UNIVERSIDADE ESTADUAL DE CAMPINAS
FACULDADE DE ENGENHARIA QUÍMICA

VALTER BRAVIM JUNIOR

# HEAT EXCHANGER NETWORK AND PLANT LAYOUT: SEQUENTIAL AND SIMULTANEOUS DESIGN WITH PIPING AND PUMPING COST CONSIDERATIONS

# TROCADORES DE CALOR: REDES E LAYOUT NA PLANTA - PROJETO SEQUENCIAL E SIMULTÂNEO CONSIDERANDO CUSTOS DE TUBULAÇÃO E BOMBEAMENTO

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# HEAT EXCHANGER NETWORK AND PLANT LAYOUT: SEQUENTIAL AND SIMULTANEOUS DESIGN WITH PIPING AND PUMPING COST ESTIMATE

# TROCADORES DE CALOR: REDES E LAYOUT NA PLANTA - PROJETO SEQUENCIAL E SIMULTÂNEO COM ESTIMATIVA DE CUSTOS DE TUBULAÇÃO E BOMBEAMENTO

Tese apresentada à Faculdade de Engenharia Química da Universidade Estadual de Campinas como parte dos requisitos exigidos para a obtenção do título de Doutor em Engenharia Química

Thesis presented to the School of Chemical Engineering of the University of Campinas in partial fulfilment of the requirements for the degree of Doctor in Chemical Engineering

Supervisor/Advisor: Roger Josef Zemp

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Prof. Dr. Roger Josef Zemp
FEQ - UNICAMP
Presidente e orientador

Dr. Marcelo Modesto da Silva
Universidade Federal do ABC - UFABC
(Videoconferência)

Dr. Luís Fernando Novazzi
Centro Universitário da FEI - Fundação Educacional Inaciana Padre Sabóia de Medeiros

Prof. Dr. Adriano Pinto Mariano
FEQ - UNICAMP

Prof. Dr. Valdir Apolinário de Freitas
Rhodia Solvay Group

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**Resumo**

A integração energética consiste de uma série de técnicas para reduzir o consumo de utilidades em processos químicos. Uma maneira sistemática para integrar energeticamente um processo químico é através do projeto de Redes de Trocadores de Calor (RTC). A estratégia de projeto normalmente adotada na literatura acadêmica considera somente os custos de utilidade e trocadores de calor. Entretanto, as RTCs também são afetadas por diferentes requerimentos do layout da planta do processo químico. Dentre estes requerimentos, a presente tese aborda a localização de trocadores de calor na planta, o acesso para manutenção dos trocadores de calor, o custo de bombeamento, o custo de tubulação e o comprimento de tubulação. O objetivo desta tese foi desenvolver um conjunto de métodos para considerar esses requerimentos no projeto da RTC. O primeiro método foi desenvolvido com programação linear. A função do primeiro método é calcular o valor mínimo que o comprimento de tubulação tem que aumentar para transportar as correntes passando por um trocador de calor em vez de transporta-las diretamente de seus equipamentos iniciais para os finais. O primeiro método pode ser utilizado para auxiliar o projeto da RTC comparando o aumento mínimo no comprimento de tubulação das possíveis combinações entre correntes. O segundo método foi desenvolvido utilizando programação linear inteira mista. A função do segundo método é calcular a localização de trocadores de calor na planta. O segundo método é orientado a encontrar as localizações que minimizam o comprimento de tubulação para transportar as correntes de seus equipamentos iniciais até os finais. O projeto da RTC é adequado para ser realizado sequencialmente com o segundo método. A característica de ser aplicado sequencialmente permitiu que capacidades adicionais fossem desenvolvidas no segundo método. Neste sentido, é possível restringir as localizações a satisfazerem um espaçamento mínimo e um máximo entre trocadores de calor. Além disso, zonas podem ser definidas dentro na planta para limitar as localizações disponíveis para o posicionamento dos trocadores de calor. O terceiro e último método desenvolvido nesta tese funciona simultaneamente com o projeto da RTC e foi desenvolvido por meio de programação não linear inteira mista. A natureza simultânea do terceiro método permite considerar os custos de tubulação e bombeamento além dos tradicionais custos de utilidades e trocadores de calor. Todos os métodos foram testados em casos de estudo com até oito correntes de processo e dez trocadores de calor. Os modelos matemáticos foram implementados em GAMS e solucionados por meio dos solucionadores CPLEX e DICOPT. Um dos casos de estudo mais demonstrativos foi modelado com 292 variáveis contínuas, 136 variáveis discretas e 375 equações. O tempo de solução foi aproximadamente 1 segundo.
**Palavras-chave:** Integração energética, Trocadores de calor, Layout, Otimização matemática, Tubulação - Projetos e construção.

## Abstract

Heat integration consists of a series of techniques to decrease the utility consumption of chemical processes. A systematic way to heat-integrate a chemical process is through the design of Heat Exchanger Networks (HEN). The design strategy usually adopted in the academic literature considers only the costs of utilities and heat exchangers. Nonetheless, the HENs are also affected by various requirements of the chemical process plant layout. Among those requirements, the present thesis addresses the location of heat exchangers in the plant, the access for the maintenance of the heat exchangers, the pumping cost, the pipe cost and the pipe length. The objective of the this thesis was to develop three methods to consider these requirements in the HEN design. The first method was developed by means of linear programming. The function of the first method is to calculate the minimum value that the length of pipe has to increase for transporting the streams passing through a heat exchanger instead of transporting them directly from their initial to their final pieces of equipment. The first method can be used to assist the HEN design by comparing the potential stream matches in terms of their minimum pipe length increase. The second method was developed by means of mixed-integer linear programming. The function of the second method is to calculate the location of heat exchangers in the plant. The second method is orientated to find the locations that minimise the pipe length for transporting the streams from their initial to their final pieces of equipment. The HEN design is suitable to be carried out in sequence with the second method. The characteristic of being applied sequentially allowed the capabilities of the second method to be further developed. The locations can be restricted to satisfying a minimum and a maximum spacing between heat exchangers. In addition, zones can be defined within the plant to limit the locations available for heat exchanger placement. The third and last method developed in this thesis functions simultaneously with the HEN design and was developed by means of mixed-integer non-linear programming. The simultaneous nature of the third method allows the consideration of the costs of pipe and pumping in addition to the usual costs of utilities and heat exchangers. All the methods were tested in case studies with up to eight process streams and ten heat exchangers. The mathematical models were implemented in GAMS and solved by means of CPLEX and DICOPT. One of the most demonstrative case studies was modelled with 292 continuous variables, 136 discrete variables and 375 equations. The solution time was approximately 1 second.


**Keywords:** Heat integration, Heat exchangers, Layout, Mathematical optimisation, Piping - Projects and Construction.

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# Chapter 1

# Introduction

## 1.1 Background

**Heat integration in chemical processes**

The energy consumption of industrial chemical processes has a major contribution to greenhouse gas emissions (YANG, 2020). Part of the energy consumption results from the supply of heating and cooling by means of utility streams. The process streams that require cooling reject heat to cold utility streams, such as cold water and refrigerant fluid. In an analogous manner, the process streams that require heating absorb heat from hot utility streams, such as steam and oil.

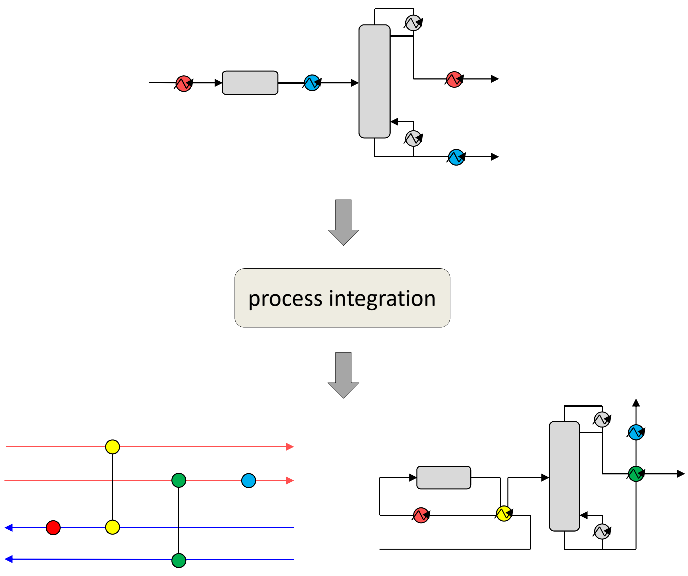
Reducing the utility consumption of the processes leads to a reduction in the energy consumption and in the operating cost. The utility consumption can be reduced by heat integration. In a heat-integrated process, the process streams that require cooling exchange heat with those that require heating. In doing so, the process streams reach their energy balance without exchanging or exchanging a smaller amount of heat with the utility streams. Figure 1.1 depicts the flowsheet of a process without and with heat integration.

The heat exchange between the process streams is performed in heat exchangers. The existence of various heat exchangers form a network, which is known as Heat Exchanger Network (HEN) (KEMP; LIM, 2020). The heat exchangers increase the capital cost of the project. Thus, the HENs are designed by making a trade-off between the costs of utilities and heat exchangers.

The design of a HEN encompasses a series of tasks, which includes:

- matching the process streams for heat exchange
- calculating the temperatures at which heat is exchanged between the matched streams;
- calculating the heat flow rate between the matched streams.

**Figure 1.1:** The flowsheet of a process without and with heat integration.




### Methods for heat integration in chemical processes

The methods for heat integration began to gain attention from the academic community in the late of the 1960s (KESLER, 1968) and early of the 1970s (HOHMANN, 1971). Currently, there is a variety of methods for heat integration available in the academic literature. Those methods can be classified in sequential and simultaneous. Sequential methods set a target for the minimum:

- consumption of utilities;
- number of heat exchange units[1];
- total heat transfer area.

Setting the targets sequentially is advantageous because it provides the information prior to designing the HEN. In this way, it possible to start designing the HEN from the most promising scenarios.

The most popular sequential method for heat integration is the pinch analysis. This method was published in the late of the 1970 (LINNHOFF; FLOWER, 1978). Based on the heat balance of the process streams and a minimum driving force for heat transfer, this method allows the user to set the targets in a simplified manner.

[1]A discussion on *heat exchange unit* is presented in Section 1.4.

By setting the targets over a range of minimum driving force for heat transfer, the user can find the targets that lead to the minimum or close-to-minimum annual costs of utilities and heat exchangers. Having set the targets, the pinch analysis features a set of rules for the user to design the HEN (SMITH, 2005).

The problem of setting the targets can be formulated as a mathematical optimisation problem. The advantage of doing that is to gain additional information about the HEN. The mathematical optimisation model provides which process streams should be matched for heat exchange and their heat flow rates. This information can be obtained with restrictions, such as to forbid or impose the heat exchange between specific process streams (BIEGLER et al., 1997).

The problem of designing the HEN can also be formulated as a mathematical optimisation problem. The objective function is defined as the minimisation of the cost of heat exchangers, which is proportional to the heat transfer area. The decision variables are defined as to match or not two process stream for heat exchange and, the temperatures of the process streams. The constraints are the heat balances (BIEGLER et al., 1997).

The major disadvantage of the sequential methods is that the target set at one step is affected by that set at the previous step. As a consequence, there is no guarantee that the HEN designed from the targets is the minimum total annual cost. The **simultaneous** methods for heat integration tackle this weakness by minimising the costs of utilities and heat exchangers simultaneously (YEE; GROSSMANN, 1990). In order to do that, the simultaneous methods allow all the process streams that require cooling to exchange heat with all the process streams that require heating and vice and versa. The process streams are matched so as to minimise the costs of utilities and heat exchangers.

All the possible matches between the process streams for heat exchange form a superstructure[2]. In the mathematical model of the superstructure, the heat flow rate between two process streams and their temperatures are represented by continuous variables. The fact that two process streams are matched for heat exchange is denoted by binary variables. The cost of heat exchangers is calculated through the heat transfer area. This results in a set of non-linear equations because of the mean temperature difference term that goes in the denominator.

For process with a large number of process streams, designing HENs through mathematical optimisation can be a difficult task. The reason why is the huge number of possible topologies for the HEN. In addition, the non-linear nature of the model makes it difficult to guarantee the convergence to the global optimum. These difficulties motivated the use of stochastic optimisation methods (LEWIN, 1998). In this sense, the most commonly used ones are based on genetic algorithm, simulated

[2]The concept of superstructure will be introduced in Section 4.1 in a detailed manner.

annealing algorithm and particle swarm optimisation algorithm (ROETZEL et al., 2019).

### Location of heat exchangers in the plant

Once the HEN is designed, it is necessary to define the **location of heat exchangers** in the plant. Heat exchangers are usually placed at ground level but can be placed in multi-floor structures (BARKER, 2017). They can be arranged individually, paired, stacked or grouped when they operate in series or parallel (BAUSBACHER; HUNT, 1993). The location of heat exchangers in the plant is a determinant of the costs of pipe and pumping required to transport the process streams (MORAN, 2016).

The pipes connect the heat exchangers to the process equipment, such as reactors, tanks, separators, distillation columns, etc. There are valves, elbows, tees and other fittings installed in the pipes. The pumps provide energy for the fluids to overcome the energy loss from the flow through the pipes. The greater is the distance between the heat exchangers connected by pipes, the greater are the costs of pipe and pumping.

Maintenance work requires that the heat exchangers are placed so that they can be accessed by large machines. For example, to remove the tube bundle of shell and tube heat exchangers, it is necessary to provide access to tube bundle extractors and gantries. Once the tube bundle is removed, it may be transported through the plant for off-site repair or to be repaired on-site (BAUSBACHER; HUNT, 1993).

The location of heat exchangers in the plant should satisfy a minimum spacing. This guarantees enough space for the operators to access the heat exchangers and carry out maintenance work. Heat exchangers should be spaced apart from other process equipment by not less than a minimum value. This is a requirement for process equipment operating at high pressure or with hazardous materials, such as compressors and reactors (BAUSBACHER; HUNT, 1993).

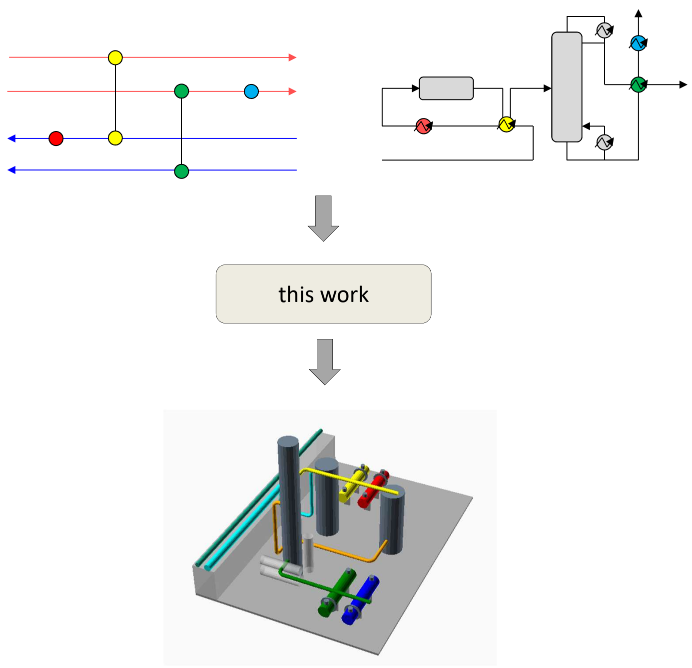
### Costs of pipe and pumping

From our viewpoint, defining the location of heat exchangers in the plant can be treated as a task of optimisation. The objective is to minimise the costs of pipe and pumping for the transport of the process streams. The maintenance requirements act as constraints on the number of locations where the heat exchangers can be placed. The amount of heat exchangers in the process increases the complexity of the optimisation task.

The costs of pipe and pumping can be calculated from the pipe length. The pipe length, in its turn, can be approximated by the distance between the pieces of equipment through which the streams pass. The location of heat exchangers in the plant allows one to calculate the distance between any two heat exchangers as well as the distance between a heat exchanger and other piece of equipment.

Typically, chemical processes contain many streams exchanging heat with each other. This characteristic has inspired us to think that defining the location of all heat exchangers can be understood as to design the layout of heat exchangers in the plant. This concept is particularly relevant for this thesis since it can be linked to HEN design. Figure 1.2 illustrates the link between the HEN and the plant layout.

**Figure 1.2:** The transition from the heat exchanger network to the plant layout.




### Designing the heat exchanger plant layout: after the HEN design and simultaneously with the HEN design

The design of the HEN defines the which streams will be matched for heat exchange, the amount of heat exchanged and the heat transfer areas. With this information, the heat exchanger plant layout is designed. Performing these task sequentially fixes to a certain extent the costs of pipe and pumping. The reason why is that, in a designed HEN, the process streams matches for heat exchange are already defined. The sequence of heat exchanges performed by the process streams is also defined. The HEN topology is fixed and cannot be modified. The best that can be done is to screen the heat exchangers over different locations.

A higher potential for the minimisation of the costs of pipe and pumping can be achieved by designing the HEN and the heat exchanger plant layout in a simultaneous manner. During the HEN design, the process stream matches for heat exchange can be modified. It is possible to modify the sequence of heat exchanges performed by the process streams. The HEN is generated in various topologies. For each topology, the heat exchanger plant layout will be designed with the objective of minimising the costs of pipe and pumping. The definitive HEN topology is the one that gives the lowest costs of pipe and pumping. The costs of utilities and heat exchangers can be included in the objective function or restricted to an upper boundary.

## 1.2 Literature Review

**Design of the layout of chemical process plants**

A well-designed layout of a chemical process plant allows the safe operation of the processes, provides adequate access to the process equipment and makes effective use of the available land. The design of the layout of a chemical process plant is carried out by engineers from different disciplines through several stages (MORAN, 2016). The output from the stages is iteratively revised and many decisions are made based on the experience of the engineers. The accuracy of the output from the conceptual stage is vital to avoid modifications during the construction stage and afterwards.

The location of the process equipment in the plant, such as towers, reactors, storage tanks, pressure vessels, heat exchangers, etc., is defined at the conceptual stage of the project (BARKER, 2017). The process equipment is usually arranged with the process flow sequence functioning as a base point. Consideration is given to grouping process equipment of the same type (BACKHURST; HARKER, 2013). The way the process equipment is arranged affects the viability of maintenance activities. The space available on the site should be used taking future expansions into consideration (BAUSBACHER; HUNT, 1993). Process equipment may have to be set up in a specific manner to fit into the space left over from previous plant expansions.

There are various factors that define the location of the process equipment in the plant (MUKHERJEE, 2021). The minimisation of pipework to interconnect the process equipment is one of the main goals. Access to the process equipment is crucial for operators to be able to carry out maintenance. The minimum recommended spacing varies with the process equipment according to its operation, maintenance and safety requirements (MEISSNER III, 2000). The spacing between process equipment of the same type should be enough to provide operator access. Greater distances are recommended when dealing with compressors, storage vessels, flares, etc. Large areas of the plant should be made available for the maintenance of some types of process equipment (BAUSBACHER; HUNT, 1993). For example, in order to load and unload catalysts into and from reactors, drop zones and handling facilities are needed.

Due to the large amount of process equipment and process streams being transported, the task of finding the location of process equipment has been performed using mathematical optimisation. The objective is to find the layout that gives the minimum cost in terms of piping, pumping, floor construction, land, safety, etc. Spacing between process equipment, nozzle orientation and other requirements are added in the form of constraints. Ejeh et al. (2018) presented a MILP methodology for the design of the layout of multi-floor chemical process plants. The process equipment was placed within sections along floors. The process equipment with common characteristics was grouped together in the same section. The methodology could handle the location of tall process equipment that spanned across multiple floors in two ways. A tall piece of equipment could be modelled as a single item or as several pseudo items stacked floor by floor.

Wang et al. (2018) developed a methodology for the design of the layout of chemical process plants using genetic algorithm along with surplus rectangle fill algorithm. The heat exchangers with common functions in the process were grouped and placed side by side with their heads pointed towards the longest edge of the site layout. The location of each group was calculated with the heat exchangers at fixed location within the groups. The pumps were placed in proper areas and restricted to being on the first floor to avoid cavitation. The columns and reactors were also restricted to being on the first floor so they could span across the higher floors.

Safety issues in the design of the layout of chemical process plants were accounted for by Caputo et al. (2015). Fatalities caused by process equipment mechanical failure were included in the objective function as a cost source. In the methodology proposed by Xu et al. (2020), the placement of the process equipment within frames was used to find the optimal layout. The process equipment location was calculated at different levels so as to minimise the length of pipe within the frames and between the frames. The heat exchangers arranged in parallel were modelled as groups instead of individual items. Tall process equipment was placed outside the frames.

### Heat exchangers: network and plant layout considerations

Amongst the process equipment commonly used within most chemical plants, process-to-process heat exchangers play a paramount role in the amount of heat recovered in the process. The consumption of utilities, such as steam and cold water, can be significantly reduced by means of the heat exchange between the process streams that require heating (cold streams) and those that require cooling (hot streams) (KEMP; LIM, 2020). To explore the potential for heat recovery, there may exist several process-to-process heat exchangers in a process, thereby forming a heat exchanger network.

The design of HEN is carried out with the goal of reaching a balance between the cost of purchasing the heat exchangers and the cost of utility consumption (SMITH, 2005). The methodologies for HEN design are well-established and available in commercial software. HENs can be designed by sequentially setting targets for the minimum utility consumption, minimum number of heat exchange units and minimum total area of heat exchange (ROETZEL et al., 2019). The HEN design can also be done by simultaneously minimising the mentioned costs. The simultaneous HEN design is modelled as an optimisation problem, which has been solved using mixed integer nonlinear programming (MINLP) and stochastic methods (PAVÃO et al., 2018; BIEGLER et al., 1997).

In process plants, heat exchangers are commonly arranged in pairs but they can also be arranged in groups or placed as individual items. Heat exchangers arranged in pair or in groups may operate in series or in parallel. Heat exchangers can be mounted at the ground level over concrete piers or in elevated structures. They are usually mounted horizontally but shell and tube heat exchangers can also be mounted vertically and attached to distillation columns (BAUSBACHER; HUNT, 1993).

Requirements of the chemical process dictate the location of some types of heat exchangers, such as thermosyphon reboilers, which are usually located adjacent to the related columns. Maintenance is also an important factor when defining the location of heat exchangers in the plant. For example, shell and tube heat exchangers should be placed with the channel end facing an access road. The purpose of doing that is to provide enough space for the tube bundle to be removed. Also, proper space should be left around the front end for bonnet removal. Similar requirements are found in other types of heat exchangers, such as plate heat exchangers and spiral heat exchangers.

### Costs of pipe and pumping of the HEN

The location of heat exchangers in the plant can be used to calculate the costs of pipe and pumping for the transport of the process streams. In the context of HEN, the costs of pipe and pumping were calculated in a small number of works. One example is that published by Pouransari and Maréchal (2014). These authors presented a method to calculate the cost of pipe of the HEN. This information was used to compare different HENs [3].

The method presented by Pouransari and Maréchal (2014) was used by Souza et al. (2016). They studied how the costs of pipe and pumping affects the topology of HENs. The cost of pumping included the pressure drop from the pipes as well as heat exchangers, which was calculated using the method proposed by Polley and Shahi (1991). A similar study was carried out by Rathjens and Fieg (2018) considering only the cost of pipe. In both works, the superstructure for HEN design was used (YEE; GROSSMANN, 1990). The minimisation of costs of pipe and pumping resulted in heat exchanges between process streams that were separated by shorter distances.

In inter-plant heat integration, process streams from different plants exchange heat with each other. The location of the heat exchangers within a single plant is not a major issue. The focus is in the plant in which the heat exchanger will be located. In a situation where the plants operate under different startup and shutdown schedules, spare heat exchangers should be provided for the process streams to reach their target temperatures (CHEW; KLEMEŠ; ALWI; MANAN, 2013). Additional heat exchangers may be needed if the operating conditions of a plant were modified, such as the process stream mass flow rate. The distance between the plants can lead to high cost of pipe (RODERA; BAGAJEWICZ, 2001).

Nair, Guo, et al. (2016) proposed a MINLP methodology for the design of inter-plant HEN with a central station for the heat exchangers to be placed. Later, Nair, Soon, et al. (2018) extended that methodology to allow the placement of heat exchangers in the plants in addition to the central station. The costs of utilities, heat exchangers, piping and pumping were considered.

Akbarnia et al. (2009) proposed a method to calculate the cost of pipe to transport a process stream through a number of heat exchangers prior to the HEN design. It was assumed that there was a pipe transporting the process stream from a pipe rack to a heat exchanger and then back to the pipe rack. The length of pipe to transport the process stream, the location of the pipe rack and heat exchanger, the piping auxiliaries and the material of construction were assumed. The expected number of heat exchange units for a process stream was determined and multiplied by the cost of pipe to transport the process stream through a single heat exchanger.

In HEN retrofit, pressure drop is a limitation for the task of defining the location of heat exchangers in the plant. The pressure drop incurred from the flow along the pipes as well as heat exchangers should not exceed the maximum allowed pressure drop fixed by the available pumping equipment. When that pressure drop lower boundary is exceeded, it is necessary to rectify the HEN topology or purchase additional pumping equipment. In this case, the cost of pumping may outweigh the economic benefits achieved from the heat recovery (CHEW; KLEMEŠ; ALWI;

[3] Detailed information on that method will be presented in Chapter 2.

