calculation, the heat load of the difference between the hot and the cold streams of the system ($450(250 + 200) - 405(165 + 240) = 45[kW]$) is equal to the difference of the cooling and heating requirement ($65 - 20 = 45[kW]$).

The potential saving is therefore of about 95 % of the present energy consumption by heat recovery.

In this example, the pinch point corresponds to a corrected temperature of $85[C]$ and is created by stream B. In real temperatures, the pinch point temperature corresponds to a temperature of $80[C]$ for the cold streams and $90[C]$ for the hot streams.

The hot and cold composite curves of the problem are displayed on figure 2 while the heat cascade is given on figure 3. We can read on the figure that the heat recovered by heat exchange is of $385[kW]$ that is distributed between a heat load of $295[kW]$ above the pinch point and $90[kW]$ below the pinch point.

	T_{in} [C]	T^*_{in} [C]	T_{out} [C]	T^*_{out} [C]	$\dot{M}cp$ [kW/C]	$\dot{Q}$ [kW]
A	20		130		-1.5	-165.0
		25		135		
B	80		140		-4.0	-240.0
		85		145		
C	160		60		+2.5	+250.0
		155		55		
D	150		50		+2.0	+200.0
		145		45		

Table 2: Corrected temperatures

T^*	ΔT	A	B	C	D	$\sum \dot{M}cp$	$\Delta\dot{Q}$	$\dot{R}^0_r$	$\dot{R}_r$
155								+0.0	**+20.0**
	10			+2.5		+2.5	+25.0		
145								+25.0	+45.0
	10		-4.0	+2.5	+2.0	+0.5	+5.0		
135								+30.0	+50.0
	50	-1.5	-4.0	+2.5	+2.0	-1.0	-50.0		
85								**-20.0**	**0.0**
	30	-1.5		+2.5	+2.0	+3.0	+90.0		
55								+70.0	+90.0
	10	-1.5			+2.0	+0.5	+5.0		
45								+75.0	+95.0
	20	-1.5				-1.5	-30.0		
25								+45.0	**+65.0**

Table 3: Problem Table Method

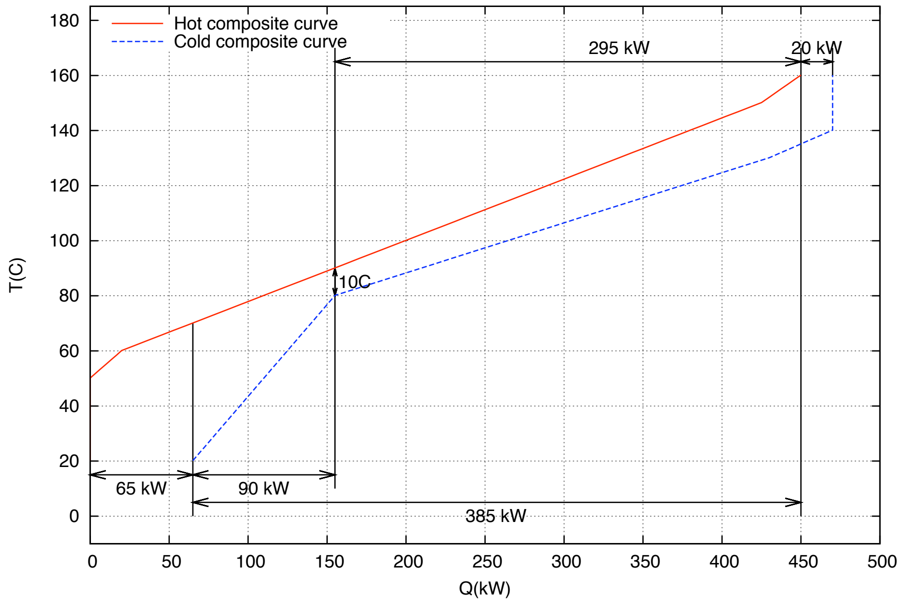


Figure 2: Hot and Cold Composite Curves

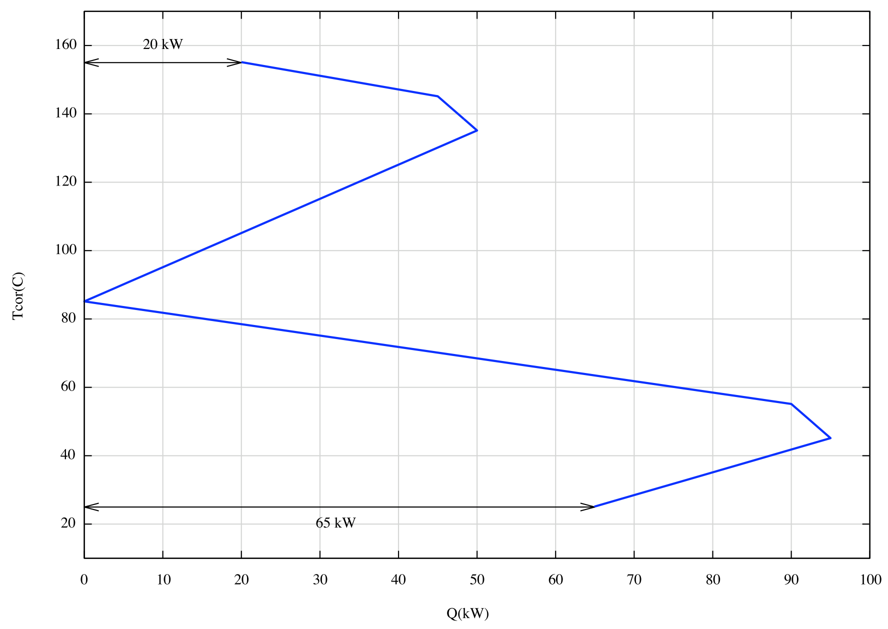


Figure 3: Grand Composite Curve

2 Pinch Design Method

Let us now calculate the heat exchanger network that realises the maximum heat recovery in the system. From the definition of the pinch point the system is divided into two sub-systems. Table 4 gives the list of the streams above the pinch point while table 6 gives the definition of the streams below the pinch point. The grid representation (Figure 4) is used to follow the progress of the design algorithm.

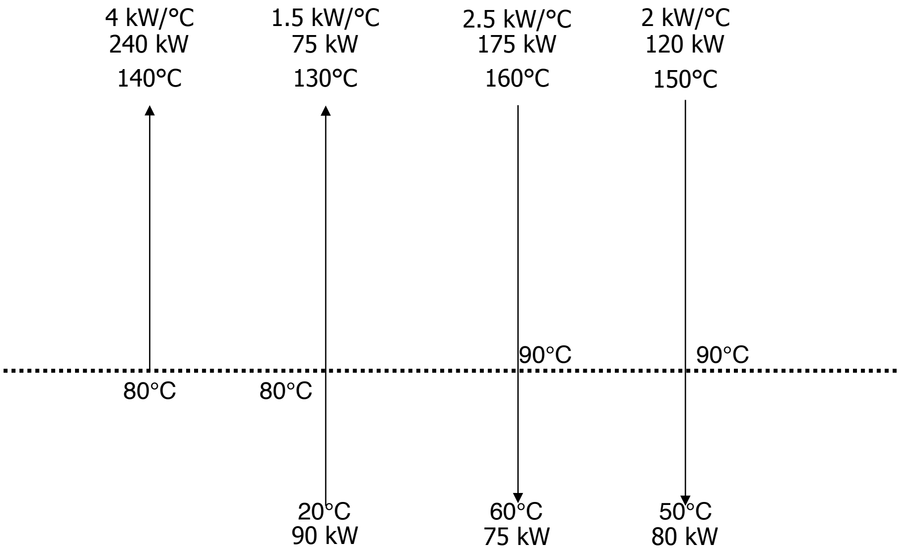


Figure 4: Grid streams representation

The heat exchanger network design algorithm is applied successively above and below the pinch point.

2.1 Heat exchanger network design above the pinch point

Above the pinch point, the goal is to cool down the hot streams to the pinch point temperature (in our case a real temperature of 90 [C] or 85 [C] when using the corrected temperature) only with the help of the cold streams above the pinch point. It should be noted that when different $\Delta T_{min}/2$ are used for the streams, it is more comfortable to apply the pinch design method using the corrected temperatures and go back to the real temperature domain once the streams matching are selected.

Cold streams						
	T_{in} [C]	T^*_{in} [C]	T_{out} [C]	T^*_{out} [C]	$\dot{M}cp$ [kW/C]	$\dot{Q}$ [kW]
A	80		130		-1.5	-75.0
		85		135		
B	80		140		-4.0	-240.0
		85		145		
Sum						-315.0
Hot streams						
C	160		90		+2.5	+175.0
		155		85		
D	150		90		+2.0	+120.0
		145		85		
Sum						+295.0
Balance						-20.0

Table 4: Streams above the pinch point

The following steps are realized

1. First, we check the number of streams rule ($N_{pinch,c} \geq N_{pinch,h}$). We have enough pinch cold streams (2) to cool down the 2 pinch hot streams to the pinch temperature.

2. Choose the biggest $\dot{M}cp$ for the hot and the cold streams above the pinch (i.e. from table 4).

3. Streams B and C are selected (figure 5).

4. Test the CP rule 1 : $\dot{M}c_{\mathrm{p}_h} \leq \dot{M}c_{\mathrm{p}_c}$. This rule is satisfied ($2.5 \leq 4.0$) .

5. Test the CP rule 2 : $\sum_{k=1,k\neq h}^{n_h} \dot{M}c_{\mathrm{p}_k} \leq \sum_{j=1,j\neq c}^{n_c} \dot{M}c_{\mathrm{p}_j}$. This rule is not satisfied ($2.0 \leq 1.5$). It is therefore necessary to split one hot stream.

6. Split the hot stream with the highest $\dot{M}cp$ to satisfy CP rule 2. The stream C is selected.

7. The split factor is calculated in order to satisfy the 'tick-off' rule with the available temperature difference. The lowest heat load is the one of stream A above the pinch point i.e. ($75[kW]$). The split factor is therefore calculated by $\dot{M}cp = \frac{\dot{Q}_{cold}}{T_{in,hot}-T_{pinch,hot}} = \frac{75}{160-90} = 1.07[kW/C]$. (see figure 6).

8. Calculate heat exchanger E1 (see table 11). The temperatures are calculated by heat balance starting from the pinch side and going towards the increasing temperatures. The heat exchange area is calculated by $A = \frac{\dot{Q}}{U\Delta T_{lm}}$.

9. Calculate the remaining problem after having placed heat exchanger E1. The new list of streams is given on table **??**. The minimum energy requirement is not modified, the heat exchanger is therefore well placed.

10. Apply the number of streams rules : $N_{pinch,c} \geq N_{pinch,h}$. $1! \geq 2$, there are not enough cold streams to cool down all the pinch hot streams to the pinch point temperature. One cold stream has to be splitted. The one with the biggest $\dot{M}cp$ is selected. (See figure 6).

11. The split factor is calculated in such a way that the tick-off rule is applied for the connexion B-C ($\dot{Q}_{E2} = 100[kW]$) and for connexion B-D ($\dot{Q}_{E3} = 120[kW]$) and that we obtain an isothermal mixing after the heat exchange. The split factor is therefore the ratio of the heat loads ($\dot{M}cp_{E2,cold} = 4 * \frac{100}{100+120} = 1.8[kW/C]$).

12. The heat exchangers E2 and E3 are calculated (see table 11 and figure 7).

13. The temperature of streams B after mixing is calculated ($80 + \frac{(100+120)}{4.0} = 135[C]$)

14. The remaining heat load ($20[kW]$) is supplied by the hot utility to reach $140[C]$ for stream B.

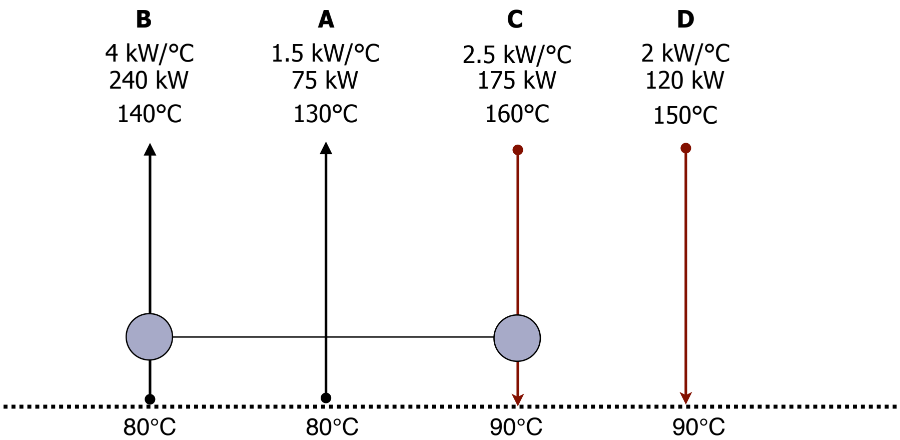


Figure 5: Pinch design method step 1 above the pinch

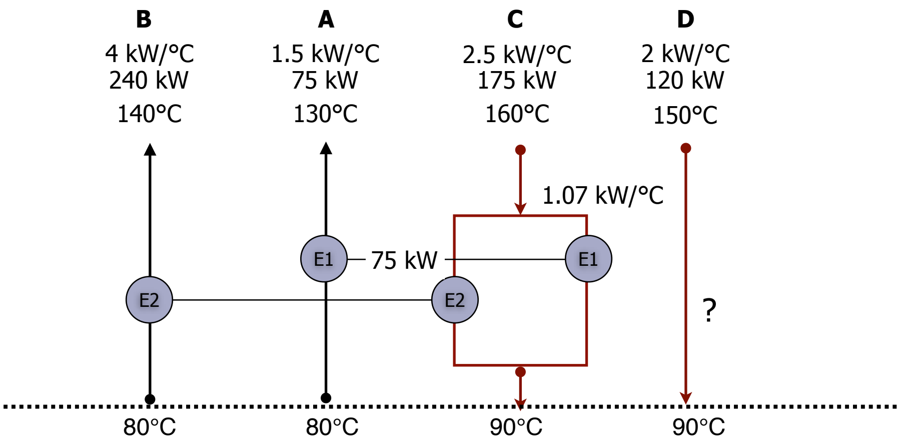


Figure 6: Pinch design method step 2 above the pinch

	T_{in} [C]	T^*_{in} [C]	T_{out} [C]	T^*_{out} [C]	$\dot{M}cp$ [kW/C]	$\dot{Q}$ [kW]
Cold streams						
A^1	20		80		-1.5	-90.0
B	80		140		-4.0	-240.0
Hot streams						
$C^{1,1}$	160		90		(2.5-1.07=+1.4286)	+100.0
$C^{1,2}$	90		60		+2.5	+75.0
D	150		50		+2.0	+200.0

Table 5: New list of streams after placement of heat exchanger E1

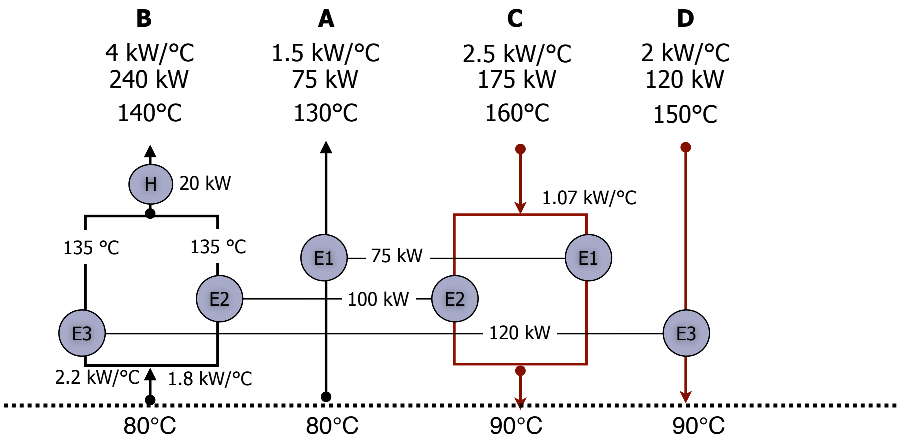


Figure 7: Pinch design method step 3 above the pinch

2.2 Heat exchanger network design below the pinch point

Below the pinch point, the goal is to heat up the pinch cold streams to the pinch temperature using the process hot streams. Below the pinch point (see table 6) only one cold stream (stream A) is present.

	T_{in} [C]	T^*_{in} [C]	T_{out} [C]	T^*_{out} [C]	$\dot{M}cp$ [kW/C]	$\dot{Q}$ [kW]
Cold stream						
A	20		80		-1.5	-90.0
		25		85		
Sum						-90.0
Hot streams						
C	90		60		+2.5	+75.0
		85		65		
D	90	50		+2.0	+80.0	
		85		45		
Sum						+155.0
Balance						+65.0

Table 6: Streams below the pinch point

The following procedure is applied :

1. The number of streams rules ($N_{pinch,h} \geq N_{pinch,c}$) is verified : we have enough hot streams to heat up the cold stream to the pinch temperature.
2. The streams with the highest $\dot{M}cp$ are selected (stream A and C).
3. The CP rule 1 below the pinch is applied $\dot{M}c_{\mathrm{p}_c} \leq \dot{M}c_{\mathrm{p}_h}$, $1.5 \leq 2.5$.
4. the CP rule 2 below the pinch ($\sum_{k=1,k\neq h}^{n_c} \dot{M}c_{\mathrm{p}_k} \leq \sum_{j=1,j\neq c}^{n_h} \dot{M}c_{\mathrm{p}_j}$) also holds since there is no additional pinch cold streams left.
5. The tick-off rule is applied to calculate the heat load of the heat exchanger. The heat load of stream C ($75kW$) is the lowest of the two and the heat exchanger E4 is calculated (see table 11).
6. The temperature of the cold stream at the inlet of E4 is calculated by $80 - \frac{75}{1.5} = 30[C]$.
7. The remaining heat load of the stream A ($90 - 75 = 15[kW]$) is satisfied by the heat exchange with stream D below the pinch. It is here not necessary to apply the CP rules since the stream A is not anymore a pinch stream after the exchange in E4.
8. The heat exchanger E5 is calculated (see table 11).
9. The remaining heat load of stream D ($80 - 15 = 65[kW]$) will be supplied by the cold utility.

2.3 Heat exchanger network design

The resulting heat exchanger network is presented on figure 8 and the corresponding heat exchangers are detailled on table 11. The number of heat exchangers is equal to the minimum number of units calculated by the formula $(4+2-1)+(3-1)=7$.

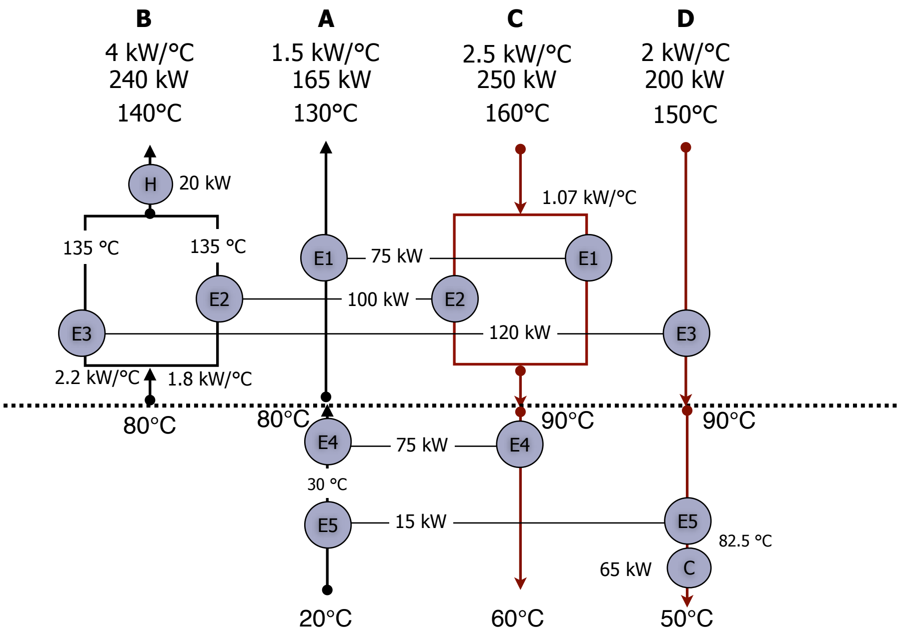


Figure 8: Heat exchanger network obtained by the pinch design method

The cost of the heat exchanger network is calculated using the cost estimation formula based on the calculated values of the heat exchanger area.

The total cost of the system is obtained from the operating cost calculated by :$OPEX = 20[kW]*2000[h/years]*0.05[CHF/kWh]/0.85 = 2352[CHF/year]$ while the capital cost is estimated at $CAPEX = 173905[CHF]/8.55[year] = 20339[CHF/year]$. The total cost is therefore of $22691[CHF/year]$ which is clearly better than the initial cost $(47647[CHF/year])$ and means 52% of cost savings).

3 Reducing the number of heat exchangers

Two techniques may be used to reduce the number of heat exchangers. Either identify the loops and then break them or/and the path following.

3.1 Loops breaking

A loop is a situation where the heat is transfered between two streams via more than one heat exchanger. In the figure 8, one identifies two loops as identified on figures 9 and 10.

1. Streams A-C via E4($75[kW]$) - E1($75[kW]$) (see figure 9)
2. Streams A-D via E5($15[kW]$) - E1($75[kW]$) - E2($100[kW]$) - E3($120[kW]$) (see figure 10)

	T_{hot} [C]	ΔT_{hot} [C]	T_{cold} [C]	ΔT_{cold} [C]	ΔT_{lm} [C]	$\dot{Q}$ [kW]	A [m^2]	Cost [CHF]
	160		90					
E1		30		10	18.20	75.0	16.5	32945
	130		80					
	160		90					
E2		25		10	16.37	100.0	24.4	43327
	135		80					
	150		90					
E3		15		10	12.33	120.0	38.9	60055
	135		80					
	90		60					
E4		10		30	18.20	75.0	16.5	32947
	80		30					
	90		82.5					
E5		43.3		62.5	61.24	15.0	1.0	4630
	30		20					
Total						385.0	97.3	173905

Table 7: Heat Exchangers obtained with Pinch Design Method

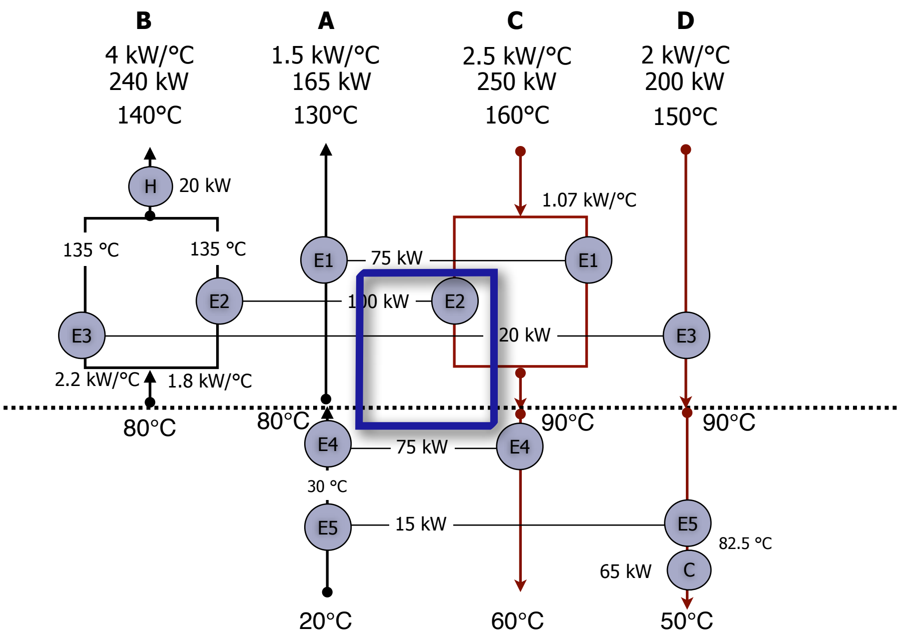


Figure 9: Identification of loop 1

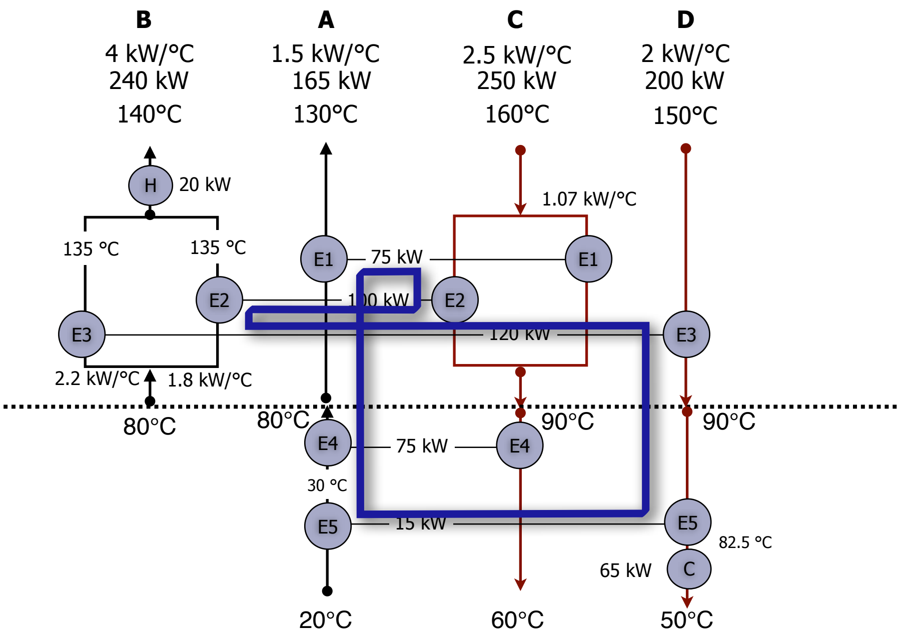


Figure 10: Identification of loop 2

3.1.1 Removing E4 using loop 1

In loop 1, the two heat exchangers have the same heat load ($75[kW]$). The heat exchanger to be eliminated is the one that is not in the split (E4), otherwize the temperature needed to reach 130 [C] for stream A would not have been sufficient on stream C. Suppressing E4 means increasing the heat load of exchanger E1 by $75[kW]$ to obtain a total heat load of $150[kW]$. The split factor of stream C has to be adapted to compensate for the suppression of E4, we use the following procedure :

1. Guess a temperature difference for E1 ($\Delta T_{cold,E1}$=4). This value should be lower than the value of $\Delta T_{min} = 10[C]$ since heat is transfered through the pinch point.
2. Calculate the outlet temperature of E1 on stream C : $T_{out^C,E1} = T_{in^A,E1} + \Delta T_{cold,E1}$ $(30 + 4 = 34[C])$.
3. Calculate the flow of stream C in exchanger E1 by : $\dot{M}_{C,E1}cp = \frac{\dot{Q}_{E1}}{T_{in^C,E1} - T_{out^C,E1}}$ $(150/(160-34) = 1.19[kW/C]$.
4. Calculate the flow of stream C in exchanger E1 by : $\dot{M}_{C,E2}cp = \dot{M}_C cp - \dot{M}_{C,E1}cp$ $(2.5 - 1.19 = 1.31[kW/C])$.
5. Calculate the outlet temperature of E2 on stream C : $T_{out^C,E2} = T_{in^A,E1} - \frac{\dot{Q}_{E2}}{\dot{M}_{C,E2}cp}$ $(160 - 100/1.31 = 83.6[C])$.

where

$\Delta T_{cold,E1}$ is the temperature difference of the cold side of heat exchanger E1.

$T_{out^C,E1}$ is the temperature at the outlet ($_{in^C,E1}$ would refer to inlet) of exchanger E1 on stream C.

$\dot{M}_{C,E1}$ is the flow of stream C in exchanger E1.

Exchangers E1 and E2 can therefore be recalculated while the other heat exchangers are not modified. The new heat exchanger network with 6 exchangers is shown on figure 11 and the corresponding table results 8 . The total cost of this new configuration is higher, although it has a smaller number of exchangers. This is mainly due to the fact that the temperature difference in the modified heat exchangers has slightly decreased (3.6 and 4 [C]) leading to much higher heat exchanger areas. The total heat exchange area has increased by 25 %.

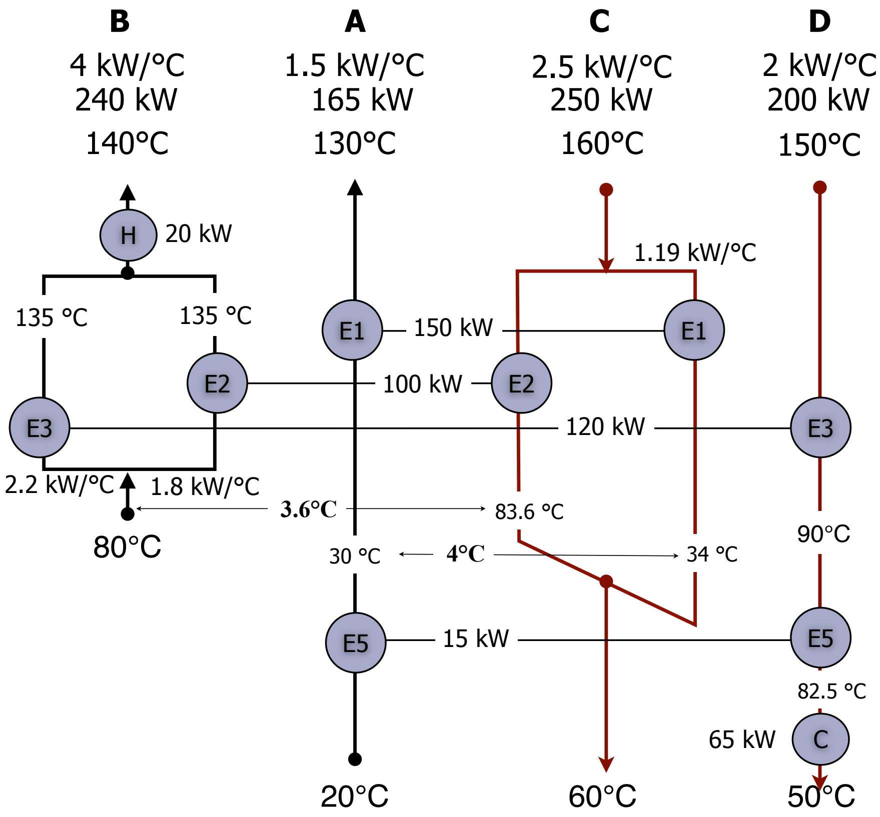


Figure 11: New heat exchanger network without E4

	T_{hot} [C]	ΔT_{hot} [C]	T_{cold} [C]	ΔT_{cold} [C]	ΔT_{lm} [C]	$\dot{Q}$ [kW]	A [m^2]	Cost [CHF]
	160		34					
E1		30		4	12.90	150.0	46.5	68043
	130		30					
	160		83.6					
E2		25		3.6	11.04	100.0	36.2	57131
	135		80					
	150		90					
E3		15		10	12.33	120.0	38.9	60055
	135		80					
	90		82.5					
E5		43.3		62.5	61.24	15.0	1.0	4630
	30		20					
Total						385.0	122.6	189859

Table 8: Heat Exchangers after loop 1 breaking to suppress E4

3.1.2 Removing E5 using loop 2

The lowest heat load of loop 2 is the heat exchanger E5 with a heat load of $(15[kW])$. The other exchangers in the loop will be affected : E5($15[kW]-15[kW]=0[kW]$) - E1($75[kW]+15[kW]=90[kW]$) - E2($100[kW]-15[kW]=85[kW]$) - E3($120[kW]+15[kW]=135[kW]$) (see figure 12). The heat exchangers have to be recalculated :

1. Flow of stream C is calculated considering the new heat load of E1. As E4 is not affected, $T_{out^C,E1}=90[C]$ is unchanged. The flow is calculated by $\dot{M}_{C,E1}cp=\frac{\dot{Q}_{E1}}{T_{in^C,E1}-T_{out^C,E1}}$ $(90/(160-90)=1.28[kW/C]$.
2. $\dot{M}_{C,E2}cp$ is deduced $(2.5-1.28=1.22[kW/C]$.
3. Flows in E2 and E3 on streams B have to be recalculated to compensate the heat load variation $(\dot{M}_{B,E2}cp=85/(135-80)=1.56)$ and $\dot{M}_{B,E3}cp=4-\dot{M}_{B,E2}cp=2.45[kW/C]$.
4. The new temperature at the outlet of E3 is already known.

The heat exchangers of the new configuration with 6 heat exchangers are given on table 9 . The total cost is now of $190001[CHF]$.

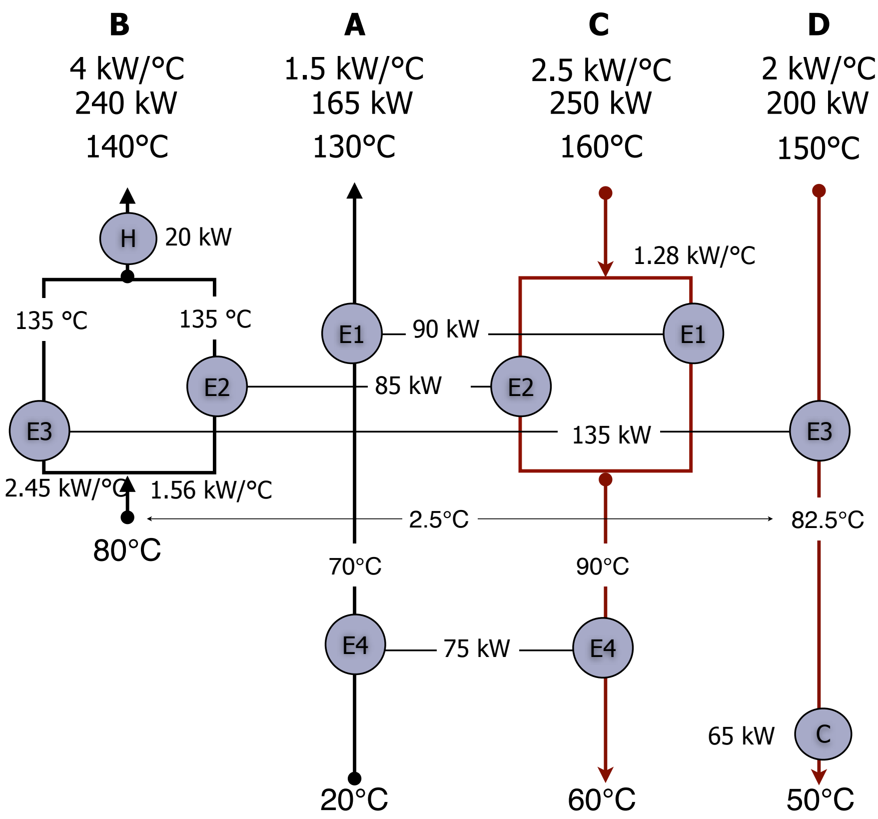


Figure 12: New heat exchanger network without E5

	T_{hot} [C]	ΔT_{hot} [C]	T_{cold} [C]	ΔT_{cold} [C]	ΔT_{lm} [C]	$\dot{Q}$ [kW]	A [m^2]	Cost [CHF]
	160		90					
E1		30		20	24.66	90.0	14.6	30238
	130		70					
	160		90					
E2		25		10	16.37	85.0	20.8	38706
	135		80					
	150		82.5					
E3		15		2.5	6.98	135.0	77.4	97211
	135		80					
	90		60					
E4		20		40	28.85	75.0	10.4	23846
	70		20					
Total						385.0	123.2	190001

Table 9: Heat Exchangers after suppression of E5 using loop2

The heat exchanger network cost is penalized by the area of the heat exchanger E3 that has a small temperature difference. Using the path following technique, the temperature difference of $10[C]$ can be reestablished. The path is identified as shown on figure 13. The new heat load of the hot utility is the heat that is transfered through the pinch i.e. $15[kW]$. The increase of the hot utility allows to recalculate the outlet temperatures of E2 and E3. $(140 - (20 + 15)/4 = 131.25$. The heat load of Exchanger E3 is of $120[kW]$. The flows in the split of stream B have to be recalculated and the heat exchangers E2 and E3 are recalculated. The total cost of the heat exchanger network is now of $145847[CHF]or145847/8.55 = 17058[CHF/year]$ but with the expense of an additional heat load of $15[kW]$. The additional utility cost is of $15[kW]*2000[h/year]*0.05[CHF/kWh]/0.85 = 1764[CHF/year]$ while the investment saving is of $20339 - 17058 = 3281[CHF/year]$. The new solution with a lower number of heat exchangers appears therefore to be more profitable with a total cost of $CAPEX(17058[CHF/year]) + OPEX(4117.6[CHF/year]) = 21176[CHF/year]$.

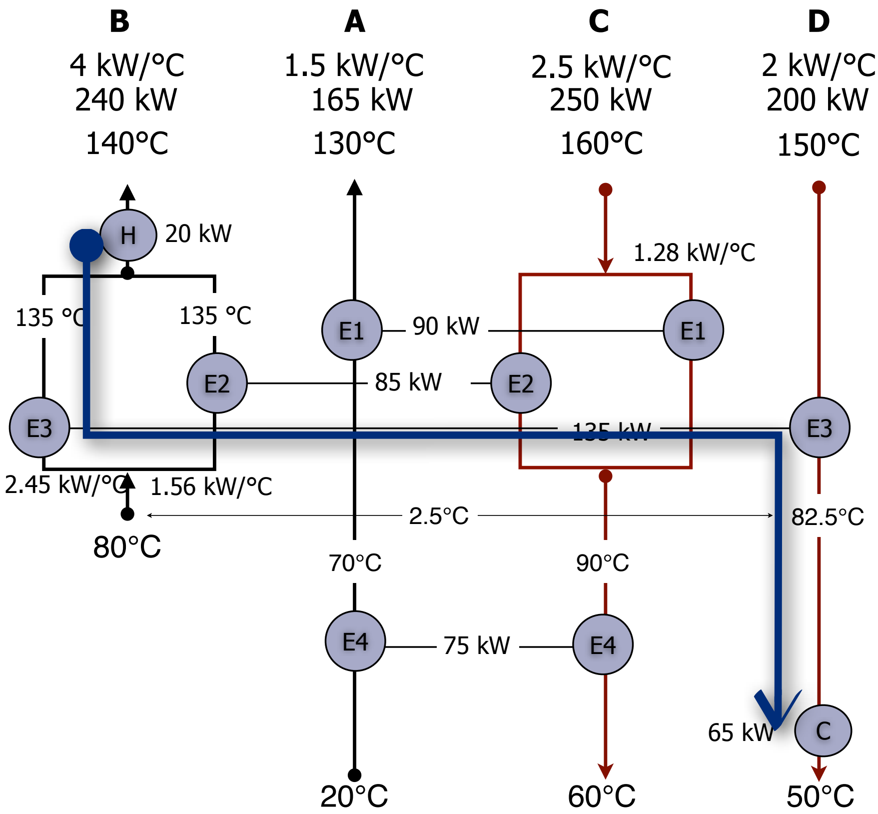


Figure 13: Path through E3 to reestablish the temperature difference

	T_{hot} [C]	ΔT_{hot} [C]	T_{cold} [C]	ΔT_{cold} [C]	ΔT_{lm} [C]	$\dot{Q}$ [kW]	A [m^2]	Cost [CHF]
E1	160 130	30	90 70	20	24.66	90.0	14.6	30238
E2	160 131.25	28.75	90 80	10	17.75	85.0	19.1	36567
E3	150 131.25	15	90 80	2.5	13.92	120.0	34.5	55196
E4	90 70	20	60 20	40	28.85	75.0	10.4	23846
Total						360	78.6	145847

Table 10: Heat Exchangers after suppression of E5 using loop2

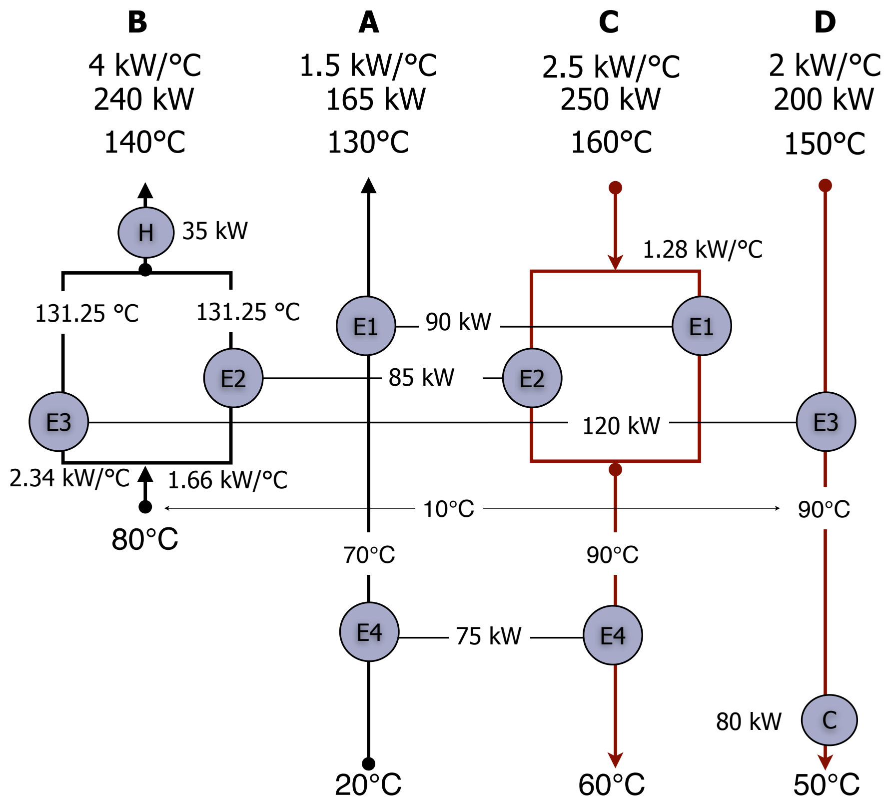


Figure 14: Heat exchanger network with higher hot utility after E5

3.2 Removing E5 using path following technique

The path following technique can also be used to remove heat exchangers. Let us consider the path as identified on figure 15. All the heat exchangers have been recalculated on table 16. The total cost of the heat exchanger is 133491$[CHF]$ giving a total cost of $CAPEX(15613[CHF/year]) + OPEX(4117.6[CHF/year]) = 19730[CHF/year]$. Which is the best solution obtained so far.

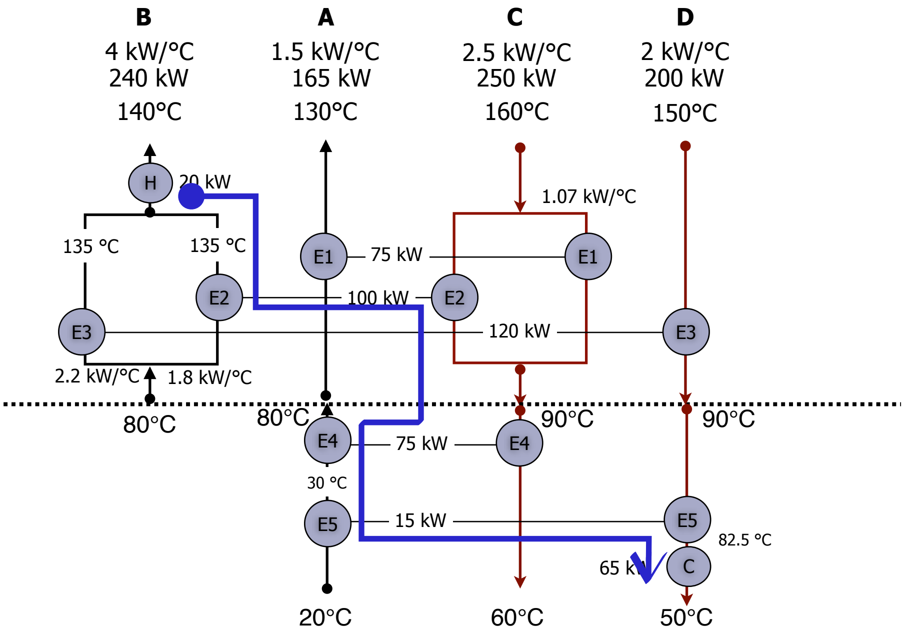


Figure 15: Path through E5

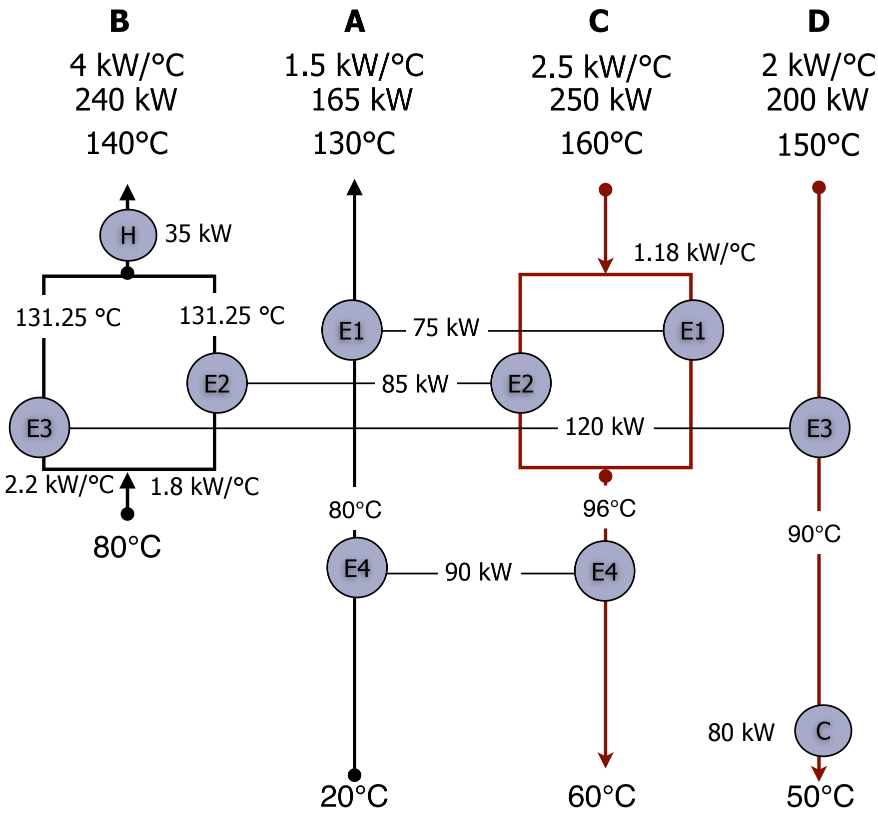


Figure 16: New heat exchanger network without E5

	T_{hot} [C]	ΔT_{hot} [C]	T_{cold} [C]	ΔT_{cold} [C]	ΔT_{lm} [C]	$\dot{Q}$ [kW]	A [m^2]	Cost [CHF]
	160		96					
E1		30		16	22.27	75.0	13.5	28585
	130		80					
	160		96					
E2		28.75		16	21.76	85.0	15.6	31719
	131.25		80					
	150		90					
E3		18.75		10	13.92	120.0	34.5	55196
	131.25		80					
	96		60					
E4		16		40	26.19	90.0	13.7	28991
	80		20					
Total						360.0	77.3	133491

Table 11: Heat Exchangers obtained with Pinch Design Method

3.3 Removing E5 and E4 using path following and loop breaking techniques

On figure 17, one can identify a new loop following E1 and E4 and linking streams A and C. E4 may be suppressed using the loop breaking techniques and the calculation method already presented before. The resulting network is presented on figure 18. The total cost of the network in this case is of 147924$[CHF]$ that is higher than the previous design. Supressing exchanger E4 is therefore not profitable due to the highest area required in E1.

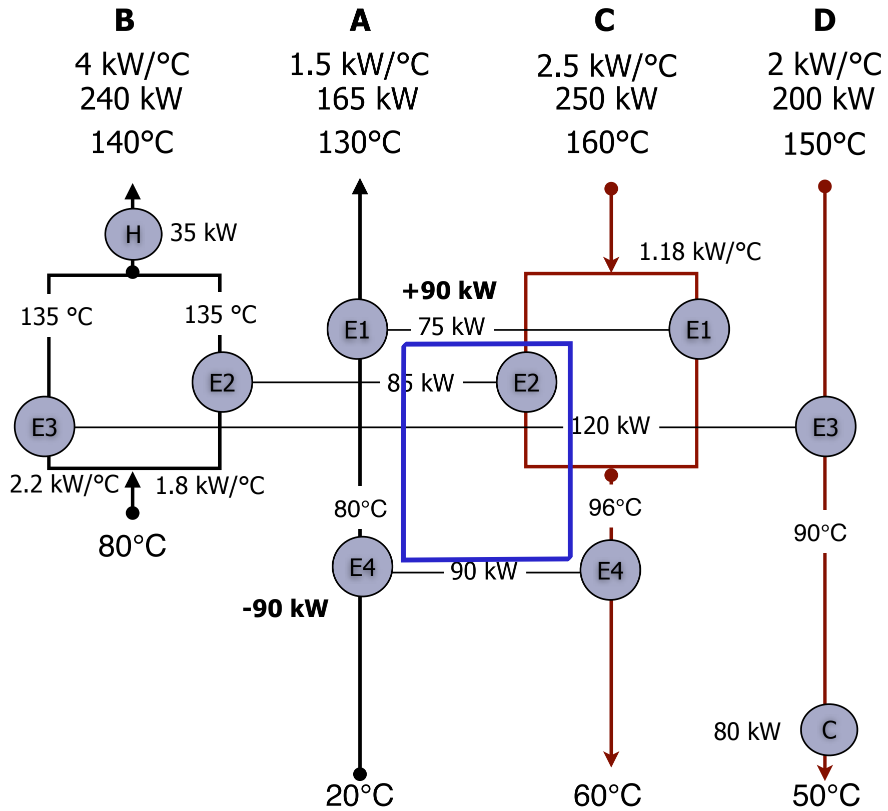


Figure 17: Loop through E1-E4

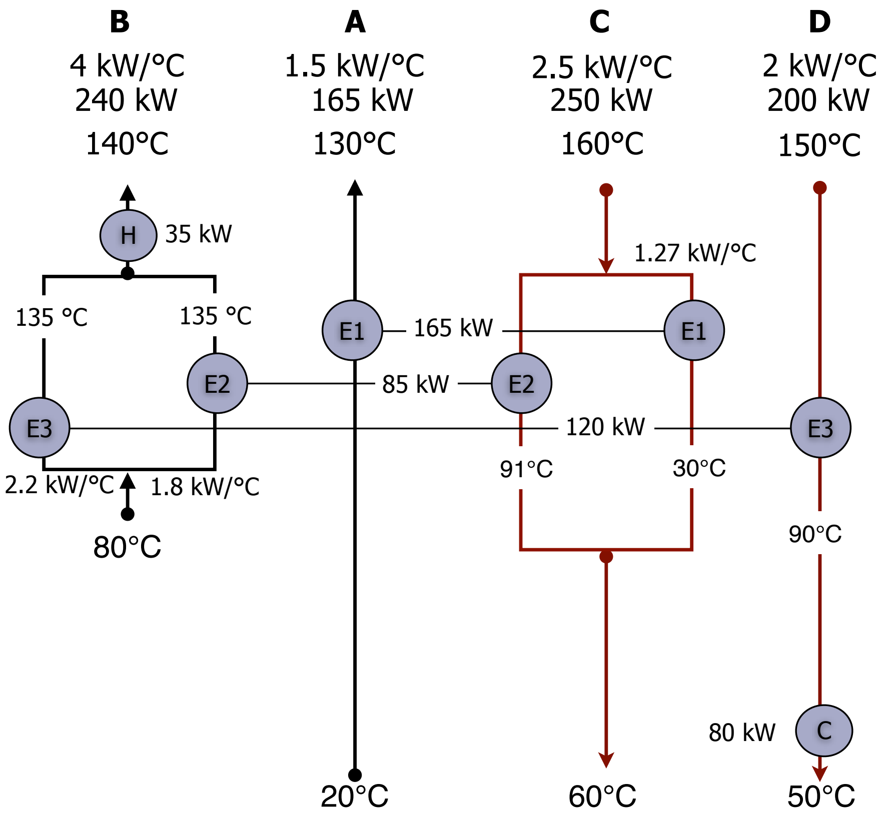


Figure 18: New heat exchanger network without E4

	T_{hot} [C]	ΔT_{hot} [C]	T_{cold} [C]	ΔT_{cold} [C]	ΔT_{lm} [C]	$\dot{Q}$ [kW]	A [m^2]	Cost [CHF]
	160		30					
E1		30		10	18.20	165.0	36.3	57165
	130		20					
	160		91					
E2		28.75		16	18.48	85.0	18.4	35563
	131.25		80					
	150		90					
E3		18.75		10	13.92	120.0	34.5	55196
	131.25		80					
Total						360.0	89.2	147924

Table 12: Heat Exchangers without E4

3.4 Removing E3 using path following technique

Using the path following technique, another path crossing heat exchanger E3 can be identified 19. The heat exchanger E3 can be suppressed at the expense of $120[kW]$ or $14119[CHF/year]$. The heat exchanger configuration is given on figure 20 and on table **??**. The total investment is the lowest obtained and corresponds to $99723[CHF]$ or $11664[CHF/year]$. Considering the penalty on the utility cost, the total cost of this solution is of $CAPEX(11664[CHF/year]) + OPEX(16471[CHF/year]) = 28135[CHF/year]$ that is more expensive than the best solutions found so far.