MANAN; REVERBERI, 2015).

In the context of HEN retrofit, Lai et al. (2019) proposed a set of heuristics to determine the location of new heat exchangers so as to reduce costs of pipe and pumping. For two streams matched for exchange heat, the heat exchanger should be placed prioritising the minimisation of the length of the pipe that transports the process stream with the higher: volumetric flow rate, viscosity, pumping head limit and positive pressure drop, in that order. Although the location of heat exchangers was defined considering various issues, no automated procedure was presented for the application of the heuristics. Besides that, the heuristics might not lead to the optimal costs of pipe and pumping.

### Motivation

This literature review reveals that there are several works that, in an incomplete manner, addressed the **heat exchanger plant layout** design. The methods used in those works are limited mainly in that they do not provide the location of each heat exchanger in the plant. This limitation is well represented in the works published by Pouransari and Maréchal (2014), Souza et al. (2016) and Rathjens and Fieg (2018).

The mentioned authors calculated the cost of pipe required to transport the process streams to exchange heat. Since, the cost of pipe is a function of the distance between the pieces of equipment, we expected that Pouransari and Maréchal (2014) calculated the location of heat exchangers in the plant. However, they did not do that. Instead, they defined a single location for each process stream in the plant. Then, they assumed that the cost of pipe was proportional to the distance between the location of the process streams being matched for heat exchange.

A consequence of not calculating the location of heat exchangers in the plant is that the cost of pipe cannot be calculated for each process stream individually. This is reflected in the method used by Pouransari and Maréchal (2014), Souza et al. (2016) and Rathjens and Fieg (2018). The cost of pipe calculated by them is associated to each pair of streams matched for heat exchange. The method that these authors used is also limited in the fact it uses a single location to the process streams. In practice, the process streams are transported from one piece of equipment to another and, thus, have multiple locations. The consequence of neglecting it is an inaccurate cost of pipe.

The location of heat exchangers in the plant was, to a certain extent, calculated in the works whose focus was on the general design of the layout of chemical process plants (EJEH et al., 2018; WANG et al., 2018; CAPUTO et al., 2015; XU et al., 2020). However, those works do not provide the location of the heat exchangers individually. Instead, they group the heat exchangers into a single block and provide the location of the block in the plant. This modelling approach does not allow the evaluation of:

- the effect of placing the heat exchangers as single items or in smaller groups on the costs of pipe and pumping;
- the change in the costs of pipe and pumping with respect to different HEN topologies.

The works on inter-plant heat integration (CHEW; KLEMEŠ; ALWI; MANAN, 2013; RODERA; BAGAJEWICZ, 2001; NAIR; GUO, et al., 2016; NAIR; SOON, et al., 2018) provide the plant where the heat exchangers are placed. However, no information is provided on the location of heat exchangers inside the plants. The method proposed by Akbarnia et al. (2009) assumed, instead of calculating, the location of heat exchangers in the plant along with the length of pipe connected to the heat exchanger.

## 1.3 Objectives

The objective of this thesis is to develop a set of methods to assist engineers in the design of the heat exchanger plant layout. There will be a method concerned with each of the following topics:

- assisting the HEN design by narrowing down the number of possible topologies.

Throughout the HEN design, there may exist many possible topologies that satisfy the energy balance of the process streams and lead to costs within the expected order of magnitude. The number of HEN topologies can be narrowed down based on the pipe length required for the transport of the process streams. A method will be developed to provide a rough estimate of the pipe length required to transport the process streams in a heat exchange. By assigning a pipe length value to each process stream match for heat exchange, this method will allow the quick reduction of the number of possible HEN topologies.

- screening the various locations of heat exchangers to evaluate the pipe length required for the transport of the process streams and the space available for maintenance work:

Once the HEN is designed, the heat exchangers can be arranged in many ways within the plant. The heat exchanger arrangements may differ from each other in terms of the pipe length required for the transport of the process streams and the space available for maintenance work. A method will be developed to design the heat exchanger plant layout that requires the minimum total pipe length. This method will allow the generation of various heat exchanger arrangements by specifying:

  - different priority levels to minimise the pipe length of the process streams;
  - the minimum allowed spacing between heat exchangers;
  - zones of the plant where the heat exchangers can be placed.

- evaluating how the HEN topology is affected by the costs of pipe and pumping.

Usually, HENs are designed by taking the costs of utilities and heat exchangers into account. However, to transport the process streams to and from the heat exchangers, pipe and pumping are required. Does the HEN topology change if the costs of pipe and pumping are taken into account throughout the design process?

In order to answer this question, a method will be developed to design the HEN and heat exchanger plant layout in a simultaneous manner. As the heat exchanger plant layout is designed, the costs of pipe and pumping are calculated. This method will allow the minimisation of the costs of pipe and pumping along with those of utility consumption and heat exchangers.

## 1.4 Essential concepts

This section is intended to introduce essential concepts used throughout this thesis. It is essential to known the commonly used terminology of HENs and the way their topologies are typically depicted. Another topic is the transport of the process streams through the plant of the process. This topic is essential for the reader to understand the type of costs of pipe and pumping we will be referring to. General information will be provided on the path along which the process streams are transported and the procedure to calculate the length of pipe required to do that. This section will also present the assumptions made to develop the methods of this thesis.

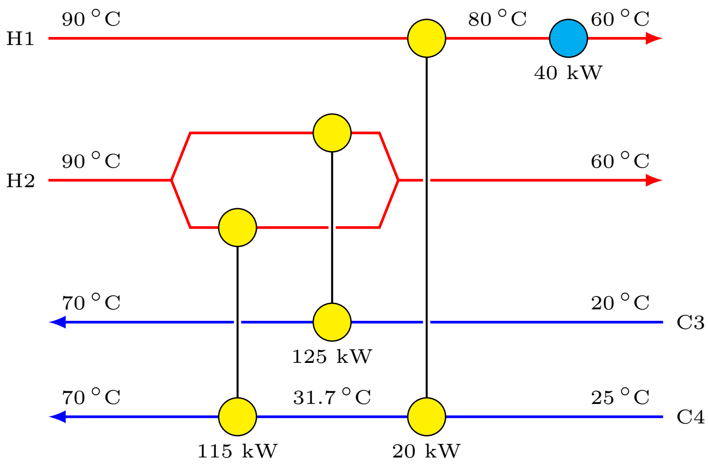
### 1.4.1 Heat exchanger network: commonly used terminology and topology

In the field of heat integration, the process streams that require cooling are usually called hot streams. In an analogous manner, the process streams that require heating are called cold streams. The temperature of the hot and cold streams before they exchange heat is usually called supply temperature. Similarly, their temperature after they exchanged their required amount of heat is called target temperature.

The HEN topology is typically depicted by the hot streams, cold streams and the heat exchangers. The hot and cold streams are represented by horizontal lines. The ends of the horizontal lines indicate the supply and target temperatures of the hot and cold streams. The heat exchange between a hot and a cold stream is represented by a vertical line linking those streams. Hot and cold **utility** streams are usually omitted from the HEN topology but they can also be shown by means of horizontal lines.

To depict the topology of a HEN, let us consider a process with two hot streams and two cold streams. Hot stream H1 exchanges heat with cold stream C3 (Figure 1.3). The other hot stream, labelled H2, exchanges heat in parallel with both the cold streams C3 and C4. At the end of hot stream H1 there is a cooler, where hot stream H1 exchanges heat with a cold utility stream. Cold stream C4 exchanges heat in series with the hot streams H1 and H2, respectively.

**Figure 1.3:** The topology of the HEN of a hypothetical process adapted from Serth and Lestina (2014)




The heat exchange between a hot and a cold stream is usually referred to as a heat exchange unit. For pure countercurrent flow, a heat exchange unit can be equivalent to a single heat exchanger (SHENOY, 1995). However, if the required heat transfer area is too large, then the construction of a single heat exchanger will be impractical. A heat exchange unit may constitute more than one heat exchanger if the streams flow in a noncountercurrent pattern.

### 1.4.2 The path along which the process streams are transported

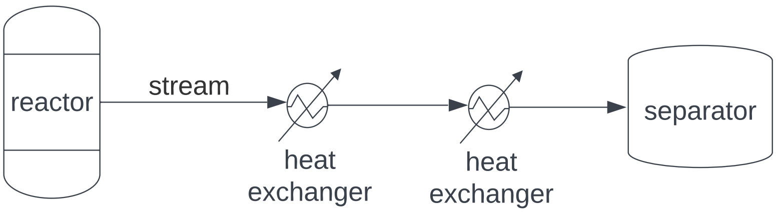
The length of pipe required to transport the process streams of the HEN will be calculated in all the mathematical models presented in this thesis. This will be done either directly in the objective function or as the basis to calculate the costs of pipe and pumping. To calculate the length of pipe it is necessary to know the path along which the process streams are transported. Each methodology presented in this thesis uses a specific stream path. However, all of the methodologies share the concept that the streams leave a piece of equipment at its target temperature, pass through one or more heat exchangers, and enters a piece of equipment at its target temperature.

The heat exchangers may be arranged in series or in parallel. If the process stream is transported to heat exchangers arranged in parallel, then there is a device installed before the heat exchangers to split the process stream into branches. Another device is installed after the heat exchangers in order to mix the branches of the process stream back. The coordinates of the piece of equipment from which a process stream leaves at its supply temperature will be called the **starting point**. In an analogous manner, the coordinates of the piece of equipment to which a process stream enters at its target temperature will be called the **endpoint**.

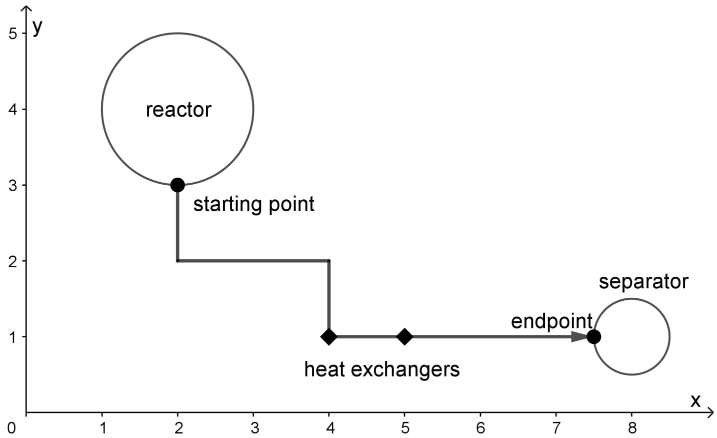
Figure 1.4a shows a hypothetical process. The stream leaves a reactor, passes through two heat exchangers in series and, enters a separator. The plant layout of this process is depicted in Figure 1.4b. The starting point of the stream is a nozzle of the reactor. Similarly, the endpoint of the stream is a nozzle of the separator.

**Figure 1.4:** A hypothetical process to illustrate the concept of starting and endpoint.

**(a)** The hypothetical process flowsheet diagram.




**(b)** The starting and endpoint of the stream marked in the equipment.




### 1.4.3 General procedure to calculate pipe length and stream path

Determining the accurate length of pipe is not straightforward. It requires information on the pipe routes considering physical obstacles, which is beyond the scope of this thesis. However, by assuming that there are not any physical obstacles and the pipes are routed orthogonally (SMITH; THOMAS, 1987), an underestimated length value can be obtained from the rectangular distance between the nozzles of the pieces of equipment being connected.

The rectangular distance between two points is defined as the absolute difference between their coordinates summed over the axes (CRAW, 2010), as shown below for the hypothetical points **A** and **B**:

$$l_{A,B,x} = |x_A - x_B| \tag{1.1}$$

$$l_{A,B,y} = |y_A - y_B| \tag{1.2}$$

$$l_{A,B} = l_{A,B,x} + l_{A,B,y} \tag{1.3}$$

### 1.4.4 General strategy to avoid the absolute function

The absolute function used to determine the rectangular distance between two points is a non-linear function (DANTZIG, 2002). In the solution of non-linear mathematical models, the solver may be trapped in a local optimum (WILLIAMS, 2013). To avoid such a type of difficulty, we will replace the absolute function with a set of linear inequalities.

To replace the absolute function, it is essential to obtain a formulation where the differences calculated from both $x$- and $y$-axis are positive values so that they can be summed to give the rectangular distance. At this point, the difficulty is that without using absolute function we may obtain either positive or negative values. To understand this, let us consider the coordinates of the hypothetical points $A$ and $B$ on the $x$-axis.

Also, let us consider that the absolute function in Equation (1.1) will be replaced with the subtraction of the coordinates of the points $A$ minus $B$ ($x_A$–$x_B$). As a result, we may obtain a positive value, if $x_A$ is greater than $x_B$ or a negative value otherwise. The same occurs if we define the inverse order for the subtraction ($x_B$–$x_A$). Thus, subtracting one coordinate from another in a single order does not guarantee we will obtain a positive sign.

To obtain the positive difference between the coordinates of the points $A$ and $B$ on the $x$-axis, we will restrict the variable for the length of pipe from $A$ to $B$ ($l_{A,B,x}$) to being greater than or equal to both the difference between $A$ and $B$ ($x_A$–$x_B$), and, $B$ and $A$ ($x_B$–$x_A$), as shown in Equation (1.4). The variable length of pipe from $A$ to $B$ on the $x$-axis will be included in the objective function. Since, the objective function will be orientated to minimisation, the variable length of pipe will be enforced to assume the minimum value possible that satisfies the inequalities.

$$l_{A,B,x} \geq x_A - x_B \tag{1.4}$$

$$l_{A,B,x} \geq x_B - x_A \tag{1.5}$$

The minimum value possible that satisfies the inequalities is exactly the positive difference between the coordinates of the points $A$ and $B$ on the $x$-axis. The same procedure presented for the $x$-axis can be used for the $y$-axis:

$$l_{A,B,y} \geq y_A - y_B \tag{1.6}$$

$$l_{A,B,y} \geq y_B - y_A \tag{1.7}$$

Having obtained the positive difference between the coordinates of the points $A$ and $B$ on the $x$- and $y$-axis from the equations (1.4) and (1.6), respectively, the rectangular distance between these points can be calculated by means of the previously presented Equation (1.3). To demonstrate the concept of rectangular distance in terms of pipe length, let us consider a process stream that passes through two heat exchangers in series. The length of pipe required to transport the process stream is equal to the summation of the rectangular distances between the coordinates of:

- the piece of equipment from which the process stream leaves at its supply temperature and the first heat exchanger;
- the two heat exchangers along the stream path;
- the last heat exchanger and the piece of equipment to which the process stream enters at its target temperature.

### 1.4.5 Assumptions to develop the mathematical models

In this thesis, the mathematical models will be developed under the following assumptions:

- only pipes that transport process streams will be considered to calculate pipe length, costs of pipe and cost of pumping
- the starting point and endpoint of the process streams[4] have coordinates known in the plant;
- a heat exchange unit of a HEN constitutes a single heat exchanger that operates with pure countercurrent flow;

[4]To simplify the writing of this thesis, from this section onwards the term *process* will be omitted from *process streams*.

- the heat exchangers, splitter and mixer devices are dimensionless pieces of equipment and represented as points in the plant layout;

- the plant is represented in a bi-dimensional Cartesian plane using the $x$- and $y$-axis.

## 1.5 Thesis outline

The remaining part of this thesis is organised into five chapters. Chapter 2 establishes a relation between the transport of the streams through the process plant and HENs. This chapter shows that exchanging heat between two streams requires additional pipe for their transportation. A method is presented to calculate the length of the additional pipes. This method can be used to assist the design of HENs by allowing the comparison of the additional pipe required by the various streams matches for heat exchange.

The chapters 3 and 4 focus on the heat exchanger plant layout design. These chapters show that the heat exchanger plant layout can be designed by minimising the total length of pipe for the transport of the streams. A method to do that is developed in Chapter 3. In addition to pipe length, the space available for maintenance work is also taken into account. Chapter 4 develops a method where pipe length is not only used to design the heat exchanger plant layout but also to design HENs.

In Chapter 5, the method developed in Chapter 4 is extended from total length of pipe to the costs of pipe and pumping. The extended method allows the HEN design by minimising the costs of pipe and pumping. Other costs can also be included in the task of designing HENs. In this way, Chapter 5 discusses the influence of the costs of pipe and pumping on the HEN design. The conclusion of this thesis is presented in Chapter 6.

# Chapter 2

# Estimate of the additional pipe length due to heat exchange

The scientific literature is rich in methods for HEN design. Those methods typically search a topology for the HEN that minimises the costs of utilities and heat exchangers. Within a single HEN, there may exist various topologies with the same costs of utilities and heat exchangers. To narrow down the number of topologies, additional characteristics of the HEN can be included in the design problem.

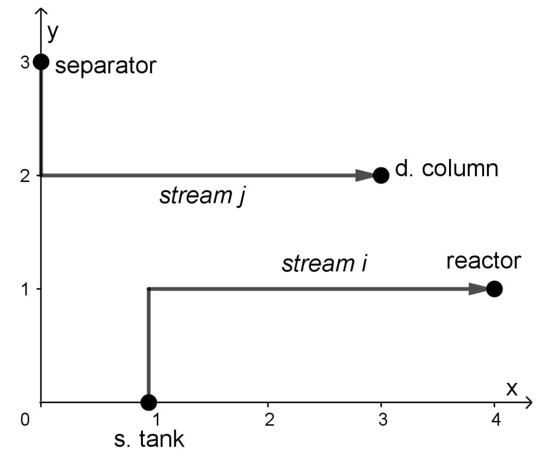
The central theme of this chapter is to to evaluate the HEN topologies in terms of the pipe length required to transport the streams. This chapter will demonstrate that transporting the streams through heat exchangers can increase the pipe length. Although not categorically stated, the idea of using pipe length to narrow down the number of HEN topologies is not novel. A review of the existing method will show how this idea works and that there is room for method extension.

The main contribution of this chapter is a method that calculates the shortest additional pipe length required to transport two streams to exchange heat with each other. The concept, the limitations and the mathematical model of this method will be presented. An example will compare this method with the existing one and its extended version.

## 2.1 The concept of additional pipe length due to heat exchange

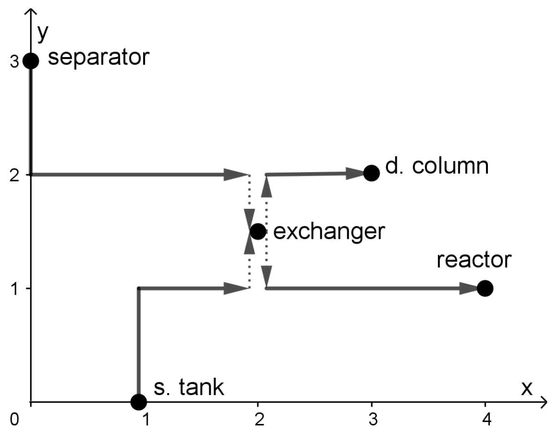
Consider a process with two streams, $i$ and $j$. Hot stream $i$ is transported from a storage tank to a reactor. Cold stream $j$ is transported from a separator to a distillation column. Figure 2.1 depicts the plant layout. By means of the plant layout, we can determine that to transport the streams from one piece of equipment to another, it is necessary to provide pipe 4 lu long. Let us now consider the process was modified. In the modified setting, hot stream $i$ needs to be cooled and cold stream $j$ needs to be heated before they reach the reactor and the distillation column, respectively.

**Figure 2.1:** The plant layout where each square has a unit length size.




A heat exchanger is then placed at the coordinates (2, 1.5). Hot stream $i$ is transported from the storage tank to the heat exchanger and, from the heat exchanger to the reactor. Cold stream $j$, in its turn, is transported from the separator to the heat exchanger and, from the heat exchanger to the distillation column. Figure 2.2 depicts the route of the pipes through the plant with the heat exchanger.

**Figure 2.2:** The plant layout of the modified process. The solid line represents the pipe routes remained from the original process. The dashed line represents the pipe routes created for the modified process.




In the modified process, the length of pipe required to transport each stream increased from 4 to 5 lu compared with the original process. Thus, the additional pipe length due to heat exchange is 1 lu for each stream. Notice that if the heat exchanger is shifted along the interval $1 \leq y \leq 2$, then the additional pipe length due to heat exchange changes with respect to the streams. However, the sum of the additional pipe length over the streams remains the same.

For example, if the heat exchanger is shifted from $y = 1.5$ to $y = 1.3$, then the additional pipe length due to heat exchange for hot stream $i$ decreases from 1 lu to 0.6 lu. In contrast, that for cold stream $j$ increases from 1 lu to 1.4 lu. The values obtained for the streams change by the same amount in opposite signs. As a result, the sum of additional pipe length due to heat exchange over the streams remains 2 lu.

This characteristic makes useful to represent the additional pipe length due to heat exchange in its summed form. In doing so, the additional pipe length due to heat exchange can be read in terms of the stream match instead of each stream individually. In a typical design of HEN, there are various matches between the streams for heat exchange. Those stream matches can be compared in terms of their additional pipe length due to heat exchange.

Another advantage is that we do not need to concern with the exact coordinates of the heat exchanger in the plant. We can assume that the heat exchanger is placed inside the intervals where the summed additional pipe length due to heat exchange is constant. For the remaining part of this chapter, we will adopt the additional pipe length due to heat exchange summed over the streams as the standard form.

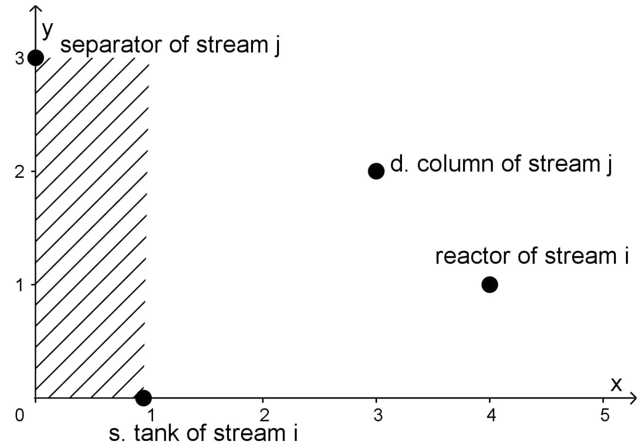
## 2.2 The existing method to estimate the additional pipe length due to heat exchange

The additional pipe length due to heat exchange was estimated by Pouransari and Maréchal (2014) to compare different topologies for the same HEN. In order to do that, these authors presented a method where each stream has a single location in the plant of the process. The additional pipe length due to heat exchange is calculated as twice the rectangular distance between the location of the streams.

The method of Pouransari and Maréchal (2014) has a few characteristic that will be discussed in the next paragraphs. The first characteristic is that each stream has a single location in the plant. In the context of HEN, we understand that there are two relevant locations for each stream apart from the heat exchanger location. One of them is the location of the piece of equipment from which a stream leaves at its supply temperature. The other one is the location of the piece of equipment to which a stream enters at its target temperature. Pouransari and Maréchal (2014) did not specify which piece of equipment the location was referring to.

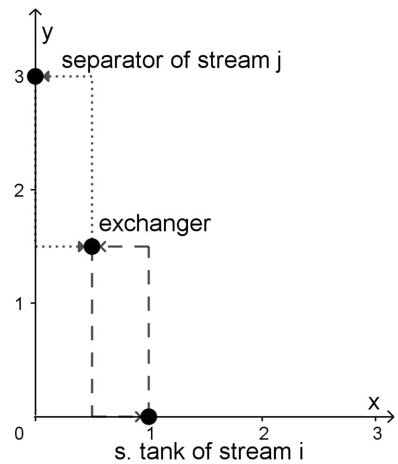
Another characteristic of the method of Pouransari and Maréchal (2014) is that no information was given on the heat exchanger location and the path of the streams. As a consequence, it is not clear the reason why to calculate twice the rectangular distance between the location of the streams. From our viewpoint, there is an inherent assumption that the heat exchanger is located somewhere inside the rectangle formed by the location of the streams. This assumption is illustrated in Figure 2.3, in which a rectangle is formed by the pieces of equipment from which the streams $i$ and $j$ leave at their supply temperatures.

**Figure 2.3:** The rectangle formed by the pieces of equipment from which the streams $i$ and $j$ leave at their supply temperatures.




Since the path of the streams was not described, we will give the method our interpretation. We first assume that the heat exchanger is located in the middle of the rectangle. That being said, the streams are transported from their pieces of equipment to the heat exchanger (Figure 2.4). After exchanging heat, the streams are transported back to their pieces of equipment. The total length of pipe required in this transportation is the additional pipe length due to heat exchange.

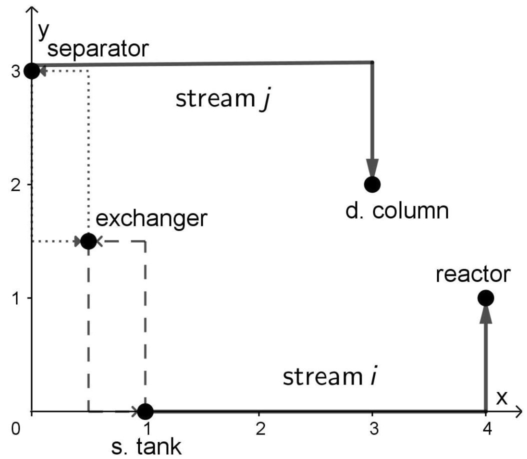
**Figure 2.4:** The path of the streams $i$ and $j$ to exchange heat.