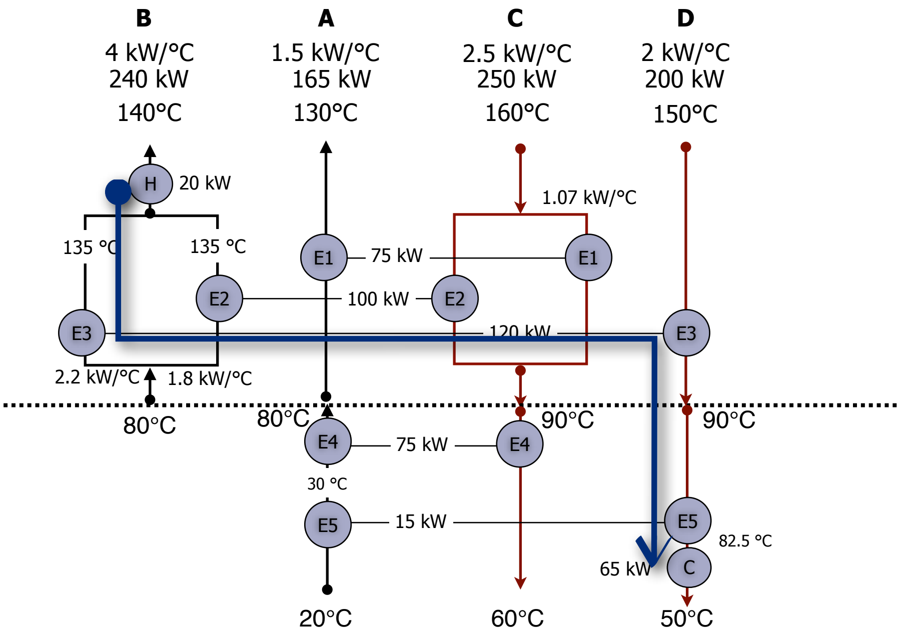


Figure 19: Path through E3

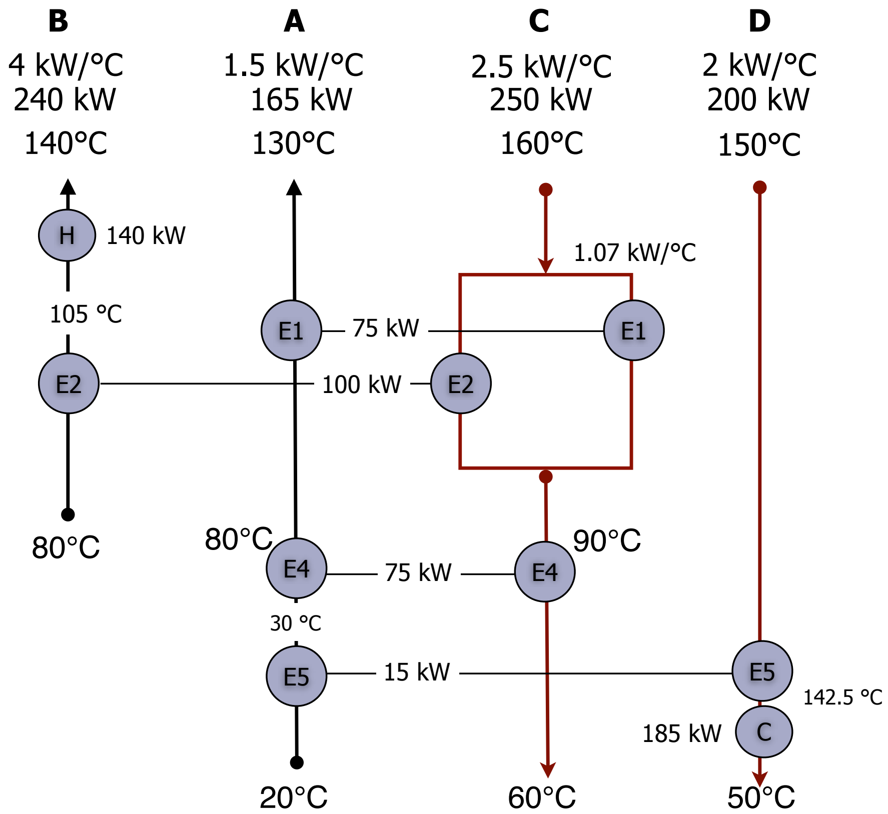


Figure 20: New heat exchanger network without E3

	T_{hot} [C]	ΔT_{hot} [C]	T_{cold} [C]	ΔT_{cold} [C]	ΔT_{lm} [C]	$\dot{Q}$ [kW]	A [m^2]	Cost [CHF]
	160		90					
E1		30		10	18.20	75.0	16.5	32945
	130		80					
	160		90					
E2		25		10	26.40	100.0	15.2	31041
	105		80					
	90		60					
E4		10		30	18.20	75.0	16.5	32947
	80		30					
	150		142.5					
E5		120		122.5	121.25	15.0	0.5	2830
	30		20					
Total						265.0	48.7	99723

Table 13: Heat Exchanger network after removing E3 with path following technique

3.5 Summary of the heat exchanger network design

Several heat exchanger network design configurations have been obtained and are summarized in table 14. These results are obtained without heat exchanger network optimization. Considering the net present value, the best solution is solution 4 (figure 15, table 16) with 6 exchangers and a $\Delta T_{min} = 10$ but with a higher operating cost.

	Invest $[CHF]$	CAPEX $[CHF/year]$	OPEX $[CHF/year]$	Total $[CHF/year]$	Nb Ex.
0	173905	20340	2352	22692	7
1	189859	22206	2352	24558	6
2	190001	22222	2352	24574	6
3	145847	17058	4118	21176	6
4	**133491**	**15613**	**4118**	**19731**	**6**
5	147924	17301	4118	21419	5
6	99723	11664	16471	28135	6

Table 14: Summary of the heat exchanger network design

LENI Systems

Process integration techniques for improving the energy efficiency of industrial processes

Part II : Integration of energy conversion systems and process efficiency improvement

Dr. François Marechal

Maître d'Enseignement et de Recherches
Industrial Energy Systems Laboratory
LENI-IGM-STI-EPFL
Ecole Polytechnique Fédérale de Lausanne
CH-1015 Lausanne
http://leni.epfl.ch

Preprints from : Marechal F., Improving process performances using pinch analysis, chapter 3.19.1.8, UNESCO Encyclopedia of Life Support Systems, ed. Christos Frangopoulos and Georges Tsatsaronis, EOLSS Publishers Co Ltd.

ÉCOLE POLYTECHNIQUE FÉDÉRALE DE LAUSANNE

3.19.1.8

Process Integration and Improvement

François Marechal

Industrial Energy Systems Laboratory, Ecole Polytechnique Fédérale de Lausanne (EPFL), CH-1015 Lausanne, Switzerland

Summary

The **pinch analysis** is a technique that has been developed to identify the possible energy recovery by counter-current heat exchange between the hot streams to be cooled down and the cold streams to be heated up in a system. The pinch analysis is based on the definition of the **minimum approach temperature** (ΔT_{min}) that represents the energy-capital trade-off between the energy savings obtained by heat exchange and the required heat exchangers investment. Pinch analysis results in the definition of the hot and cold composite curves and the pinch point location. Systematically analysing the results of the pinch analysis allows on one hand to improve the energy performances of the process by modifying the operating conditions of the process units and integrating in an optimal manner the energy conversion technologies that will convert the purchased energy into process energy requirements. Based on the hierarchical representation of the unit operations of the process, the pinch point location allows one to draw recommendations on the value operating conditions (pressure, temperatures) of successively the reactors, the separation units and the waste treatment units. The analysis of the grand composite curve allows then to identify the possible energy conversion technologies to be used to supply the energy requirement of the system. The Carnot composite curves are used to analyse the integration of the polygeneration units that implement heat pumping and combined production of power and heating-cooling requirements. Due to the high level of integration of the system, optimisation based methods are used to compute the interactions between options in the integrated system and to select the best process configurations out of possible options. The analysis of the optimisation results is made using integrated composite curves that represent the integration of a specified sub-system with the rest of the system.

Keywords

Process integration, Pinch Analysis, Rational Use of Energy, Heat Recovery, Heat Exchangers, Minimum approach temperature, Heat Exchanger Network, Minimum Energy Requirement, Heat load distribution.

Glossary

Nomenclature

Capital letters corresponds to total quantities while lower case letters correspond to specific values.

Cel^+	$[MU/kJ]$	the selling price of electricity;
Cel^-	$[MU/kJ]$	the import price of electricity;
$C1_w$	$[MU/year]$	the maintenance cost of the technology w
$c2_w$	[MU/s/unit of w]	the proportional cost of using one unit of the technology w
c	[MU/unit]	Specific cost of one unit, unit may be kg, kJ,...
c_p	[kJ/(kg K)]	Specific heat at constant pressure
$\dot{m}$	[kg/s]	Mass flowrate
ΔT_{min}	[K]	Minimum approach temperature in a heat exchanger
$\dot{E}$	$[kW]$	Exergy
$\dot{E}^-_{grid}$	$[kW]$	the net production of electricity;
$\dot{E}^+_{grid}$	$[kW]$	the net import of electricity;
ICF_w	[MU]	the fixed investment cost when selecting the technology w in the final configuration;
ICP_w	[MU/unit of w]	the proportional investment cost when selecting the technology w in the final configuration;
i		the annualisation interest rate
n_{years}	[year]	the expected life of the equipment
n		the number of process streams;
n_w		the number of technologies proposed in the utility system superstructure;
$\dot{q}_{wr}$	$[kW/$unit of w]	the heat load of the technology w in the temperature interval r for a given reference flowrate (unit of w), ($\dot{q}_{wr} > 0$ for a hot stream) ;
$\dot{Q}$	[kW]	Heat load or heat exchanged
$\dot{Q}^+$	[kW]	Hot utility requirement
$\dot{Q}^-$	[kW]	Cold utility requirement
$\dot{Q}_{ir}$	$[kW]$	the heat load of the process stream i in the temperature interval r ($Q_{ir} > 0$ for hot streams);
R_r	$[kW]$	the heat cascaded from the temperature interval r to the lower temperature intervals (r=1,$n_r + 1$);
T	[K]	Temperature
T^*	[K]	Corrected temperature
$\dot{W}$	$[kW]$	Work or electrical power
$\dot{w}_w$	$[kW/$unit of w]	the mechanical power produced by the reference flowrate of technology w; ($\dot{w}_w > 0$ for a mechanical power producer);
t	$[s/year]$	the total annual operation time;
$\dot{W}c$	$[kW]$	the overall mechanical power required by the process; ($\dot{W}c < 0$ if the overall balance corresponds to a mechanical power production);
Δh^0	$[kJ/kg]$	Specific heat formation of a fuel based on gaseous reference state, is also equal to the lower heating value of the fuel
Δk^0	$[kJ/kg]$	Specific exergetic value of a fuel
η_{Carnot}	[-]	Efficiency with respect to the reversible Carnot cycle
η_d		the efficiency of the electrical drives in the plant;
η_g		the efficiency of the electrical generator in the plant;

Sub and superscripts

$^{(+)}$	referring to the temperature at which the heat is supplied to the system
$^{(-)}$	referring to the temperature at which the heat is extracted from the system
$^{+}$	entering the system
$^{-}$	leaving the system
$_{h}$	hot streams
$_{c}$	cold streams
$_{grid}$	delivered to or from the electrical grid
$_{u}$	referring to the utility system

Abbreviations

MER	Minimum Energy Requirement
HEN	Heat exchanger network
MI(N)LP	Mixed Integer (Non) Linear Programming
EMAT	Exchanger Minimum Approach Temperature
HRAT	Heat Recovery Approach Temperature
MU	Monetary units
HRB	Heat Recovery Boiler

Contents

1 Introduction **4**

2 Pinch analysis and process improvement **4**

3 Utility integration **11**
3.1 Using the Grand composite curve 11
3.2 Combining pinch analysis and exergy concepts 13

4 Methodology for designing integrated utility systems **22**
4.1 Gas turbine and combustion system 24
4.2 Steam network 29
4.3 Refrigeration cycles 33
4.4 Heat pumps 34
4.5 Handling non linear cost functions 38

5 Using the exergy depletion as the objective function **39**

6 Representing the integration of the utility system **42**

7 Final remarks concerning the process optimisation and the utility system integration **47**

1 Introduction

In a production site (Figure 1), industrial processes aim at converting raw materials into valuable products and by-products. To do so, raw materials are processed through a list of processing steps or interconnected process unit operations that are driven by energy. The purchased energy will be first converted and distributed to be supplied in the suitable form to the process unit operations. This is typically done by a succession of conversion and heat transfer steps. Conceptually speaking, improving the energy efficiency of the system means to maximize the horizontal streams (conversion of raw materials in to products and by-products) while minimising the vertical ones (conversion of purchased energy into useful energy). By balance, this will result in the minimisation of the waste streams and losses of the system.

In a generic manner, one may consider that each process in the system is typically made of three steps : a preparation step, a reaction or conversion step and a post treatment step. Both preparation and post treatment steps use typically separation units and produce unreacted or out of specification materials that will be either recycled, introducing processing loops, or processed in a waste management unit. Each process unit operation is realised at given temperature and pressure levels and often require heat exchange to drive the operation. The heat transfer requirements are defined as hot streams to be cooled down and cold streams to be heated up. Pinch analysis has been developed to identify the possible heat recovery by heat exchange between these hot and cold streams and compute the minimum energy requirement of the system. In this chapter, we will analyse how pinch analysis will assist in defining the most appropriate process units operating conditions that will maximise the energy efficiency of the whole system. The analysis will not only concern **process improvement** where operating condition of one given process configuration or design are adapted to improve the energy efficiency of the system but also the **conceptual process design** of new processes where the unit operations have to be defined and arranged and for which the operating conditions have to be optimised. By extension, the method will also concern **process retrofit** where an existing process configuration will be improved by investing new equipment and optimising the operating conditions of the process units integrated in the new configuration. Knowing the improved minimum energy requirement of the system, we will then present how pinch analysis will help to determine how heating and cooling requirements will be supplied by converting purchased energy into useful energy in well integrated energy conversion units.

2 Pinch analysis and process improvement

In order to identify the way the process energy requirement can be reduced, a hierarchical approach can be used. The process can be seen as an onion with different layers (Figure 2). The heart of the process is the reaction layer. For improving the Minimum Energy Requirement (MER) of the process, we will first concentrate on the chemical reactions including catalysis aspects, their operating conditions and the type of reactor. The second layer will concern the separation units (distillations, stripping, absorptions, membranes,...) and the recycle streams. The third layer includes the support of production (like catalysts or solvent), then comes the waste treatment units.

The next layer defines the energy conversion and distribution known as being part of the utility system and that aims at satisfying the process energy requirements at a minimum cost using purchased energy.

Once all the previous layers are fixed, the complete list of streams for the heat recovery exchangers is fixed and the synthesis of the heat exchanger network can be realised.

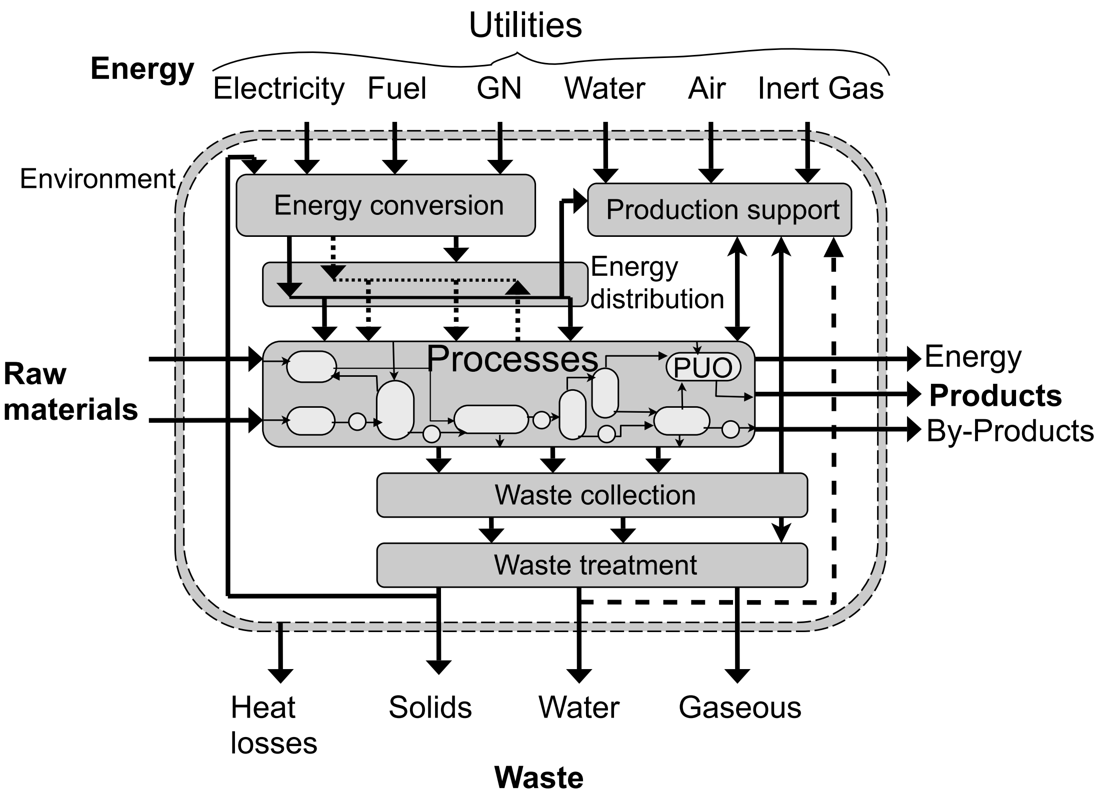


Figure 1: Production site : representation as an energy conversion system

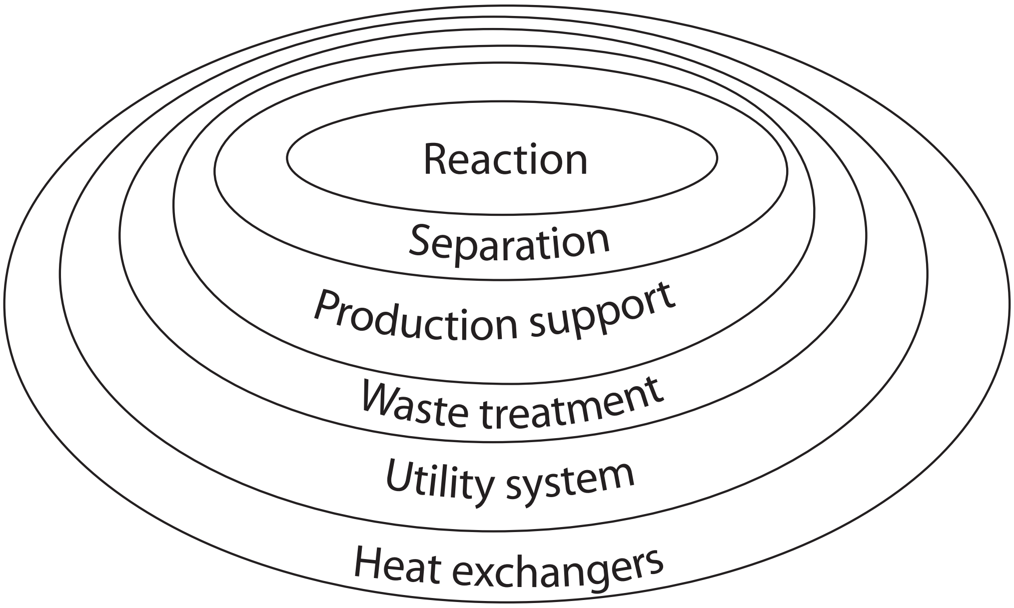


Figure 2: The onion representation of the process for energy efficiency improvement

Resulting from the calculation of the minimum energy requirement and the assumption of the ΔT_{min}, the pinch point temperature divides the heat transfer requirements in two groups:

- above the pinch point temperature, the heat transfer requirements define a heat sink and an additional amount of heat is needed to satisfy the requirements.

- below the pinch point temperature, the heat transfer requirements define a heat source that requires heat to be transferred to the environment.

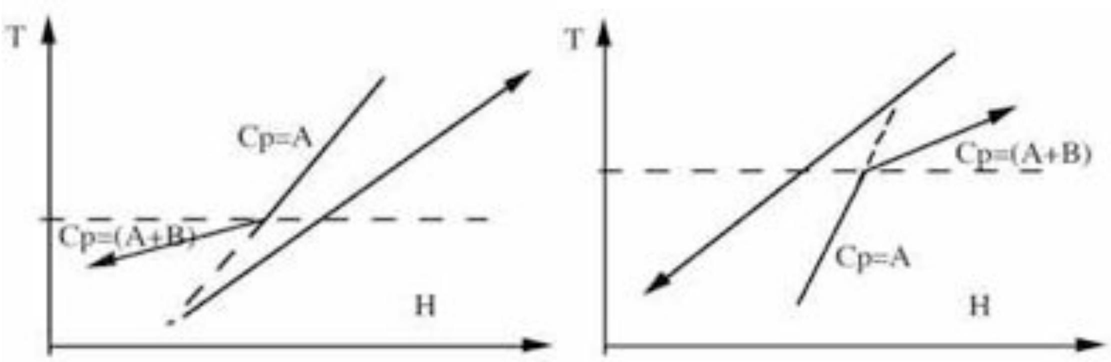
Being related to the fluid convective heat transfer coefficient, the pinch temperature is a stream dependent value corresponding to a single value in the corrected temperatures space (see chapter Pinch analysis for details). It will be the key driver for identifying profitable process modifications. Modifying process unit operations will aim at relocating the heat transfer requirement across the pinch temperature while delivering the same function in the process. This means to identify hot streams below the pinch point the conditions of which could change, in order to be relocated above the pinch point (i.e. moved from a heat source to a heat sink) and to identify cold streams above the pinch point the temperature of which could be modified to be relocated below the pinch point in order to profit from the excess of energy available.

This is known as the "plus-minus" principle (Figure 4):

1. Relocate hot streams from below to above the pinch point
2. Relocate cold streams from above to below the pinch point

In this analysis, the most important streams are obviously the streams that are near the pinch point since the temperature change will require fewer modifications in the operating conditions.

At this stage, it is worth to note that the pinch point is always created by the inlet conditions of a stream. It can be the inlet of a stream or of the segments created if a fluid phase change occurs.



(a) Hot streams (b) Cold streams

Figure 3: The pinch is introduced by the inlet temperature

Referring to the onion diagram, changing the temperature level of a requirement is typically obtained by changing the operating conditions of the process unit operations while keeping as a general goal the efficiency of the conversion of raw materials to products and as constraints the final products to be delivered by the process. Among the important streams to be considered, the heat of reactions typically introduce near vertical lines in the composite curves, when an exothermic reaction occurs below the pinch temperature or an endothermic reaction occurs above the pinch temperature, one could imagine to modify the operating pressure or temperature or even change the reactor type, i.e. change from an adiabatic reactor to a heat transfer type reactor. Another option would be to realise the reaction in several steps in order to maintain the temperature as low as possible (if possible below the pinch point) for endothermic reactions and as high as possible (if possible above the pinch point) for exothermic reactions. This is for example what is realised when adding a pre-reformer in a steam methane reforming process for hydrogen production. In such process, the pinch temperature is the high temperature of the reforming reaction. By realising the reformation in two steps, the heat of reaction of the pre-reformer becomes a cold stream below the pinch temperature while reducing in the same time the heat requirement of the remaining reforming reaction above the pinch temperature. This can be understood as a chemical heat pump that uses heat from the heat source to reduce the requirement in the heat sink above the pinch point temperature.

When fluid phase change occurs near the pinch point, changing the pressure of the phase change may be used to relocate a requirement around the pinch point. Decreasing the pressure of a fluid to be evaporated or increasing the pressure of a stream to be condensed will relocate respectively a cold stream from above to below the pinch temperature or a hot stream from below to above the pinch temperature. The changes (especially the pressures) obviously must remain compatible with the process unit operations in the flowsheet.

Not only the temperature level, but also the heat cascade has to be considered in this analysis.The grand composite curve of the process (Figure 5) gives useful insight in order to evaluate the interest of modifying the operating conditions. Each modification will be efficient if it does not create a new pinch point, otherwise part of the expected energy savings will not be realised. Considering that the pinch point divides the system into two independent sub-systems, the ap-

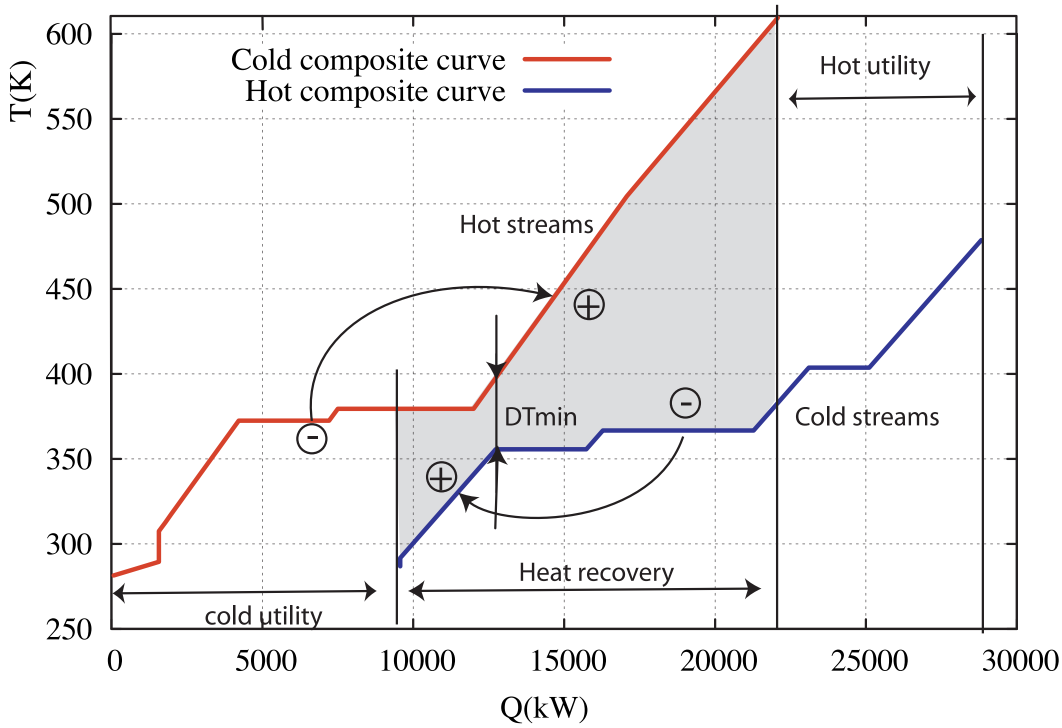


Figure 4: The composite curves for identifying key streams and process unit operations

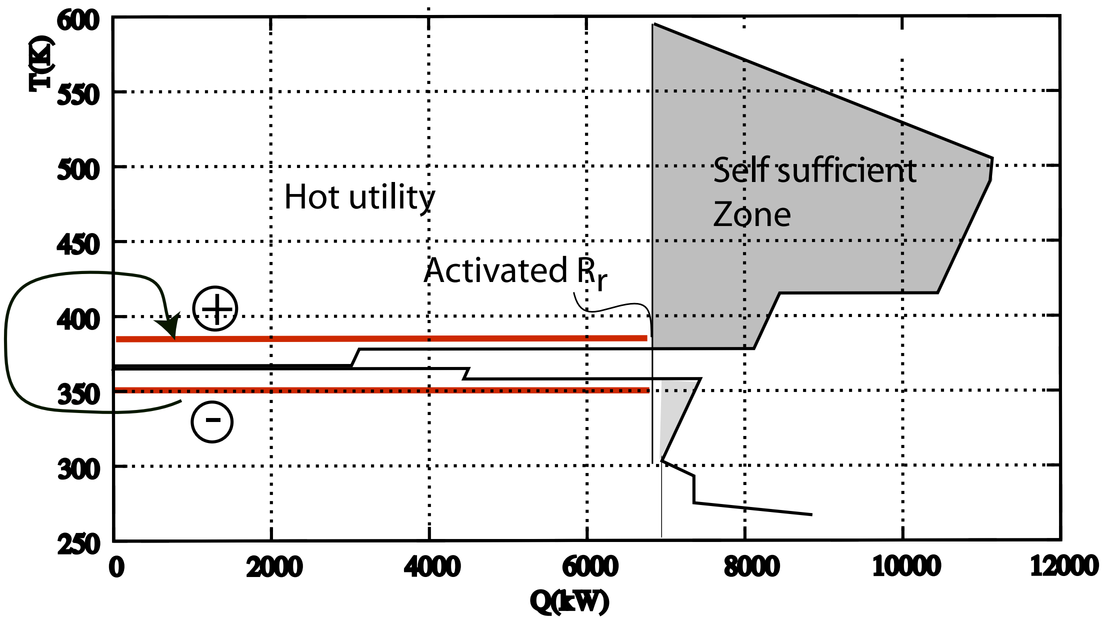


Figure 5: The Grand composite curves for computing the plus-minus heat load

plication of the "plus-minus" principle will have the effect of adding heat to and subtracting heat from the corresponding sub-systems. The **grand composite curve** is the plot of the heat cascaded as a function of the corrected temperatures defined by $((R_r, T_r^*), \forall r = 1, ..., n_r + 1)$. Since the heat cascade has still to be satisfied, the maximum heat $(\dot{Q}^{(+)})$ that can be subtracted from one temperature interval $(r^{(-)})$ and send back to another $(r^{(+)})$ will be obtained by solving Eq. 1. This equation is also valid for the transfer of the cold streams from above to below the pinch point. The mechanism that is explained here is illustrated graphically in the Figure 5.

$$\dot{Q}^{(+)}_{r^{(+)}} = min((\min_r(R_r), \forall r = n_r + 1, ..., r^{(+)}), (\min_r(R_r), \forall r = r^{(-)}, ..., 1)) \qquad (1)$$

In the onion diagram, the operating pressure in the separation units will be of major importance. Distillation column typically introduces two streams with nearly constant temperature : the reboiler defines a cold stream with a higher temperature, while the condenser defines a hot stream at a lower temperature. Changing the operating pressure of one distillation column will allow one to change the temperature of the requirement around the pinch point. Decreasing the column pressure will allow one to shift the reboiler from above to below the pinch point, while increasing the pressure will transfer the condenser from below to above the pinch point. This would be easy to do when the column is fed with a liquid. Considering that most of the separation systems proceed with more than one column, changing the pressure of one column with respect to the other will allow to save energy by heat integration. An example of such a modification is given in Figure 6 where the pressure of the first column is increased to relocate its condenser above the pinch point. This modification assumes that the hydrodynamics of column 1 has been verified with the new conditions. The modification leads as well to a modification of the pinch point location.

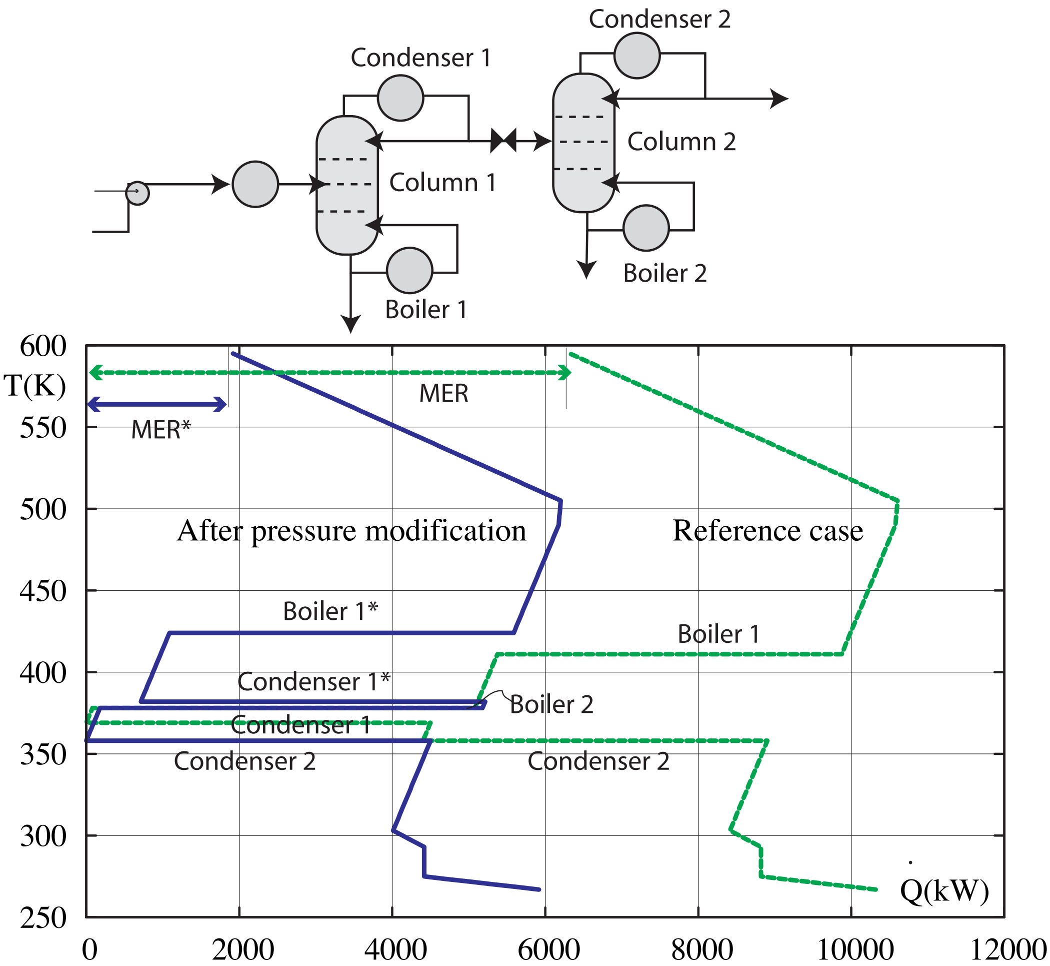


Figure 6: Modifying column pressure in a sequence of two distillation columns (the pressure of column 1 is increased)

3 Utility integration

The calculation of the MER assumes that the hot and cold utility will have the appropriate temperature levels to supply the energy to the process. In reality, the energy requirement is satisfied by converting, in energy conversion units, energy resources into useful energy that will be delivered to the process by means of heat exchangers or heat transfer. The streams resulting from the conversion of the resources will therefore be added to the hot and cold streams of the system and their enthalpy-temperature profiles will be considered for the calculation of the heat cascade. In contrast to the process streams that have constant temperatures and flowrates, the utility system streams have more degrees of freedom. Their temperatures, pressures and flowrates will be chosen in order to minimise the cost of satisfying the process energy requirement.

3.1 Using the Grand composite curve

The grand composite curve (Figure 7) gives information on the temperatures and flows conditions to be met by the utility system. It describes as a function of the temperature, the way energy has to be supplied to or removed from the system. The process grand composite curve is divided into three zones : above the process pinch point, the system is a heat sink to be supplied by a hot utility. Below the process pinch point and above the ambient temperature, the process is a heat source. The heat has to be evacuated from the process by a cold utility or used in another energy consuming sub-system like another process or a district heating system. Below the ambient temperature, the process requires refrigeration. Considering the heat cascade formulation, the grand composite curve represents the value of the pinch constraints. Any combination of utility streams that will maintain the $R_r \geq 0$ constraints will be feasible. The flowrate of the utility streams will be determined in order to minimise the cost of the utility (Eq. 2), therefore the flow of the cheapest stream will be maximised until it activates a heat cascade constraint and so on with the next cheapest. In addition, the number of utility streams to be considered should be limited in order to minimise the investment of the conversion units and the heat exchangers.

$$CU = time_{year} \sum_{u=1}^{n_u} \dot{m}_u cu_u \qquad (2)$$

Where

cu_u	$[MU/kg]$	is the cost of the utility stream u
$\dot{m}_u$	$[kg/s]$	is the flowrate of utility u

Considering the hot utility requirement above the pinch point temperature (Figure 7), the grand composite curve defines the process as a "global" cold stream to be heated up by the hot utilities. As a cold stream should have an increasing enthalpy-temperature profile, the so-called self sufficient zones will be ignored. In such zones, the heat delivered by the hot streams balances the requirements of the cold ones. The goal of the hot utility integration will be to define a hot stream (or a set of hot streams) whose enthalpy-corrected temperature diagram (hot composite curve) will always be above the Grand composite curve above the pinch temperature. Similarly, below the pinch, the Grand composite curve of the process will appear as a hot stream to be cooled down by the cold utility.

Enthalpy-temperature profiles of utility streams are obtained by calculating the conversion of the purchased energy resource (typically a fuel) in the unit operation that produces the required hot and cold stream. This calculation is first made for nominal conditions (eg one unit of fuel). The resulting enthalpy-temperature profile of the utility stream is then used to determine the flowrate in the conversion unit such that the utility profile will always be above the process

Grand composite curve as illustrated on figure 7. The minimum utility flow is computed by Eq. 3 that defines the intersection of the two curves in the corrected temperature space.

$$\dot{m}_w = \max_{k=1,n_k} \frac{R_{n_k+1} - R_k}{c_{p_w} * (T_w^{in} - max(T_w^{out}, T_k^* + \Delta T_{min}/2_w)} \tag{3}$$

with
- $\dot{m}_w$ — the flowrate of utility w $[kg/s]$;
- T_w^{in}, T_w^{out} — respectively the inlet and the minimum outlet temperature of the utility stream w;
- c_{p_w} — the specific heat of the utility w $[kJ/kg/K]$
- $\Delta T_{min}/2_w$ — the contribution to the minimum approach temperature of the utility stream w;
- (R_k, T_k^*) — for k=1,$n_k + 1$ the values of the heat cascade, R_{n_k+1} is the MER of the process;

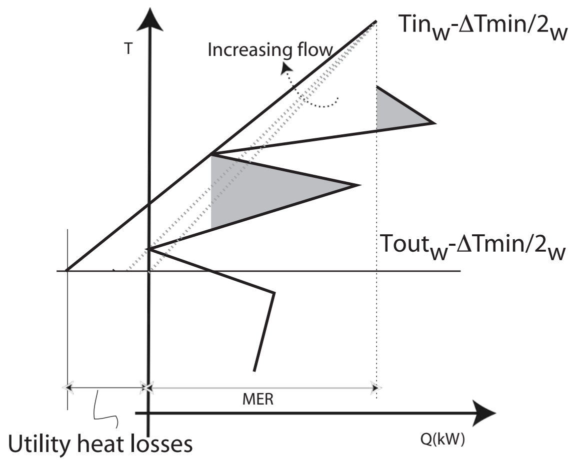


Figure 7: Computing the flowrate of the hot utility

It should be mentioned that in the example of Figure 7, the heat delivered by the hot utility is higher than the MER. The temperature T_k defining the intersection between the utility curve and the grand composite curve is called a **utility pinch point**. In the example, it differs from the process pinch point. Additional heat is therefore available from the hot utility and may be later on valorised. This situation often occurs in high temperature processes like steam methane reforming where the high temperature endothermic reaction defines a utility pinch point resulting from the integration of the combustion flue gases to satisfy the heat of reaction. In such process, the heat excess available in the flue gases at lower temperatures is used to produce high pressure steam that will be expanded in a condensing turbine to transform the excess heat into useful mechanical power by combined heat and power production.

The same graphical method applies to the cold utility and the refrigeration requirements. The graphical definition of the utility flowrate has been widely used to define the flowrates in the steam system. In this case, the steam condensation defines an horizontal segment (constant temperature) whose length defines the steam floware.

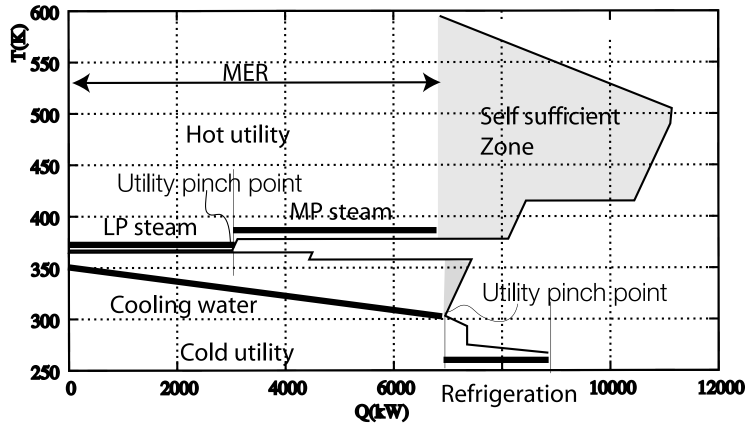


Figure 8: Grand composite curve for the definition of the utility system

The example on Figure 8 shows the integration of a hot utility system made of two hot streams : one steam condensation at low pressure (LP steam) and one at medium pressure (MP steam). The flow of the cheapest stream (LP steam) has been maximised by activating a heat cascade constraint that defines a **utility pinch point**. Below the pinch point, the flow of the cooling water has been minimised by maximising its outlet temperature. Below the ambient temperature, the refrigeration requirement is satisfied by a fluid that evaporates with a constant temperature. The cost of this utility stream is defined by the mechanical power consumption of the refrigeration compressor. To be rigourous, the heat load of the hot stream of the refrigeration cycle should be integrated in the system therefore requiring an additional flow of the cooling water.

The grand composite curve defines the enthalpy temperature profile to which the utility system has to be integrated. Above the pinch point, it defines a cold stream to be heated up by a hot utility, below the pinch point and above the ambient temperature the process is a hot stream to be cooled down by a cold utility. Below the ambient temperature, the grand composite defines a hot stream to be cooled down by a refrigeration system. The feasibility rule of the utility integration is that the grand composite curve of the utility system should envelop the process grand composite curve. Resulting from the linear nature of the heat cascade constraints, the optimal integration of the utilities will result in the definition of utility pinch points (intersection between the utility composite and the process composite). Each of them corresponding to the optimal use (maximum feasible flowrate) of the cheapest utility stream.

3.2 Combining pinch analysis and exergy concepts

The integration of the utility system concerns the conversion of purchased energy into useful energy for the process. The efficiency of the conversion concerns not only the supply of the heat requirement but also the polygeneration aspects like combined heat and power production, heat pumping and refrigeration cycle integration. When considering the different temperature levels at which one will supply the heat to the process, we consider implicitly the quality of the energy supplied to the process.

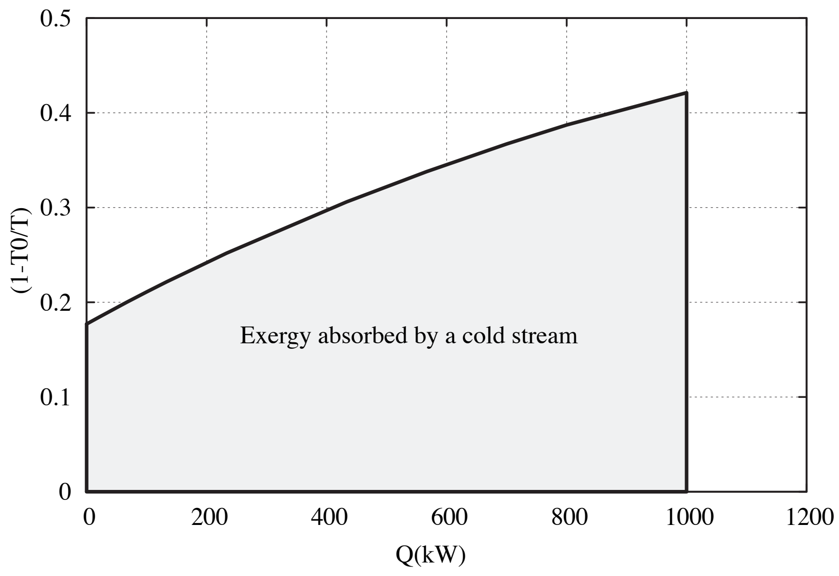


Figure 9: Exergy received by a cold stream heated up from 350 (K) to 500 (K)

The exergy analysis is a thermodynamic concept that helps understanding and analysing the efficiency of energy conversion systems (see Article Exergy and Thermodynamic Analysis). The exergy approach is used to represent in a coherent way both the quantity and the quality of the different forms of energy considered and is therefore useful when analysing the way energy is converted to satisfy the needs of the process.

Considering the constant c_p assumption used for computing the composite curves, the exergy delivered ($\dot{E}$) by a hot stream delivering heat ($\dot{Q}$) from T_r to T_{r+1} is computed by : $\dot{E} = \dot{Q}(1 - \frac{T_0}{T_{lm}})$ where T_{lm} is the logarithmic mean of temperatures computed by $T_{lm} = \frac{T_r - T_{r+1}}{ln(\frac{T_r}{T_{r+1}})}$ and T_0 is the ambient temperature, all temperatures being expressed in [K]. In a heat-Carnot factor ($(1 - \frac{T_0}{T})$) diagram (Figure 9), the area between the exchange curve and the heat axis represents the exergy delivered.

Therefore, when plotting the composite curves against the Carnot factor, we define the **Carnot composite curves**. The ambient temperature divides the Carnot composite curves into two parts. Above the ambient temperature, the area between the X-axis (positioned at Y=0 and the cold composite curve (Figure 10) represents the minimum exergy required to heat up the process cold streams above the temperature of the environment. One may consider that this requirement corresponds to the work that would be consumed by reversible heat pumps that would heat up the cold streams using the ambient heat as cold source. Below the ambient temperature, the area between the Carnot composite of the cold streams and the X-axis represents the exergy delivered by the cold streams below the temperature of the environment. By analogy, one may consider that this is the work that would be delivered by reversible Rankine cycles that use the environment as a hot source and the cold streams of the process as a cold source. For the hot streams, the area between the Carnot composite curve and the X-axis represents for positve Carnot factors the exergy that is delivered by the hot streams (i.e. the work that would be delivered by reversible Rankine cycles using the hot streams as hot source and the environment

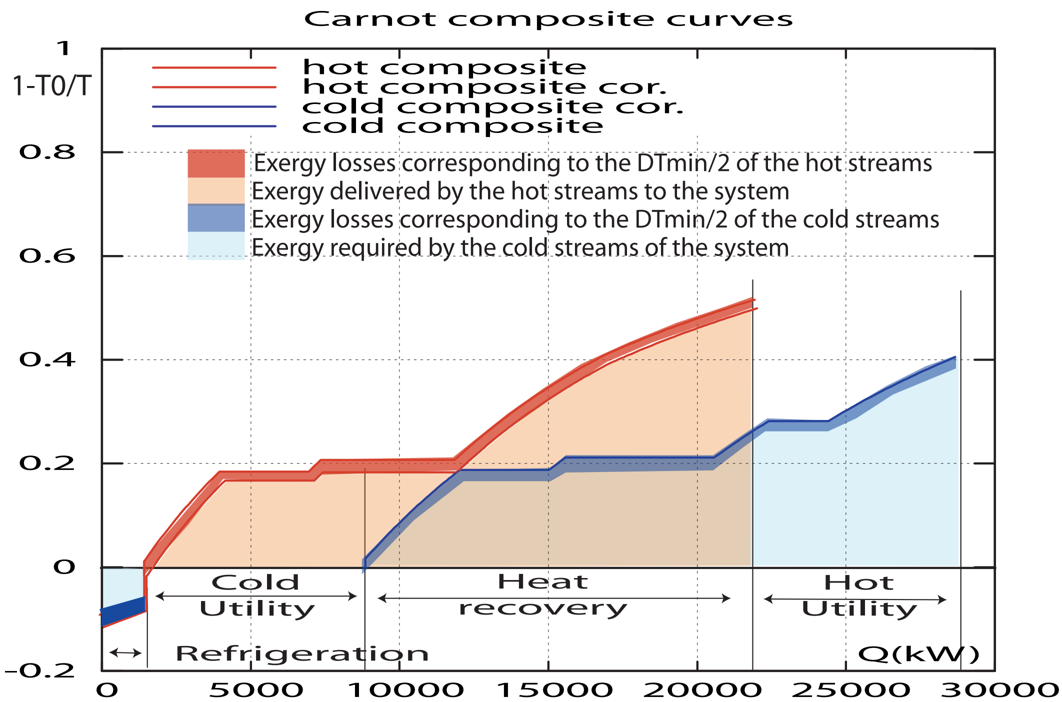


Figure 10: Carnot Composite Curves of a process

as a cold source) while for negative Carnot factors, the area corresponds to the minimum exergy required to refrigerate the hot streams. The difference between the two areas represents the net exergy that would be destroyed by the heat transfer that realises the maximum heat recovery by heat exchange.

This exergy destruction can be divided into two parts : an exergy conversion opportunity part and inevitable exergy destruction. The inevitable exergy destruction results from the ΔT_{min} assumption, it corresponds to the shift from real to corrected temperatures that are obtained by shifting vertically the hot streams downwards and the cold streams upwards by their respective $\Delta T_{min}/2_i$. This destruction is associated to the convective heat transfer coefficient of the hot and cold streams and the resulting investment in heat exchangers and is supposed to result from an energy-capital trade-off analysis. When the corrected temperatures are used to compute the Carnot composite curves, the difference of the area between the composite curves and the X-axis correspond to an amount of exergy that remains available for polygeneration after having accepted the $\Delta T_{min}/2$ values.

The Carnot grand composite curve is established on the basis of the corrected temperatures, it describes as a function of the temperature, the way energy has to be supplied to or removed from the system. Above the pinch point, the system is a heat sink to be supplied by a hot utility (Figure 11). In this figure, it can be seen that even if the hot and cold streams are in balance in a self-sufficient zone, there is exergy available due to the temperature difference between the hot and the cold streams in the zone.

Below the pinch temperature and above the ambient temperature, the subsystem is a heat

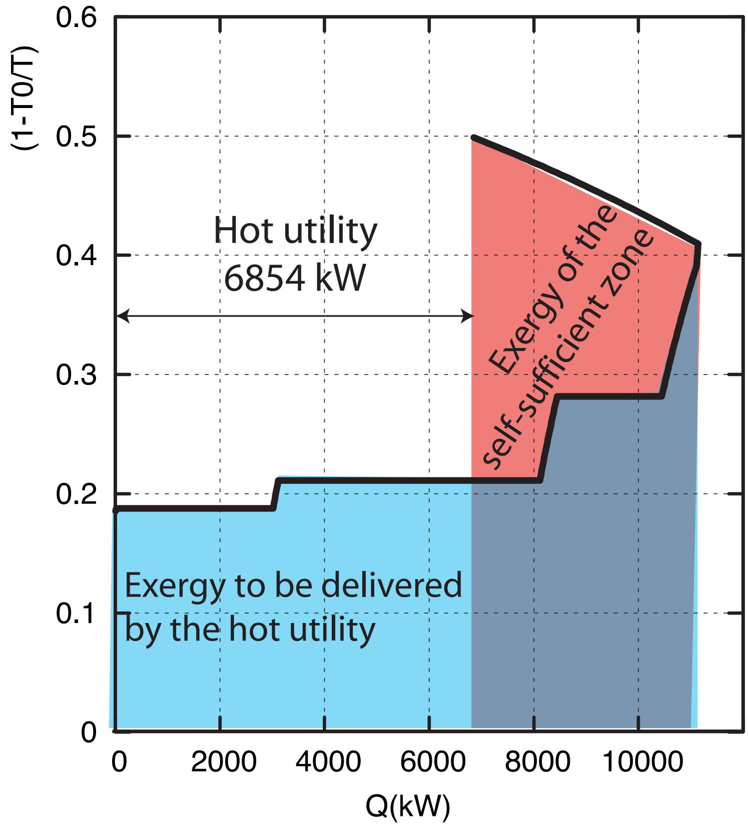


Figure 11: Carnot Grand composite curve above the process pinch point

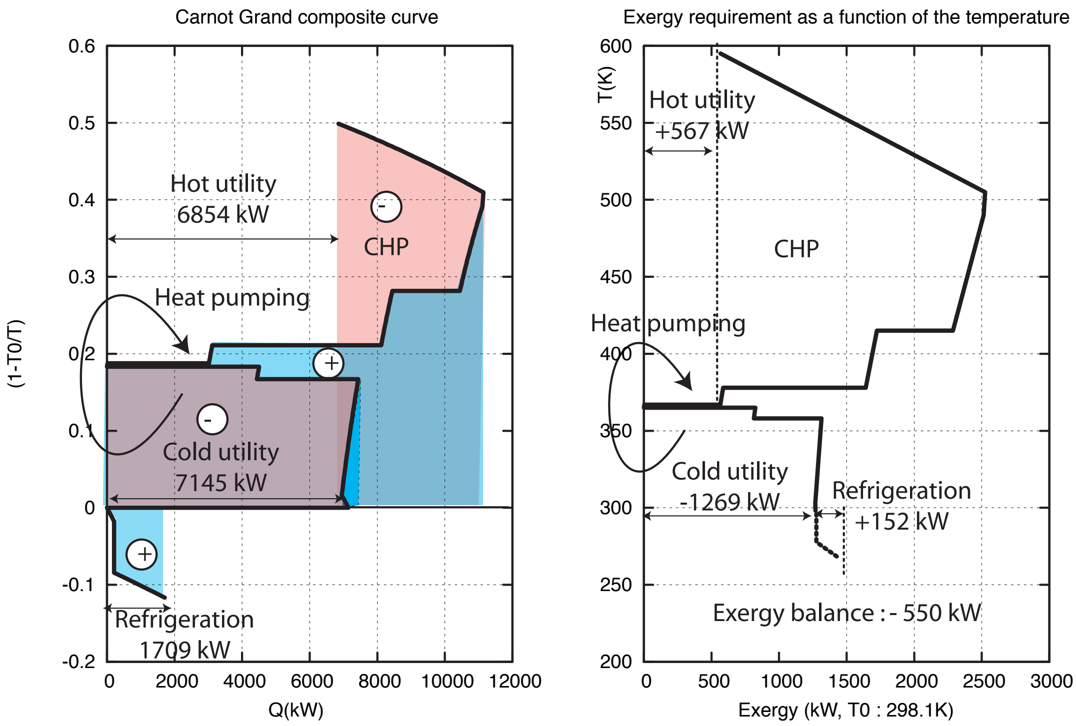


Figure 12: Carnot Grand composite curve of the process. (+) refers to exergy to be delivered to the process, while (-) refers to exergy delivered by the process

source that delivers exergy. Below the ambient temperature, the subsystem requires refrigeration. The Carnot grand composite curve that combines the hot utility and the cold utility requirements (Figure 12, left) has therefore deficit and surplus exergy area according to the temperature level.

The Temperature-Exergy grand composite curve on figure 12, right represents the cumulated exergy required by the process as a function of the temperature. The overall exergy balance shows an overall exergy availability. This exergy is made available by the process streams through heat transfer. However, from the definition of the pinch point, the three sub-systems have to be considered and both the energy and exergy balances have to be realised by properly integrating combined heat and power devices (Rankine cycles), while exergy transfer between sub-systems will be possible by heat pumping.

The integration of the utility system that supplies the heating and cooling requirements may be represented in the Carnot integrated composite curves of the utility system (Figure 13) where the utility system Grand composite is drawn against the Grand composite of the process. When working in the Carnot factor dimension, the area between the two curves represents the exergy destroyed in the system and to be used by heat pumps or combined heat and power units. In the figure, the shaded area represents the exergy of the purchased fuel. First, the conversion creates losses in the combustion so that only part of the puchased exergy becomes available for the process through heat exchange with the combustion gases. After considering the exergy losses corresponding to the $\Delta T_{min}/2$, the exergy delivered by the combustion gases satisfies the needs of the process in terms of energy. However the temperature difference between the combustion gases and the process streams shows an important exergy loss represented by the area between the utility and the process composites.

The combined heat and power production principle will be used to reduce this exergy loss by converting the available exergy partly into work.