In Figure 2.4, the length of pipe required to transport each stream from its piece of equipment to the heat exchanger is 2 lu. The same length of pipe is required to transport each stream from the heat exchanger back to its piece of equipment. As a result, the total length of pipe required to transport each stream is 4 lu, or it is 8 lu for both streams.

If the heat exchanger is placed in another location, then one of the pipes will be longer than the other one. However, as long as the heat exchanger is inside the rectangle, the total length of pipe for both streams remains 8 lu. Once the streams are back to their locations, they can be transported to the pieces of equipment to which they enter at their target temperatures (Figure 2.5). The heat exchanger does not affect the pipe length anymore.

**Figure 2.5:** The path of the streams $i$ and $j$ after they exchanged heat.




## 2.3 Extension of the existing method

The method presented by Pouransari and Maréchal (2014) may overestimate the additional pipe length due to heat exchange. The problem lies in the fact that this method defines a single location for each stream. As a consequence, a single value of rectangular distance can be calculated between the location of two streams. As mentioned earlier, from our viewpoint, each stream has two locations. Thus, it is possible to combine the location of the streams so as to calculate four values of rectangular distance. In doing so, we are able to take the shortest of the four values as the additional pipe length due to heat exchange.

In order to understand that, let us consider a process where the streams $i$ and $j$ exchange heat with each other. We will assume that the location defined for the streams corresponds to the pieces of equipment from which they leave at their supply temperatures. The rectangular distance between these pieces of equipment determines the additional pipe length due to heat exchange. This situation was previously depicted in Figure 2.3.

Now, let us include the pieces of equipment to which the streams enter at their target temperatures. By doing that, each stream will have two locations. For the sake of simplicity, we will refer to the location of the streams by means of the following terminology:

- starting point: the piece of equipment from which the stream leaves at its supply temperature;
- endpoint: the piece of equipment to which the stream enters at its target temperature.

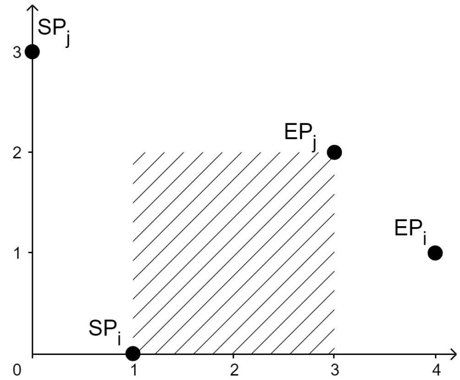
Since the starting and the endpoint are defined as the location of the streams, it is possible to calculate the rectangular distance between:

- the starting points of the streams $i$ and $j$ (Figure 2.3);
- the starting point of hot stream $i$ and the endpoint of cold stream $j$ (Figure 2.6a);
- the endpoint of hot stream $i$ and the starting point of cold stream $j$ (Figure 2.6b);
- the endpoints of the streams $i$ and $j$ (Figure 2.6c).

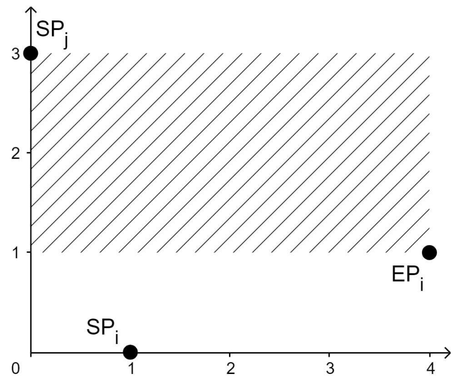
The four values of rectangular distance can be compared. The longest value is obtained from the rectangular distance between the endpoint of hot stream $i$ and the starting point of cold stream $j$ (Figure 2.6b), which is 6 lu. The value obtained from the rectangular distance between the starting points of both streams is 4 lu (Figure 2.3). The same value is obtained from the rectangular distance between the starting point of hot stream $i$ and the endpoint of cold stream $j$ (Figure 2.6a). The rectangular distance between the endpoints of both streams is 2 lu (Figure 2.6c), which is the shortest value. The additional pipe length due to heat exchange can be calculated by multiplying the shortest value of rectangular distance by two.

**Figure 2.6:** The rectangles formed by the starting and endpoints of the streams $i$ and $j$.

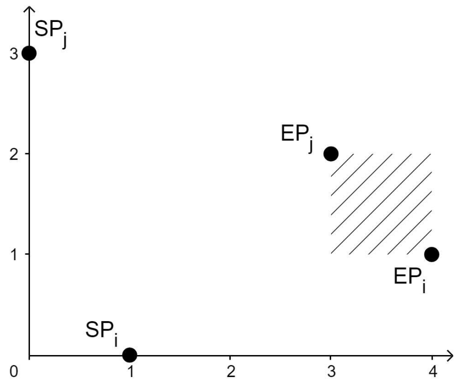
**(a)** Hot stream $i$*'s* starting point and cold stream $j$*'s* endpoint.




**(b)** Hot stream $i$*'s* endpoint and cold stream $j$*'s* starting point.




**(c)** Streams $i$ and $j$*'s* endpoints.




# 2.4 The Shortest Length Method

## 2.4.1 Concept

The method presented by Pouransari and Maréchal (2014) is limited in that it defines a single location for each stream. Based on our interpretation, that location is either the starting or the endpoint of the streams. We showed that by not including both starting and endpoints, three values of rectangular distance are missed. As a consequence, the additional pipe length due to heat exchange may be overestimated. To overcome this limitation, we proposed to calculate the rectangular distance between the starting and endpoints of two streams in the four possible combinations. The shortest value of the four can be multiplied by two and taken as the additional pipe length due to heat exchange.

Although a shorter value can be found by considering both the starting and endpoints of the streams, that value may still be overestimated. In this chapter, we will show how to find the shortest additional pipe length due to heat exchange. The central idea is to calculate the rectangular distance between the routes from the starting to the endpoints instead of the starting and endpoints individually. In doing so, we will be able to find the coordinates in the plant where the streams come closest to each other as they are transported by pipes. By placing the heat exchanger at those coordinates, the additional pipe length will be the shortest possible.

The sequence of steps mentioned above will be called the Shortest Length Method (SLM). To introduce the SLM, let us consider the streams $i$ and $j$ with the starting and endpoints defined in the plant (Figure 2.7). The streams $i$ and $j$ will be transported by pipes from their starting to their endpoints. There are many routes that the pipes can take to do that, a number of them is depicted in Figure 2.8.

**Figure 2.7:** The plant layout with the starting and endpoints of the streams $i$ and $j$.

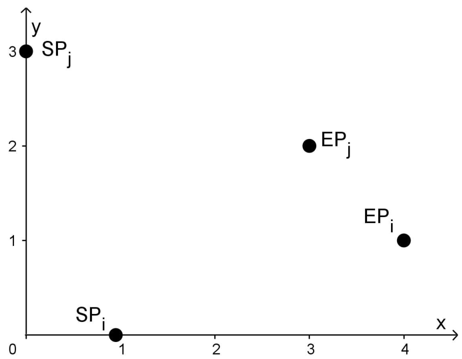
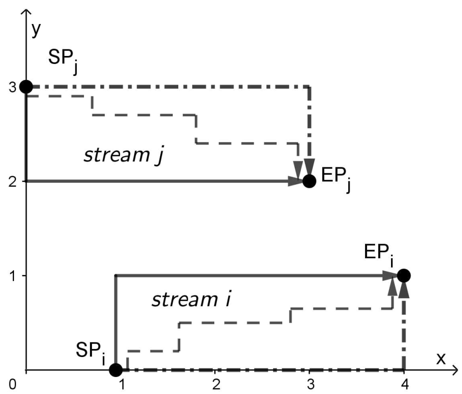



**Figure 2.8:** Three routes of same length connecting the starting and endpoints of the streams $i$ and $j$.




Since the coordinates of the starting and endpoints of the streams are known, we are able to determine the intervals on the rectangular axes along which the pipes will be routed. For example, hot stream *i's* starting point is located at the coordinates (1, 0) and endpoint at the coordinates (4, 1). Thus, the pipe that transports hot stream $i$ will be routed along the intervals:

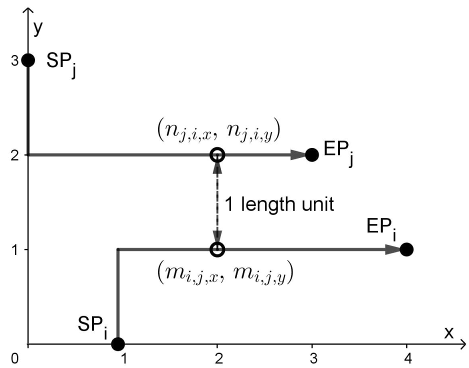
$$1 \leq x \leq 4$$
$$0 \leq y \leq 1$$

In an analogous manner, the coordinates of cold stream *j's* starting point is (0, 3) and endpoint is (3, 2). Any route taken by the pipe to transport cold stream $j$ will be encompassed by the intervals:

$$0 \leq x \leq 3$$
$$2 \leq y \leq 3$$

Having the intervals where the pipes will be routed, we are able to find which are the coordinates where the streams $i$ and $j$ come closest to each other. In order to that, we should minimise the rectangular distance between the coordinates encompassed by the intervals. The result is that hot stream $i$ comes closest to cold stream $j$ at the coordinates (2, 1). Cold Stream $j$ does the analogous at the coordinates (2, 2)[1]. Figure 2.9 depicts these coordinates in the plant layout. The rectangular distance between these coordinates is 0 lu on the $x$-axis and 1 lu on the $y$-axis.

**Figure 2.9:** The coordinates where the streams $i$ and $j$ come closest to each other.




Let us now consider that the streams $i$ and $j$ will exchange heat with each other. The pipes will transport the streams $i$ and $j$ from their original routes to the heat exchanger. Once the streams $i$ and $j$ exchanged heat, the pipes will transport them back to their original routes. We know the coordinates where the streams $i$ and $j$ come closest to each other. If we place the heat exchanger between those coordinates, then the additional pipe length due to heat exchange will be the shortest possible.

[1]Notice that the streams $i$ and $j$ pass along the same interval on the $x$-axis, which ranges from $x = 1$ to $x = 3$. Therefore, any coordinate inside this interval would be acceptable. In an arbitrary manner, we picked the coordinate $x = 2$.

The heat exchanger should be placed at the coordinate $x = 2$ and at any coordinate from $y = 1$ to $y = 2$. We will assume that the heat exchanger is at the coordinates (2, 1.5). The plant layout with the heat exchanger placed at these coordinates was previously shown in Figure 2.2. The length of pipe required to transport hot stream $i$ from its original route to the heat exchanger and back to its original route is 1 lu. The same value will be required to do the analogous for cold stream $j$. Thus, the additional pipe length due to the heat exchange between the streams $i$ and $j$ is 2 lu. This value is equal to twice the rectangular distance between the coordinates where the streams $i$ and $j$ come closest to each other.

The coordinates of the heat exchanger affects the pipe length for the transport of each stream individually. However, as long as the heat exchanger is located between the coordinates where the streams $i$ and $j$ come closest to each other, the sum of the pipe length for the transport of each stream remains the same. To conclude this section, we can state the SLM as follows:

*the additional pipe length due to heat exchange is equal to twice the rectangular distance between the coordinates where the streams come closest to each other.*

### 2.4.2 Limitation

The SLM is limited in that it does not calculate the location of heat exchangers in the plant. As a consequence, it is not possible to calculate the shortest additional pipe length due to **multiple** heat exchanges, such as heat exchangers in series or in parallel. To understand that, we shall remember that the SLM assumes that after leaving the heat exchanger, the streams are transported back to their original routes.

This assumption allows the SLM to calculate the additional pipe length due to heat exchange from the coordinates on routes of the streams. Thus, the location of heat exchangers in the plant is not required. In multiple heat exchanges, after leaving the heat exchanger, the streams are transported to the next heat exchanger and so forth. To calculate the pipe length from one heat exchanger to another, the location of heat exchangers in the plant is required.

There is another consequence of not calculating the location of heat exchangers in the plant. The additional pipe length due to heat exchange is given in terms of the stream matches. In this way, there is no information about how many length units of additional pipe is required for the streams individually. It would be possible to obtain this information by calculating the rectangular distance between the location of heat exchangers in the plant the coordinates on the route of the streams. It is worth mentioning that the other methods discussed in this chapter present the same limitations as those of the SLM.

### 2.4.3 Mathematical model

This section presents a linear programming model to calculate the shortest additional pipe length due to heat exchange. The assumptions previously presented in Subsection 1.4.5 are applied here. The input data is:

- a set of hot streams;
- a set of cold streams;
- a set of rectangular axes.

The parameters are:

- the hot and cold streams matched for heat exchange;
- the coordinates of the starting and endpoints of the hot and cold streams;

The linear programming model will be composed of:

**decision variables:** the coordinates of the streams;

**dependent variables:** the rectangular distance between the coordinates of the streams;

**constraints:** the coordinates of the streams should be encompassed by their starting and endpoints;

**objective function:** the minimisation of the rectangular distance between the coordinates of the streams.

#### Input data and parameters

The set of hot streams and cold streams will be represented by the indices $\{i = 1, \ldots, n_i \,|\, i \in \mathrm{H}\}$ and $\{j = 1, \ldots, n_j \,|\, j \in \mathrm{C}\}$, respectively. The matches between the the streams $i$ and $j$ for heat exchange will be represented by the parameter $\mathrm{k}_{i,j}$:

$$\mathrm{k}_{i,j} = \begin{cases} 1, & \text{if stream } i \text{ is matched with stream } j \text{ for heat exchange;} \\ 0, & \text{otherwise.} \end{cases} \tag{2.1}$$

$$i \in \mathrm{H}, \quad j \in \mathrm{C}$$

The coordinates of the starting point and the endpoint of stream $i$ on the axes $\{a = x,\, y \,|\, a \in \mathrm{A}\}$ will be represented by the parameters $\mathrm{SP}_{i,a}$ and $\mathrm{EP}_{i,a}$, respectively. The nomenclature for cold stream $j$ is analogous.

**Variables and constraints**

Let us assume that the streams $i$ and $j$ exchange heat with each other ($\mathrm{k}_{i,j} = 1$). The coordinates where hot stream $i$ comes closest to cold stream $j$ will be represented by the variable $m_{i,j,a}$. Those coordinates should be encompassed by the starting and endpoint of hot stream $i$:

$$m_{i,j,a} \geq \min\left(\mathrm{SP}_{i,a}; \mathrm{EP}_{i,a}\right) \tag{2.2}$$

$$m_{i,j,a} \leq \max\left(\mathrm{SP}_{i,a}; \mathrm{EP}_{i,a}\right) \tag{2.3}$$

$$i \in \mathrm{H}, \quad j \in \mathrm{C} \mid \mathrm{k}_{i,j} = 1, \quad a \in \mathrm{A}$$

An analogous constraint applies to the coordinates where cold stream $j$ comes closest to hot stream $i$, which will be represented by the variable $n_{j,i,a}$:

$$n_{j,i,a} \geq \min\left(\mathrm{SP}_{j,a}; \mathrm{EP}_{j,a}\right) \tag{2.4}$$

$$n_{j,i,a} \leq \max\left(\mathrm{SP}_{j,a}; \mathrm{EP}_{j,a}\right) \tag{2.5}$$

$$i \in \mathrm{H}, \quad j \in \mathrm{C} \mid \mathrm{k}_{i,j} = 1, \quad a \in \mathrm{A}$$

The rectangular distance between the coordinates where the streams $i$ and $j$ come closest to each other will be represented by the variable $L_{i,j}$ and calculated by:

$$L_{i,j,a} \geq m_{i,j,a} - n_{j,i,a} \tag{2.6}$$

$$L_{i,j,a} \geq n_{j,i,a} - m_{i,j,a} \tag{2.7}$$

$$i \in \mathrm{H}, \quad j \in \mathrm{C} \mid \mathrm{k}_{i,j} = 1, \quad a \in \mathrm{A}$$

$$L_{i,j} = \sum_{a \in \mathrm{A}} L_{i,j,a} \tag{2.8}$$

$$i \in \mathrm{H}, \quad j \in \mathrm{C} \mid \mathrm{k}_{i,j} = 1$$

**Objective function**

The additional pipe length due to heat exchange will be given by the minimisation of twice the rectangular distance between the coordinates where the streams $i$ and $j$ come closest to each other:

$$\mathrm{Min} \sum_{i \in \mathrm{H}} \sum_{j \in \mathrm{C}} 2\, L_{i,j} \tag{2.9}$$

## 2.5 Application of the SLM

The present section shows the application of the SLM and compares it with the other two methods discussed in this chapter. All the computations will be carried out using the General Algebraic Modelling System (GAMS) software version 24.0.2 with CPLEX solver.

### 2.5.1 Case Study

To demonstrate the application of the SLM, we will adapt the bio-ethanol-to-gasoline process from (SERTH; LESTINA, 2014). The process consists of four hot streams and three cold streams with the main equipment shown in Figure 2.10. A plant layout will be assumed for the process, as depicted in Figure 2.11. The coordinates of the starting and endpoints are shown in Table 2.1. The streams are matched for heat exchange according to the HEN shown in Figure 2.12. Based on this information, the additional pipe length due to heat exchange will be calculated using the:

**Pouransari and Maréchal (2014)'s method:** to apply this method it is necessary to define a single location for each stream in the plant. We will define the location of each stream as the coordinates of their starting points;

**extended Pouransari and Maréchal (2014)'s method:** this method makes the calculations from the starting as well as the endpoint of the streams;

**SLM:** this method makes the calculations from the coordinates bounded by the starting and endpoints of the streams.

**Figure 2.10:** The adapted process flowsheet for the SLM.

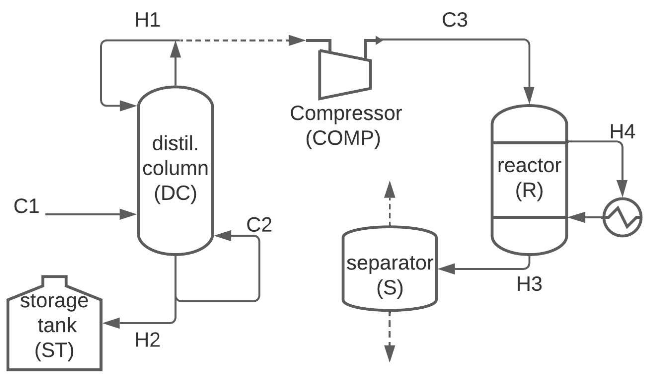


**Figure 2.11:** The plant layout assumed for the process of the SLM. The points labelled with the name of the streams and the indices $i$ and $f$ represent the starting and endpoints, respectively.

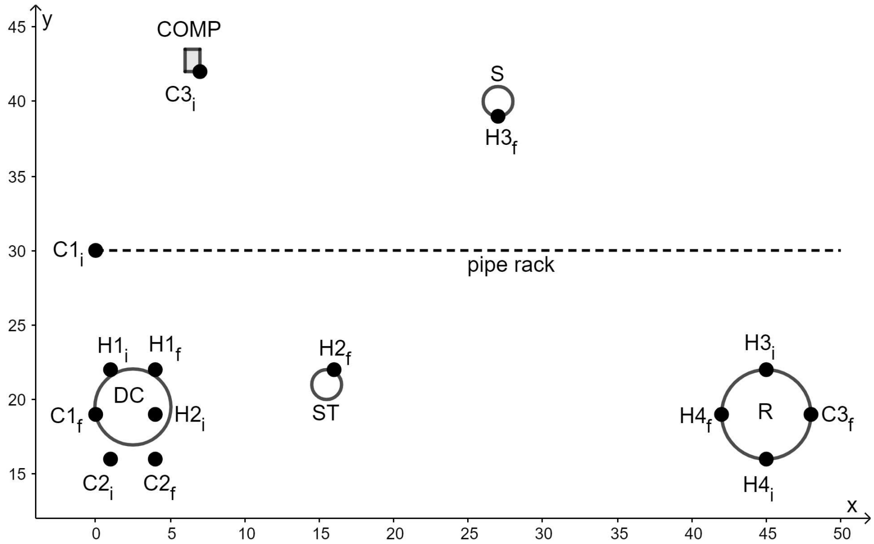
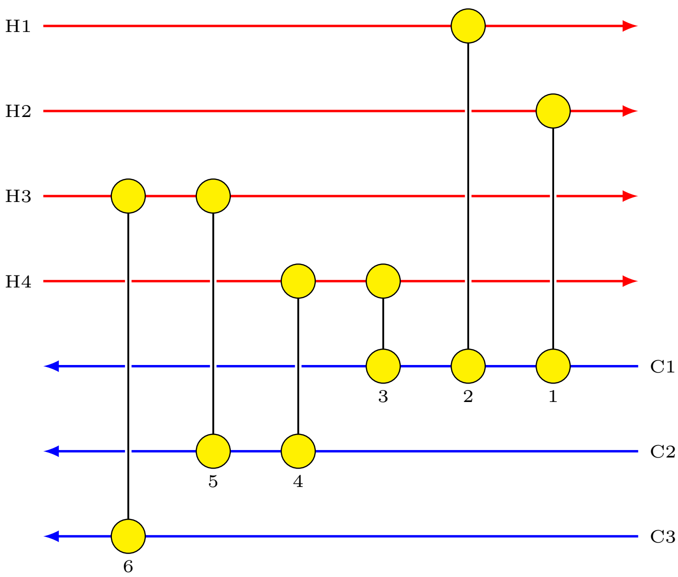



**Table 2.1:** The coordinates of the starting ($SP$) and endpoints ($EP$) of the streams for the SLM.

| stream | $SP$ | | $EP$ | |
|---|---|---|---|---|
| | $x_\mathrm{i}$ | $y_\mathrm{i}$ | $x_\mathrm{f}$ | $y_\mathrm{f}$ |
| H1 | 1 | 22 | 4 | 22 |
| H2 | 4 | 19 | 16 | 22 |
| H3 | 45 | 22 | 27 | 39 |
| H4 | 45 | 16 | 42 | 19 |
| C1 | 0 | 30 | 0 | 19 |
| C2 | 1 | 16 | 4 | 16 |
| C3 | 7 | 42 | 48 | 19 |

**Figure 2.12:** The topology of the HEN for the SLM.



### 2.5.2 Results and discussion

To compare the results, the additional pipe length due to heat exchange calculated by each method will be summed over the stream matches. The results are shown in Table 2.2. The total additional pipe length due to heat exchange is 470 lu calculated from the starting points of the streams only. By including the endpoints of the streams, this number decreases to 286 lu. The SLM decreases this number further to 228 lu.

**Table 2.2:** The additional pipe length due to heat exchange obtained from the SLM, the Pouransari and Maréchal (2014)'s method and its extended version.

| stream match | heat exchanger | using only starting points | using starting points and endpoints | SLM |
|---|---|---|---|---|
| H2-C1 | 1 | 30 | 8 | 8 |
| H1-C1 | 2 | 18 | 8 | 2 |
| H4-C1 | 3 | 118 | 84 | 84 |
| H4-C2 | 4 | 88 | 82 | 76 |
| H3-C2 | 5 | 100 | 92 | 58 |
| H3-C3 | 6 | 116 | 12 | 0 |
| total | | 470 | 286 | 228 |

#### Matches H4-C1 and H3-C3

The longest additional pipe length due to heat exchange was obtained from the starting points of the streams H4 and C1, which is 118 lu. The starting points of the streams H3 and C3 gives the second longest one: 116 lu. By including the endpoints, the additional pipe length due to the heat exchange between the streams H4 and C1 falls from 118 lu to 84 lu. The inclusion of the endpoints causes a similar but sharper decrease for the streams H3 and C3. Their additional pipe length due to heat exchange decreases from 116 lu to 12 lu.

A decrease of such magnitude was found by calculating the rectangular distance between the starting point of stream H3 and the endpoint of stream C3. The difference in magnitude between the results found by considering only starting points is explained by the plant layout. The starting points of the streams H3 and C3 are located in different pieces of equipment. In contrast, the starting point of stream H3 and the endpoint of stream C3 are located in the same pieces of equipment, which is the reactor.

By using the SLM, the additional pipe length due to the heat exchange between the streams H4 and C1 does not change compared to that obtained by including the endpoints. The SLM estimates a null additional pipe length due to the heat exchange between the streams H3 and C3. This result means that streams H3 and C3 share the same route in the plant layout as they are transported between their starting and endpoints.

**Matches H4-C2 and H3-C2**

Stream C2 exchanges heat in series with the streams H4 and H3, respectively. The additional pipe length due to heat exchange obtained from the starting points of these streams is 88 lu and 100 lu. By considering the endpoints, the additional pipe length due to heat exchange decreases by 6 lu and 8 lu. The magnitude of the decrease caused by the endpoints is not meaningful.