Above the pinch point temperature, the mechanical power ($\dot{W}$) produced by the combined heat and power system will extract energy out of the system. To close the energy balance, the work produced will be compensated by supplying additional heat energy above the pinch temperature. Considering $\eta_{conv} = \frac{\dot{Q}_{above}}{\dot{m}_{fuel}\Delta h^0_{fuel}}$, the efficiency of converting purchased energy into heat above the pinch temperature, the efficiency of the mechanical power production will be given by : $\eta_{CHP} = \frac{\dot{W}}{\eta_{conv}} = \frac{\dot{W}\cdot\dot{m}_{fuel}\Delta h^0_{fuel}}{\dot{Q}_{above}}$. Below the pinch point and above the ambient temperature, the work produced by converting the available exergy will reduce the energy to be extracted from the corresponding sub-system. Therefore, below the process pinch temperature, the work production corresponds to a reduction of the cooling requirement.

In both cases, the combined production of mechanical power is limited by the efficiency of the conversion system and the available exergy, i.e. by the activation of utility pinch points.

Resulting from this analysis, it is important to note that the pinch temperature is the key decision factor for the combined heat and power production. If purchased energy, or heat above the pinch point is used in a combined heat and power production, the heat produced has to be delivered above the pinch temperature. When it is delivered below the process pinch temperature, there is no advantage through the integration. The combined heat and power unit becomes a power unit only since the heat is added to the cooling requirements when looking at the integrated system that considers that all the heat recovery exchanges will be realised.

Combined heat and power production is widely used in the process industry. It is typically implemented by steam networks (Figure 14) that allow not only to exchange heat between process sections or processes in a production site but also to implement combined heat and power production. In such systems, steam is used as a heat transfer and conversion fluid. It will receive heat from process hot streams or purchased energy conversion units and supply the received heat to cold streams of the process. The temperature difference between the hot and the cold streams will correspond to different steam pressures that allows mechanical power production by expansion in turbines.

Steam network is used to distribute heat in industrial sites and produce shaft work by expansion. Steam network will therefore allow for a site scale process integration by allowing the indirect heat recovery between process streams and the production of mechanical power.

Referring to the grand composite obtained when balancing the utility and the process streams (Figure 15), the steam levels will be identified by integrating rectangles between the balanced composite curve and the temperature axis. In such rectangles, the horizontal lines correspond to evaporation and condensation levels. Maximising the height of the rectangle corresponds therefore to maximising the expansion ratio while maximising the basis means maximising the steam flow expanded in the turbine. When the Carnot factor is used as the Y-axis, the area of the rectangles represents the expected work to be produced by the expansion in the steam turbine. In the example, although the pressure in the first rectangle could be higher, it has been limited for due to practical constraints.

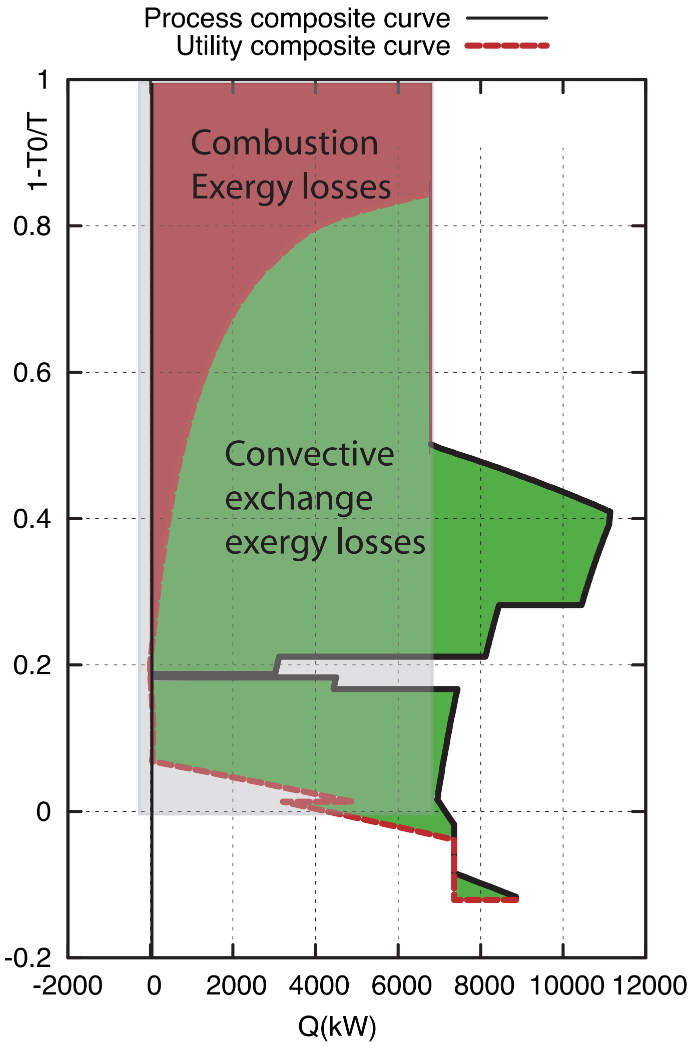


Figure 13: Carnot integrated composite curve of the utility system

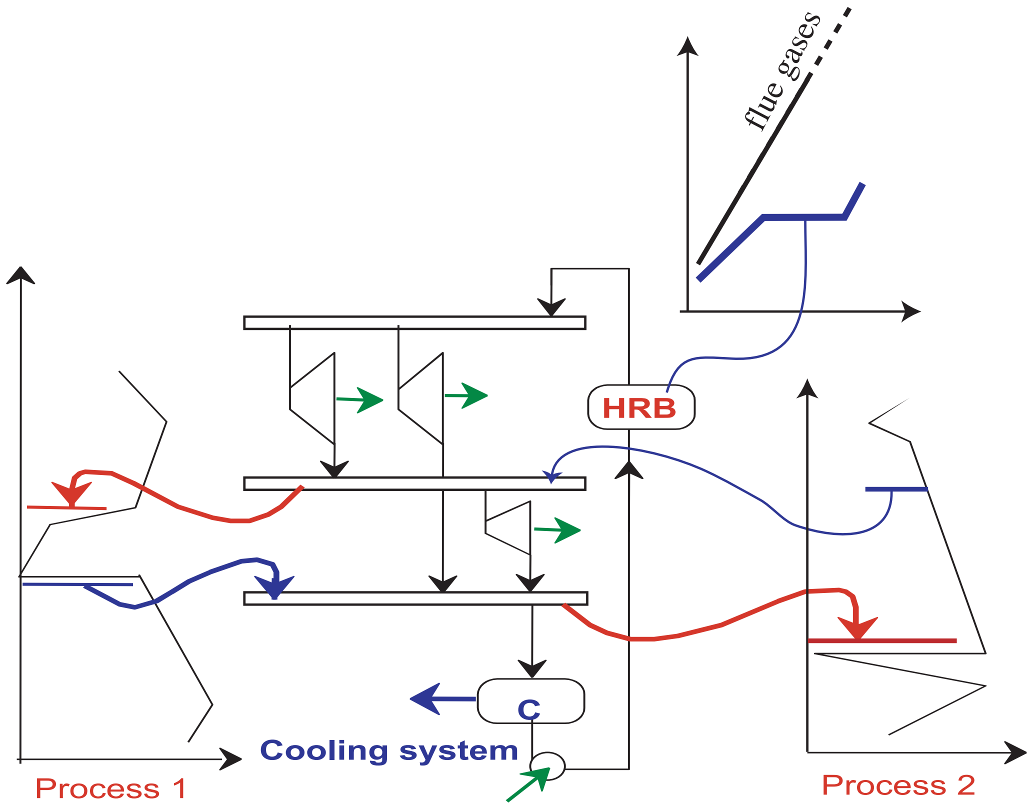


Figure 14: Steam distribution network as a way of realising process streams heat exchange and converting available exergy from a process

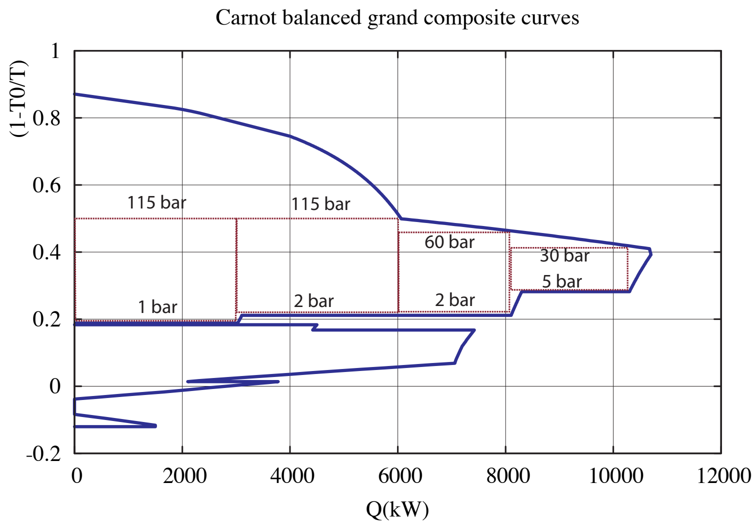


Figure 15: Choice of the optimal pressure levels

4 Methodology for designing integrated utility systems

The integration of the energy conversion system aims at defining the complete list of heat exchanging streams including the hot and cold streams of the utility sub-system, prior to any heat exchanger network design. The graphical method is limited to cases where utility streams do not interact. When utility streams interact as it is the case when steam is produced in a boiler, expanded in a turbine before being condensed to heat up process streams, the graphical method can not be used anymore since modifying the flow of one utility stream influences the flows of the other. In this case, optimisation methods will be used to determine the flowrates of the utility streams so that they will minimise the cost of the energy requirement and/or make the best use of the available exergy. The calculation of the best system will start with a utility system superstructure that will include the possible conversion technologies that are envisaged. Although it is possible to set up a generic problem that would state and solve the problem in an automatic manner, it is more convenient to proceed by successive iterations, keeping in mind that learning from one step will result in new problem definitions and perhaps news ideas for the integration of alternative energy conversion technologies and that the pinch point location is used to identify how to modify the process operating conditions.

From the analysis of the grand composite curve and the possible placement of heat pumping and combined heat and power units, one may define a utility system superstructure that includes the possible energy conversion technologies that are able to supply the energy requirement of the process. Knowing the temperatures, pressure, power and nominal flows of the energy conversion units, the optimal flowrate of each of the utility streams will be computed by solving the mixed integer linear programming (MILP) problem defined by Eq. 4. This method will select the equipment in the superstructure and determine their optimal operating flowrates in the integrated system. Two variables are therefore associated with any utility technology w : the integer variables y_w represent the presence of the technology w in the optimal configuration and f_w represents its level of utilisation ($fmin_w \leq f_w \leq fmax_w$). For this calculation, the objective function is the total cost including the operating and the linearised annualised investment costs, both expressed in Monetary Units (MU) /year.

$$\begin{aligned}\min_{R_r,y_w,f_w} \sum_{w=1}^{n_w}(C2_w f_w) + Cel^+\dot{E}^+_{grid} - Cel^-\dot{E}^-_{grid}) * t \\ + \sum_{w=1}^{n_w}(C1_w y_w) \\ + \frac{i(1+i)^{n_{years}}}{(1+i)^{n_{years}}-1}\sum_{w=1}^{n_w}(ICF_w y_w + ICP_w f_w)\end{aligned} \tag{4}$$

subject to

Heat balance of the temperature intervals

$$\sum_{w=1}^{n_w} f_w \dot{q}_{w,r} + \sum_{i=1}^{n} \dot{Q}_{i,r} + R_{r+1} - R_r = 0 \qquad \forall r = 1,...,n_r \tag{5}$$

Electricity consumption:

$$\sum_{w=1}^{n_w} f_w \dot{w}_w + \eta_d \dot{E}^+_{grid} - \dot{W}c \geq 0 \tag{6}$$

Electricity exportation

$$\sum_{w=1}^{n_w} f_w \dot{w}_w + \eta_d \dot{E}^+_{grid} - \frac{\dot{E}^-_{grid}}{\eta_g} - \dot{W}c = 0 \tag{7}$$

Other additional constraints

$$\sum_{w=1}^{n_w} a^i_x f_w + c^i_w y_w + \sum_{k=1}^{n_x} d^i_k x_k - b_i \gtreqless 0 \forall i = 1, ..., n_e \tag{8}$$

$$xmin_k \leq x_k \leq xmax_k \qquad \forall k = 1, ..., n_x \tag{9}$$

Existence of operation w during the time period p:

$$fmin_w y_w \leq f_w \leq fmax_w y_w \qquad {}^{\forall w=1,...,n_w}_{y_w \in 0,1} \tag{10}$$

Thermodynamic feasibility of the heat recovery and utility systems

$$Wel \geq 0, Wel_s \geq 0 \tag{11}$$

$$R_1 = 0, R_{n_{r+1}} = 0, R_r \geq 0 \qquad \forall r = 1, ..., n_{r+1} \tag{12}$$

with	y_w	the integer variable associated with the use of the technology w;
	R_r	the heat cascaded from the temperature interval r to the lower temperature intervals (r=1,$n_r + 1$) $[kW]$;
	n	the number of process streams;
	$\dot{Q}_{i,r}$	the heat load of the process stream i in the temperature interval r ($Q_{ir} > 0$ for hot streams) $[kW]$;
	$\dot{E}^-_{grid}$	the net production of electricity $[kW]$;
	$\dot{E}^+_{grid}$	the net import of electricity $[kW]$;
	η_d	the efficiency of the electrical drives in the plant;
	η_g	the efficiency of the electrical generator in the plant;
	$C1_w$	the fixed cost of using the technology w [€$/year$];
	$C2_w$	the proportional cost of using the technology w. This value is defined in [€$/s$];
	ICF_w	the fixed cost related to the investment of using technology w; ICF_w is expressed in monetary units [€] and refers to the investment cost of the combustion and cogeneration equipments as defined above as well as to the other equipments considered in the utility system (turbines, heat pumps, refrigeration systems,...);
	ICP_w	the proportional investment cost of the technology w, ICP_w [€] allows to account for size effect in the investment;
	n_w	the number of technologies proposed in the super configuration of the utility system;
	$\dot{q}_{wr}$	the heat load of the technology w in the temperature interval r for a given reference flowrate, $q_w r > 0$ for a hot stream$[kW]$;
	f_w	the multiplication factor of the reference flowrate of the technology w in the optimal situation;
	$\dot{w}_w$	the mechanical power produced by the reference flowrate of technology w; $\dot{w}_w < 0$ for a mechanical power consumer and > 0 for a producer $[kW]$;

Cel^+	the selling price of electricity [MU/kJ];
Cel^-	the electricity cost at import [MU/kJ];
t	the total annual operation time [$s/year$] ;
$\dot{W}c$	the overall mechanical power needs of the process; $Wc < 0$ if the overall balance corresponds to a mechanical power production [kW];
x_k	the (n_x) additional variables used in the additional equations of the technology models;
a^i_w, c^i_w	respectively the coefficients of the multiplication factor and the integer variables of technology w in the constraint i in the effect models;
d^i_k, b_i	respectively the coefficients of the additional variables and the independent term in the constraint i in the effect models;
$xmin_r, xmax_r$	respectively the minimum and maximum bounds of x_r;
$fmin_w, fmax_w$	the minimum and maximum values accepted for f_w.
i	the annualisation interest rate
n_{years}	the expected life of the equipment [year]

The method presented may be applied to any kind of energy conversion technologies. It is based on the assumption that the operating conditions have been defined for each of the equipment concerned and that only the flowrates are unknown. This is a limiting assumption but it allows to solve most of the problems of energy conversion integration mainly because non linearities may usually be solved by discretising the search space. The method has been further adapted to compute the optimal integration of steam networks, to incorporate restricted matches constraints, to integrate refrigeration cycles and Organic Rankine Cycles as well as heat pumps. It has been applied to integrate new technologies likes the partial oxidation gas turbine, or to design new type of power plants by introducing the concept of isothermal gas turbines.

4.1 Gas turbine and combustion system

In order to demonstrate the ability of the formulation to tackle complex problems, the model for computing the integration of gas turbines and combustion will be given in more detail . The purpose is to explain how to formulate the problem as a linear problem even if the models appear to be non linear. The model represents the integration of the gas turbine including its partial load operation, the possible post-combustion of the gas turbine flue gas, the use of different fuels in the gas turbine and in the post combustion, and of course the integration of conventional combustion in a radiative furnace with possible air enrichment or air preheating. The post combustion and the partial load models are required because there is no possibility of identifying a gas turbine model whose heat load will perfectly match the heat requirement of the process. The principle of the integration are illustrated on figure 16.

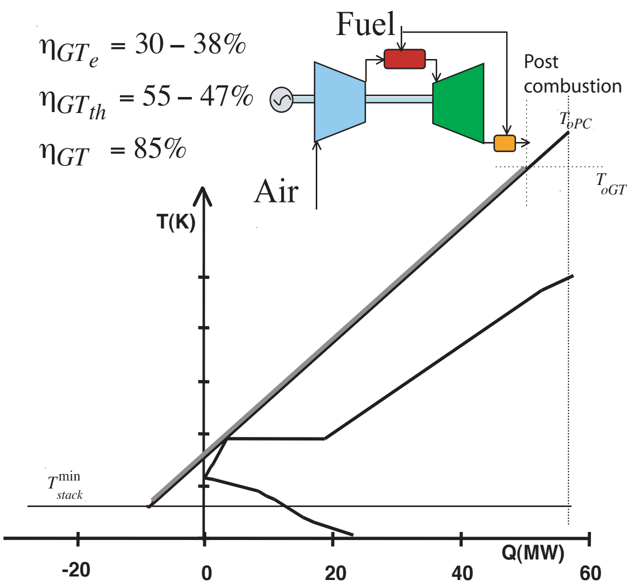


Figure 16: Integration of a gas turbine with post combustion to a process

The following integration constraints are added to the aforementioned problem. Hot stream corresponding to the flue gas of a gas turbine g :

$$Q_g^{gt} = f_g * \dot{m}_g * cpf_g * (TOT_g - Tstack_g) \qquad \forall g = 1, n_g \quad (13)$$

where

$\dot{m}_g$	is the flue gas flowrate at the outlet of the gas turbine g in nominal conditions. These values result from the simulation of the gas turbine g;
Q_g^{gt}	is the total heat load of the fumes from the gas turbine g;
cpf_g	is the mean cp of the flue gas at the outlet of the gas turbine g;
TOT_g	is the temperature of the flue gas of the gas turbine g;
$Tstack_g$	is the stack target temperature accepted for the outlet of the gas turbine g after heat recovery;
f_g	is level of utilisation of gas turbine g, $f_g^{min} * y_g \leq f_g \leq y_g * f_g^{max}$;
y_g	is the integer variable representing the use or not (1,0) of the gas turbine g;
$f_g^{min(max)}$	is the minimum (maximum) level of utilisation of the gas turbine g;
n_g	is the number of gas turbines proposed in the utility system super configuration.

Hot stream corresponding to the post combustion (heat available for convective heat exchange)

$$Q_g^{pc} = f_g^{pc} * \dot{m}_g * cpf_g * (Trad - TOT_g) \qquad \forall g = 1, n_g \quad (14)$$

where	$Trad$	is an arbitrary temperature used in the combustion model and representing the limit of the radiative exchange;
	f_g^{pc}	is the fraction of the nominal gas turbine flue gas flowrate used for post combustion;
	Q_g^{pc}	is the heat load supplied by the flowrate fraction of the flue gas flowrate of gas turbine g used in the post combustion device.

Fuel consumption in the gas turbine g

$$\sum_{c=1}^{n_{cgt}} f_c^g * LHV_c - (y_g * FCI_g + f_g * FCP_g) = 0 \qquad \forall g = 1, n_g \quad (15)$$

where	n_{cgt}	is the number of fuels available for combustion in the gas turbines;
	LHV_c	the lower heating value of the fuel c;
	f_c^g	the flowrate of the fuel c in the gas turbine g;
	$y_g * FCI_g + f_g * FCP_g$	is the linearised fuel consumption of gas turbine g as a function of its level of utilitisation.

Electricity production with the gas turbines W_{gt}

$$W_{gt} - \sum_{g=1}^{n_g} (y_g * WI_g + f_g * WP_g) = 0 \qquad (16)$$

where $y_g * WI_g + f_g * WP_g$ is the linearised mechanical power production of the gas turbine g as a function of its level of utilisation.
The parameters for the linearisation are computed by simulation considering the partial load operation of the gas turbine. For each gas turbine g, the unknowns are f_g, y_g, f_g^{pc} while the other parameters are obtained from the thermo-economic models. The quality of the linearisation will mainly depend on the range in which the partial load operation is expected to happen in the optimal situation.
The operating costs OC_{gt} and the investment costs IC_{gt} of the selected gas turbines are computed by :

$$\sum_{g=1}^{n_g} (y_g * OCI_g + f_g * OCP_g) - OC_{gt} = 0 \qquad (17)$$

$$\sum_{g=1}^{n_g} y_g * ICI_g - IC_{gt} = 0 \qquad (18)$$

where	$y_g * OCI_g + f_g * OCP_g$	is the linearised maintenance cost of gas turbine g as a function of its level of utilisation ;
	$y_g * ICI_g$	is the investment cost of gas turbine g from the data base catalog;

The fraction of the flue gas of the gas turbine used in the post combustion is limited to the level of utilisation of the gas turbine g.

$$f_g^{pc} \leq f_g \qquad \forall g = 1, n_g \quad (19)$$

The combustion model is made of different equations: (20) includes different terms representing the oxygen balance required by the combustion of the fuels and the oxygen supplied by air and post combustion flue gas.

$$\sum_{g=1}^{n_g} f_g^{pc} * \dot{m}_g * x_g^{O_2} + f_{air} * x_{air}^{O_2} + \sum_{a=1}^{n_a} f_a * \dot{m}_a * x_a^{O_2} - \sum_{c=1}^{n_c} f_c^c * \kappa_c^{O_2} \geq 0 \qquad (20)$$

where

$x_g^{O_2}$	is the oxygen content of the flue gas at the outlet of the gas turbine g
$x_{air}^{O_2}$	is the oxygen content of the ambient air
f_{air}	is the amount of air used by the combustion in the system
f_c^c	is the flowrate of fuel c used in combustion, ($f_c \leq f_c^{max}$), its specific cost is C_c
$\dot{m}_c^f$	is the fumes floware resulting from the combustion of fuel c
cp_c^f	is the mean specific heat of the fumes resulting from combustion. This cp is considered between $Trad$ and $Tstack$
n_c	is the number of fuels that can be used in the system including those for firing the gas turbine (n_{cgt})
$\kappa_c^{O_2}$	is the oxygen requirement per unit of fuel c. For practical reasons, the oxygen requirement includes the minimum oxygen excess for this fuel
$x_a^{O_2}$	is the oxygen content of the enriched air stream leaving the air separation unit a
$\dot{m}_a$	is the flowrate of enriched air leaving the air separation unit a in nominal conditions
f_a	is the level of utilisation of air separation unit a, $f_a^{min} * y_a \leq f_a \leq f_a^{max} * y_a$
y_a	is the integer variable representing the use or not (1,0) of the air separation unit a
$f_a^{min(max)}$	is the minimum (maximum) level of utilisation of the air separation unit a
n_a	is the number of air separation units considered in the system

Fuel consumption balance of any fuel c that might be used either in gas turbine either in standard combustion.

$$f_c^c + \sum_{g=1}^{n_g} f_c^g - f_c = 0 \qquad (21)$$

where f_c is the overall consumption of fuel c

High temperature balance : radiative exchange model above $Trad$

$$\sum_{c=1}^{n_c}(f_c*(LHV_c+(cp_{air}*\frac{\kappa_c^{O_2}}{x_{air}^{O_2}}*(Trad-T_0))))-f_{air}*cp_{air}*(Trad-T_0)$$
$$-\sum_{a=1}^{n_a}f_a*\dot{m}_a*cp_a*(Trad-TO_a)+Q_{prh}-Q_{rad}=0 \quad (22)$$

Low temperature balance : convective exchange below $Trad$

$$f_{air}*cp_{air}*(Trad-Tstack)+\sum_{c=1}^{n_c}(f_c*(\dot{m}_c^f*cpf_c-cp_{air}*\frac{\kappa_c^{O_2}}{x_{air}^{O_2}})*(Trad-Tstack))$$
$$+\sum_{a=1}^{n_a}f_a*\dot{m}_a*cp_a*(Trad-Tstack)-Q_{cnv}=0 \quad (23)$$

where
- Q_{rad} is the total amount of heat available above $Trad$
- Q_{cnv} is the total amount of heat available from $Trad$ to $Tstack$
- LHV_c is the lower heating value of the fuel c. This value is the value computed by simulation of the combustion using the minimum accepted value of the oxygen content in the fumes
- T_0 is the reference temperature used for computing the LHV
- T_{air}^{in} is the inlet temperature of air
- Q_{prh} is the heat load of air preheating, the existence of the air preheating equipment is defined by an integer variable y_{prh} and the following equation : $y_{prh}Q_{prh}^{min} \leq Q_{prh} \leq y_{prh}Q_{prh}^{max}$. The investment cost of the air preheating device is computed by linearising the air preheater cost by $IC_{prh} = ICF_{prh}y_{prh} + ICP_{prh}Q_{prh}$
- cp_a is the mean specific heat of the enriched air leaving unit a at a temperature of TO_a.

Table 11 gives the values needed to compute the integration of some typical fuels used in the industry including some renewable fuels (wood and biogas).

4.1.1 Air preheating : outlet temperature calculation

When combustion is considered, the air preheating plays the role of a chemical heat pump, it is used to pump waste heat available below the pinch point and make it available above the pinch point (by an increase of the adiabatic temperature of combustion). The effect of air preheating is however limited to the preheating of the stoechiometric air flow. When the flowrate is higher, the adiabatic temperature of combustion decreases and the benefit of air preheating is lost for the part corresponding the the excess air. When combined heat and power is considered using steam network, the process pinch point does not anymore define the maximum preheating temperature. In this case, the combustion air may be preheated up to the highest condensation pressure of steam because this steam will produce mechanical power before being used as a preheating stream. The preheating heat load will become available at a higher temperature to produce an additional amount of steam at the highest pressure or to increase the superheating temperature. The air preheating temperature is therefore unknown and its optimal value has to be computed.

When heat cascade is considered, computing the optimal preheating temperature is a non trivial task, mainly because the temperature is used to generate the list of the heat cascade constraints. This makes the problem non linear and discontinuous (i.e. according to the temperature the stream will appear or not in a given heat cascade constraint). Some techniques have been proposed to solve this problem as a non linear programming (in our case mixed integer) problem using smooth approximation techniques (e.g. [1]), this approach is explained in more details in another chapter. Another approach consists in keeping the linear programming formulation by discretizing the temperature range in which the air preheating will take place in n_i intervals of ΔT. The air preheating stream is therefore defined by a list of cold streams from T_i^{air} to $T_{i+1}^{air} = T_i^{air} + \Delta T$ and by adding the following constraints :

$$f_{air} \geq fa_i \qquad \forall i = 1, ..., n_i \tag{24a}$$

$$Q_{prh} = \sum_{i=1}^{n_i} fa_i cp_{air,i}(T_{i+1} - T_i) \tag{24b}$$

with fa_i the flowrate of air preheated from T_i to T_{i+1}.
$cp_{air,i}$ the specific heat capacity of the air flowrate between T_i to T_{i+1}.

In the combustion model, the optimal temperature calculation model is also used to compute the outlet temperature of the air and enriched air preheating, fuel preheating as well as to compute the outlet temperature at the stack. This calculation is made in two steps :

1. solve the model and compute the optimal flowrates in each interval (fa_i);
2. compute the resulting temperature To_{n_i} by solving from i =1 to n_i ,:
   $To_i = \frac{(fa_{i-1} - fa_i)To_{i-1} + fa_i * T_{i+1}}{fa_{i-1}}$
   with $To_0 = Ta_{in}$ the inlet temperature of the stream a.

The precision of the model is related to the size of the discretizing temperature intervals. A compromise between the precision required for the equipment sizing and the number of variables is therefore required. A similar formulation is also used to compute the optimal temperature of the gas turbine flue gas after heat recovery.

This systematic choice has been made to keep the robustness advantage of the MILP formulation.

4.2 Steam network

The steam networks play a very important role in most of the industrial process plants. They are the major interface between the utilities and the process streams while allowing the combined production of heat and power. Furthermore, by transferring heat from one process stream to another, the steam network will be used to reduce the energy penalty resulting from restricted matches. The importance of steam network reveals also for site scale process integration since the steam network will be the way to transfer heat from one process to another.

Targeting of the integration of the steam network is an important part of the integration of the energy conversion technologies. In the first attempt to study the integration of energy conversion technologies, the idea has been to consider steam as being a constant temperature stream that supplies or extracts heat from the process. This gave an easy way of understanding the multiple utilities integration from the Grand composite curve analysis. When designing the site scale steam networks, Dhole and Linnhoff [5] introduced the total site integration concept. They defined the total site composite curves to represent the integration of chemical plants that are composed of several processes and that may be integrated. The purpose of these curves is to identify the way energy has to be transferred from one plant section to another using the steam network. This method assumes that heat recovery inside the process has already been performed before allowing for exchanges between processes. The construction of the total site composite curve is explained on top of figure 17 for the integration of two processes whose Grand composite curves are given on the left. After eliminating the pockets (self sufficient zones), the hot utility and cold utility profiles are composed to build the hot and cold site profiles. The exchange between the processes is then realised using the steam network as heat transfer medium. If this approach is convenient from the graphical point of view, it can not be applied when considering the integration of steam networks in practice. Two major defaults should be removed. First, the pockets can not be ignored because these may hide heat exchange potentials. This is demonstrated in the centre of the figure where the integrated composite curves of process 1 versus process 2 shows the energy saving that could be obtained when exchanging heat between the two processes. In this example, the energy saving mainly results from the integration inside the pockets. In reality, the hot and cold composite curves of the whole system should be considered. The second drawback of the total site approach is that it ignores the combined production of mechanical power in the steam network. This is shown in the bottom of the figure where steam is produced at high pressure in process 2 to be used at high pressure in process 1 and low pressure steam produced in process 1 is used at low pressure in process 2. In between, the steam is expanded in a back pressure turbine to produce mechanical power. From the exergy analysis, the potential for combined production of mechanical power is proportional to the size of the pockets in the exergy Grand composite curve. These can not therefore be ignored from heat integration and CHP perspective. In real steam networks, the steam production and condensation cannot be considered at a constant temperature. One should consider the preheating and the superheating for the steam production and the desuperheating, condensation, liquid undercooling for the steam condensation. Furthermore, the maximisation of the mechanical power production will be obtained by optimising the heat exchanges within the steam network, e.g. by condensing low pressure steam to preheat high pressure water of the steam network. The MILP formulation presented here above may be extended to define a more precise model of the steam network. The formulation has been given in [12]. It is based on the definition of a steam network superstructure (figure 18). This superstructure has been first proposed by Papoulias and Grossmann [14]. It has been adapted to account for the temperature-enthalpy profiles of the steam production (i.e. preheating, vaporisation and superheating) and consumption (desuperheating, condensation and undercooling). One of the difficulties of this formulation has been to guarantee a coherency between the heat and the mass balances while using linear equations. This has been obtained by a special formulation of the mass balances of the steam network headers. Hot and cold streams of the steam network are considered in the heat cascade model and the contributions of the steam expansion in turbines are added in the mechanical power balances. It should be mentioned that using this model, the optimal flowrate in the steam network will be determined considering also the possibility of exchanging heat between the streams of the steam network. This leads to an optimised steam network configuration where steam draw offs are used to preheat the high pressure water. The model may therefore be used as a tool for a rapid prototyping of complex steam cycles in conventional power plants [9].

The integrated composite curves of the steam network is used to represent the results of its integration. On the figure, the overall mechanical power production corresponds to the balance between the hot and cold streams of the system. It is made of the contribution of the expansions between the different pressure levels in the superstructure. A post processing analysis will be required in order to decide the best configuration of the steam turbine(s) : single turbine with multiple draw off or multiple back-pressure turbines. The use of the process pinch point as a

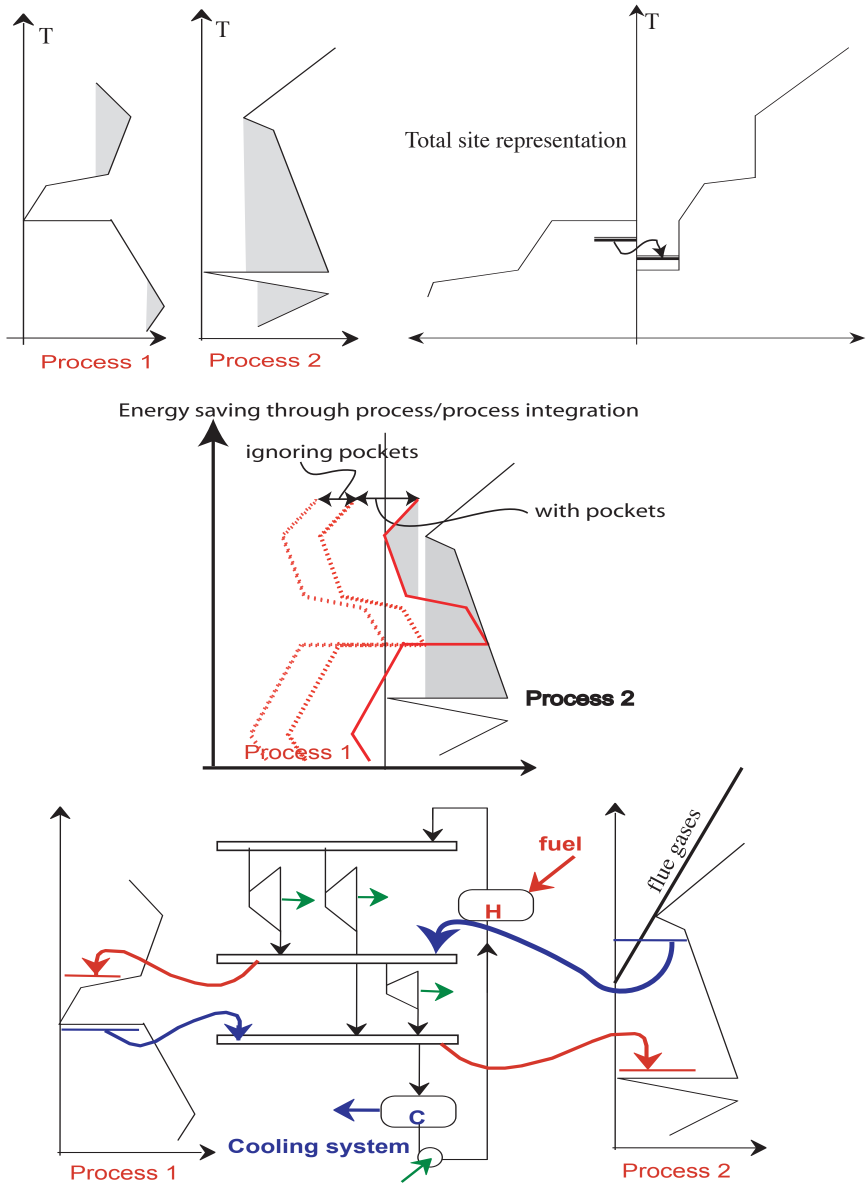


Figure 17: Total site integration and steam network

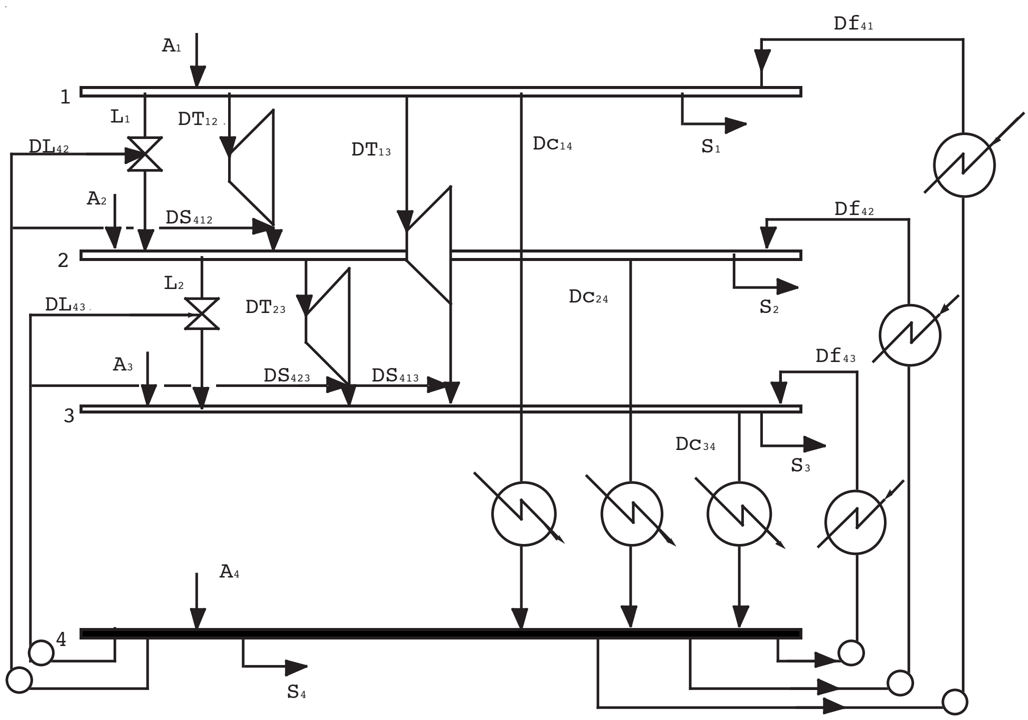


Figure 18: Superstructure of a steam network including 3 production/usage levels and one condensing level (deaerator)

reference to locate the zero heat (temperature axis) is used to verify the appropriate placement of the steam headers and the combined production of mechanical power. Combined heat and power is well located when it takes heat above the pinch point and sends it above or when it takes heat below and sends it to the cold utility [16]. When the combined heat and power production satisfies the rules for appropriate placement above the pinch point, the part of the integrated composite curve of the steam network that will appear in the left side of the temperature axis should correspond to the mechanical power production. This would indicate the the additional energy required by the system is equal to the mechanical power produced. If not, the rules for appropriate placement are not satisfied and it is recommended to understand the reason for this penalty.

4.3 Refrigeration cycles

The refrigeration cycles are used as a cold utility below the ambient temperature. The major principle of the refrigeration cycle is to use the compression power to change the temperature level of the streams. A simple refrigeration cycle is presented on figure 19, it is composed of one compressor, one evaporator (at low temperature), one condenser (at higher temperature) and a valve. From the process integration point of view, the refrigeration cycle is defined by one hot and one cold streams and by the corresponding mechanical power consumption. The optimal flowrate will be determined by the MILP formulation presented above. The temperature levels obtained from the Grand composite curve analysis will usually define the type of fluid to be used but other considerations will have to be taken into account like the environmental aspects (CFC refrigerants) or safety (flammability). The use of fluids already in use in the process plant is also an important criteria. The efficiency of the integration of one refrigeration cycle depends on its compression ratio and on the flowrate. It will depend also on the structure of the cycle and on the possibility of combining cycles with different refrigerants or with different pressure levels. The problem is therefore highly combinatorial since refrigerants, structures, pressure levels and flowrates have to be optimized. A graphical approach based on the exergy analysis has been proposed in [10]. This approach illustrates the methodology of the integration of complex refrigeration systems. A non linear programming model has been proposed in [4]. The method presented by Marechal and Kalitventzeff [13] shows the extension of the MILP formulation to tackle this complex problem. The method first identifies the most important temperature levels in the Grand composite curve using a MILP formulation. The systematic integration of the cycles with the possible refrigerants is made applying heuristic rules. From this first selection, the remaining cycles for which the refrigerants, the configuration, the temperature levels and the mechanical power are known are added in the energy conversion system superstructure and the best configurations are sorted out by solving the MILP problem. When several cycles compete, integer cuts constraints are added to the problem to systematically generate ordered set of solutions. The integer cut constraint is used to avoid the generation of an already known solution when solving the MILP problem. The restriction of the kth solution is obtained by adding the following constraint

$$\sum_{w=1}^{n_w} (2 * y_w^k - 1) * y_w \leq (\sum_{w=1}^{n_w} y_w^k) - 1 \qquad \forall k = 1, ..., n_{sol} \tag{25}$$

with y_w^k the value of y_w in the solution k
n_{sol} the number of solutions

The use of an integer cut constraint is an important tool when solving utility system integration. The systematic generation of multiple solutions allows the comparison of the proposed utility system configurations using different criteria (not accounted in the definition of objective function) and to perform a sensitivity analysis to uncertain problem parameters like the cost of energy or the investment.

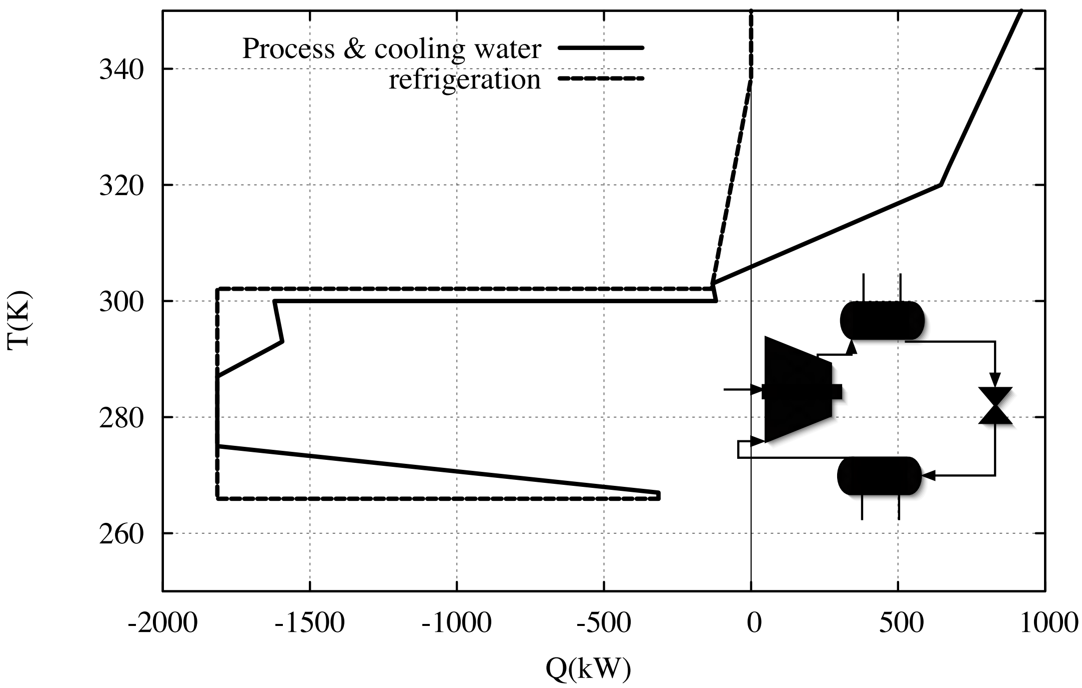


Figure 19: Integrated composite curve of a single stage refrigeration cycle

4.4 Heat pumps

4.4.1 Heat pump description

Mechanical compression cycle heat pumps The most common heat pump is the mechanical compression cycle illustrated in Figure 20. A fluid (typically a refrigerant) is evaporated by cooling a process stream ($\dot{Q}^{(-)}$). Using the mechanical power ($\dot{W}^{+}$), the pressure is changed and the heat is send back by condensing the evaporated fluid at a higher pressure and temperature. Considering the temperature of evaporation ($T^{(-)}$) and of condensation ($T^{(+)}$), the mechanical power may be approximated by applying an efficiency (η_{Carnot}) to the reversible work of heat pumping (26).

$$\dot{W}^{+} = \dot{Q}^{(-)} \frac{(T^{(+)} - T^{(-)})}{T^{(-)}.\eta_{Carnot}} \tag{26}$$

The first thermodynamic principle indicates us $\dot{Q}^{(+)} = \dot{W}^{+} + \dot{Q}^{(-)}$.

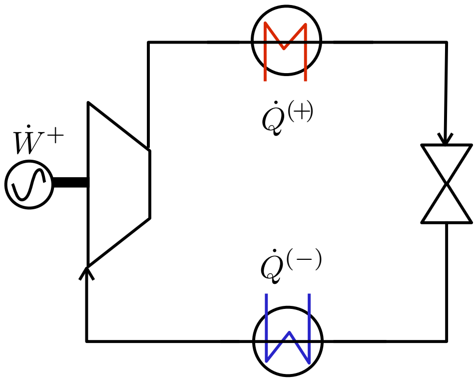


Figure 20: Representation of a compression heat pump

Mechanical vapour recompression Heat pumping effect can also be obtained by mechanical vapour recompression. A hot stream initially to be condensed below the pinch temperature will be relocated partly in the heat sink by using a compressor that will raise the condensing temperature above the pinch point. As the vapour come from the process, such a system is called open or semi-open system (depending on where this vapour is condensed).

Absorption heat pumps and heat transformers Absorption heat pumps and heat transformers proceed with three streams. Absorption heat pumps (Figure 21) have two cold streams and one hot stream. The cold streams at low temperature will take heat in the exothermic subsystem below the pinch temperature. The heat pumped will be recovered above the pinch point in the condenser and the absorber while a cold stream corresponding to the boiler at high temperature will drive the heat pumping effect. The benefit of the absorption heat pump is therefore the heat load pumped below the pinch. When optimising the flowrate of the absorption heat pump, there will be three potential pinch points, each one associated to the evaporation and the condensation levels of the heat pump. It should be noted that absorption heat pumps may profit from self-sufficient zones above the pinch point to maximise their flowrate and therefore the amount of heat pumped in the system. In heat transformers (Figure 22), the cold streams are at the medium temperature while one hot stream will be at low temperature and one at high temperature. Therefore the optimal integration of the heat transformer will be obtained by driving the heat below the pinch temperature to relocate part of the heat in the high temperature hot stream above the pinch point. The heat transformer will maximise its flowrate by using the energy of self-sufficient zones below the pinch point.

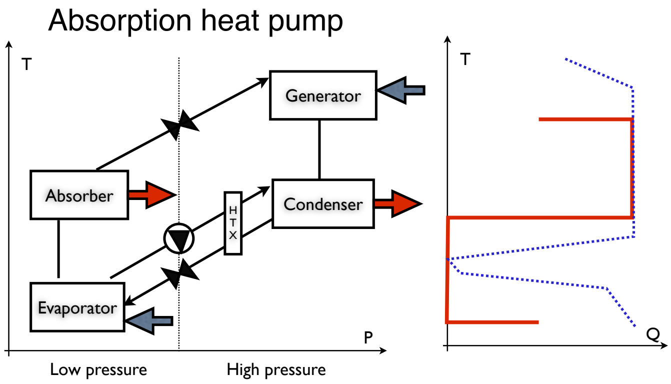


Figure 21: Absorption heat pump

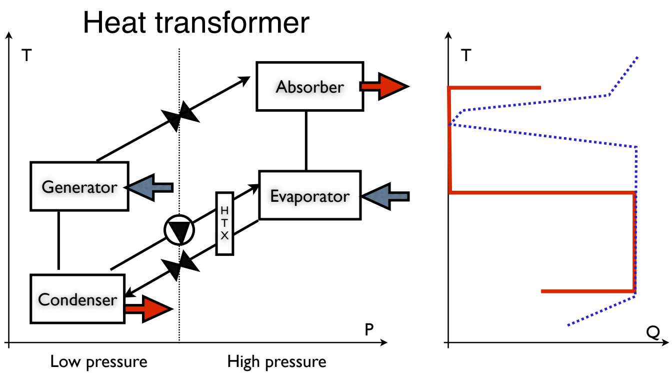


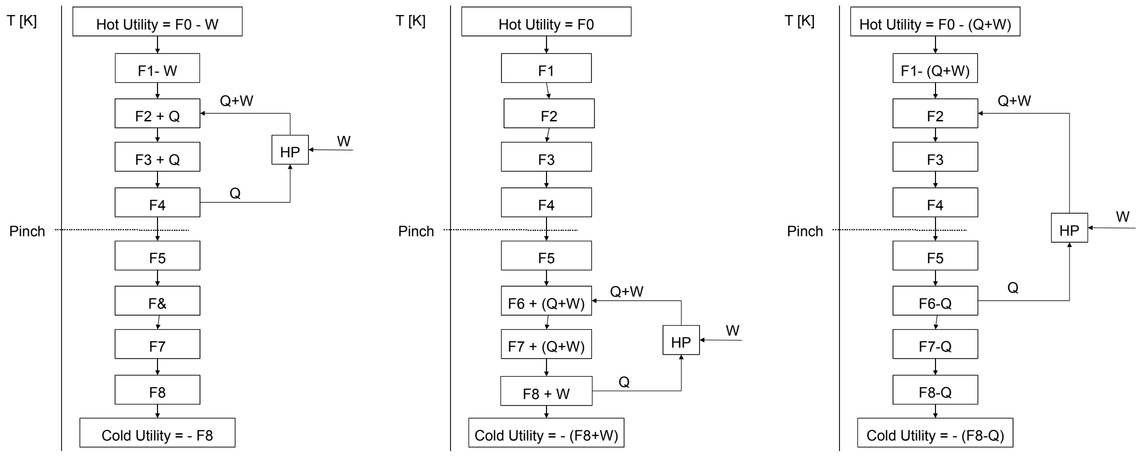
Figure 22: Heat Transformer

4.4.2 Integration

The advantage of a heat pump is the fact that is modifies the temperature level of a heat source to make it available at a higher more useful temperature. A heat pump will therefore be attractive when the heat source is a free source (e.g. the environment or waste heat) and when the heat can be delivered to satisfy an energy requirement of the process.

Pinch analysis identifies the possible heat recovery between the hot and the cold streams. It also defines the enthalpy-temperature profile of the process heat source and the process heat sink. The heat source will become the cold source of the heat pump (i.e. the hot stream of the evaporation) and the heat sink profile defines the energy requirements of the process that defines the hot source (cold stream of the condenser). From the definition of the "plus-minus" principle, it can be seen that the only feasible possibility for appropriately integrating a heat pump in the system with a heat exchanger network is to introduce a new cold stream below the pinch temperature. This stream will receive heat from a hot stream of the heat source sub-system and send the heat back after compression in the heat sink sub-system by introducing a new hot stream above the pinch point.

In figure 23, demonstration is made of correct heat pump placement rule.



(a) (b) (c)

Figure 23: Heat pump placement

case a It is indeed equivalent to an electric heater since both $\dot{Q}^{(+)}$ and $\dot{Q}^{(-)}$ concern the same sub-system, the difference $\dot{W}^{+}$ being the only energy input in the sub-system.

case b The electric power of the heat pump is added to the exothermic sub-system, therefore it will just increase the cooling requirement of the system.

case c This result to a net hot utility recovery of $\dot{Q}$ +W to find zero flow at the pinch point, and a cold utility recovery of $\dot{Q}$ at the expense of $\dot{W}^{+}$ of mechanical power.

The use of mechanical power to rise heat in the thermal cascade is energetically interesting when it rises heat from below to above the pinch point.

This means that a heat pump may be profitable in one process configuration and not when the same process is considered as integrated in the production site.
The optimal integration of the heat pump system will try to maximise the heat load ($\dot{Q}^{(-)}$), while minimising the mechanical power ($\dot{W}^{+}$). The temperature lift has therefore to be minimised. The mechanical power of the heat pump can be estimated using the Carnot factor computed from the temperatures of $\dot{Q}^{(-)}$ and $\dot{Q}^{(+)}$, therefore the integration of the heat pump may be calculated by Eq. 27.

$$\dot{Q}^{(+)}_{r^{(+)},r^{(-)}} = \min((\min_r R_r, \forall r = n_r+1, ..., r^{(+)}), ((1+\frac{(T^{(+)} - T^{(-)})}{T^{(-)}.\eta_{Carnot}})(\min_r R_r, \forall r = r^{(-)}, ..., 1))) \tag{27}$$

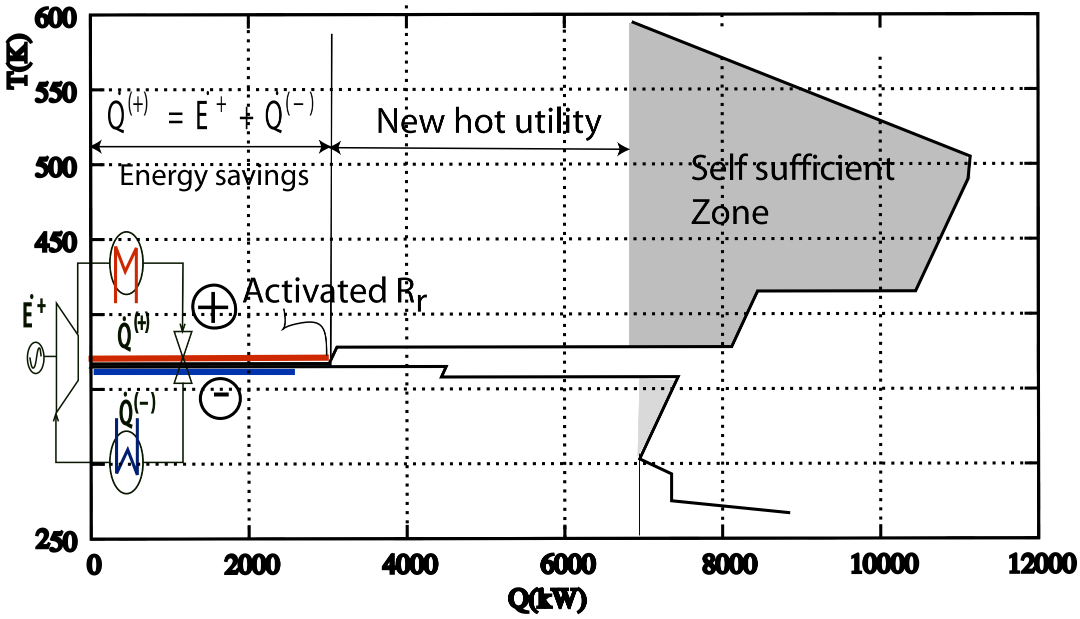


Figure 24: Representation of a heat pump integration

with

η_{Carnot}	the efficiency of the heat pump with respect to the reversible heat pump
$T^{(+)}$	the temperature of the hot stream of the heat pump that supplies heat to the process at the temperature $T_{r(+)}$ of the heat cascade
$T^{(-)}$	the temperature of the cold stream of the heat pump that takes heat from the process at the temperature $T_{r(-)}$ in the heat cascade

In reality, the hot and cold streams in the condenser and the evaporator do not have a constant enthalpy-temperature profile. The equation 27 will therefore be adapted to account for such heat transfer profiles. In such situation, more detailed models applying linear programming methods (e.g. Eq. 4) will be used.
An example of heat pump integration is shown on figure 24.