That magnitude is of the order of the rectangular distance between nozzles of a single piece of equipment. For example, stream C2 is the boil-up stream of the distillation column. The rectangular distance between its starting and endpoint is 3 lu. Another example is stream H4. This stream starts and ends in the reactor. It is used to remove heat from the reactor by circulating a cooling jacket. The distance between its starting and endpoint is 6 lu.

By using the SLM, the additional pipe length due to the heat exchange between the streams C2 and H4 decreases again by 6 lu. The small reduction obtained from the SLM is explained by the plant layout. Since the streams C2 and H4 start from and end in a single piece of equipment, they are not transported through a large portion of the plant. Consequently, they do not create regions where their routes come closer to each other.

The situation is different for the match between the streams H3 and C2. Stream H3 is transported from the reactor to the separator. As stream H3 is transported, it comes closest to the pipe that transports stream C2 in the surroundings of the distillation column. The SLM detects that stream H3 comes closest to stream C2 and calculates the additional pipe length due to heat exchange. The result is 58 lu. This result is 34 lu shorter than that found by including the endpoints.

**Matches H1-C1 and H2-C1**

The additional pipe length due to the heat exchange calculated from the starting points of the streams H1 and C1 is 30 lu and, the streams H2 and C1 is 18 lu. When the endpoints are included, the additional pipe length due to heat exchange is reduced to 8 lu for both stream matches. This result is achieved from the endpoint of stream C1 and the starting points of the streams H1 and H2.

The results obtained from the SLM shows that, the additional pipe length due to the heat exchange between the streams H1 and C1 decreases further from 8 lu to 2 lu. The analysis of the routes of the streams H1 and C1 reveals where they come closest to each other. This occurs when stream H1 is at its starting point (1, 22) and stream C1 is being transported between its starting and endpoint, more specifically at the coordinates (0, 22).

The additional pipe length due to heat exchange obtained from the SLM for the match between the streams H2 and C1 is 8 lu. As mentioned in the earlier (Section 2.4), the SLM calculates shortest the additional pipe length due to heat exchange. The value obtained from the SLM is the same as that obtained by including the endpoints. This means that depending on how the equipment is arranged in the plant, considering both the starting and endpoints may be enough to find the shortest additional pipe length due to heat exchange.

## 2.6 Conclusion

This chapter introduced the concept that installing a heat exchanger in the plant may increase the pipe length required for transporting the streams. The magnitude of that increase was termed by us the additional pipe length due to heat exchange. Three methods were presented to calculate the additional pipe length due to heat exchange.

The method published by Pouransari and Maréchal (2014) defines a single location for each stream in the plant. Based on our understanding, the defined location is either the starting or the endpoint of the stream. The heat exchanger is assumed to be somewhere between the location of the streams. To exchange heat, the streams are transported from their locations to the heat exchanger. After exchanging heat, the streams are transported from the heat exchanger back to their locations. The additional pipe length due to heat exchange is calculated by taken twice the rectangular distance between the location of the streams.

The Section 2.3 showed that defining a single location for each stream may overestimate the additional pipe length due to heat exchange. In order to overcome this limitation, we proposed to define two locations for each stream, which corresponds to their starting and the endpoint. This allowed us to calculate the rectangular distance not only between the starting points of the streams but also between the starting point of one stream and the endpoint of the other stream and, the endpoints of the streams. The additional pipe length due to heat exchange will be the shortest of the four rectangular distances.

The main contribution of this chapter is the shortest length method (SLM). This method finds the coordinates where two streams come closest to each other as they are transported from their starting to their endpoints. By assuming that the heat exchanger is placed between the coordinates found, the additional pipe length due to heat exchange is the shortest possible.

The SLM is not able to handle multiple heat exchangers. In addition, it does not calculate the location of heat exchangers in the plant. The other methods discussed in this chapter are not able to do that either. A linear programming model was presented for the implementation of the SLM. Its application was demonstrated for a six-stream-match HEN by means of the General Algebraic Modelling System (GAMS). The results obtained from the SLM were compared with those obtained from the other methods discussed in this chapter.

A method to overcome the mentioned limitations will be presented in the next chapter. The novel method will calculate the location of heat exchangers in the plant by using mathematical optimisation. The objective will be to minimise the length of pipe for the transport of the streams. Unlike the SLM, which gives only the additional pipe length, the novel method will give the entire pipe length. Unlike the SLM, which is based on the concept of additional pipe length, the novel method will calculate the entire length of pipe. Multiple heat exchanges will be allowed, which includes in series as well as parallel heat exchange.

# Chapter 3

# Sequential HEN and plant layout design

The academic literature is abundant in methods for designing HENs. Once the HEN is designed, the heat exchangers have to be installed in the plant. The location where the heat exchangers will be installed is defined taking into consideration a list of requirements, which includes the cost of pipe for transporting the streams, spacing between heat exchangers, maintenance and safety. As a consequence, not all the locations of the plant are available for placing heat exchangers.

The heat exchangers can be arranged in the plant in different layouts while satisfying the mentioned requirements. The layouts may differ from each other in adequacy of the requirements. For example, grouping many heat exchangers may reduce the total pipe length. In contrast, placing heat exchangers separated may facilitate the access for the operators to carry out maintenance. The high number of heat exchangers in the chemical processes increases the complexity of the layout.

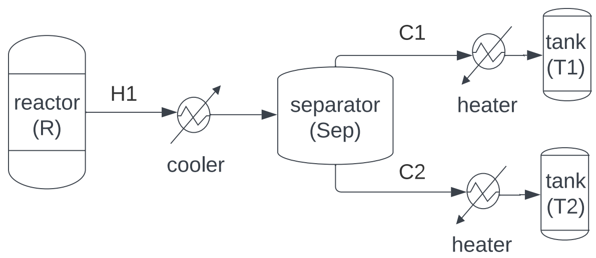
The issues discussed above motivated us to develop a method for the automated design of the heat exchanger plant layout. In the first section of this chapter, we will consider only the problem of the pipe length. In the next sections, we will include the problems of minimum equipment spacing and of which locations in the plant are available for heat exchanger placement. We will support the development of the method with examples.

## 3.1 Design of the heat exchanger plant layout: pipe length minimisation

Let us consider a process with hot stream H1 and the cold streams C1 and C2. Hot stream H1 leaves reactor R and is cooled before entering separator Sep. The cold streams C1 and C2 leave the separator, are heated, and stored in the tanks T1 and

T2, respectively. The process flowsheet is depicted in Figure 3.1.

**Figure 3.1:** The flowsheet of the three-stream process.




The reactor, the separator and the tanks have location defined in the plant of the process (Figure 3.2). The coordinates of the starting ($SP$) and endpoints ($EP$) are shown in Table 3.1. The minimum length of pipe required to transport a stream from its starting to its endpoint can be determined from the rectangular distance between their coordinates. This specific value of pipe length will be referred to as the *baseline* length of pipe. The baseline length of pipe required to transport each of these streams is 5 length units (lu).

**Figure 3.2:** The plant layout of the three-stream process.

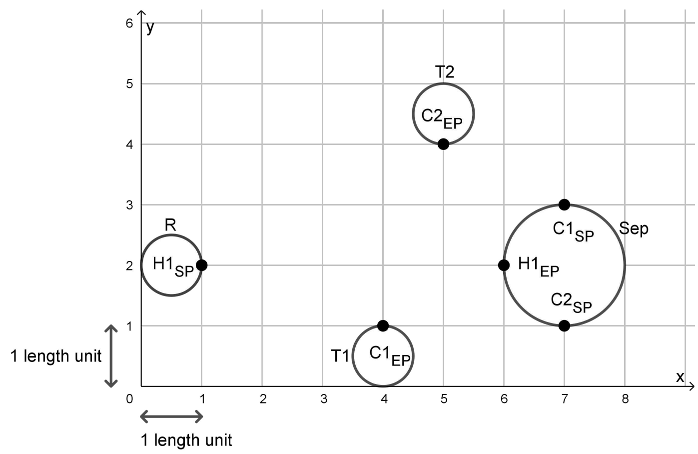



**Table 3.1:** The coordinates of the starting and endpoints.

| stream | starting point ($SP$) | | endpoint ($EP$) | |
|---|---|---|---|---|
| | $x$ | $y$ | $x$ | $y$ |
| H1 | 1 | 2 | 6 | 2 |
| C1 | 7 | 3 | 4 | 1 |
| C2 | 7 | 1 | 5 | 4 |

In a HEN proposed for this process, stream H1 exchanges heat with streams C1 and C2 in heat exchangers E1 and E2, respectively (Figure 3.3). The heat exchangers E1 and E2 should be placed in the process plant so as to minimise the length of the pipes used to transport the streams. By visual inspection, heat exchanger E1 can be placed, for example, at coordinates (4, 2) and heat exchanger E2 at coordinates (5, 2) while still keeping the pipes with the baseline length (Figure 3.4).

**Figure 3.3:** The topology of the HEN for the process with three streams.

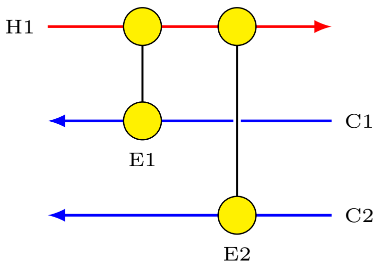
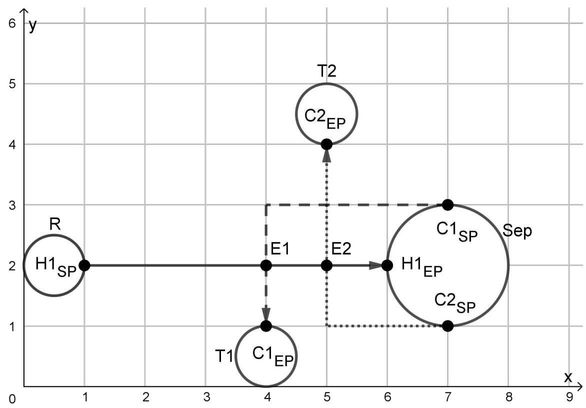



**Figure 3.4:** The heat exchangers E1 and E2 arranged so that the pipes have baseline length.




Depending on how the process equipment is set up in the plant it is not always possible to arrange the heat exchangers in such way the pipes have the baseline length. To illustrate this situation, let us consider that the endpoint of stream C1 was relocated from coordinates (4, 1) to (3, 4), as depicted in Figure 3.5. The baseline length of pipe remains 5 lu for each stream. Again, by visual inspection, the location of the heat exchangers that minimises the length of the pipes is heat exchanger E1 at any coordinate inside the region formed by the intervals $3 \leq x \leq 6$ and $2 \leq y \leq 3$ and heat exchanger E2 at coordinates (5, 2). The region for the location of heat exchanger E1 that minimises the length of pipe is hatched in Figure 3.6.

**Figure 3.5:** The plant layout with the endpoint of stream C1 at coordinates (3, 4).

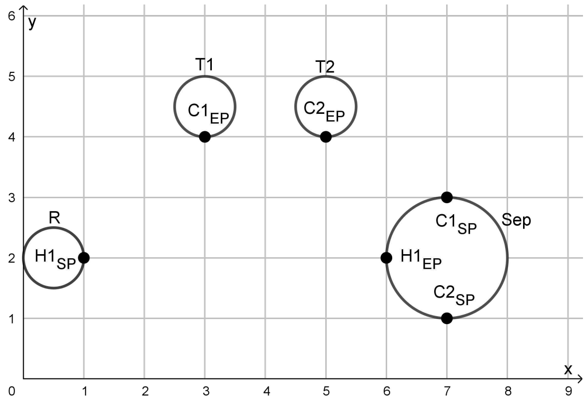
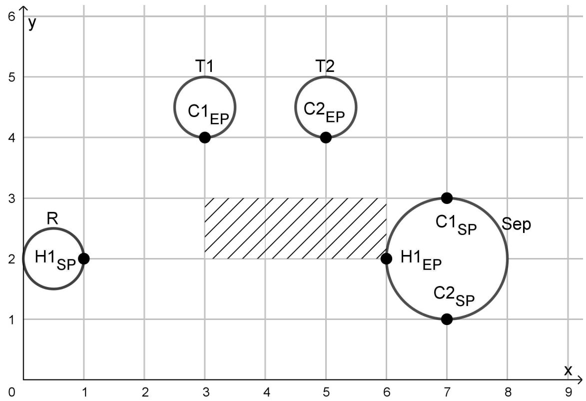



**Figure 3.6:** The intervals where heat exchanger E1 should be placed to minimise the pipe length.




As long as heat exchanger E1 is inside the hatched region and heat exchanger E2 is at coordinates (5, 2), the summation of the length of the pipes that transport streams H1 and C1 is 12 lu. For example, by placing heat exchanger E1 at coordinates (3, 2.5) the pipes that transport streams H1 and C1 are both 6 lu long (Figure 3.7). Since the baseline length of pipe for each stream is 5 lu, we can conclude that it is necessary to add 1 lu of pipe for the streams H1 and C1 to be transported. The pipe that transports stream C2 is still 5 lu long. Thus, the total length of pipe is 17 lu, 2 lu longer than the total baseline length of pipe.

With the sole purpose to minimise the length of pipe, the heat exchangers can be placed over each other as well as at a starting or and endpoint. Let us consider that heat exchangers E1 and E2 are at coordinates (6, 2), the same coordinates as the endpoint of stream H1. The pipe that transports stream C1 is 7 lu long and those that transport streams H1 and C2 are both 5 lu long (Figure 3.8). The total length of pipe remains 17 lu.

**Figure 3.7:** The arrangement of the heat exchangers in which the length of the pipes is longer than the baseline length of pipe.

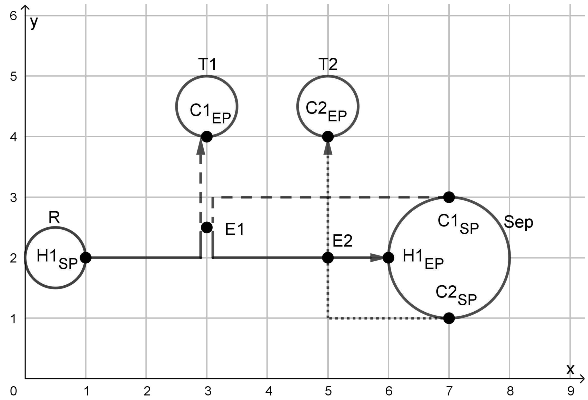
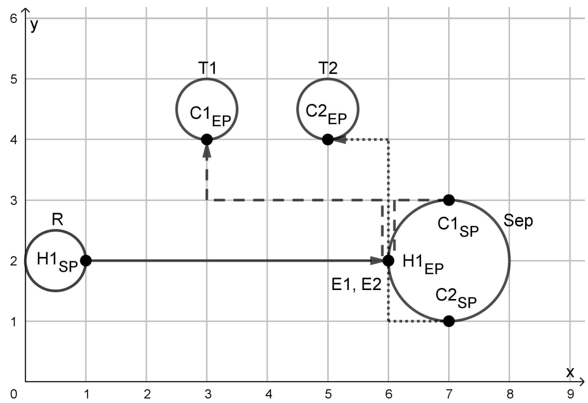



**Figure 3.8:** The heat exchangers overlapping each other and also an endpoint.




A simple process has been presented. The location of the heat exchangers that minimises the length of the pipes could be found in a straightforward manner. For a process with a greater number of streams, heat exchangers and a complex equipment layout, this task can be difficult. A hot stream may exchange heat with several cold streams and vice versa. The optimal location not always follow the sense of placing a heat exchanger as near as possible the routes that lead to the baseline length of pipe. The complexity increases with the presence of splitter devices, mixer devices and pipes with different cost per unit length.

This chapter is dedicated to automatising the design of the heat exchanger plant layout. In order to do that, we will formulate the design of heat exchanger plant layout as an optimisation problem. Each heat exchanger will have a variable to describe its coordinate in the plant of the process. The coordinates of the heat exchangers will be used to calculate the length of pipe for transporting the streams. The solution of the optimisation problem will be the coordinates of the heat exchangers that minimises the total length of pipe.

### 3.1.1 Mathematical model

A linear programming model will be developed. The input data is the set of hot streams, cold streams and the set of rectangular axes. The parameters are the HEN topology and the coordinates of the starting and endpoints of the streams. The decision variables are the coordinates of the heat exchangers, splitter devices and mixer devices. The dependent variables are the length of the pipes to transport the streams. The constraints are the equations to calculate the length of the pipes. The objective function is defined as the minimisation of the summation of the length of the pipes. The assumption previously presented in Subsection 1.4.5 are applied here.

#### Input data and parameter: the set of streams and HEN topology

A HEN can be represented by a set of H hot streams, C cold streams and U devices, such as heat exchangers, splitters and mixers[1]. The hot streams will be identified by the index $i = 1, \dots, n_i$. Analogously, the cold streams will be identified by the index $j = 1, \dots, n_j$, and the devices by the index $u = 1, \dots, n_u$. The description of the topology of a HEN will be done by specifying the ordinal position of each device on each stream.

The ordinal position of device $u$ on hot stream $i$ will be denoted by the parameter $q_{i,u}$:

$$q_{i,u} = \begin{cases} 1, & \text{if } u \text{ is the first device on hot stream } i \\ 2, & \text{if } u \text{ is the second device on hot stream } i \\ 3, & \text{if } u \text{ is the third device on hot stream } i \\ \vdots & \\ n, & \text{if } u \text{ is the } n^{th} \text{ device on hot stream } i \\ 0, & \text{if } u \text{ does not exist on hot stream } i \end{cases} \tag{3.1}$$

$$i \in \mathrm{H}, \quad u \in \mathrm{U}$$

The value of the parameter $q_{i,u}$ can also be understood in terms of the existence of device $u$ on hot stream $i$: if device $u$ exists on hot stream $i$, then the parameter $q_{i,u} > 0$, otherwise, the parameter $q_{i,u} = 0$. The procedure used to describe the devices on the hot streams can also be applied to the cold streams, with the ordinal position of device $u$ on cold stream $j$ denoted by the parameter $q_{j,u}$.

[1] Notice that in this mathematical model the term *device* will also refer to heat exchangers.

**Parameter: identifying whether a device is a heat exchanger**

A device in a HEN can be a heat exchanger, a splitter or a mixer. A heat exchanger is on a hot and a cold stream simultaneously, whereas a splitter and a mixer are either on a hot stream or on a cold stream. This characteristic can be used to sort the devices that are heat exchangers from those that are splitters and mixers. In doing so, it will be possible to create equations that apply only to heat exchangers.

The procedure to identify if a given device $u$ is a heat exchanger is to sum the parameter $q_{i,u}$ over all the hot streams $i$ and the parameter $q_{j,u}$ over all the cold streams $j$. If both summations are greater than zero, then device $u$ is a heat exchanger, otherwise it is either a splitter or a mixer. The procedure described above will be implemented using the parameter $k_u$, which is equal to one if device $u$ is a heat exchanger and to zero otherwise:

$$k_u = \begin{cases} 1, & \text{if} \sum_{i \in \mathrm{H}} q_{i,u} > 0 \,\text{and} \sum_{j \in \mathrm{C}} q_{j,u} > 0 \\ 0, & \text{otherwise} \end{cases} \qquad u \in \mathrm{U} \tag{3.2}$$

**Input data and parameter: the rectangular axes and the coordinates of the starting and endpoints**

The coordinates of the starting and endpoints of the streams will be given on the set of rectangular axes A with index $a$. The coordinates of the starting and endpoint of hot stream $i$ will be represented by $SP_{i,a}$ and $EP_{i,a}$, respectively. To represent the coordinates of the starting and endpoint of cold stream $j$, the index $i$ should be replaced with the index $j$.

**Decision variables: coordinates of the heat exchangers, splitters and mixers**

Having described which are the devices that the hot and cold streams pass through using the parameters $q_{i,u}$ and $q_{j,u}$, we can proceed to the coordinates of those devices on the rectangular axes. The coordinate on axis $a$ of device $u$ that hot stream $i$ passes through will be represented by the variable $c_{i,u,a}$. Similarly, the coordinate of device $u$ that cold stream $j$ passes through will be represented by the variable $c_{j,u,a}$.

A heat exchanger is a device that exists on a hot and a cold stream simultaneously ($k_u = 1$). Thus, its coordinate on a rectangular axis will be represented by two variables, $c_{i,u,a}$ and $c_{j,u,a}$, which must have the same value:

$$c_{i,u,a} = c_{j,u,a} \tag{3.3}$$

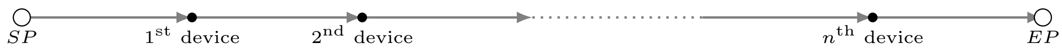
$$i \in \mathrm{H}, \quad j \in \mathrm{C}, \quad u \in \mathrm{U} \quad a \in \mathrm{A}$$

**Dependent variables: the logic to calculate the length of the pipes**

The length of a pipe will be calculated from the summation of the lengths of all the segments that compose the pipe. The length of a pipe segment will be determined by the rectangular distance between the devices being connected. The rectangular distance will be calculated using the set of linear inequalities previously defined in Subsection 1.4.4.

The pipe that transports a process stream is composed of $n+1$ segments, where $n$ is the number of devices on the process stream. The pipe segments from the starting point ($SP$) to the first device, then from the first to the second device and so forth up to the pipe segment from the $n^{th}$ device to the endpoint ($EP$) are depicted in Figure 3.9.

**Figure 3.9:** Segments of the pipe that transports a process stream.



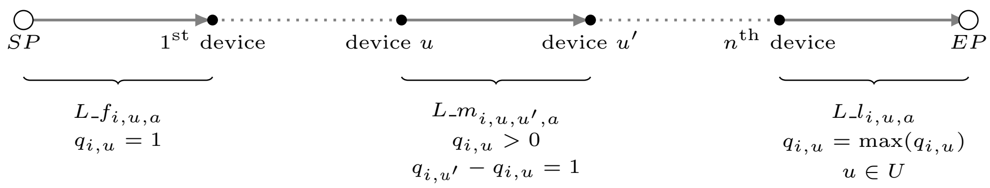

The length of the pipe segments will be represented by variables with indices for the hot and cold streams ($i$ or $j$), the connected devices ($u$ and $u'$) and the axis ($a$). For hot stream $i$, the length of the pipe segment from the starting point to the first device will be represented by the variable $L\_f_{i,u,a}$, where the index $u$ refers to the device with parameter $q_{i,u} = 1$. The length of a pipe segment that connects two adjacent devices will be represented by the variable $L\_m_{i,u,u',a}$, where the indices $u$ and $u'$ refer to the devices that satisfy the following adjacency criteria:

(i) the devices exist on hot stream $i$: $q_{i,u} > 0$ and $q_{i,u'} > 0$;

(ii) device $u$ is placed one ordinal position before device $u'$ on hot stream $i$. This criterion can be verified by subtracting the ordinal position of the devices: $q_{i,u'} - q_{i,u} = 1$.

The length of the pipe segment from the $n^{th}$ device to the endpoint will be represented by the variable $L\_l_{i,u,a}$, where the index $u$ refers to the device with parameter $q_{i,u} = \max\limits_{u \in \mathrm{U}} \left(q_{i,u}\right)$. The segments of the pipe that transports hot stream $i$ along with the variables that represent the lengths are depicted in Figure 3.10.

**Figure 3.10:** Variables for the length of the pipe segments.




### Constraints: the equations to calculate the length of the pipes

We will firstly present the equations to calculate the length of pipe for the transport of hot stream $i$. The length of the pipe segment from the starting point to the first device will be given by the rectangular distance between the coordinates of the starting point and device $u$ with the parameter $q_{i,u} = 1$:

$$L\_f_{i,u,a} \geq SP_{i,a} - c_{i,u,a} \tag{3.4}$$
$$L\_f_{i,u,a} \geq c_{i,u,a} - SP_{i,a} \tag{3.5}$$

$$i \in \mathrm{H}, \quad u \in \mathrm{U}, \quad a \in \mathrm{A}$$

The length of a pipe segment that connects two adjacent devices will be given by the rectangular distance between the coordinates of the devices $u$ and $u'$ with the parameters $q_{i,u} > 0$ and $q_{i,u'} - q_{i,u} = 1$:

$$L\_m_{i,u,u',a} \geq c_{i,u,a} - c_{i,u',a} \tag{3.6}$$
$$L\_m_{i,u,u',a} \geq c_{i,u',a} - c_{i,u,a} \tag{3.7}$$

$$i \in \mathrm{H}, \quad u,\ u' \in \mathrm{U} \mid u' > u, \quad a \in \mathrm{A}$$

The length of the pipe segment from the $n^{th}$ device to the endpoint will be given by the rectangular distance between the coordinates of device $u$ with the parameter $q_{i,u} = \max\limits_{u \in \mathrm{U}} \left(q_{i,u}\right)$ and the endpoint:

$$L\_l_{i,u,a} \geq c_{i,u,a} - EP_{i,a} \tag{3.8}$$
$$L\_l_{i,u,a} \geq EP_{i,a} - c_{i,u,a} \tag{3.9}$$

$$i \in \mathrm{H}, \quad u \in \mathrm{U}, \quad a \in \mathrm{A}$$

The length of the pipe that transports hot stream $i$ will be calculated from the summation of the length of all pipe segments over the rectangular axes:

$$L_i = \sum_{a \in \mathrm{A}} L\_f_{i,u,a} + \sum_{u,u' \in \mathrm{U}} \sum_{a \in \mathrm{A}} L\_m_{i,u,u',a} + \sum_{a \in \mathrm{A}} L\_l_{i,u,a} \tag{3.10}$$

$$i \in \mathrm{H}$$

To calculate the length of the pipe that transports cold stream $j$, the index $i$ in the equations (3.4) to (3.10) should be replaced with the index $j$. This would yields the following equation to calculate the length of the pipe that transports cold stream $j$:

$$L_j = \sum_{a \in \mathrm{A}} L\_f_{j,u,a} + \sum_{u,u' \in \mathrm{U}} \sum_{a \in \mathrm{A}} L\_m_{j,u,u',a} + \sum_{a \in \mathrm{A}} L\_l_{j,u,a} \tag{3.11}$$

$$j \in \mathrm{C}$$

### Objective function

The objective function will be defined as the minimisation of the summation of the length of all the pipes:

$$\min \left( \sum_{i \in \mathrm{H}} L_i + \sum_{j \in \mathrm{C}} L_j \right) \tag{3.12}$$

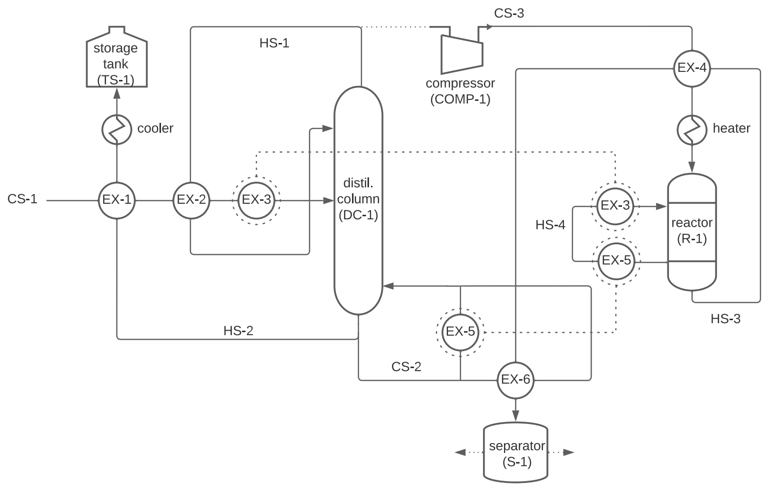
### Mathematical model outline

The location of the heat exchangers can be calculated by minimising the objective function shown in Equation (3.12) that is subjected to the constraints shown in the equations (3.3) to (3.11). The decision variables are the coordinates of the heat exchangers on the rectangular axes ($c$). The dependent variables are the length of the pipe segments ($L\_f$, $L\_m$ and $L\_l$) and the length of the pipes ($l$). The parameters are the coordinates of the starting points ($SP$), endpoints ($EP$) and the ordinal positions of the heat exchangers on the process streams ($q$). The presented mathematical model is linear.

## 3.2 Example: the Base Case

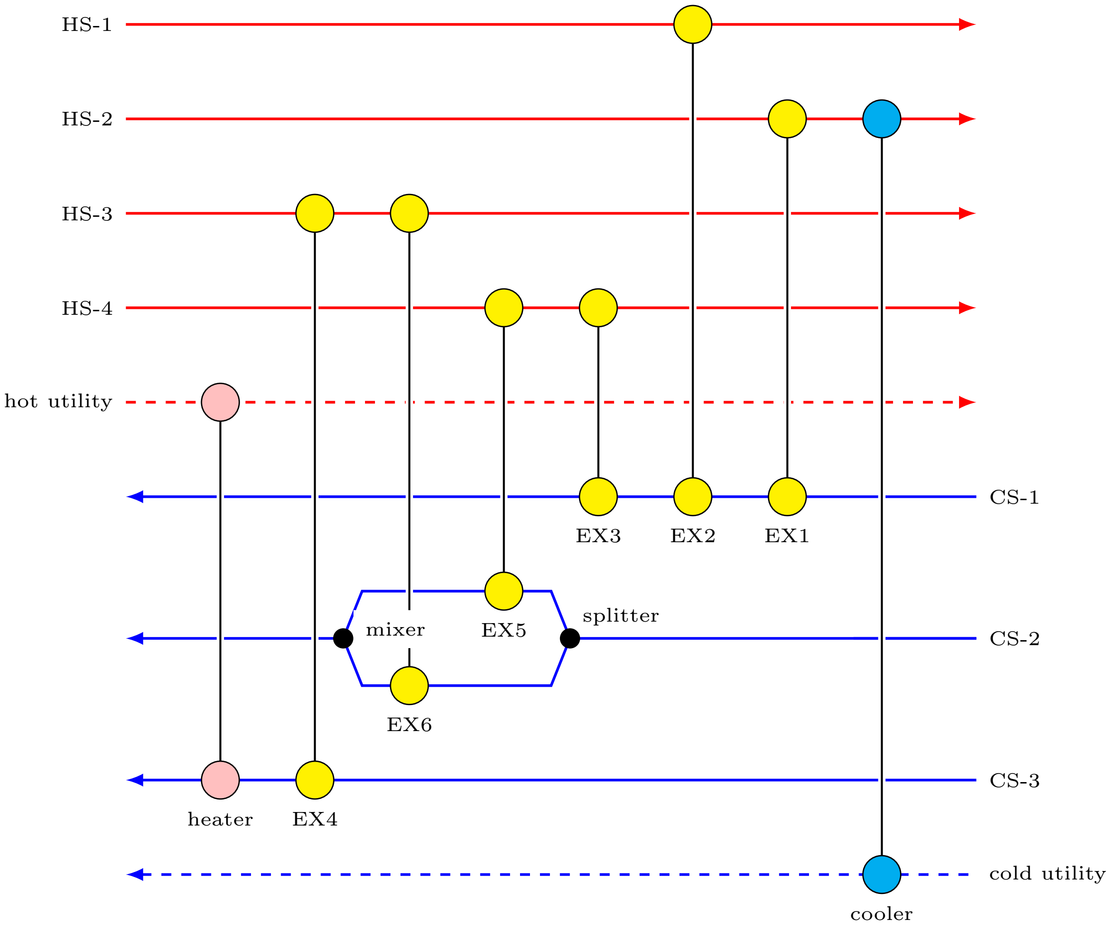
The linear model presented above will be applied to calculate the location of the heat exchangers of the process adapted from Serth and Lestina (2014). The process consists of four hot streams and three cold streams. Stream CS-1 leaves a pipe rack and is heated prior to entering a distillation column (DC-1). The streams CS-2 and HS-1 are the reboiling and the condensed reflux streams of the distillation column, respectively. Stream HS-2 leaves the bottom of the distillation column, is cooled, and stored in a tank (ST-1). Stream CS-3 leaves a compressor (COMP-1), is heated, and feeds a reactor (R-1). Stream HS-3 leaves the reactor, is cooled, and enters a separator (S-1). Stream HS-4 removes heat from the reactor, is cooled, and re-enters the reactor cooling jacket (Figure 3.11).

**Figure 3.11:** The flowsheet of the process for the Base Case.




The process has six process-to-process heat exchangers, including a condenser (heat exchanger EX-2) and two reboilers (heat exchangers EX-5 and EX-6). There is a splitter device and a mixer device installed along stream CS-2 for it to pass through the reboilers in parallel. A heater is installed along stream CS-3 and a cooler along stream HS-2 (Figure 3.12).

**Figure 3.12:** The topology of the HEN for the Base Case.




A simplified layout will be assumed for the plant of process. The simplified plant shows the main process equipment along with the starting and endpoint of the streams (Figure 3.13). The coordinates of the starting and endpoints associated with their respective process equipment are presented in Table 3.2. In addition, Table 3.2 shows the minimum length of pipe necessary for each stream to be transported from its starting to its endpoint (baseline length of pipe).

**Figure 3.13:** The assumed layout of the plant of the process for the Base Case.

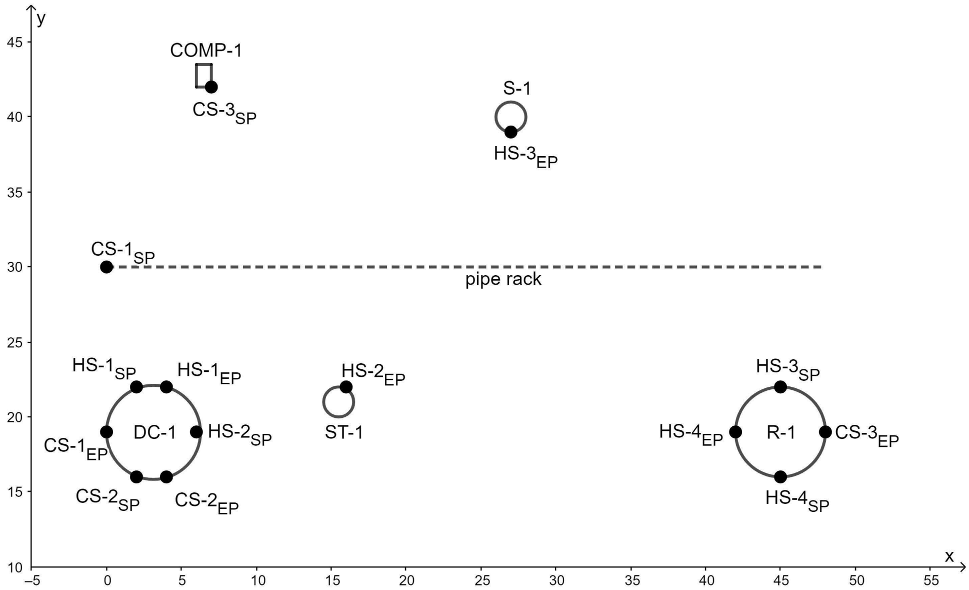



**Table 3.2:** The coordinates of the starting points, the endpoints and the baseline length of pipe for the Base Case.

| stream | starting point ($SP$) | | | endpoint ($EP$) | | | baseline |
|---|---|---|---|---|---|---|---|
| | $x$ | $y$ | equipment | $x$ | $y$ | equipment | length of pipe |
| HS-1 | 2 | 22 | DC-1 | 4 | 22 | DC-1 | 2 |
| HS-2 | 6 | 19 | DC-1 | 16 | 22 | ST-1 | 13 |
| HS-3 | 45 | 22 | R-1 | 27 | 39 | S-1 | 35 |
| HS-4 | 45 | 16 | R-1 | 42 | 19 | R-1 | 6 |
| CS-1 | 0 | 30 | pipe rack | 0 | 19 | DC-1 | 11 |
| CS-2 | 2 | 16 | DC-1 | 4 | 16 | DC-1 | 2 |
| CS-3 | 7 | 42 | COMP-1 | 48 | 19 | R-1 | 64 |
| total | | | | | | | 133 |

The mathematical model for the location of heat exchangers requires the description of the HEN topology in an algorithm readable form. The topology of the HEN depicted in Figure 3.12 is described by specifying the ordinal position of the devices on the hot streams (the parameter $q_{i,u}$) and cold streams (the parameter $q_{j,u}$), as previously stated in Equation (3.1). The value of the parameters $q_{i,u}$ and $q_{j,u}$ can be shown using a matrix, where the rows represent the hot and cold streams ($i$ or $j$), the columns represent the devices ($u$) and the matrix elements represent the ordinal position of the devices on the hot and cold streams. The matrices 3.13 and 3.14 show the values of the parameters $q_{i,u}$ and $q_{j,u}$, respectively. The label of the streams are shown beside their corresponding rows. Analogously, the label of the devices are aligned with their corresponding columns. The labels of the splitter device, mixer device, heater and cooler are shortened to sp., mi., he. and co., respectively.

$$
\begin{array}{rc}
 & \begin{array}{cccccccccc} \text{EX-1} & \text{EX-2} & \text{EX-3} & \text{EX-4} & \text{EX-5} & \text{EX-6} & \text{sp.} & \text{mi.} & \text{he.} & \text{co.} \end{array} \\
\begin{array}{r} \text{HS-1} \\ \text{HS-2} \\ \text{HS-3} \\ \text{HS-4} \\ \text{hot util.} \end{array} &
\begin{bmatrix}
 & 1 & & & & & & & & \\
1 & & & & & & & & & 2 \\
 & & & 1 & & 2 & & & & \\
 & & 2 & & 1 & & & & & \\
 & & & & & & & & 1 &
\end{bmatrix}
\end{array}
\tag{3.13}
$$

$$
\begin{array}{rc}
 & \begin{array}{cccccccccc} \text{EX-1} & \text{EX-2} & \text{EX-3} & \text{EX-4} & \text{EX-5} & \text{EX-6} & \text{sp.} & \text{mi.} & \text{he.} & \text{co.} \end{array} \\
\begin{array}{r} \text{CS-1} \\ \text{CS-2} \\ \text{CS-3} \\ \text{cold uti.} \end{array} &
\begin{bmatrix}
1 & 2 & 3 & & & & & & & \\
 & & & & 2 & 2 & 1 & 3 & & \\
 & & & 1 & & & & & 2 & \\
 & & & & & & & & & 1
\end{bmatrix}
\end{array}
\tag{3.14}
$$

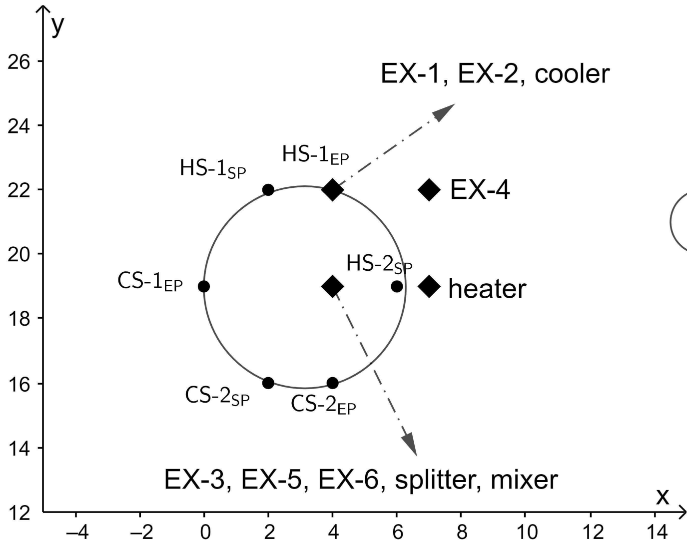
The coordinates of the heat exchangers, the splitter and the mixer devices in the plant will be calculated by minimising the summation of the length of the pipes that transport the hot and cold streams. The pipes that carry the utility streams will not be included in the objective function. All the computations will be carried out by using the General Algebraic Modelling System (GAMS) software version 24.0.2 with CPLEX solver version 12.5.

### Simulation results and discussion

The mathematical model for the solution of this problem consists of 100 variables, 136 equations and was solved in less than 1 s. The plant with the heat exchangers in the optimal location is depicted in Figure 3.14. The heat exchangers EX-1, EX-2 and the cooler overlap each other at coordinates (4, 22). The heat exchangers EX-3, EX-5, EX-6 along with the splitter and mixer devices are at coordinates (4, 19). Heat exchanger EX-4 and the heater are at coordinates (7, 22) and (7, 19), respectively.

To transport the streams from their starting points, passing through the heat exchangers, to their endpoints, it is necessary to add 146 lu of pipe in comparison with the situation without heat exchangers, which is 133 lu. Thus, the total length of pipe with the heat exchangers placed in the plant is 279 lu. This result can be achieved with the heat exchangers in different arrangements. In other words, there is a multiplicity of solutions that lead to the same objective value.

**Figure 3.14:** The heat exchangers in the optimal location for the Base Case.




Although mathematically optimal, the obtained solution is infeasible from the practical view point. The present model has no restriction on the spacing between the process equipment. As a consequence, the heat exchangers may be placed over each other and also over other equipment. For example, heat exchanger EX-1 was placed at the same coordinates as the endpoint of stream HS-1. Also, heat exchanger EX-5 was placed over the distillation column. In the next section, a way to overcome this limitation is presented.

The problem presented in this example was solved by allowing the heat exchangers to be placed anywhere in the plant, including over each other. Thus, the obtained objective value (279 lu) can be regarded as the least total length of pipe necessary for the streams to be transported with the heat exchangers placed in the plant. There is no other heat exchanger arrangement that leads to a shorter total length of pipe.

## 3.3 Minimum spacing between process equipment

Chemical process plants require the process equipment to be spaced. In this way, the operators can access the instrumentation for performance monitoring, carry out maintenance and leave the workplace in an emergency (BAUSBACHER; HUNT, 1993; MORAN, 2016). In this section, we will develop a mathematical model for the heat exchangers to being spaced from each other so as to meet a minimum spacing. This mathematical model will be extended for the heat exchangers to being spaced from splitter devices, mixer devices, starting points and endpoints. This section ends with an example to demonstrate the application of the minimum spacing restriction.

### 3.3.1 Mathematical model

To restrict the heat exchangers to being spaced from each other at least a minimum value, the spacing between every two heat exchangers has to be calculated. The key information to do that are the coordinates of the heat exchangers. We defined a heat exchanger as a device $u$ with parameter $k_u = 1$ (Equation 3.2). The coordinates of a device can be obtained with respect to the hot stream by the variable $c_{i,u,a}$ as well as the cold stream by the variable $c_{j,u,a}$ (Equation 3.3). To develop this mathematical model, we will use only the variable corresponding to the hot stream.

The spacing between two heat exchangers can be calculated from the absolute difference between their coordinates. The absolute difference between the coordinates of two heat exchangers can be constrained in a straightforward way to be greater than or equal to the minimum spacing, with the minimum spacing set as a positive number. Nonetheless, the use of the absolute function results in a non-linear formulation (DANTZIG, 2002).

To avoid using the absolute function, one may consider using a pair of inequalities to obtain the positive difference between the coordinates of two heat exchangers. In this strategy, a variable, say $s$, is constrained to being greater than or equal to the positive and also the negative difference between the coordinates of two heat exchangers. This is shown as follows for the heat exchangers E1 and E2 with coordinates $c_{\mathrm{E1}}$ and $c_{\mathrm{E2}}$, respectively:

$$
\begin{aligned}
s &\geq c_{\mathrm{E1}} - c_{\mathrm{E2}} \\
s &\geq c_{\mathrm{E2}} - c_{\mathrm{E1}}
\end{aligned}
$$

The variable $s$ has to be included in the objective function. In this way, the variable $s$ will satisfy the inequalities by assuming the minimum value possible, which is the positive difference between the coordinates of two heat exchangers.

This strategy was effective earlier to calculate the length of a pipe segment (Subsection 1.4.4). However, for the current situation, this strategy does not work. The reason why is that we will not include the variable $s$ in the objective function. Including the variable $s$ in the objective function means minimising the spacing between the heat exchangers.

Being outside of the objective function, the variable $s$ can satisfy the inequalities by assuming any value greater than the positive difference between the coordinates of two heat exchangers. If this happens, then the variable $s$ does not represent the actual spacing between two heat exchangers. As a consequence, it is not possible to restrict the heat exchangers to being spaced from each other so as to meet a minimum spacing.

As an alternative to the use of the absolute function, the spacing between two heat exchangers will be calculated from the simple difference between their coordinates. Thus, the spacing on axis $a$ between heat exchanger $u$ that is on hot stream $i$ and heat exchanger $u'$ that is on hot stream $i'$ is given by:

$$s_{u,u',a} = c_{i,u,a} - c_{i',u',a} \tag{3.15}$$

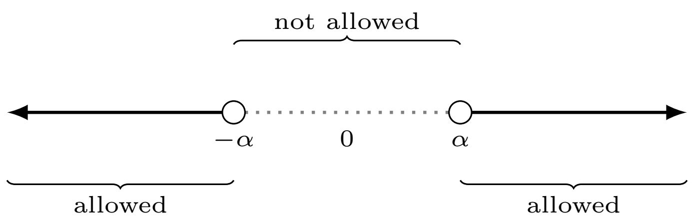
$$i,\ i' \in \mathrm{H} \mid i' \geq i, \quad u,\ u' \in \mathrm{U} \mid u' > u, \quad a \in \mathrm{A}$$

where indices $i$ and $i'$ can refer to the same hot stream.

The major difficulty of using the simple difference is that the result of the spacing between two heat exchangers may be a positive or a negative number. If it is a positive number, then any negative spacing between two heat exchangers is not acceptable, no matter how far the heat exchangers are from each other in the physical plant.

A way of dealing with the possibility of both positive and negative signs is by setting a positive ($\alpha$) and also a negative minimum spacing ($-\alpha$) while restricting two heat exchangers to meeting only one of them. If two heat exchangers meet the positive minimum spacing, then the negative one is cancelled, and vice and versa. The spacing between two heat exchangers can thus be any value encompassed in the open intervals $(-\infty, -\alpha]$ and $[\alpha, \infty)$ (Figure 3.15).

**Figure 3.15:** Number line with the intervals allowed and not allowed for the spacing between two exchangers.




To restrict the heat exchangers $u$ and $u'$ to meet either the positive or negative minimum spacing on axis $a$, the binary variable $w_{u,u',a}$ will be used. If the binary variable $w_{u,u',a}$ is equal to zero, then the heat exchangers $u$ and $u'$ are restricted to meeting the positive minimum spacing on axis $a$. In addition, the negative negative minimum spacing is cancelled. The opposite happens if the binary variable $w_{u,u',a}$ is equal to one.

As the heat exchangers are represented in the plant as dimensionless devices, two heat exchangers can be prevented from being physically placed over each other by applying the minimum spacing restriction to at least one of the rectangular axes. Every two heat exchangers may have to meet the minimum allowed spacing on different rectangular axes. For example, consider the heat exchangers E1, E2 and E3. A minimum spacing restriction of 1 lu is applied to the $x$-axis for the heat exchangers E1 and E2. For the heat exchangers E1 and E3, the minimum spacing restriction is applied to the $y$-axis. This leads to the following outcome:

- the heat exchangers E1 and E2 have to meet the minimum spacing on the $x$-axis but are allowed to be spaced from each other at any distance on the $y$-axis, even at the same $y$-coordinate (Figure 3.16);
- The same occurs with the heat exchangers E1 and E3 but with the $x$-axis and $y$-axis switched over (Figure 3.17).

**Figure 3.16:** The heat exchangers E1 and E2 at the same coordinate on the $y$-axis.

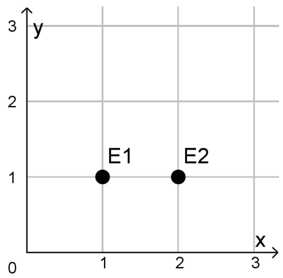
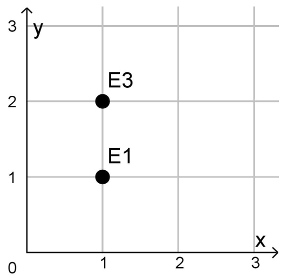



**Figure 3.17:** The heat exchangers E1 and E3 at the same coordinate on the $x$-axis.




To denote whether the minimum spacing restriction for the heat exchangers $u$ and $u'$ will be applied to axis $a$, the binary variable $v_{u,u',a}$ will be used. If the binary variable $v_{u,u',a}$ is equal to one, then the minimum spacing restriction is applied to axis $a$, otherwise, it is not. To enforce the minimum spacing restriction to be applied to at least one of the rectangular axis, the following constraint will be used:

$$\sum_{a \in \mathrm{A}} v_{u,u',a} \geq 1 \tag{3.16}$$

$$u,\ u' \in \mathrm{U} \mid u' > u$$

As we discussed a few paragraphs above, the spacing between the heat exchangers $u$ and $u'$ ($s_{u,u',a}$) on axis $a$ will be calculated from the simple difference between their coordinates. The heat exchangers $u$ and $u'$ will restricted to being spaced from each other on axis $a$ either the positive ($w_{u,u',a} = 0$) or the negative minimum spacing ($w_{u,u',a} = 1$). Since we have declared all the rectangular axes using the index $a$, the minimum spacing restriction may be applied to axis $a$ ($v_{u,u',a} = 1$) or may not be applied to axis $a$ ($v_{u,u',a} = 0$). These constraints will be expressed by means of:

$$s_{u,u',a} - \mathcal{M}\left(1 - w_{u,u',a}\right) \leq -\alpha\, v_{u,u',a} \tag{3.17}$$

$$s_{u,u',a} + \mathcal{M}\, w_{u,u',a} \geq \alpha\, v_{u,u',a} \tag{3.18}$$

$$u,\ u' \in \mathrm{U} \mid u' > u, \quad a \in \mathrm{A}$$

where the minimum spacing $\alpha$ is a positive constant and $\mathcal{M}$ is a large positive constant.

**Minimum spacing between a heat exchanger and a splitter or a mixer device**

The mathematical model used to set a minimum spacing between two heat exchangers can be adapted to a heat exchanger and a splitter device or a mixer device. In order to do that, the index $u'$ in the equations (3.15) to (3.18) has to be restricted to represent only the devices with parameter $k_{u'} = 0$. This is the only modification necessary to restrict a heat exchanger and a splitter or a mixer device, which are on a hot stream, to meeting a minimum spacing. For a splitter or mixer that is on a cold stream it is also necessary to replace the index $i'$ with the index $j$.