4.5 Handling non linear cost functions

In the problem formulation, linear cost are needed. When the whole range of sizes is covered by the model, a piece-wise linearisation technique may be used to represent non linear cost function. The generic investment cost function $C(S) = C_{ref} * (\frac{S}{S_{ref}})^b$ will be approximated by a set of segments (figure 25) defined by the following set of constraints (28) in the linear optimisation

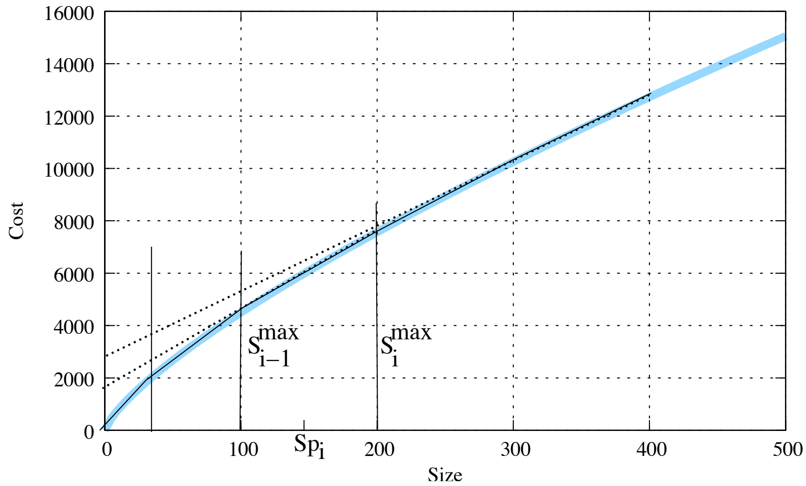


Figure 25: piecewize linearisation of cost function (exponent 0.75)

problem.

$$
\begin{aligned}
&C = \sum_{i=1}^{n_{segments}} \{(C(S_{i-1}^{max}) - \frac{C(S_{i-1}^{max}) - C(S_{i}^{max})}{S_{i-1}^{max} - S_{i}^{max}} * S_{i-1}^{max}) * y_i + \frac{C(S_{i-1}^{max}) - C(S_{i}^{max})}{S_{i-1}^{max} - S_{i}^{max}} * Sp_i\} \\
&S = \sum_{i=1}^{n_{segments}} Sp_i \\
&y_i * S_{i-1}^{max} \leq Sp_i \leq y_i * S_{i}^{max} \qquad \forall i = 1, ..., n_{segments} \\
&y_i \in \{0, 1\}
\end{aligned}
\tag{28}
$$

with	$C(S)$	the installed cost of the equipment of size S
	S_i^{max}	The maximum size in segment i
	Sp_i	the size of the equipment in segment i
	y_i	the integer variable used to select the segment i

The linearization by segments is also applicable for performances indicators of technologies like power and efficiency of a gas turbine. The linear formulation may in this case be extended to account for piece-wice linearized of non linear functions.

5 Using the exergy depletion as the objective function

Due to the linear nature of the problem formulation, the use of the energy cost as an objective function may reveal some difficulties. When the cost of fuel and electricity is such that the electrical efficiency of a cogeneration unit is attractive without the use of heat (i.e. when Eq. 29

holds), there is an economical interest to produce electricity even without cogeneration. In this case, the linear programming procedure leads to a situation where the cogeneration unit is used at its maximum. This situation usually does not occur when the investment costs are properly accounted in the objective function and when the cost of the different forms of energy are coherent with respect to the electrical efficiency. Nevertheless, the relative price of the different forms of energy will influence the technology selection and their level of usage in the integrated system.

$$\eta_{el} = \frac{\dot{E}^-_{grid}}{\dot{M}^+_{fuel}\Delta h^0_{fuel}} \geq \frac{c^+_{fuel}}{c^-_{grid} + \frac{1}{\tau}\frac{inv_{unit}}{t}} \tag{29}$$

with

c^+_{fuel}	[MU/kJ]	the specific cost of the fuel used in the cogeneration unit
c^-_{grid}	[MU/kJe]	the specific cost of the electricity sold to the grid
inv_{unit}	[MU/kWe]	the specific investment of the cogeneration unit computed per unit of power delivered
t	[s]	the annual operation time of the cogeneration unit
$\tau = \frac{(1+i)^{n_{years}}-1}{i(1+i)^{n_{years}}}$	[year]	is the annualising factor of the investment

The minimisation of the exergy depletion (Eq. 30) is an alternative way of formulating the objective function of the optimal energy conversion units integration. We will define the exergy depletion ($\dot{L}$) as being the sum of the exergy destruction in the system and the exergy losses released to the environment. Therefore, we will seek here for the maximum exergy recovery regardless of the investment in the energy conversion units.

$$\underset{\dot{R}_r, y_w, f_w}{Min} \sum_{w=1}^{n_w} \dot{L}_w = \sum_{w=1}^{n_w} (f_w * (\sum_{f=1}^{n_{fuel,w}} \dot{m}_{f,w}\Delta k^0_f + \dot{e}^+_w - \sum_{r=1}^{n_r} (\dot{e}q^-_{w,r})_{\Delta T_{min}} - \dot{e}^-_w)) \tag{30}$$

with

$\dot{e}_w^+$	the specific consumption of electricity of the energy conversion unit w
$\dot{e}_w^-$	the specific production of electricity of the energy conversion unit w
$\sum_{f=1}^{n_{fuel,w}} \dot{m}_{f,w} \Delta k_f^0$	the exergy consumed as fuel resources to produce the hot and cold streams of the energy conversion unit w in nominal conditions;
ns_w	the number of streams of unit w;
$(\dot{e}q_{w,r}^-)_{\Delta T_{min}}$	the heat exergy supplied by the hot and cold streams of the conversion unit w in the temperature interval r in its nominal conditions. $(\dot{e}q_{w,r}^-)_{\Delta T_{min}}$ is given by eq. 31. For this calculation, the temperatures used are the corrected temperatures, therefore, $(\dot{e}q_{w,r}^-)_{\Delta T_{min}}$ includes the exergy destruction due to the stream's contributions $(\Delta T_{min}/2_s)$ to the ΔT_{min} assumption.

$$(\dot{e}q_{w,r}^-)_{\Delta T_{min}} = \sum_{s=1}^{ns_w} \dot{q}_{s,r}^-(1 - \frac{T_0}{T_{lmr}^*}) \tag{31}$$

where T_{lmr}^* is the logarithmic mean temperature of interval r
$T_{lmr}^* = \frac{T_{r+1}^* - T_r^*}{ln(\frac{T_{r+1}^*}{T_r^*})}$ when $T_{r+1}^* \neq T_r^*$ and $T_{lmr}^* = T_r^*$ otherwize

By balance, the term $\dot{L}_w$ represents the exergy depletion related to the production of the hot and cold streams of the unit w, it includes the exergy destroyed in the conversion process of unit w and the exergy losses released to the surroundings.

Using this formulation, it is possible to define the set of energy conversion technologies that minimises the exergy depletion of the system. Considering the exergy made available for heat transfer by the hot and cold streams of the process and the exergy value of the fuel consumed in the system to satisfy the process requirements and possibly export electricity to the grid, the efficiency of the energy conversion system (η_{ex}) may be defined by Eq. 32.

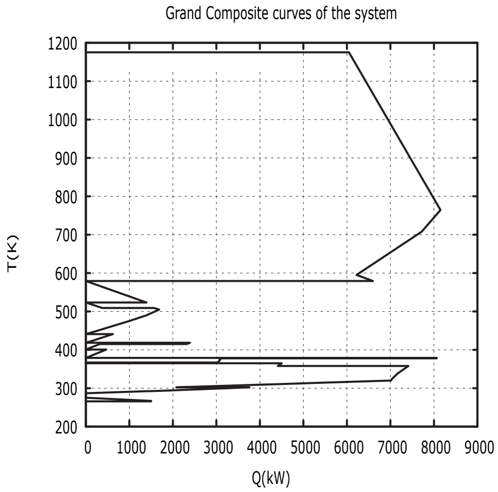
$$\eta_{ex} = \frac{\dot{E}q_{cold_a} + \dot{E}q_{hot_r} + \dot{E}_{grid}^-}{\dot{E}_{grid}^+ + \sum_{w=1}^{n_w} (f_w * (\sum_{f=1}^{n_{fuel,w}} \dot{m}_{f,w} \Delta k_f^0)) + \dot{E}q_{cold_r} + \dot{E}q_{hot_a}} \tag{32}$$

with
- $\dot{E}q_{cold_a}$ the exergy required by the cold streams of the process above the ambient temperature;
- $\dot{E}q_{hot_r}$ the exergy required by the hot streams of the process below the ambient temperature;
- $\dot{E}q_{cold_r}$ the exergy delivered by the cold streams of the process below the ambient temperature;
- $\dot{E}q_{hot_a}$ the exergy delivered by the process hot streams above the ambient temperature.

The optimisation problem formulation above does not consider the investment of new equipment. It would be possible to introduce the aspects related to the investment by adding a grey exergy term in the calculation of $\dot{L}_w$.

6 Representing the integration of the utility system

The composite curves resulting from the integration of the energy conversion units are known as the **balanced composite curves**. An example of such curves is given on figure 26. This representation is characterised by a number of pinch points, one being the process pinch point, the others corresponding to the maximum use of the cheapest utility to satisfy the process requirement. Composite curves are used to represent the heat transfer in the system, the temperatures being a good geographical identifier in complex systems. By the presence of multiple pinch points, the balanced composite curves are however difficult to read.



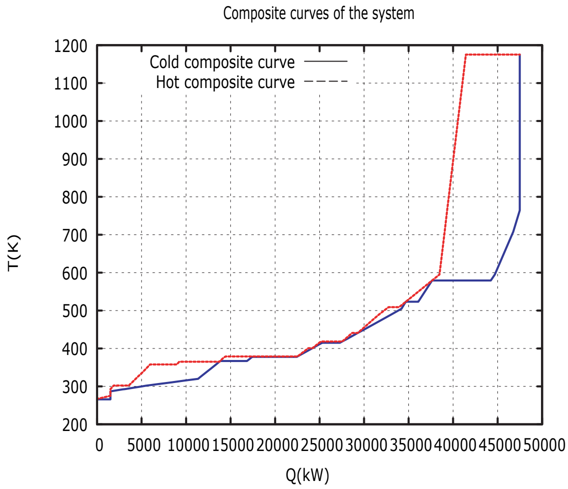


Figure 26: Balanced composite curves of the process with boiler, steam and cooling system

The **integrated composite curves** will give a better visualisation of the system integration. Their aim is to represent the optimisation results in order to understand the integration of sub-systems. This representation is based on the definition of sub-systems : processes, boiler house, refrigeration cycle, steam network, heat pump, utility system,... even very detailed sub-systems may be considered like one or several existing heat exchangers.

The integrated composite curves of a sub-system are obtained by subtracting from the grand composite curve of the overall system, the grand composite of the sub-system under study. The next step is to mirror the sub-system curve. The two curves intersect at the pinch points of the balanced composite curves. In order to locate the Y zero axis, it is convenient to consider the process pinch point location as a reference. From a mathematical point of view, the integrated composite curves are computed using Eq. 33 and 34.

The set of streams is divided into two sub-sets. The first (sub-set A) defines the sub-system whose integration should be visualised, the second (sub-set B) being formed by all the other streams. The sub-set B will be referred to as being the reference set.
The sub-set B will be represented by a curve (RB_k,T_k^*), where T_k^* is the kth corrected temperature of the heat cascade and RB_k is computed by :

$$RB_k = R_{ref} + \sum_{r=k}^{n_k} \sum_{s=1}^{N_B} Q_{sr} \qquad \forall k = 1, ..., n_k \tag{33}$$

with	N_B	the number of streams in sub-system B;
	RB_k	the heat cascaded at the corrected temperature T_k^* in the sub-set B ;
	Q_{sr}	the heat load of stream s that belongs to sub-set B in the temperature interval r
	R_{ref}	the enthalpy reference that defines the position of the temperature axis (see below).

The opposite curve (RA_k,T_k^*), corresponding to the sub-set A, is computed to make the balance between the sub-set B and the grand composite curve . It defines the integration of the streams of sub-set A with the others (reference sub-set B):

$$RA_k = R_{ref} - R_{n_k+1} - \sum_{r=k}^{n_k} \sum_{s=1}^{N_A} Q_{sr} \qquad \forall k = 1, ..., n_k \tag{34}$$

with	N_A	the number of streams in sub-set A;
	R_{n_k+1}	the additional energy that can not be provided by the proposed utility set.

The R_{n_k+1} has been introduced to obtain a general definition, when the utilities are well integrated $R_{n_k+1} = 0$.
The value R_{ref}, that appears in the definition of the two curves, defines the position of the temperature axis on the energy axis. The value of R_{ref} is computed by considering that the curve of set B (the reference set) will intercept the temperature axis at the process pinch point temperature ($T_{k_p}^*$). The latter being identified by computing the heat cascade where only the process streams are considered. If k_p refers to the process pinch point, R_{ref} is computed by Eq.35

$$R_{ref} = \sum_{r=k_p}^{n_k} \sum_{s=1}^{N_B} Q_{sr} \tag{35}$$

Using this definition, the temperature axis divides the energy range into two parts, the positive values correspond the energy concerned with the set B integration, while the negative values refer to the energy involved in the set A integration. The utility composite curves are used to visualise the way the heat of the utility system is transfered to satisfy the requirements of the process.
When using the Carnot factor as Y-axis, the area between the two curves is proportional to the exergy destruction in the system. Figure 27 compares two utility systems used to satisfy the requirements of a process. On the left, the use of combustion, cooling and refrigeration system shows a much higher exergy destruction in the heat transfer (area between the two curves) than the figure on the right that represents the integration of an energy conversion system that includes three heat pumps, one steam cycle with 3 pressure levels, one refrigeration cycle and a cooling water system. The system on the left has an efficiency (η_{ex} in Eq. 32) of 35% while the one on the right has an efficiency of 72.6%.

> One heuristic rule resulting from the exergy analysis is to try to reduce to a minimum the area between the hot and cold composite curves of the integrated systems including the energy conversion system.

The integrated composite curves representation is especially useful when analysing the integration of cycles. In this case, the difference between the hot and the cold streams corresponds to the mechanical power that closes the energy balance. The representation will therefore be used to verify the integration of steam networks or refrigeration cycles and to confirm the appropriate placement of the cycles in the process integration.

The Figure 28 shows, for example, the integrated composite curve of a steam network. It should be mentioned that the choice of the process pinch point location as a reference for locating the temperature axis allows to verify the appropriate choice of the steam pressure levels. The energy balance of the hot and cold streams of the steam network is the net mechanical power production. When appropriately placed, it corresponds to a supplement of energy to be supplied to the process. The fact that it appears on the left of the temperature axis proves that the steam network characteristics are appropriate for the optimal production of mechanical power. In the Carnot factor ordinate, the area between the two curves gives an indication of the quality of the exergy use of the corresponding sub-system.

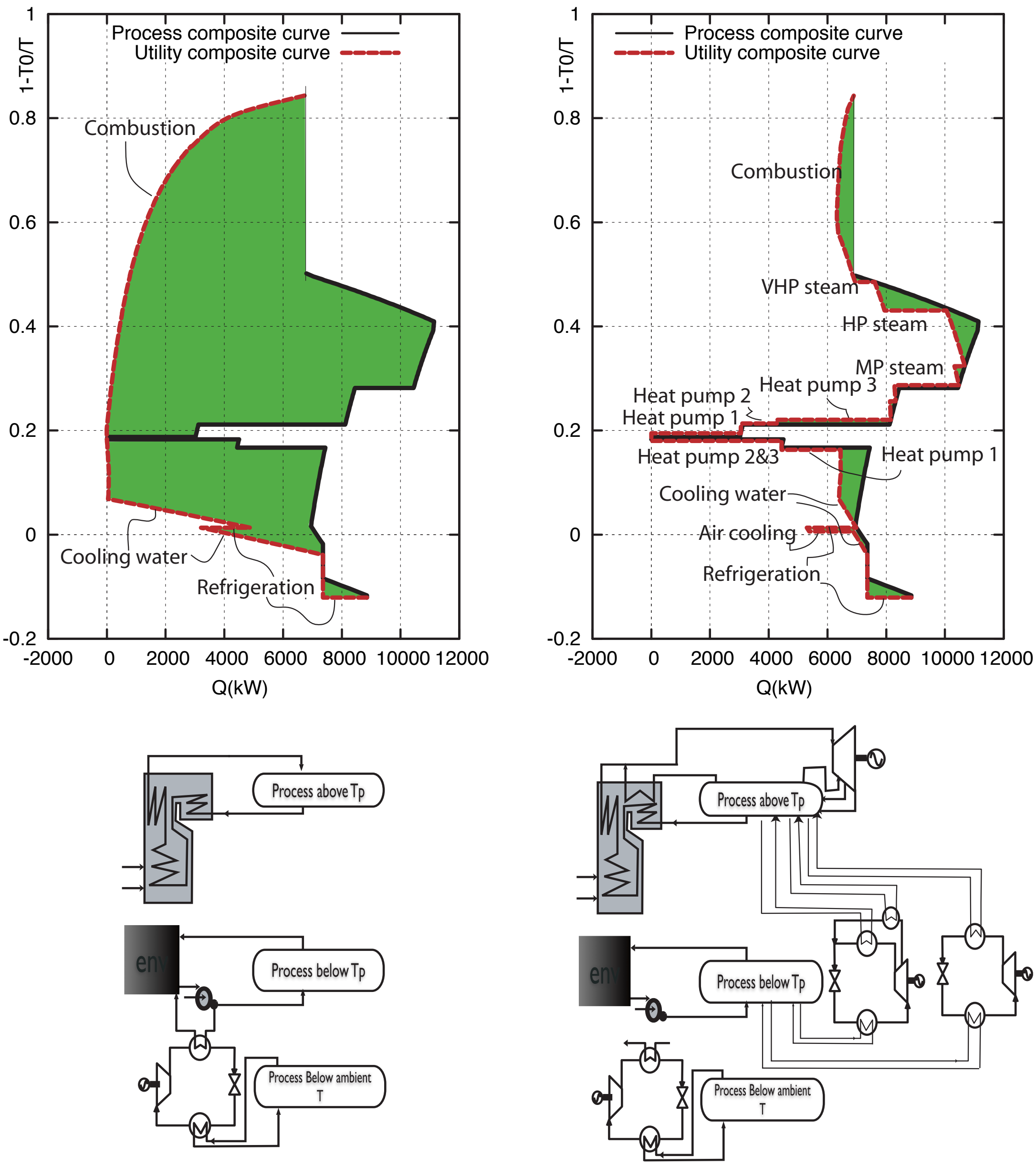


Figure 27: Carnot integrated composite curves of the utility systems

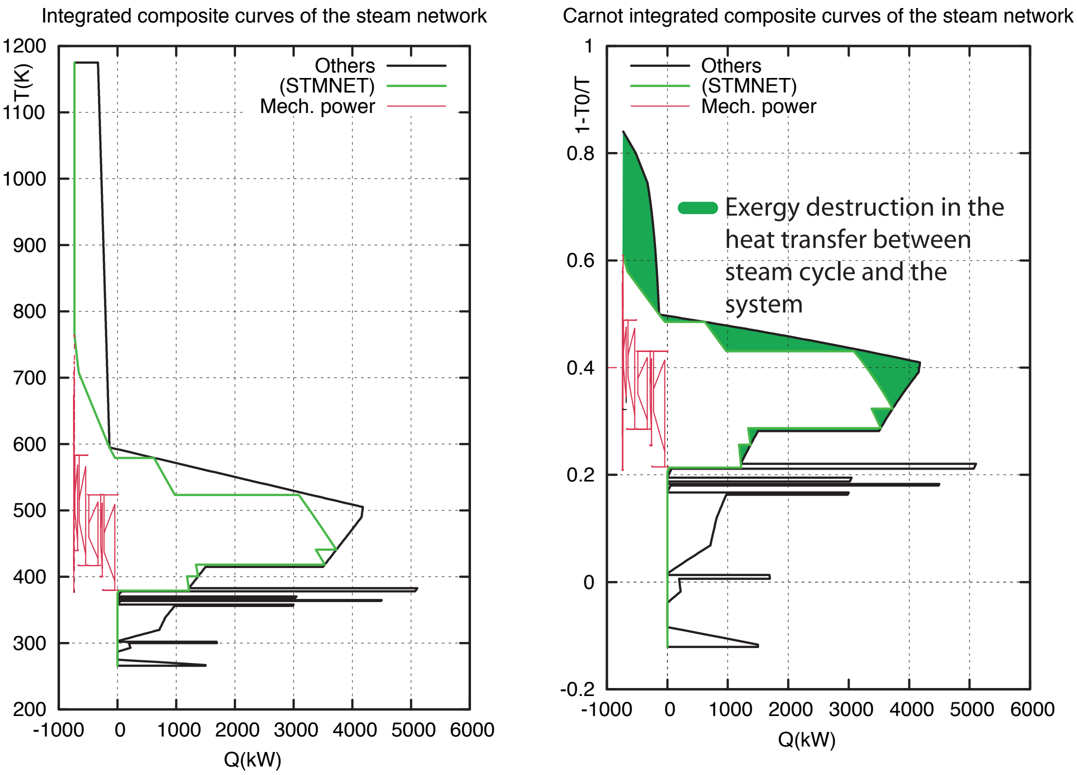


Figure 28: Integrated composite curves of the steam network

7 Final remarks concerning the process optimisation and the utility system integration

Pinch analysis theory has been developed on the basis of graphical techniques and mainly deals with energy requirementss. The integration of the utility system requires new methods that remain based on the same principles, however the use of a computer aided method allows to better represent the complex interactions in the system.

The computer aided methodology for integrating the energy conversion system (utility) in chemical production sites combines the use of graphical techniques and mixed integer optimisation. Considering that the problem formulation is not always known from the beginning, it is important to use such methods as a support to engineer's creativity rather than as a push-button method. The MILP (Mixed Integer Linear Programming) technique is a robust problem formulation and solving method for the process engineers. Many complex problems in process design and operation can be formulated and efficient solutions may be identified. The utility integration allows to capture in a easy and quick way the major aspects of the integration and to identify the most important options while eliminating the less attractive ones.

The use of a computer aided methodology changes the way the problem is solved and the way the tools will be used. A three steps procedure is applied :

Analyse Using the analysis of the composite curves and the identification of the pinch point, several process improvement options may be proposed. In the same time, the analysis of the grand composite curve and the Carnot composite curves will be applied to define the options for polygeneration (combined heat and power, refrigeration and heat pumping) as well as to identify the purchased energy conversion units.

Generate Together with the process modification options, the energy conversion system superstructure will be defined and a mixed integer mathematical programming formulation will be used to select the most profitable options that will define the final system configuration. If the problem is typically solved as a linear problem, some authors have also proposed mixed integer non linear optimisation strategies to simultaneously optimise process operating conditions and utility integration.

Evaluate From the numerical results of the mathematical programming tools, graphical representations are used to evaluate the results and understand the optimisation results. The graphical representation will support the engineers decision and help them propose other system improvement options. For this task, the use of the Carnot composite curves and the integrated composite curves allows one to better understand the combined heat and power production options and the integration between sub-systems.

The utility system integration defines additional streams that have to be considered together with the process streams, not only because their interface with the process will be heat exchangers but also because these streams will be used to solve integration issues like start-up and shut-down procedures, flexibility and control issues. In most of the cases, the utility streams, like hot water or steam, will also play the role of heat transfer fluids between process streams allowing one to overcome restricted matches constraints and topology restriction problems.

When analysing process designs to improve its energy efficiency, the optimal integration of the utility system allows to transform energy minimisation problems into energy cost or exergy losses minimisation. It offers a way of considering the energetic problem as a whole adopting a system vision for the use of energy in the process.