**Minimum spacing between a heat exchanger and a starting or endpoint**

A heat exchanger can be restricted to being spaced from a starting or an endpoint of a hot as well as a cold stream by means of the same mathematical model employed for the spacing between two heat exchangers. Each pair formed by a heat exchanger and the starting point of a hot stream, the endpoint of a hot stream, the starting point of a cold stream, and the endpoint of a cold stream requires an individual formulation which:

- calculates the spacing between the heat exchanger and the starting or endpoint from the simple difference between their coordinates (Equation 3.15);
- restricts the heat exchanger to being spaced from the starting or endpoint on a rectangular axis so there is at least the positive or the negative minimum spacing between them (the equations 3.17 and 3.18);
- applies the minimum spacing restriction to at least one of the rectangular axes (Equation 3.16).

### 3.3.2 Example: minimum spacing between process equipment

In this example, the minimum spacing restriction will be applied to the heat exchangers of the process introduced in the Base Case (Section 3.2). The layout of the heat exchangers in the plant will be calculated in two situations. In the first one, the heat exchangers have to be spaced from each other as well as the starting and endpoints at least 1 lu. In the second one, the minimum spacing between a heat exchanger and a starting or an endpoint will be increased to 4 lu. The values of minimum spacing used in this example are demonstrative and do not reflect the engineering practice. No spacing restriction will be applied to the splitter and mixer.

**Simulation results and discussion**

The mathematical model generated for the solution of this problem consists of 1132 continuous variables, 688 discrete variables and 1340 equations. The time elapsed to solve the mathematical model was less than 1 s. For a minimum spacing of 1 lu, the total length of pipe with the heat exchangers placed in the plant of the process is the same as that calculated in the Base Case, which is 279 lu. Nonetheless, in the solution presented for the Base Case some heat exchangers overlapped each other and also other process equipment. As previously discussed, the objective value calculated in the Base Case can be achieved with the heat exchangers arranged in different ways. The minimum spacing restriction can thus be used to narrow down the number of arrangements towards those where the process equipment is spaced in a suitable manner.

For a minimum spacing between heat exchangers and starting as well as endpoints of 4 lu, the total length of pipe increased from 279 lu to 289 lu. This result indicates that in the problem being solved, the more spaced the heat exchangers are from the equipment they are connected to, the more pipe is necessary for the transport of the streams.

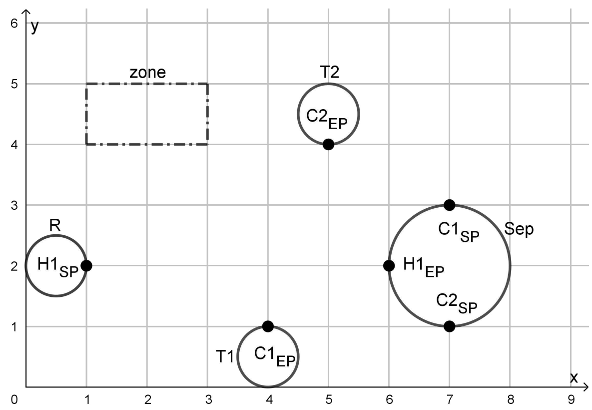
## 3.4 Heat exchanger placement in specific zones of the plant

Thus far, we have developed a method suitable for a situation in which all the locations of the plant are available for placing the heat exchangers. The heat exchangers tend to be placed near their starting and endpoints so as to minimise the length of pipe. From a mathematical point of view, a specific location minimises the length of pipe. However, in practice, that location may not provide adequate conditions for the heat exchanger maintenance and, therefore, it should not be available for placing the heat exchangers.

The objective of this chapter is to equip the user with a procedure that avoids placing the heat exchangers in locations with inappropriate conditions for maintenance. In order to do that, we will develop a mathematical model that allows the user to define zones in the plant where heat exchanger placement is permitted. To demonstrate the effect of this mathematical model, let us consider the three-stream process introduced in Section 3.1.

The process consists of hot stream H1, the cold streams C1 and C2 and, the heat exchangers E1 and E2. The location of these heat exchangers should be defined in the plant of the process (Figure 3.18). Let us consider that due to maintenance reasons, the heat exchangers can be placed only in the zone defined from $x = 1$ to $x = 3$ and $y = 4$ to $y = 5$. Besides that, the heat exchangers have to be spaced at least 1 lu from each other.

**Figure 3.18:** The zone of the plant where the placement of the heat exchangers E1 and E2 is allowed.

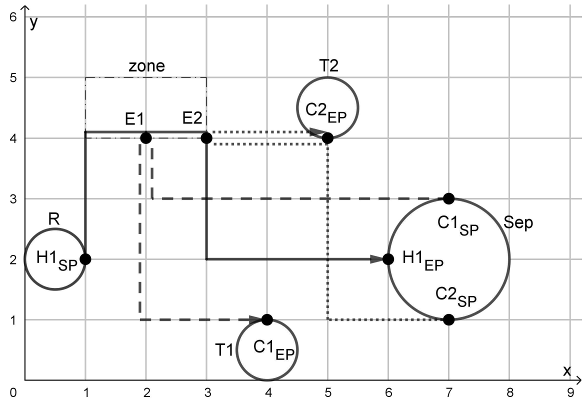



The coordinates of the heat exchangers in the specified zone have to be defined so as to minimise the length of the pipes that transport the streams H1, C1 and C2. By visualising Figure 3.18, we can realise that for the heat exchangers to be located as close as possible to the starting and endpoints of the streams, they should be placed at the coordinate $y = 4$. In the same manner, it is possible to infer that the higher along the x-axis the heat exchangers are, the closer they are to the starting and endpoints of the streams C1 and C2. In addition, since stream H1 passes through the heat exchangers E1 and E2 in series, heat exchanger E1 has to be placed before heat exchanger E2 along the $x$-axis.

Based on this information, the location of the heat exchangers that minimises the length of pipe is: heat exchanger E1 at coordinates (2, 4) and heat exchanger E2 at coordinates (3, 4). By placing the heat exchangers at these coordinates, the pipes that transport the streams are: stream H1's pipe is 9 lu long, stream C1's pipe is 11 lu long and stream C2's pipe is 9 lu long (Figure 3.19). In comparison with the heat exchanger arrangement that we obtained in Section 3.1, the demand for pipe increased by 14 lu.

This short example demonstrated a procedure to restrict the locations of the plant available for placing heat exchangers. Let us now develop the mathematical model of for the automated implementation of this procedure.

**Figure 3.19:** The plant layout of the three-stream process with the heat exchangers placed inside a zone.




### 3.4.1 Mathematical model

A zone of the plant where the heat exchangers can be placed will be defined by set their lower and upper boundaries on the rectangular axes. The lower and upper boundaries of zone $z$ on axis $a$ will be represented by the parameters $low_{z,a}$ and $upp_{z,a}$, respectively. A number of zones can be specified in the plant for placing the heat exchangers. To denote which zone of the set of zones Z a heat exchanger will be placed, the binary variable $r_{u,z}$ will be used:

$$r_{u,z} = \begin{cases} 1, & \text{if heat exchanger } u \text{ is inside zone } z; \\ 0, & \text{otherwise.} \end{cases} \tag{3.19}$$

$$u \in \mathrm{U}, \quad z \in \mathrm{Z}$$

A heat exchanger has to be placed in one of the specified zones:

$$\sum_{z \in \mathrm{Z}} r_{u,z} = 1 \tag{3.20}$$

$$u \in \mathrm{U}$$

The coordinates of a heat exchanger have to be enclosed within the lower and upper boundaries of the zone where the heat exchanger will be placed:

$$c_{i,u,a} \geq \sum_{z \in \mathrm{Z}} low_{z,a}\, r_{u,z} \tag{3.21}$$

$$c_{i,u,a} \leq \sum_{z \in \mathrm{Z}} upp_{z,a}\, r_{u,z} \tag{3.22}$$

$$i \in \mathrm{H}, \quad u \in \mathrm{U}, \quad a \in \mathrm{A}$$

A zone of the plant may have a minimum and a maximum allowed number of heat exchangers placed in it. The minimum and maximum number will be represented by the constants $\lambda_z$ and $\kappa_z$, respectively:

$$\sum_{u \in \mathrm{U}} r_{u,z} \geq \lambda_z \tag{3.23}$$

$$\sum_{u \in \mathrm{U}} r_{u,z} \leq \kappa_z \tag{3.24}$$

$$z \in \mathrm{Z}$$

It is important to point out that to apply the presented equations only to heat exchangers and not to splitter and mixer devices, the user should restrict the index $u$ to represent only devices with parameter $k_u = 1$.

### 3.4.2 Example: heat exchanger placement in specific zones of the plant

In this example, the plant of the process introduced in the Base Case (Section 3.2) will be redesigned with three zones where heat exchanger placement is allowed (Figure 3.20). The boundaries of the zones are presented in Table 3.3. The location of the heat exchangers will be calculated with heat exchanger placement allowed in one zone at a time. The minimum spacing restriction will be not applied to the heat exchangers and other equipment either.

**Figure 3.20:** The zones $z_1$, $z_2$ and $z_3$ of the plant where heat exchanger placement is allowed.

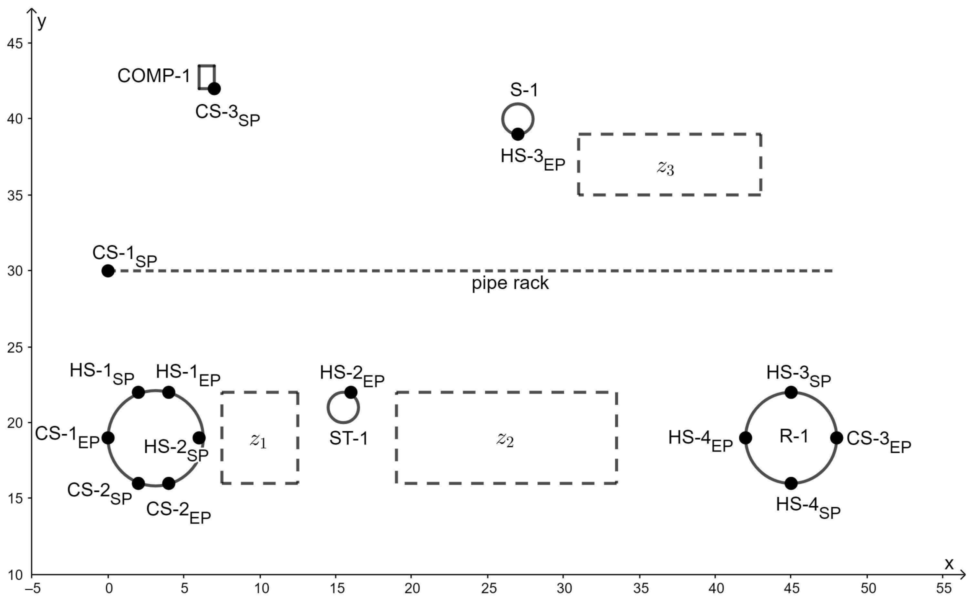


**Table 3.3:** Lower and upper boundaries of the zones $z_1$, $z_2$ and $z_3$.

| zone | lower boundary | | upper boundary | |
|---|---|---|---|---|
| | $x$ | $y$ | $x$ | $y$ |
| $z_1$ | 7.5 | 16 | 12.5 | 22 |
| $z_2$ | 19 | 16 | 33.5 | 22 |
| $z_3$ | 31 | 35 | 43 | 39 |

### Simulation results and discussion

The mathematical model generated for the solution of this problem is composed of 108 continuous variables, 8 discrete variables and 176 equations. The model was solved in less than 1 s. The total length of pipe calculated with heat exchanger placement allowed in zones $z_1$, $z_2$ and $z_3$ is 282 lu, 311 lu and 487 lu, respectively. These results mean an increase in the total length of pipe in comparison with that calculated in the Base Case, which is 279 lu. A similarity between the arrangement obtained with the heat exchangers placed in the different zones and also in the plant of the Base case is that they all have heat exchangers overlapping each other.

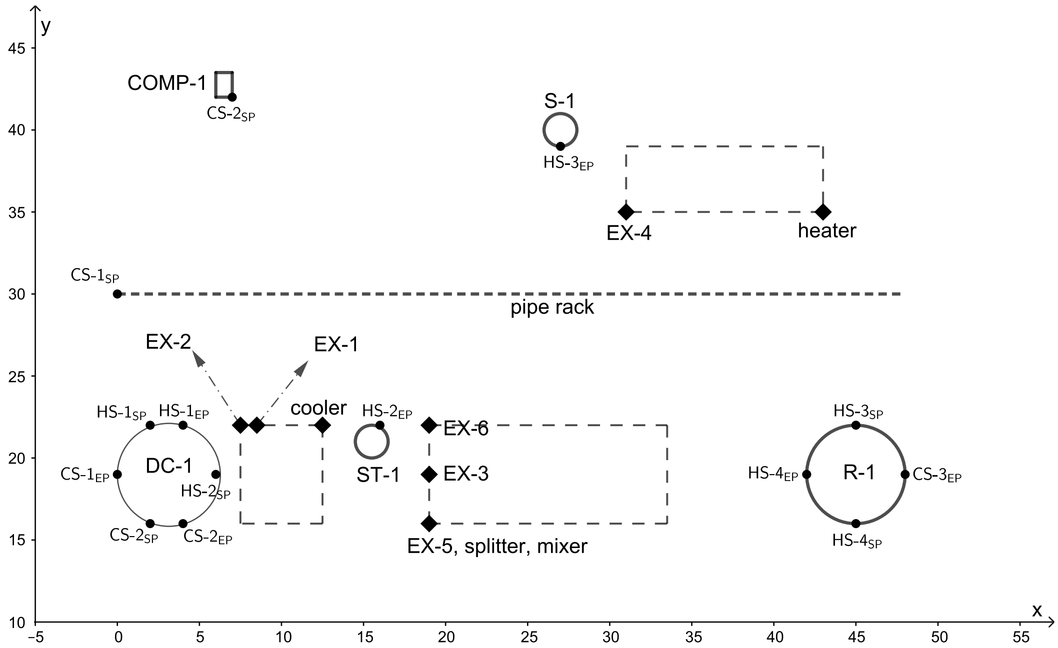
## 3.5 Example: combining the minimum spacing restriction with heat exchanger placement in specific zones of the plant

This example combines the minimum spacing restriction with heat exchanger placement in specific zones of the plant. The problem introduced in the Base Case (Section 3.2) will be solved with the heat exchangers restricted to being spaced at least 1 lu from each other. We will allow the placement of heat exchangers in all the three zones shown in Figure 3.20. A maximum of three heat exchangers can be placed in a single zone of the plant. We will not restrict the spacing between a heat exchanger and a starting point, an endpoint, a splitter and a mixer.

### Simulation results and discussion

The mathematical model to solve this problem consists of 292 continuous variables, 136 discrete variables and 375 equations. The solution was obtained in less than 1 s. The plant layout with the heat exchangers placed in the three zones $z_1$, $z_2$ and $z_3$ is depicted in Figure 3.21.

**Figure 3.21:** The plant layout with the heat exchangers placed in the three zones $z_1$, $z_2$ and $z_3$.




The heat exchangers EX-1 and EX-2, and the cooler are located in zone $z_1$. The heat exchangers EX-3, EX-5 and EX-6 are located in zone $z_2$. Heat exchanger EX-4 and the heater are located in zone $z_3$. The coordinates of the heat exchangers in the zones are shown in Table 3.4. The splitter and mixer are placed at the same coordinates as heat exchanger EX-5.

**Table 3.4:** Coordinates of the heat exchanger in the zones.

| heat exchanger | $x$ | $y$ | zone |
|---|---|---|---|
| EX-1 | 8.5 | 22 | |
| EX-2 | 7.5 | 22 | $z_1$ |
| cooler | 12.5 | 22 | |
| EX-3 | 19 | 19 | |
| EX-5 | 19 | 16 | $z_2$ |
| EX-6 | 19 | 22 | |
| EX-4 | 31 | 35 | $z_3$ |
| heater | 43 | 35 | |

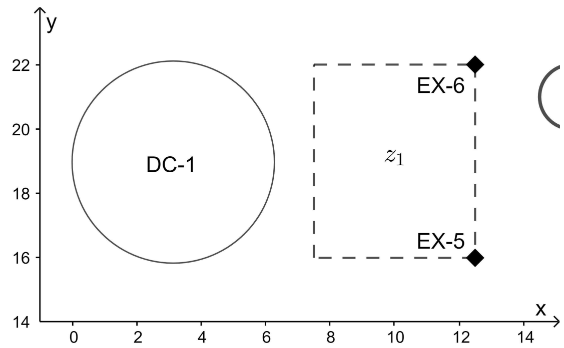
The total length of pipe for the transport of the streams with the heat exchangers placed in the plant is 310 lu. In comparison with the Base Case, the total length of pipe increased by 31 lu. From the practical view point, the reboilers EX-5 and EX-6 are located considerably far from the distillation column, which may represent an issue for the operation of the distillation column (BAUSBACHER; HUNT, 1993). In addition, it is desirable that the heat exchangers EX-1, EX-2 and EX-3 be grouped within the same zone because stream CS-1 passes through them all in series. To calculate the location of the heat exchangers taking the mentioned aspects into account, we will discuss some strategies.

### Strategies to place the reboilers EX-5 and EX-6 nearer the distillation column

There are a number of ways to obtain the reboilers nearer the distillation column while optimising the location of the other heat exchangers. The reboilers can be restricted to zone $z_1$ by fixing the binary variables $r_{\mathrm{EX}-5,z_1} = 1$ and $r_{\mathrm{EX}-6,z_1} = 1$. As a result of doing so, heat exchanger EX-1 and the cooler are replaced from zone $z_1$ to $z_2$ so as to meet the limit of up to three heat exchangers placed in a single zone. The other heat exchangers stay in the same zones.

The reboilers EX-5 and EX-6 are at the coordinates (12.5, 16) and (12.5, 22), respectively. The location of the reboilers corresponds to the border of zone $z_1$ that is furthest away from the distillation column (Figure 3.22). By restricting the reboilers to zone $z_1$, they are placed at the coordinates that minimise the total length of pipe within zone $z_1$. Those coordinates are not necessarily the ones nearest to the distillation column.

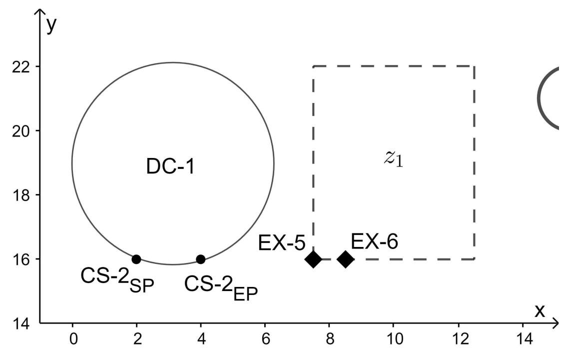
**Figure 3.22:** The reboilers located in the border of zone $z_1$ that is furthest away from the distillation column.




An alternative for we to obtain the reboilers even nearer the distillation column is to prioritise the minimisation of the length of the pipe connecting the distillation column to the reboilers. In order to do that, the length of the pipe that transports stream CS-2 will be multiplied by an assumed factor of ten in the objective function.

The results show that the reboilers are located in zone $z_1$ as near as possible to the starting and endpoint of stream CS-2 in the distillation column (Figure 3.23). The coordinates of the reboilers EX-5 and EX-6 are (7.5, 16) and (8.5, 16), respectively. All the other heat exchangers stay in the same zone except for heat exchanger EX-1 and the cooler that are replaced from zone $z_1$ to $z_2$.

**Figure 3.23:** The reboilers located as near as possible the starting and endpoint of stream CS-2.




The analysis of the impact of the presented strategies on the total length of pipe reveals that the nearer the reboilers are to the distillation column, the greater the total length of pipe is. The total length of pipe with the reboilers far from the distillation column is 310 lu (Figure 3.21). By restricting the reboilers EX-5 and EX-6 to zone $z_1$, the total length of pipe increases to 352 lu. Prioritising the minimisation of the length of the pipe connecting the distillation column to the reboilers results in a total length of pipe of 362 lu.

### Strategy to group the heat exchangers EX-1, EX-2 and EX-3 in the plant

In order to discuss a strategy to group heat exchangers, let us consider that for two or more heat exchangers to be grouped, the spacing between them must not exceed a maximum value along each rectangular axis. To group the heat exchangers $u$ and $u'$, the simple difference between their coordinates on all the rectangular axes $a$ ($s_{u,u',a}$) will have to be less than or equal to the positive maximum spacing ($\beta$) and greater than or equal to the negative maximum spacing ($-\beta$):

$$s_{u,u',a} \leq \beta \tag{3.25}$$

$$s_{u,u',a} \geq -\beta \tag{3.26}$$

$$u,\ u' \in \mathrm{U} \mid u' > u, \quad a \in \mathrm{A}$$

To calculate the location of heat exchangers, the maximum spacing between the heat exchangers EX-1, EX-2 and EX-3 will be assumed 2 lu. To obtain the reboilers nearer the distillation column, the length of the pipe that transports stream CS-2 will be multiplied by an assumed factor of ten in the objective function.

The results show the heat exchangers EX-1, EX-2 and EX-3 grouped in zone $z_2$ at the coordinates (19, 21), (19, 22) and (19, 19), respectively. The reboilers EX-5 and EX-6 are located in zone $z_1$ at the coordinates (7.5, 16) and (7.5, 17), respectively, along with the cooler at the coordinates (12.5, 21). Zone $z_3$ is populated with heat exchanger EX-4 and the heater. The total length of pipe with the heat exchangers placed in the plant is 369 lu. In comparison with the case in which the heat exchangers EX-1, EX-2 and EX-3 were grouped and the reboilers EX-5 and EX-6 could be located in any zone (Figure 3.21), the total length of pipe increased by 59 lu.

## 3.6 Conclusion

In this chapter, we developed a method to design the heat exchanger plant layout. The method requires a chemical processes with defined HEN. The design of the heat exchanger plant layout was modelled as a mathematical optimisation problem. The decision variables are the coordinates of the heat exchangers in the plant. The objective function was defined as the minimisation of the total length of pipe.

The method calculates the coordinates of each heat exchanger and the length of pipe for the transport of each stream. The user can attribute a weight to the length of each pipe in the objective function. The weight can be understood as a cost of pipe per unit length. The user can also set a minimum spacing between heat exchangers. Another feature is that zones can be defined in the plant to restrict the number of available locations.

We used the method to solve a case study with seven process streams and six heat exchangers. In order to implement the method, we used the General Algebraic Modelling System. The heat exchanger plant layout was designed a number of times for different restrictions of minimum spacing and zones defined in the plant. The results were obtained within seconds.

The application of the method is limited to chemical processes with defined HEN. This means that the choice of which streams will be matched to exchange heat cannot be affected by the length of pipe required to transport the streams. The sequence of the heat exchanges that will be performed by the streams is not affected by the length of pipe either.

Another limitation is that it is not possible to investigate whether there is another HEN topology that satisfies the energy balance of the streams while requiring a shorter total length of pipe. In the next two chapters, we will tackle these limitations by designing the HEN and heat exchanger plant layout in a simultaneous manner.

# Chapter 4

# Simultaneous design of HEN and heat exchanger plant layout: pipe length

The transport of the streams through heat exchangers is done by pipes and pumping. The pipes connect the heat exchangers to other process equipment. The pumping equipment is used to overcome the energy loss and provide head. The costs of pipe and pumping are usually calculated after the design of the HEN is finished.

At that point, regardless of the costs of pipe and pumping, the HEN topology is completely defined. The best that can be done is to optimise the location of heat exchangers in the plant. This approach misses the opportunity to evaluate the costs of pipe and pumping of the potential HEN topologies. A more comprehensive approach calculates the costs of pipe and pumping during the design of the HEN. This can be achieved by designing the heat exchanger plant layout simultaneously with the HEN.

The objective of this chapter is to develop a method that designs the HEN and heat exchanger plant layout in a simultaneous manner. Throughout the design, many HEN topologies will be generated. For each one, the location of heat exchangers in the plant will be calculated. The locations will be defined so as to minimise the total length of pipe, which indirectly minimises the costs of pipe and pumping. The definitive HEN topology will be the one that requires the shortest total length of pipe.

To develop the method, the concept of superstructure will be used. A superstructure allows different criteria to be minimised simultaneously. There are many mathematical models of superstructures for HEN. Traditionally, our research group has used mathematical optimisation to design HENs and is licensed to use GAMS software. These characteristics led us to choose the superstructure proposed by Yee and Grossmann (1990).

We will firstly present the concept, mathematical model and application of the superstructure for HEN. Then, we will proceed to develop the superstructure for heat exchanger plant layout. The combination of these two superstructures will result in the method for simultaneous design of HEN and heat exchanger plant layout. A case study will be presented to demonstrate the application of the method.

## 4.1 The superstructure for HEN

In the superstructure proposed by Yee and Grossmann (1990), all the streams can be matched for heat exchange. The streams can exchange heat in series as well as in parallel. The heat flow rate between the streams is calculated with respect to an objective function, such as the minimisation of the costs of utilities and heat exchangers. The streams are matched for heat exchange so as to optimise the objective function.

### 4.1.1 Designing a superstructure for HEN

To design an example of the superstructure, let us consider a process with two hot streams and two cold streams. The hot and cold streams will be identified by the following nomenclature:

- the two hot streams are identified by the labels $i$ and $i'$;
- the two cold streams are identified by the labels $j$ and $j'$.

A superstructure composed of two stages will show all the matches between the hot and a cold streams for heat exchange. The two stages will be identified by the labels $k$ and $k'$, respectively. The superstructure is depicted in Figure 4.1.

The hot streams firstly run stage $k$ whereas the cold streams stage $k'$. As the streams run, they are split into a number of branches. The number of branches corresponds to the number of streams of the other type. If there are two hot streams and two cold streams, then each hot stream is split into two branches. In an analogous manner, each cold stream is split into two branches. Figure 4.2 depicts the hot and cold streams being split at the stages $k$ and $k'$, respectively.

**Figure 4.1:** The two-stage superstructure with two hot and two cold streams.

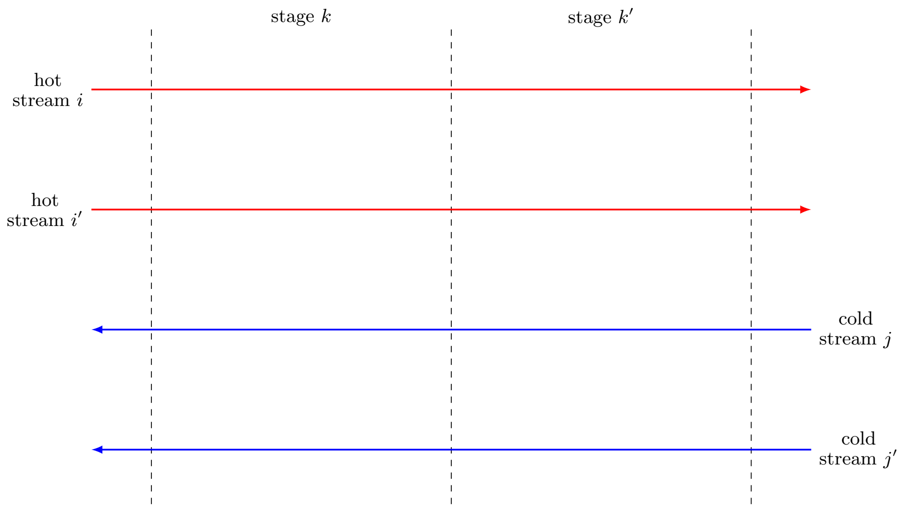
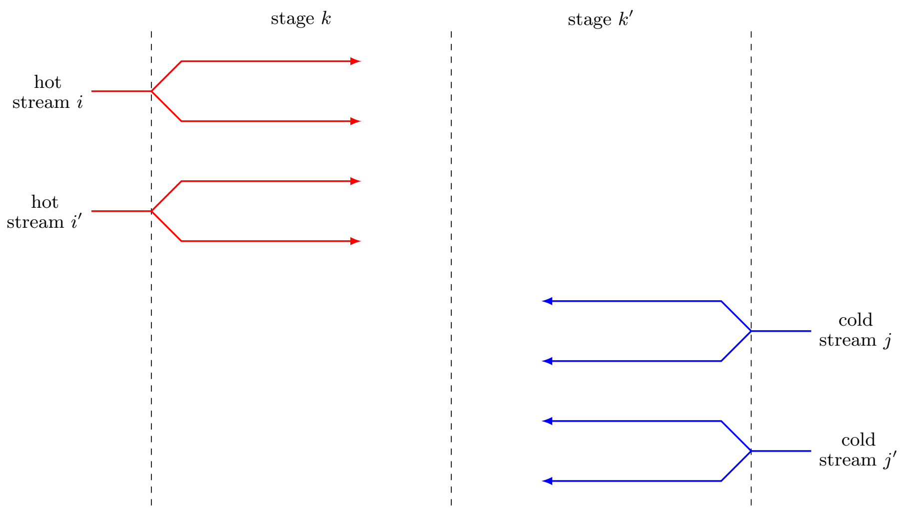



**Figure 4.2:** The hot and cold streams being split as they enter a stage.




After a stream is split, each stream branch passes through a heat exchanger. There, the stream branch exchanges heat with the branch of another stream. In doing so, all the hot streams can exchange heat in parallel with all the cold streams. The heat exchangers are identified by the labels of the hot stream, the cold stream and the stage where it is placed (Figure 4.3).

After passing through the heat exchangers, the branches of the stream are mixed before leaving the stage. In this way, they form a single stream again. Next, the hot streams run stage $k'$ whereas the cold streams run stage $k$. The steps of splitting a stream into branches, exchanging heat between the branches of different streams, and mixing the branches of the stream are performed again (Figure 4.4).

**Figure 4.3:** The hot and cold streams being mixed at a stage.

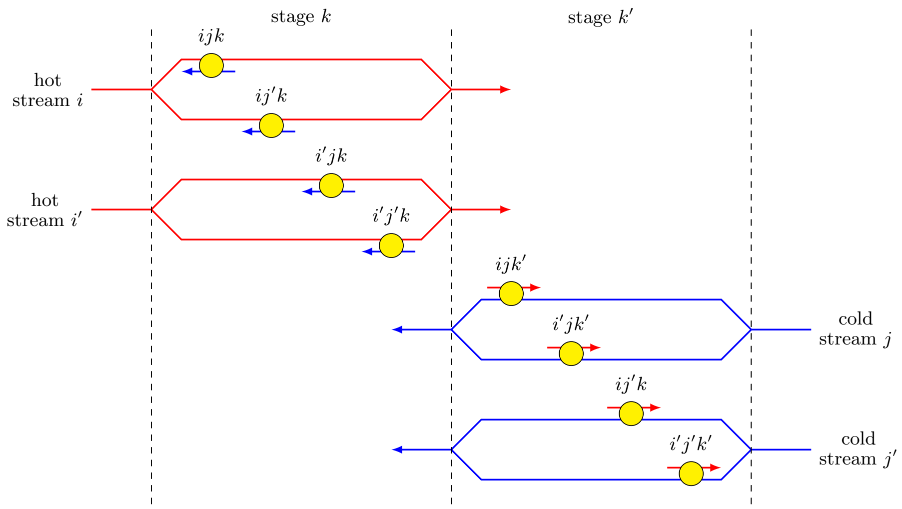
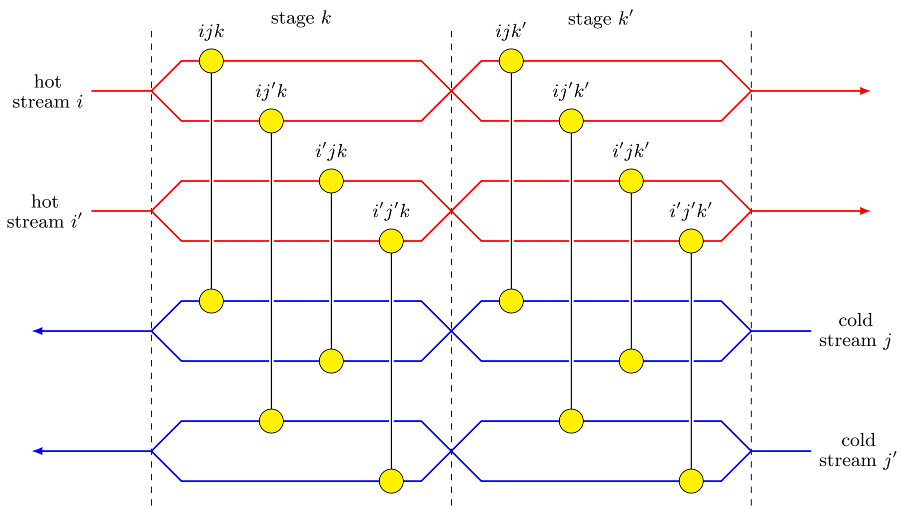


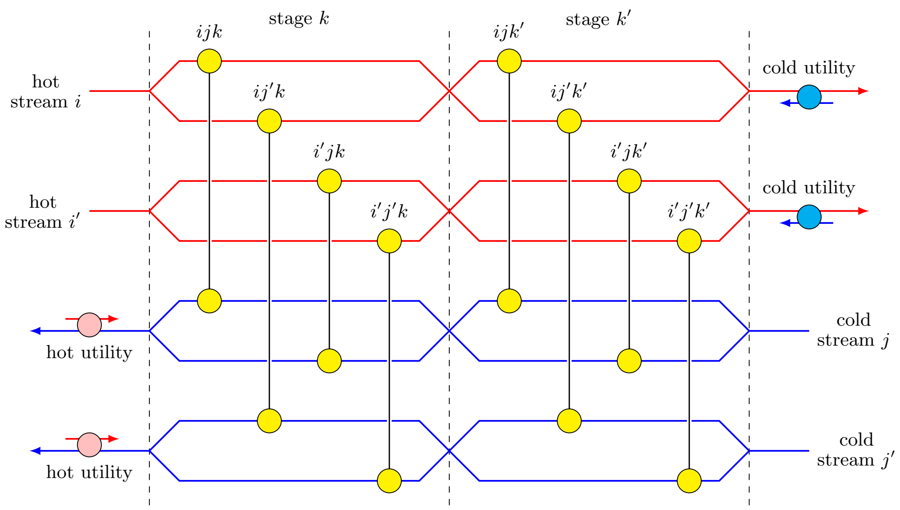

**Figure 4.4:** The hot and cold streams being split, exchanging heat and being mixed at every stage.




Performing these steps at the stages successively allows the hot and cold streams to exchange heat in series. Usually, the number of stages is defined so that a stream can exchange heat in series with all streams of the other type. In this way, the number of stages corresponds to the greatest between the number of hot streams and the number of cold streams. Since in this example there are two hot and two cold streams, we defined the superstructure with two stages.

After the hot streams leave stage $k'$, they exchange heat with a utility stream. The cold streams do the same as they leave stage $k$ (Figure 4.5). The superstructure in its entire form allows all the possible heat exchanges between the hot and cold streams in series as well as in parallel. In this way, the superstructure is ready to be optimised.

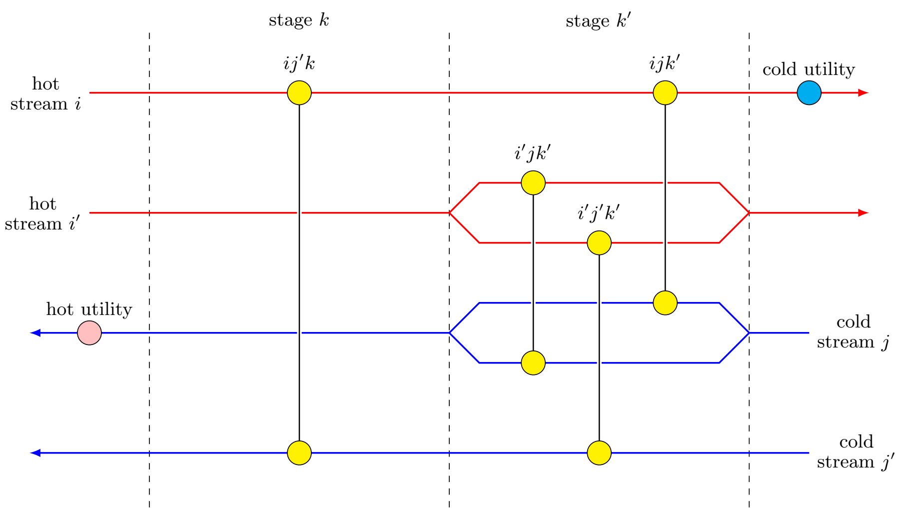
**Figure 4.5:** The hot and cold streams exchanging heat with the utility streams after running all the stages.




The optimisation can be understood as a process in which the matches between the streams that optimise the objective function are retained in the superstructure, whereas the other ones are excluded from the superstructure. To provide a hypothetical example, let us assume that the optimised superstructure has the topology depicted in Figure 4.6, in which:

- hot stream $i$ exchanges heat with the cold streams $j'$ and $j$ in series;
- hot stream $i'$ exchanges heat in parallel with the cold streams $j$ and $j'$ in parallel at stage $k'$;
- cold stream $j$ exchanges heat with a hot utility stream.

**Figure 4.6:** The optimised topology of the superstructure for HEN.



### 4.1.2 Mathematical model

This mathematical model requires the input of the set of hot streams, cold streams, utility streams and stages. The parameters are the supply and target temperatures of the hot and cold streams along with their heat capacity flow rate. The parameters also include the inlet and outlet temperatures of the utility streams.

The decision variables denote whether two streams are matched for heat exchange. The matches include the heat exchange between the hot and cold streams, the hot streams and cold utility, and the hot utility and cold stream. The dependent variables are the heat flow rate between the streams and their temperatures through the stages.

The constraints are linear and mixed-integer linear. The objective function is defined as the minimisation of the costs of utilities and heat exchangers. The cost of heat exchangers defined in the objective function in a direct manner is a non-linear equation. The indirect definition of the cost of heat exchangers can be expressed in a mixed-integer linear form, as will be shown in the next Subsection (4.1.3). Given this overview, let us present the mathematical model of the superstructure.

#### Input data

Let us consider:

- a set H of hot streams represented by the index $i$;
- a set C of cold streams represented by the index $j$;
- a single hot utility stream;
- a single cold utility stream;
- a set of stages K represented by the index $k$. The stages are numbered in the superstructure from left to right in ascending order.

#### Parameters

The streams are provided with their supply temperature ($T\_supply$), target temperature ($T\_target$) and mass flow rate ($\dot{m}$). By assuming a constant specific heat ($c_\mathrm{P}$), we can compute the heat flow rate ($\dot{q}$) required to take the streams from their supply to their target temperatures:

$$\dot{q}_i = \dot{m}cp_i \left(T\_supply_i - T\_target_i\right) \tag{4.1}$$

$$i \in \mathrm{H}$$

$$\dot{q}_j = \dot{m}cp_j\,(T\_target_j - T\_supply_j) \tag{4.2}$$

$$j \in \mathrm{C}$$

where $\dot{m}cp$ is the heat flow rate capacity.

**The overall heat balance**

The overall heat flow rate from the supply to the target temperature of a stream is equal to the contribution of the heat flow rate through all the stages ($\dot{q}$) plus the heat flow rate from/to the utility stream ($\dot{q}\_cooler$ or $\dot{q}\_heater$):

$$\dot{q}_i = \sum_{j\in\mathrm{C}}\sum_{k\in\mathrm{K}} \dot{q}_{i,j,k} + \dot{q}\_cooler_i \tag{4.3}$$

$$i \in \mathrm{H}$$

$$\dot{q}_j = \sum_{i\in\mathrm{C}}\sum_{k\in\mathrm{K}} \dot{q}_{i,j,k} + \dot{q}\_heater_j \tag{4.4}$$

$$j \in \mathrm{C}$$

**The temperature of the streams**

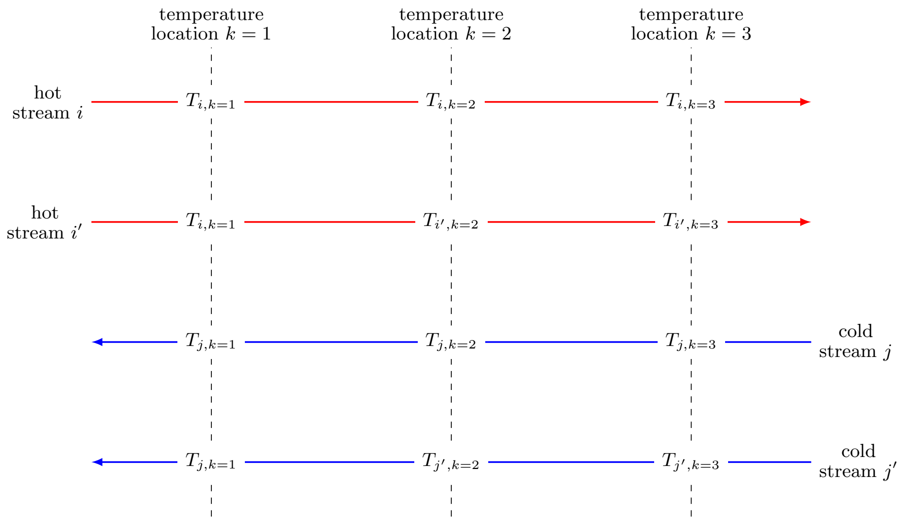
The temperature of the streams is calculated as they enter a stage (before splitting the stream) and leave a stage (after mixing the branches of the stream). Since the stages are adjacent to each other, the temperature of a stream as it leaves a stage is the same temperature as the stream enters the next stage. In this way, in a superstructure with two stages, the temperature of the streams is calculated in three locations. Those locations are depicted in Figure 4.7 and identified by the labels *temperature location* $k = 1$, $k = 2$ and $k = 3$.

A rule for the number of temperature location in a superstructure can be generalised as follows: in a superstructure with $n_k$ stages, the temperature of the streams is calculated in $n_k + 1$ locations.

**The heat balance at each stage**

Inside a stage, a stream may be matched with one or more streams of the other type for heat exchange (Figure 4.8). The heat balance at a stage can be obtained from the sum of all heat flow rates. By assuming that the branches of a stream leave the heat exchangers at the same temperature, we can express the heat flow rate with a single temperature difference term:

**Figure 4.7:** The temperature locations.



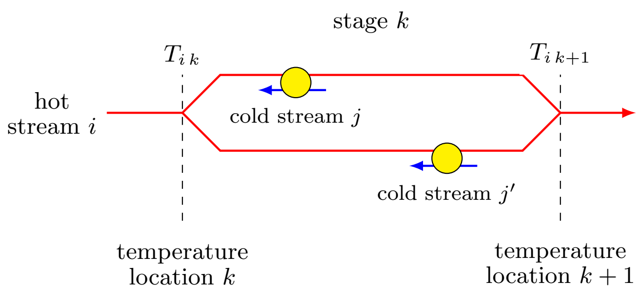

$$\dot{m}cp_i \left(t_{i,k} - t_{i,k+1}\right) = \sum_{j \in \mathrm{C}} \dot{q}_{i,j,k} \tag{4.5}$$

$$i \in \mathrm{H}, \quad k \in \mathrm{K}$$

$$\dot{m}cp_j \left(t_{j,k} - t_{j,k+1}\right) = \sum_{i \in \mathrm{H}} \dot{q}_{i,j,k} \tag{4.6}$$

$$j \in \mathrm{C}, \quad k \in \mathrm{K}$$

**Figure 4.8:** Hot stream $i$ exchanging heat inside stage $k$.




### Constraint on the superstructure inlet temperature

The temperature of the hot streams calculated at the first temperature location ($k = 1$) is equal to their supply temperature. In an analogous manner, the temperature of the cold streams calculated at the last temperature location ($k = n_k + 1$) is equal to their supply temperature:

$$t_{i,k=1} = T\_supply_i \tag{4.7}$$

$$i \in \mathrm{H}, \quad k \in \mathrm{K}$$

$$t_{j,k=n_k+1} = T\_supply_j \tag{4.8}$$

$$j \in \mathrm{C}, \quad k \in \mathrm{K}$$

**Monotonic decrease of temperature**

In the superstructure, the temperature of the streams is highest at the temperature location $k = 1$ and lowest at the temperature location $k = n_k + 1$. The monotonic decrease of the temperature of the hot and cold streams through the temperature locations can be guaranteed by the following constraints:

$$t_{i,k} \geq t_{i,k+1} \tag{4.9}$$

$$i \in \mathrm{H}, \quad k \in \mathrm{K}$$

$$t_{j,k} \geq t_{j,k+1} \tag{4.10}$$

$$j \in \mathrm{C}, \quad k \in \mathrm{K}$$

**Constraint on the superstructure outlet temperature**

After the hot and cold streams ran through all the stages of the superstructure, they may or may not have reached their target temperatures. In this case, the temperature of the hot streams at the last temperature location ($k = n_k + 1$) is greater than or equal to their target temperature. In an opposite manner, the temperature of the cold streams at the first temperature location ($k = 1$) is less than or equal to their target temperature:

$$t_{i,k=n_k} \geq T\_target_i \tag{4.11}$$

$$i \in \mathrm{H}, \quad k \in \mathrm{K}$$

$$t_{j,k=1} \leq T\_target_j \tag{4.12}$$

$$j \in \mathrm{C}, \quad k \in \mathrm{K}$$

**The use of utilities**

The heat flow rate between a hot stream and a cold utility stream is calculated from the temperature of the hot stream at the last temperature location ($k = n_k + 1$)

and its target temperature (Figure 4.9). This procedure is also applied for the heat exchange between a cold stream and a hot utility stream (Figure 4.10). For the cold streams, the temperatures to be used are the target temperature of the cold stream and its temperature at the first temperature location ($k = 1$):

$$\dot{q}\_cooler_i = \dot{q}_i(t_{i,k=n_k+1} - T\_target_i) \tag{4.13}$$
$$i \in \mathrm{H}, \quad k \in \mathrm{K}$$

$$\dot{q}\_heater_j = \dot{q}_j(T\_target_j - t_{j,k=1}) \tag{4.14}$$
$$j \in \mathrm{C}, \quad k \in \mathrm{K}$$

**Figure 4.9:** Hot stream $i$ exchanging heat with a cold utility stream from its temperature at the last location to its target temperature.

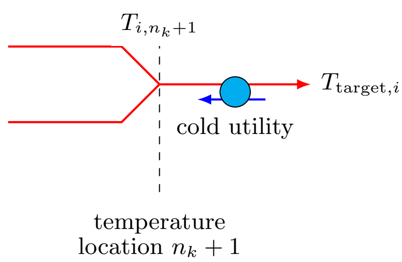
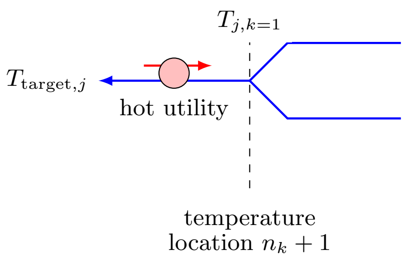



**Figure 4.10:** Cold stream $j$ exchanging heat with a hot utility stream from its temperature at the first location to its target temperature.



### The utility consumption

The utility consumption is defined as the summation of the heat flow rate between process streams and the utility streams. In this sense, the cold utility consumption ($\dot{q}_{\textit{cold-utility}}$) can be obtained by summing the heat flow rate between hot stream $i$ and the cold utility stream over the set of hot streams H. The hot utility consumption ($\dot{q}_{\textit{hot-utility}}$), in its turn, can be obtained in an analogous manner for cold stream $j$ over the set C:

$$\dot{q}_{\textit{cold-utility}} = \sum_{i \in \mathrm{H}} \dot{q}\_cooler_i \tag{4.15}$$

$$\dot{q}_{\textit{hot-utility}} = \sum_{j \in \mathrm{C}} \dot{q}\_heater_j \tag{4.16}$$

### Logical constraints to calculate the heat flow rate

To denote whether hot stream $i$ and cold stream $j$ are matched for heat exchange at stage $k$, we will use the binary variable $\sigma_{i,j,k}$:

$$\sigma_{i,j,k} = \begin{cases} 1, & \text{if hot stream } i \text{ and cold stream } j \text{ are matched for heat} \\ & \text{exchange other at stage } k; \\ 0, & \text{otherwise.} \end{cases} \tag{4.17}$$

$$i \in \mathrm{H}, \quad j \in \mathrm{C}, \quad k \in \mathrm{K}$$

If hot stream $i$ and cold stream $j$ are matched for heat exchange at stage $k$, then the heat flow rate is equal to a value between zero and the smallest involved heat load. The heat load of a stream was previously defined in equations (4.1) and (4.2). If hot stream $i$ and cold stream $j$ are not matched for heat exchange at stage $k$, then the heat flow rate is equal to zero. These constraints can be expressed by:

$$\dot{q}_{i,j,k} - \Omega_{i,j}\, \sigma_{i,j,k} \leq 0 \tag{4.18}$$

$$\dot{q}_{i,j,k} \geq 0 \tag{4.19}$$

$$i \in \mathrm{H}, \quad j \in \mathrm{C}, \quad k \in \mathrm{K}$$

where $\dot{q}_{i,j,k}$ represents the heat flow rate and $\Omega_{i,j}$ compares the heat load of hot stream $i$ and cold stream $j$ and takes the smallest of them by means of:

$$\Omega_{i,j} = \max(\dot{q}_i, \dot{q}_j) \tag{4.20}$$

$$i \in \mathrm{H}, \quad j \in \mathrm{C}$$

To denote whether a hot or a cold stream is matched with a utility stream for heat exchange, we will use the binary variable $\sigma\_td$ in the same manner as shown in Equation (4.17). If a hot or a cold stream is matched with a utility stream for heat exchange, then the heat flow rate is equal to a value between zero and the heat load of the hot or cold stream ($\dot{q}_i$ or $\dot{q}_j$), otherwise, the heat flow rate is equal to zero:

$$\dot{q}\_cooler_i - \dot{q}_i\,\sigma\_td_i \leq 0 \tag{4.21}$$

$$\dot{q}\_cooler_i \geq 0 \tag{4.22}$$

$$i \in \mathrm{H}$$

$$\dot{q}\_heater_j - \dot{q}_j\,\sigma\_td_j \leq 0 \tag{4.23}$$

$$\dot{q}\_heater_j \geq 0 \tag{4.24}$$

$$j \in \mathrm{C}$$

### 4.1.3 Objective function

As mentioned in the background section of this thesis (Section 1.1), the objective of designing a HEN includes to reduce the utility consumption and therefore its cost. The heat exchange between the process streams requires the purchase and installation of heat exchanger devices. Thus, the decrease in the cost of utilities is accompanied by an increase in the cost of heat exchangers. By solving this mathematical model, these costs are minimised simultaneously.