Gas engines, lean burn		
Generator eff.	$[-]$	$\eta_{gen} = 0.015 \cdot \ln(P) + 0.8687,\ P \leq 845\ [kW]$ $\eta_{gen} = 0.0013 \cdot \ln(P) + 0.9611,\ P > 845\ [kW]$
Mechanical eff.	$[-]$	$\eta_{mec} = 0.2419 \cdot P^{0.0653}$
Engine cooling @ 90°C	$[-]$	$\eta_{th,refr} = 0.7875 \cdot P^{-0.1682}$
Heat from combusiton gases	$[-]$	$\eta_{th,echap} = 0.1556 \cdot P^{0.0513}$
NO_x emissions	$[mg/m^3N]$	250
CO emissions	$[mg/m^3N]$	650
Installed cost	[?]	$C_{inst} = -0.0266 \cdot P^2 + 578.84 \cdot P + 208174$
Catalysts cost	[?]	$85 \cdot P^{0.8}$
Maintenance cost	$[?/kWh]$	$C_{maint} = 0.0407 \cdot P^{-0.2058} + 0.0034$
Life time	$[hours]$	48000

Table 3: Gas engine, lean burn configurations. P : power in (kW)

________________ ________________ ________________ ________________ ______________

Diesel engines		
Generator eff.	$[-]$	$\eta_{gen} = 0.015 \cdot \ln(P) + 0.8687,\ P \leq 845\ [kW]$ $\eta_{gen} = 0.0013 \cdot \ln(P) + 0.9611,\ P > 845\ [kW]$
Mechanical eff.	$[-]$	$\eta_{mec} = 0.0131 \cdot \ln(P_{mec}) + 0.3452$
Engine cooling @ 90¡C	$[-]$	$\eta_{th,refr} = 0.2875 \cdot P_{mec}^{-0.0139}$
Heat from combusiton gases	$[-]$	$\eta_{th,echap} = 0.5433 \cdot P_{mec}^{-0.1026}$
NO_x	$[mg/m^3N]$	100
CO	$[mg/m^3N]$	400
Installed cost	[?]	$C_{inst} = -0.0266(\frac{3}{4}P)^2 + 578.84\frac{3}{4}P + 208174,\ P \leq 1000\ [kW]$ $C_{inst} = 1147.62 \cdot P^{0.9009},\ P > 1000\ [kW]$
Catalyst	[?]	$136 \cdot P$
Maintenance	$[?/kWh]$	$C_{maint} = 0.0407 \cdot P^{-0.2058}$
Lifetime	$[hours]$	48000

Table 4: Diesel engines

— ________________

Aeroderivative gas turbines		
$\eta_{generator}$	$[-]$	$\eta_{gen} = 0.015 \cdot \ln(P) + 0.8687,\ P \leq 845\ [kW]$ $\eta_{gen} = 0.0013 \cdot \ln(P) + 0.9611,\ P > 845\ [kW]$
$\eta_{elec.}$	$[-]$	$\eta_{el} = 0.0439 \cdot \ln(P) - 0.0684$
η_{heat}	$[-]$	$\eta_{th,echap} = 0.838 \cdot P^{-0.0587}$
NO_x	$[mg/m^3N]$	80
CO	$[mg/m^3N]$	50
Turbine	[?]	$C_{ach} = 1516.41 \cdot P^{0.8503}$
Reference cost	[?]	$ref = \frac{C_{ach}}{(0.0503 \cdot \ln(\frac{P}{1000}) + 0.3208)}$
Recovery boiler	[?]	$C_{chaud} = (0.1255 - 0.0004 \cdot \frac{P}{1000}) \cdot ref$
Connection Charges	[?]	$C_{conn} = (0.09318 - 0.00011 \cdot \frac{P}{1000}) \cdot ref$
Instrumentation	[?]	$C_{ins} = (0.0494 - 0.0047 \cdot \ln(\frac{P}{1000})) \cdot ref$
Civil engineering	[?]	$C_{G.C.} = (0.1232 - 0.0005 \cdot \frac{P}{1000}) \cdot ref$
Engineering	[?]	$C_{eng} = (0.1211 - 0.000735 \cdot \frac{P}{1000}) \cdot ref$
Others	[?]	$C_{div} = (0.1403 - 0.0142 \cdot \ln(\frac{P}{1000}) \cdot ref$
Installed	[?]	$C_{inst} = \sum_i C_i$
Maintenance	$[?/kWh]$	$C_{maint} = 0.085 \cdot P^{-0.3081}$
Lifetime	$[hour]$	48000

Table 5: Aeroderivative gas turbines

Heavy duty gas turbines		
$\eta_{generator}$	$[-]$	$\eta_{gen} = 0.015 \cdot \ln(P) + 0.8687,\ P \leq 845\ [kW]$ $\eta_{gen} = 0.0013 \cdot \ln(P) + 0.9611,\ P > 845\ [kW]$
$\eta_{elec.}$	$[-]$	$\eta_{el} = 0.0187 \cdot \ln(P) + 0.1317$
η_{heat}	$[-]$	$\eta_{th,echap} = 0.7058 \cdot P^{-0.0315}$
NO_x	$[mg/m^3N]$	50
CO	$[mg/m^3N]$	50
Turbine	[?]	$C_{ach} = 4786.5 \cdot P^{0.7338},\ P \leq 50'000$ $C_{ach} = 2977.0 \cdot P^{0.7791}$
Reference cost	[?]	$ref = \frac{C_{ach}}{(0.0503 \cdot \ln(\frac{P}{1000}) + 0.3208)}$
Recovery boiler	[?]	$C_{chaud} = (0.1255 - 0.0004 \cdot \frac{P}{1000}) \cdot ref$
Connection Charges	[?]	$C_{conn} = (0.09318 - 0.00011 \cdot \frac{P}{1000}) \cdot ref$
Instrumentation	[?]	$C_{ins} = (0.0494 - 0.0047 \cdot \ln(\frac{P}{1000})) \cdot ref$
Civil engineering	[?]	$C_{G.C.} = (0.1232 - 0.0005 \cdot \frac{P}{1000}) \cdot ref$
Engineering	[?]	$C_{eng} = (0.1211 - 0.000735 \cdot \frac{P}{1000}) \cdot ref$
Others	[?]	$C_{div} = (0.1403 - 0.0142 \cdot \ln(\frac{P}{1000}) \cdot ref$
Installed	[?]	$C_{inst} = \sum_i C_i$
Mantenance	$[?/kWh]$	$C_{maint} = 0.085 \cdot P^{-0.3081}$
Lifetime	$[hour]$	55000

Table 6: Thermo-economic model of the heavy duty gas turbines

Type	ΔT_{lift} °C	Tc_{max} °C	Power kW	a $/kW_{th}$	b	Carnot eff. %
Electric compression heat pump	45	110	10-3000	814	-0.327	45%
Mechanical vapour recompression	30	200	250-50000	663.5	-0.3925	45%
Absorption Heat pump NH_3	50	150	5-60000	810.2	-0.3154	1.4
Absorption Heat pump $liBr$	50	150	5-60000	810.2	-03154	1.6
Absorption heat transformer $LiBr/H_2O$	50	150	250-4000	1164.8	-0.288	0.45
Thermal Vapour recompression	20	180	15-50000	268.56	-0.4832	-

Table 7: Thermo-economical characteristics of industrial heat pumps systems, The sizing parameter is the heat delivered Q_{th} in kW_{th}, the cost is expressed in $/kW_{th}$ computed by $Investment = a * (Q_{th})^b$, source contribution of TNO in [6]

PAFC		
η_{el}	$[-]$	0.35-0.4 [17]
η_{th}	$[-]$	0.25
NO_x	$[ppm]$	0 [8]
CO	$[pppm]$	0 [8]
Installed cost	$[/kW]$	4000[17]
Maintenance	$[cts/kWh]$	1 [17]
Life time	$[h]$	40000 [11]

Table 8: Phosphoric acide Fuel Cells

SOFC		
η_{el}	$[-]$	$\sim$ 0.5 [17]
η_{th}	$[-]$	0.35
NO_x	$[ppm]$	¡ 0.2 [7]
CO	$[pppm]$	0 [7]
Installed cost	$[/kW]$	450 (long term) - 1500 [2]
Maintenance	$[cts/kWh]$	1[17]
Life time	$[h]$	¿ 20000 [15]

Table 9: Solid Oxide Fuel cells

PEMFC		
η_{el}	$[-]$	$\sim$ 0.3 - 0.4 [17]
η_{th}	$[-]$	0.5-0.45
NO_x	$[ppm]$	-
CO	$[pppm]$	-
Installed cost	$[/kW]$	$\sim$ 500 (long term) [3] - 1000[17]
Maintenance	$[cts/kWh]$	1 [17]
Life time	$[h]$	87600 [3]

Table 10: Proton Exchange Membranes

Table 11: Values for some typical fuels

	LHV kJ/kg	UHV kJ/kg	T_{ad} °K	T_{stack} °K	O_2 kg_{O_2}/kg	CO_2 kg_{CO_2}/kg	Remarks
CH_4	50001.2	55784.1	2646.5			2.74	Pure methane
Natural Gas	39680		2270	374		2.18	87% CH_4
Light fuel oil	45316		2425	440		3.40	
Heavy fuel oil	44500		2423	441		3.38	
Coal (lignite)	25450		2111	438		2.03	
Wood	18900		2185.43		1.4112		
Wood (oak)	17769		4059.52		1.35	1.81	
Biogas	13358		4069.9		1.07	1.47	50%CO_2 and 50% CH_4

- Natural Gas composition: 87% Methane, 13% N_2
- Light Fuel Oil: C-86.2% mass, H-12.4% mass, S-1.4% mass
- Heavy Fuel Oil: C-86.1% mass, H-11.8% mass, S-2.1% mass
- Lignite: C-56.52%, H-5.72%, O-31.89%, N-0.81%, S-0.81%, Ash-4.25%
- Wood (oak) composition:

$$\begin{aligned} C\ 49.5\%w &= C_1 \\ H\ 6\%w &= H_{1,4} -> C_1H_{1,4}O_{0,7} \\ O\ 44.6\%w &= O_{0,7} \end{aligned}$$

Table 12: Minimum energy requirements of the process

Heating requirement	6854	kW	Above 365 K
Cooling requirement	6948	kW	Between 365 K and 298 K
Refrigeration requirement	1709	kW	Below 298 K (lowest T =267 K)

Table 13: Refrigeration cycle characteristics

Refrigerant		R717	Ammonia		
Reference Flowrate		0.1	kmol/s		
Mechanical power		394	kW		

	P (bar)	T_{in} (°K)	T_{out} (°K)	Q kW	$\Delta Tmin/2$ (°K)
Hot stream	12	340	304	2274	2
Cold stream	3	264	264	1880	2

Table 14: Steam cycle characteristics

	P (bar)	T (°K	Comment
HP2	92	793	superheated
HP1	39	707	superheated
HPU	32	510	condensation
MPU	7.66	442	condensation
LPU	4.28	419	condensation
LPU2	2.59	402	condensation
LPU3	1.29	380	condensation
DEA	1.15	377	deaeration

Table 15: Results of the energy conversion system integration for different options

Option	Fuel kW_{LHV}	GT kWe	CHP kWe	Cooling kW	Heat Pump kWe
Boiler	7071	-	-	8979	-
Boiler+ steam	10086		2957	9006	-
GT+steam	16961	5427	2262	9160	-
Boiler+ heat pump	-	-	-	2800	485
Boiler + steam + heat pump	666	-	738	2713	496

Table 16: Overall energy consumption of the different options based on 55% fuel equivalent for electricity

Option	Fuel kW_{LHV}	Net Electricity kWe	Total consumption kW_{LHV}
Boiler	7071	371	7746
Boiler+ steam	10086	-2481	5575
GT+steam	16961	-7195	3879
Boiler+ heat pump	-	832	1513
Boiler + steam + heat pump	666	125	893

Table 17: Characteristics of the heat pump system, based on R123 as working fluid

Option	P_{low} (bar)	T_{low} (°K)	P_{high} (bar	T_{high} (°K)	COP -	kWe
Cycle 3	5	354	7.5	371	15	130
Cycle 2	6	361	10	384	12	323
Cycle 0	6	361	7.5	371	28	34

Bibliography

[1] Duran M. A. and Grossmann I.E. Simultaneous optimization and heat integration of chemical processes. *AIChE Journal, Vol. 32, no1, January 1986.*, (55), 1986a.

[2] Teagan P. Yang Y. Carlson E.J. Cost modelling of sofc technology. *1rst International conference on fuel cell Development and deployement*, www.ctfuelcell.uconn.edu/fcic/pdf/fcic-program-oral-4a.2.pdf, 2004.

[3] Thomas C.E. Cost analysis of stationary fuel cell systems including hydrogen co-generation. Technical report, Directed Technologies, December 1999, available online at www.directedtechnologies.com, 1999.

[4] Seider W.D. Colmenares T.R. Synthesis of cascade refrigeration systems integrated with chemical processes. *Computers and Chemical Engineering*, 1989.

[5] V.R. Dhole and B. Linnhoff. Total site targets for fuel, co-generation emissions, and cooling. *Computers and Chemical Engineering*, 17(161):s101–s109, 1992.

[6] Arpentinier et al. Exsys ii: an expert system for optimal insertion of intensified energy saving technologies (iest) in the industrial processes. - publishable final report of project joe3-ct97-0070. Technical report, EU commission, 2001.

[7] Crawley G. Operational experience with siemens-westinghouse cogeneration experience. *Fuel cell today*, 17 october 2001.

[8] Taylor W.R. Holcomb F.h. Reduced emissions from cogeneration, applications of dod fuel cell power plant fleet. *DoD fuel cell (www.dodfuelcell.com (29/06/2004))*, 2004.

[9] B. Kalitventzeff, M. D. Dumont, and F. Marechal. Process integration techniques in the development of new energy technologies : application to the isothermal gas turbine. *Keynote lecture, paper F3.1, Proceedings of CHISA 98, 13th International Congress of Chemical and Process Engineering, Praha, 23-28 August 1998*, (205), 1998.

[10] B. Linnhoff and V.R. Dhole. Shaft work targeting for subambient plants. *Aiche Spring Meeting, Houston, Paper 34d*, April 1989.

[11] Cropper M. Why is interest in phosphoric acid fuel cells falling. *Fuel cell today*, 8 october 2003.

[12] F. Marechal and B. Kalitventzeff. Targeting the optimal integration of steam networks. *Computers and Chemical Engineering*, 23:s133–s136, 1999.

[13] F. Marechal and B. Kalitventzeff. A tool for optimal synthesis of industrial refrigeration systems. *Computer Aided Chemical Engineering, ed. R. Gani, S.B. Jorgensen, Ed. Elsevier*, 9:457–463, May 2001.

[14] S.A. Papoulias and Grossmann I.E. A structural optimization approach in process synthesis - i. utility systems. *Computers and Chemical Engineering*, 7(6):695–706, 1983a.

[15] Colson-Inam S. Solid oxide fuel cells - ready to market ? *Fuel cell today*, 07 January 2004.

[16] D.W. Townsend and B. Linnhoff. Heat and power networks in process design. part 1: Criteria for placement of heat engines and heat pumps in process networks. *AIChE Journal, Vol. 29, no 5, september 1983.*, 29(5):742–748, 1983.

[17] California Energy Commission : www.energy.ca.gov.

LENI Systems

Process integration techniques for improving the energy efficiency of industrial processes

Estimating the cost of equipment

Dr. François Marechal

Maître d'Enseignement et de Recherches
Industrial Energy Systems Laboratory
LENI-IGM-STI-EPFL
Ecole Polytechnique Fédérale de Lausanne
CH-1015 Lausanne
http://leni.epfl.ch

Based on the document Memo: Cost Functions and the LENI database, Methodology basics of cost computation for usage with LENI's cost database, F.Palazzi

ÉCOLE POLYTECHNIQUE FÉDÉRALE DE LAUSANNE

Abstract

This document presents the basic cost computation formulae for study estimates of industrial processes as outlined in *Turton et al.* [1]. The guidelines for developing and using the cost database at LENI are specified.

1 Equipment Cost Estimation: The Basics

1.1 Capital Costs Estimating Classifications

The capital cost is the cost related to the purchase of process equipment and its installation on a production site. Estimation of the capital cost of a plant can be preformed at different levels, depending to the accuracy needed. *Turton et al.* [1] classify the estimate types in five categories:

1. Order of magnitude estimate:
   - Cost information on the complete process, based on scaling from an existing plant of the same type
   - Block flow diagram level
   - Accuracy: +40% ; -20%
2. Study estimate:
   - Cost computation based on the list of the major equipment found in the process
   - Each piece of equipment is roughly sized
   - Process flow diagram level
   - Accuracy: +30% ; -20%
3. Preliminary design estimate:
   - The cost computation includes the equipment cost and the layout cost (piping, instrumentation. . .)
   - Accurate sizing of the equipment is performed
   - Accuracy: +25% ; -15%
4. Definitive estimate:
   - Preliminary specifications for equipment, utilities, instrumentation, off-sites
   - Accurate sizing of the equipment is performed
   - Accuracy: +15% ; -7%
5. Detailed estimate:
   - Complete engineering of the process, off-site utilities. Vendor quotes for all major items. . .
   - Accuracy: +6% ; -4%

The following cost estimation methodology addresses point 2: the **study estimate level**. It is the basis of the equipment cost database at LENI.

1.2 Equipment rough cost estimate

The rough cost estimate of an equipment is determined through sizing and scaling the equipment from a reference case and/or using sizing heurisics and scaling parameters from literature data. The parameters influencing the cost of an equipment piece are:

- Size (capacity)
- Material
- Pressure
- Inflation (time)

The global cost of an equipment in a plant can be decomposed into four contributions:

- Direct expenses: purchase, material, installation
- Indirect expenses: freight, taxes, overhead, engineering expenses
- Contingencies and Fees: unforeseen circumstances, contractor fee, ...
- Auxiliary facilities: site development, land buildings, offsites, utilities, ...

The terminology developed by *Turton et al.* [1] and adapted for the database refers to the costs presented below. Their computation will be detailed in the following paragraphs.

- Bare module cost (C_{BM}): Direct and indirect expenses related to the unit: material, construction, freight, taxes, overheat, engineering expenses, installation.
- Total module cost (C_{TM}): Bare module cost plus the contingencies and fees due to unforeseen circumstances and fees during the plant commissioning.
- Cost for auxiliary facilities (C_{AUX}): Sum of the expenses for site development, buildings, offsites, utilities.
- Grass roots cost (C_{GR}): Sum of the total module cost plus the cost for auxiliary facilities. This cost corresponds to the overall capital investment of a new plant without considering the expenses for land.

1.3 The bare module cost C_{BM}

1.3.1 Cost details

The Bare-Module cost C_{BM} of an equipment is defined as the sum of direct and indirect costs:

$$C_{BM} = C_d + C_i \tag{1}$$

The direct costs C_d include the purchase C_p, material C_m, and installation C_{ins} costs:

$$C_d = C_p + C_m + C_{ins} \tag{2}$$

The indirect cost C_i include the freight C_f, the construction overhead C_o, and contractor engineering C_e expenses:

$$C_i = C_f + C_o + C_e \tag{3}$$

All other costs than the purchase cost contributing to the bare-module cost are assumed to be directly proportional to the purchase cost. They are taken into consideration by introducing the bare-module factor F_{BM}:

$$C_{BM} = F_{BM} \cdot C_p \tag{4}$$

1.3.2 Factors influencing the bare module cost

Influence of size. As *size* we intend a specific attribute of the equipment that gives an indication of the capacity. The attribute can be the volume of the equipment, the area of a piece, the power delivered or the quantity of reactant treated, among others. The size of the equipment influences the *purchase cost*. The purchase cost can be obtained scaling from a reference case, considering that the reference case and the computed case are simultaneous:

$$C_p(A) = C_{p,ref} \cdot \left(\frac{A}{A_{ref}}\right)^{\gamma} \tag{5}$$

$C_{p,ref}$: purchase cost of the reference case
A : equipment attribute
A_{ref} : equipment reference attribute
γ : capacity exponent, the typical value of this coefficient is 0.6 the range is between 0.6 and 0.8 for most of the equipment.

Alternatively, the purchase cost can be computed with an empiric formula derived from a cost database of existing equipment. We give here an example:

$$C_p = 10^{k_1 + k_2 log A + k_3 (log A)^2} \tag{6}$$

k_i : empirical constants derived from the cost database

Effect of time: inflation. The effect of inflation on equipment cost is computed using cost indexes. Two cost indexes are generally used

- Marshall and Swift Equipment cost index
- Chemical Engineering Plant cost index

The values of both indexes can be found at the end page of Chemical Engineering.

The purchased cost of an equipment of same size at different time is given by:

$$C_p = C_{p,ref} \cdot \frac{I_t}{I_{t,ref}} \tag{7}$$

$C_{p,ref}$: cost of the equipment at the reference time
$I_{t,ref}$: cost index for the reference year
I_t : cost index for the actual year

Effect of pressure and material. The bare module cost for an equipment type is influenced by the operating pressure and by the choice of the construction material:

$$F_{BM} = B_1 + B_2 \cdot F_m \cdot F_p \tag{8}$$

B_i : constants computed on the base of existing equipment cost databases
F_m : material factor, describes the effect of material change on the bare-module cost. If there is no material change $F_m = 1$
F_p : pressure factor, describes the effect of operating pressure change on the bare-module cost.

The pressure factor can be computed for an equipment type using empiric formulae derived from cost databases:

$$F_p = 10^{c_1 + c_2 \cdot log\tilde{P} + c_3 \cdot (log\tilde{P})^2} \tag{9}$$

c_i : empirical constant derived from the existing plant cost database
$\tilde{P}$: pressure difference from atmospheric pressure e.g. pressure expressed in bar gauge (barg)

Turton and al. give pressure factors expressions for different equipment types in appendix A. The pressure range of these expressions have to be kept in mind. When deriving such an expression indicate the pressure range of application in the matlab file.

Summary: calculating the bare module cost. All the influences on the bare module cost are accounted for following these steps:

- Determine the sizing parameter(s) and determine the purchase cost C_p for the reference year.
- Actualise the purchase cost to the actual year with the Marshall and Swift Index.
- Determine the operating pressure and material of construction and account for it with the pressure and material factors F_p and F_m.
- Account for the influence of installation and indirect cost and determine the bare module factor F_{BM}.
- Calculate the bare module costs.

1.4 The total module cost C_{TM}

The contingencies and fees during the construction of a plant due to unforeseen circumstances like weather, design changes, price increases, etc. are taken into consideration by factoring the bare module costs. The cost of a single process equipment including the bare module cost and the associated contingencies and fees is termed total module cost C_{TM}:

$$C_{TM} = (1 + \alpha_1)C_{BM} \tag{10}$$

As a rule of thumb, contingencies and fees amount to 15% and 3% of the bare module cost, respectively [1]. A value of 0.18 is thus used as default for α_1.

If the purpose of the cost calculation is to estimate the cost associated with small to moderate expansions or alterations to an existing facility, the total investment is calculated by summing the total module cost of all new equipment.

1.5 The cost for auxiliary facilities and new plants

New plants require site development and the construction of auxiliary facilities. These expenses are assumed to be proportional to the bare module cost of all equipment at base case conditions C^*_{BM} (i.e. carbon steel construction and ambient pressure). The grass roots cost is thus calculated according to:

$$C_{GR} = \sum C_{TM} + \sum (\alpha_2 C^*_{BM}) \tag{11}$$

It is recommended to use 0.35 as a default value for α_2 [1].

2 Guidelines for the cost database at LENI

2.1 Generalities

Due to the stringent need for the consistency of data and its underlying assumptions, it is crucial to follow general conventions of the database:

- The cost function database uses the terminology and definitions of section 1 of this document.
- Clearly indicate the source of information necessary for the cost estimation. Whenever possible, add the reference as pdf to the reference folder. Name the file as *firstauthor_secondauthor_year.pdf.*
- Write the cost function in Matlab language adopting the structure as described in section 2.2.

2.2 Structure of a cost function

The cost functions of the database are structured as follows:

```
function Cost = cost_EquipmentName(c, arg1, arg2, ...)
```

- `c` is a structure defining cost computation constants and must contain the following fields:

`c.Year_Index_MS`	Marshall and Swift Index
`c.f_TotalModule`	α_1
`c.f_GrassRoot`	α_2

  These field can be set (or completed) by calling `c=cost_defaults` before invoking the cost functions the first time.

- The arguments `arg1`, `arg2`, ... are specific to the costing function, i.e. parameters related to capacity, sizing heuristics, etc.
- The function output `Cost` is a structure with the fiels:

`.BM`	Bare module cost of the equipment
`.TM`	Total module cost of the equipment
`.GR`	Specific grass roots cost associated to the equipment, i.e. $\alpha_1 C_{TM} + \alpha_2 C^*_{BM}$ (cf. equation 11)
`.Unit`	Monetary unit as a string structure: `Cost.Unit = {'USD'};` (or `{'EUR'}`, `{'CHF'}`)

An example is given in the Appendix.

Bibliography

[1] Turton, R., Bailie, R.C., Whiting, W.B., Shaeiwitz, J.A., *Analysis, synthesis, and design of chemical processes.* Prentice Hall, New Jersey, 1998.

[2] *Cost/capacity relationships part 2* Process engineering, April, 1998, pp 16

LENI Systems

Economic evaluation

Sales (CHF/year)

Products Electricity

$$\int_{t_0}^{t_f} \left(\sum_{P=1}^{n_P} \dot{M}_P^-(t) c_P^-(t) + E^-(t) c_e^-\right) dt$$

Operating costs (CHF/Year)

Raw material Fuel Electricity Waste

$$-\int_{t_0}^{t_f} \left(\sum_{R=1}^{n_R} \dot{M}_R^+(t) c_R^+(t) + \sum_{F=1}^{n_F} \dot{M}_F^+(t) c_F^+(t) + E^+(t) c_e^+ + \sum_{W=1}^{n_W} \dot{M}_W^-(t) c_W^-(t)\right) dt$$

Men power + Maintenance + Taxes (CHF/year)

$$-\int_{t_0}^{t_f} \sum_{cat=1}^{n_{cat}} (N_{cat}(t) c_{cat}(t)) dt$$

$$-\sum_{i=1}^{n_{equipment}} Maint_i$$

$$-\sum_{t=1}^{n_{tax}} Taxes_t$$

Investments (CHF/year)

$$-\frac{1}{\tau} \sum_{i=1}^{n_{equipments}} I_i$$

Anualising the investment

- **Why ?**
  - The Investment value to be compared with annual income (savings)
  - Money value of today to be compared with future income

Anualising a value

- **Value of the investment (I) after n years under an interest rate of i**

$$I^*(i,n) = I(1+i)^n$$

where	$I =$	Investment [CHF]
	$i =$	Interest rate
	$n =$	expected life time of the unit
	$I^*(i,n) =$	Value of I after n years with an interest rate of i

Value of an annual income B after n years with an interest rate of i

$$B^*(i,n) = \sum_{r=1}^{n} B(1+i)^{r-1} = B\frac{(1+i)^n - 1}{i}$$

Anualising a value

- **Present value V of an annual income B after n years under an interest rate of i**

$$V^*(i,n) = B^*(i,n) \rightarrow V[CHF] = B\frac{(1+i)^n - 1}{i(1+i)^n}$$

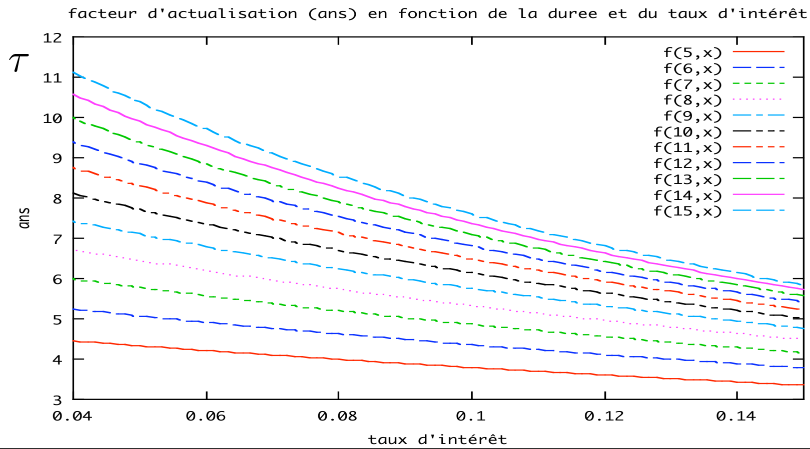
- **The annual expenditure of an investment I can be compared with the money invested today**

$$I^*(i,n) = IC^*(i,n) \rightarrow IC[CHF/year] = I\frac{i(1+i)^n}{(1+i)^n - 1}$$

Project evaluation

- **Net present value** $$NPV[CHF] = B\frac{(1+i)^n - 1}{i(1+i)^n} - I \geq 0$$
- **Annualised Cost**

$$P[CHF/year] = B - I\frac{i(1+i)^n}{(1+i)^n - 1} = B - \frac{I}{\tau} \geq 0$$



Retrofit project : benefit vs incremental invest

Sales

$$\Delta Sales[CHF/year]$$

Products Electricity

$$\int_{t_0}^{t_f} \left(\sum_{P=1}^{n_P} \Delta \dot{M}_P^-(t) c_P^-(t) + \Delta E^-(t) c_e^-\right) dt$$

Operating costs

$$\Delta Cost[CHF/year]$$

Raw material Fuel Electricity Waste

$$\int_{t_0}^{t_f} \left(\sum_{R=1}^{n_R} \Delta \dot{M}_R^+(t) c_R^+(t) + \sum_{F=1}^{n_F} \Delta \dot{M}_F^+(t) c_F^+(t) + \Delta E^+(t) c_e^+ + \sum_{W=1}^{n_W} \Delta \dot{M}_W^-(t) c_W^-(t)\right) dt$$

Benefit

$$\Delta B[CHF/year] = \Delta Sales[CHF/year] - \Delta Cost[CHF/year]$$

Investments (CHF/year)

$$\Delta IC[CHF/year] = \Delta I\frac{i(1+i)^n}{(1+i)^n - 1}$$

Project profitability

Rate of Return [years] $= \frac{\Delta I}{\Delta B}$

Net Present Value [\$] $= \Delta B \frac{(1+i)^{n_y}-1}{i(1+i)^{n_y}} - \Delta I \geq 0$

Annualised Profit $[\frac{\$}{year}]$ $= \Delta B - \Delta I \frac{i(1+i)^{n_y}}{(1+i)^{n_y}-1} = \Delta B - \frac{I}{\tau} \geq 0$

Internal interest rate [%] $i^* | \Delta B \frac{(1+i^*)^{n_y}-1}{i^*(1+i^*)^{n_y}} - \Delta I = 0$

ΔI	[\$]	Additional investment
ΔB	$[\frac{\$}{year}]$	Annual expected profit from the investment
n_y	[year]	Expected life time of the equipment