The minimisation of the costs of utilities and heat exchangers can be done either indirectly or directly in the objective function. As the name suggests, the direct approach calculates the costs of utilities and heat exchangers in an direct manner. The cost of heat exchangers is dependent on the heat transfer area. The temperature variables of the streams go in the denominator of the heat transfer area term. As a result, the direct approach results in a mixed-integer **non-linear** programming model.

In the indirect approach, the attributes of the HEN, such as the utility consumption, total heat transfer and number of heat exchangers can be restricted to an upper boundary or minimised in the objective function. The indirect approach is enough for we to demonstrate the application of the superstructure for HEN. Thus, it will be presented in the remaining part of this chapter. The direct approach will be addressed in the next chapter.

**Indirect approach**

To indirectly minimise the cost of utilities, a target or an upper limit can be set on the utility consumption. The value to be set can be obtained through the composite curve analysis or the problem table algorithm (SMITH, 2005; BIEGLER et al., 1997). The cold utility consumption ($\dot{q}_{cold\text{-}utility}$) and hot utility consumption ($\dot{q}_{hot\text{-}utility}$) will be set to an upper limit by means of the following inequalities:

$$\dot{q}_{cold\text{-}utility} \leq \zeta_{cold\text{-}utility} \tag{4.25}$$

$$\dot{q}_{hot\text{-}utility} \leq \zeta_{hot\text{-}utility} \tag{4.26}$$

where the parameters $\zeta_{hot\text{-}utility}$ and $\zeta_{cold\text{-}utility}$ represent the maximum hot and cold utility consumption, respectively.

By setting an upper limit on the utility consumption, the heat balance of the process streams will be satisfied by exchanging heat with each other. There may exist many topologies for the HEN that satisfy the heat balance of the streams. Some of them may feature many heat exchangers. To avoid HENs with such a topology, the objective function can be defined as the minimisation of the number of heat exchangers. In doing so, the cost of heat exchangers is minimised in an **indirect** manner.

The number of heat exchangers can be obtained from the mathematical model of the superstructure. To denote whether two process streams are matched for heat exchange, the binary variable $\sigma_{i,j,k}$ is used. In a similar manner, the match between a process stream and a utility stream is denoted by the variable $\sigma\_td_i$ for the hot utility and, the variable $\sigma\_td_j$ for the cold utility. A match between two streams may require one or more heat exchangers. By minimising the number of stream matches, we will be minimising the number of heat exchangers:

$$\min \sum_{i \in \mathrm{H}} \sum_{j \in \mathrm{C}} \sum_{k \in \mathrm{K}} \sigma_{i,j,k} + \sum_{j \in \mathrm{C}} \sigma\_td_j + \sum_{i \in \mathrm{H}} \sigma\_td_i \tag{4.27}$$

## 4.2 Example: a process for the conversion of bio-ethanol to gasoline

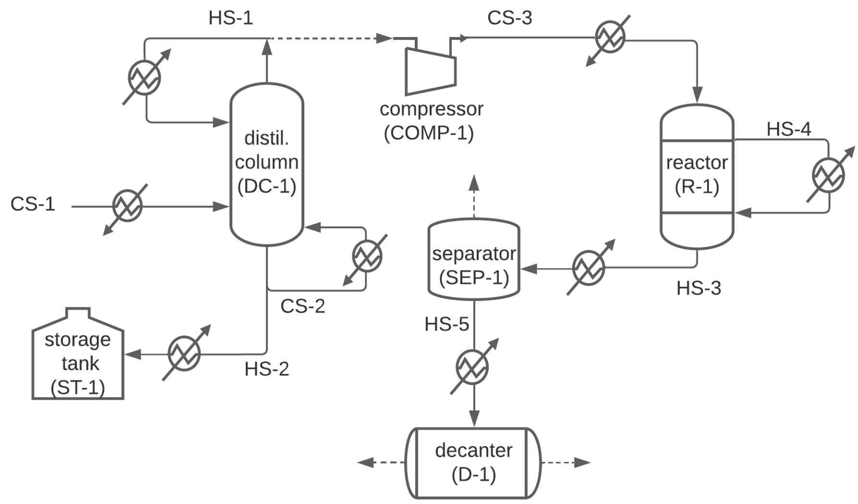
In this section, we will show the application of the superstructure for HEN design. Let us consider the chemical process adapted from Serth and Lestina (2014) and Aldridge et al. (1984). The goal of this process is to convert bio-ethanol into gasoline. There are five streams that require cooling and three streams that require heating. The streams with their supply and target temperatures and heat loads are shown in Table 4.1.

**Table 4.1:** The supply and target temperatures and the heat loads of the streams for the conversion of bio-ethanol into gasoline process.

| stream | | $T_{supply}$ (°C) | $T_{target}$ (°C) | $\dot{q}$ (kW) |
|---|---|---|---|---|
| hot stream | HS-1 | 89.0 | 80.7 | 147.7 |
| | HS-2 | 99.9 | 45.0 | 483.1 |
| | HS-3 | 300.0 | 110.0 | 836.0 |
| | HS-4 | 290.0 | 200.0 | 576.0 |
| | HS-5 | 110.0 | 20.0 | 126.0 |
| cold stream | CS-1 | 25.0 | 91.9 | 682.4 |
| | CS-2 | 99.0 | 100.0 | 1170.7 |
| | CS-3 | 238.7 | 321.7 | 74.7 |

The process flowsheet diagram is depicted in Figure 4.11. Stream CS-1 is heated and transported to Distillation Column DC-1. Stream CS-2 is removed from the bottom of the distillation column, vaporised in a reboiler and returned to the distillation column. Stream HS-2 is the bottom product of the distillation column and needs to be cooled before being stored in Tank ST-1. Stream HS-1 is removed from the top of the distillation column, condensed and refluxed to the distillation column.

**Figure 4.11:** The flowsheet of the process for the conversion of bio-ethanol into gasoline.




Stream CS-3 is the top product of the distillation column. After leaving the distillation column, this stream is compressed (COMP-1), heated and transported to Reactor R-1. The reactor product stream, Stream HS-3, is cooled and transported to Separator SEP-1. A mixture of gasoline and water (Stream HS-5) is removed from the separator and cooled prior to entering Decanter DEC-1. A thermal fluid identified as Stream HS-4 is used to remove heat from the reactor. The thermal fluid circulates the reactor jacket, passes through a cooling system and re-enters the reactor jacket.

With the information presented above, we will study the scenarios for heat integration of the process. In order to do that, we will use the pinch method, whose a brief description was given in the background section of this thesis (Section 1.1). Based on the targets output from the pinch method, the mathematical model of the superstructure will be solved to design the HEN of the process. All the computations will be carried out by using the General Algebraic Modelling System (GAMS) software version 40.4.0 with CPLEX solver version 22.1.0.

**Process pinch analysis**

To obtain the target for utility consumption and total heat transfer area, we performed a pinch analysis over a range of minimum driving force for heat transfer. The minimum costs of utilities and heat exchangers was found for a minimum driving force for heat transfer of 13.0 °C. The obtained target for the minimum hot utility consumption is 28 kW and that for the minimum total heat transfer area is 400 $m^2$.

**Design of the HEN through superstructure**

To obtain a heat-integrated process, the hot utility consumption will be restricted to being not greater than the target output from the pinch method, which is 28 kW. The objective function will be defined as the minimisation of the number of heat exchangers. The number of stages of the superstructure will be defined so that a single stream can exchange heat in series with all the other streams. Based on the number of hot and cold streams in the process, a hot stream can exchange heat in series with up to three cold streams. On the other hand, a cold stream can exchange heat in series with up to five hot streams. Thus, the number of stages for this example is five.

**Results and discussion**

The designed HEN has six process-to-process heat exchangers, two coolers and one heater, which totals nine heat exchangers. Based on the heat flow rate and stream temperatures, we can calculate the temperature differences in the heat exchangers and heat transfer area in a post-run environment. The heat transfer area will be calculated by means of:

$$A = \frac{\dot{q}}{U\, LMTD}$$

The log mean temperature difference ($LMTD$) will be approximated by Paterson (1984):

$$LMTD = \frac{2}{3}\sqrt{\theta_1 \theta_2} + \frac{\theta_1 + \theta_2}{6}$$

where $\theta$ is the difference between the temperature of the hot and cold stream at one end of the heat exchanger. To calculate the global heat transfer coefficient ($U$), we will assume that all the streams have a heat transfer coefficient of $200\,\mathrm{W\,m^{-2}\,K}$.

The total heat transfer area is approximately $990\,\mathrm{m}^2$. There is a heat exchanger involving the the bottom product and the feed of the distillation column (the streams HS-2 and CS-1) that has a heat transfer area of around $697\,\mathrm{m}^2$. This value is significantly greater than the total heat transfer area of $400\,\mathrm{m}^2$ estimated from the pinch analysis. Furthermore, this heat exchanger operates with tight temperature differences of 8.0 °C and 2.7 °C at the left and right ends, respectively.

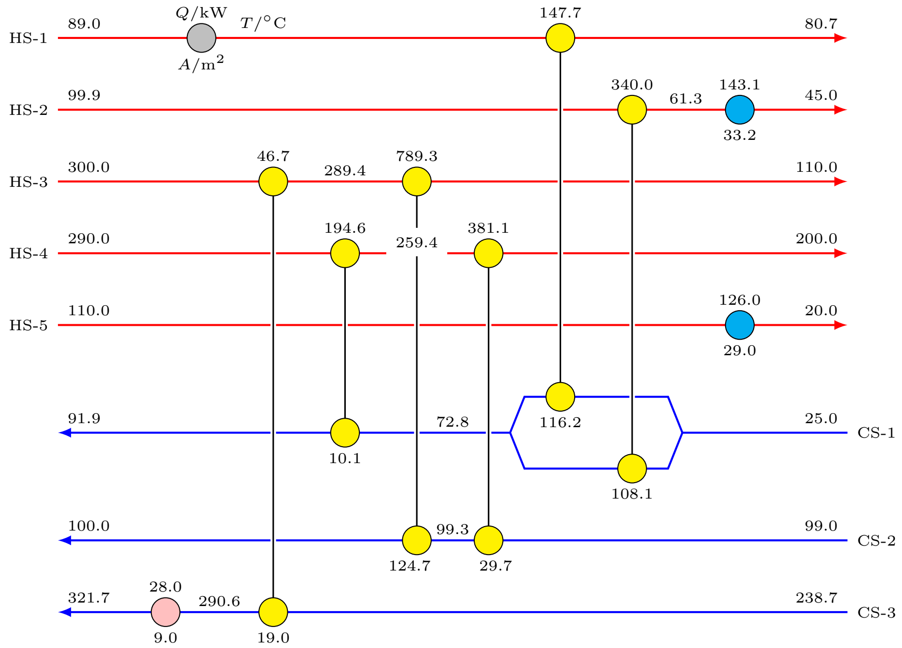
These results can be explained by the fact that the heat transfer area was not minimised in the objective function. The feasible solutions found in the search procedure were evaluated only in terms of the number of heat exchangers. There were multiple optima solutions with different heat transfer areas. Coincidentally, the one returned by the solver has a heat exchanger with a large heat transfer area.

This analysis suggests that there is room for we to achieve a HEN that requires a smaller total heat transfer area. In order to do that, we will restrict the number of heat exchangers to being less than or equal to nine. Although the previous obtained HEN is also structured with nine heat exchangers, by adding a constraint to the mathematical model, the solver will be perturbed and may move from an optimum solution to another.

The obtained HEN has nine heat exchangers. The HEN topology with the heat flow rates, heat transfer areas and stream temperatures is shown in Figure 4.12. The total heat transfer area is approximately $409\,\mathrm{m}^2$, which is to some extent close to that estimated from the pinch method. The smallest temperature difference in a single process-to-process heat exchanger is around 9.4 °C and involves the reactor product (Stream HS-3) and the stream that leaves the compressor (Stream CS-3).

A further step in this example is to restrict the number of heat exchangers to being less than or equal to eight. The result of doing that shows it is impossible to design a HEN that operates with 28 kW in hot utility and less than nine heat exchangers. In other words, if we want a HEN with fewer heat exchangers it is necessary to allow a higher utility consumption.

**Figure 4.12:** Topology of the HEN designed by minimising the number of heat exchangers for a hot utility consumption fixed at 28 kW. The number of heat exchangers was restricted to being less than or equal to nine.



## 4.3 The superstructure for heat exchanger plant layout design: pipe length

### 4.3.1 Concept

A superstructure for HEN design allows all the possible matches between the streams for heat exchange. In the design of HENs, the streams are matched to optimise the objective function and satisfy the constraints. For the matched streams, the heat flow rate is calculated from their temperatures, mass flow rate and specific heat. In contrast, for the not matched streams, a null heat flow rate is assigned, as if the streams did not exchange heat (YEE; GROSSMANN, 1990).

The concept of the superstructure inspired us to think that for each possible match between the streams there is a heat exchanger in the plant. The coordinates of the heat exchangers should be defined to minimise the length of the pipes. The length of the pipes should be calculated based on whether the streams associated to the heat exchanger are matched or not in the HEN.

For example, if a hot and a cold stream are matched, then the length of pipe for their transport should be calculated from the coordinates of the associated heat exchanger. In contrast, if the hot and cold streams are not matched, then the length of pipe for their transport should be zero. In this case, the coordinates of the heat exchanger does not affect the objective value.

The set of ideas presented above is the basis for we to formulate a superstructure. By calculating the length of the pipes with the heat exchangers at all the possible coordinates, we are able to find the optimal set of coordinates. Thus, we obtain the layout of the heat exchangers in the plant.

### 4.3.2 Mathematical model

The set of ideas presented in this section will be expressed in a mixed-integer linear programming model. The assumptions to develop the mathematical model were made in the first chapter of this thesis, Subsection 1.4.5.

#### Input data

The input data is the set of hot streams, cold streams and stages of the superstructure. This data was input into the mathematical model of the superstructure for HEN, as follows:

- a set H of hot streams $i$;
- a set C of cold streams $j$;

- a set K of stages $k$;

Input data is also the set A of rectangular axes $a$, where $a = x, y$.

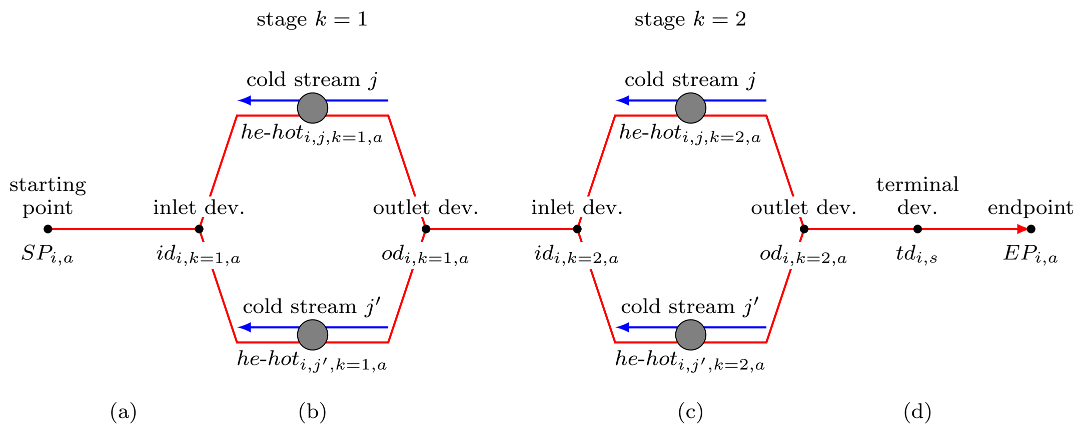
### Parameters

The parameters are the coordinates of the starting and endpoints of the streams. The coordinate of the starting and endpoint of hot stream $i$ on the axis $a$ will be represented by the parameters $SP_{i,a}$ and $EP_{i,a}$, respectively. To represent the coordinates of the starting and endpoint of cold stream $j$ the index $j$ should be replaced with the index $j$.

### Decision variables

The decision variables are the coordinates of the heat exchangers and other devices, which will be introduced as this section progresses. To define the decision variables, let us consider a process with one hot stream and two cold streams. The hot stream will be represented by the index $i$ and the cold streams by the indices $j$ and $j'$. Let us also consider a two-stage superstructure that allows all the possible matches between the hot and cold streams for heat exchange. The stages of the superstructure will be represented by the indices $k = 1$ and $k = 2$. The topology of hot stream $i$ in the superstructure is depicted in Figure 4.13.

**Figure 4.13:** The topology of hot stream $i$ in the superstructure.




As depicted in Part **(a)** of Figure 4.13, hot stream $i$ is first located in its starting point. Then, hot stream $i$ leaves from its starting point to stage $k = 1$. At the entrance of this stage $k = 1$, there is an inlet device (Part **(b)** of Figure 4.13). The inlet device is a splitter if the stream exchanges heat in parallel, otherwise, it is a simple joint between two pipe segments. The coordinate of the inlet device of stage $k$ on the axis $a$ will be represented by the variable $id_{i,k,a}$.

Inside stage $k = 1$, there are two heat exchangers for hot stream $i$ to exchange heat with each cold stream individually (Part **(b)** of Figure 4.13). The coordinate on the axis $a$ of the heat exchanger that hot stream $i$ exchanges heat with cold stream $j$ at stage $k$ will be represented by the variable $ex_{i,j,k,a}$.

At the exit of stage $k = 1$, there is an outlet device (Part **(b)** of Figure 4.13). The outlet device is a mixer if the stream exchanged heat in parallel, otherwise, it is a simple joint between two pipe segments. The coordinate on the axis $a$ of the outlet device of stage $k$ will be represented by the variable $od_{i,k,a}$.

The sequences of pieces of equipment along which hot stream $i$ passed at stage $k = 1$ is repeated at stage $k = 2$ (Part **(c)** of Figure 4.13). After the outlet device of stage $k = 2$, there is the terminal device (Part **(d)** of Figure 4.13). If hot stream $i$ needs to reject a surplus of heat, then the outlet device is a cooler, otherwise, it is a simple joint between two pipe segments. The procedure to denote whether hot stream $i$ needs extra cooling was presented in the mathematical model of the superstructure for HEN, Equation (4.21).

The coordinate of the terminal device on the axis $a$ will be denoted by the variable $td_{i,a}$. A general rule about the terminal device is that it always exists after the outlet device of the last stage of this superstructure. At the end of hot stream $i$ there is its endpoint. The pieces of equipment and the variables to represent their coordinates are summarised in Table 4.2.

**Table 4.2:** The pieces of equipment in the superstructure and the variables to denote the coordinates in the plant.

| piece of equipment for hot stream $i$ | variable coordinate on the axis $a$ |
|---|---|
| starting point | $SP_{i,a}$ |
| inlet device of stage $k$ | $id_{i,k,a}$ |
| heat exchanger for hot stream $i$ and cold stream $j$ at stage $k$ | $ex_{i,j,k,a}$ |
| outlet device of stage $k$ | $od_{i,k,a}$ |
| terminal device | $td_{i,a}$ |
| endpoint | $EP_{i,a}$ |

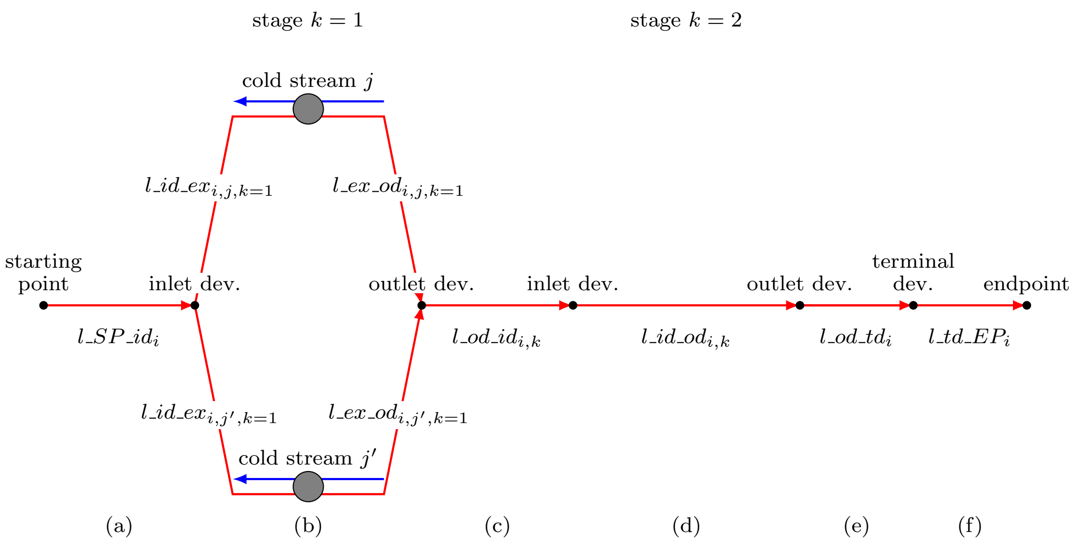
Cold stream $j$ passes through pieces of equipment of the same type as those for hot stream $i$. The variable to represent the coordinate on the axis $a$ of a heat exchanger is indexed by both hot stream $i$ and cold stream $j$ ($ex_{i,j,k,a}$). The coordinate on the axis $a$ of the inlet and outlet devices of stage $k$ will be represented by the variables $id_{j,k,a}$ and $od_{j,k,a}$, respectively. The coordinate on the axis $a$ of the terminal device will be represented by the variable $td_{j,a}$.

### Dependent variables and constraints: hot stream $i$

The dependent variables are the length of the pipes to transport the streams. The constraints are the equations to calculate the length of the pipes. To define the dependent variables and constraints, let us consider the process introduced in this section. There are one hot stream $i$ and two cold streams $j$ and $j'$. We will modify the superstructure of hot stream $i$ previously shown in Figure 4.13 as follows:

- hot stream $i$ exchanges heat in parallel with both cold streams at stage $k = 1$;
- hot stream $i$ does not exchange heat with any cold stream at stage $k = 2$;
- the topology of the modified superstructure is depicted in Figure 4.14.

**Figure 4.14:** The topology of the modified superstructure of hot stream $i$.




As shown in Figure 4.14, there are a number of pipes to transport hot stream $i$ from one piece of equipment to another. The length of each pipe will be calculated from the rectangular distance between the coordinates of the pieces of equipment being connected. The procedure to do that was described in the first chapter of this thesis, Subsection 1.4.4.

Part **(a)** of Figure 4.14 shows that hot stream $i$ is transported from its starting point to the inlet device of the first stage. The length of pipe required to connect the starting point to the inlet device of the first stage will be represented by the variable $L\_SP\_id_i$ and calculated by:

$$L\_SP\_id_{i,a} \geq SP_{i,a} - id_{i,k=1,a} \tag{4.28}$$

$$L\_SP\_id_{i,a} \geq id_{i,k=1,a} - SP_{i,a} \tag{4.29}$$

$$i \in \mathrm{H}, \quad a \in \mathrm{A}$$

$$L\_SP\_id_i = \sum_{a \in \mathrm{A}} L\_SP\_id_{i,a} \tag{4.30}$$

$$i \in \mathrm{H}$$

As the pipe that transports hot stream $i$ passes through the inlet device of stage $k = 1$, it is split into two pipe branches (Part **(b)** of Figure 4.14). The pipe branches transport hot stream $i$ to exchange heat in parallel with the cold streams $j$ and $j'$. In a more detailed description, each pipe branch transports hot stream $i$ (or a fraction of its mass flow rate) from the inlet device to a heat exchanger, and from that heat exchanger to the outlet device. The length of each pipe branch depends on whether hot stream $i$ is matched with the corresponding cold stream for heat exchange.

At this point, we shall remember the mathematical model of the superstructure for HEN, which was presented in the previous section (Section 4.1). If the hot stream $i$ and cold stream $j$ are matched for heat exchange at stage $k$, then the binary variable $\sigma_{i,j,k}$ is equal to one. For the case where these streams are not matched for heat exchange, the binary variable $\sigma_{i,j,k}$ is equal to zero. We can understand this binary variable as the link between the superstructure for the HEN design and the superstructure being developed here.

The binary variable $\sigma_{i,j,k}$ will be used to calculate or not the length of the pipe connected to the heat exchangers. If hot stream $i$ is matched with cold stream $j$ for heat exchange, then the length of the pipe branch is equal to the rectangular distances from the inlet device to the heat exchanger plus from the heat exchanger to the outlet device. If hot stream $i$ is not matched with cold stream $j$ for heat exchange, then the pipe branch has a null length regardless of the coordinates of the heat exchanger.

The length of the pipe branch that transports hot stream $i$ from the inlet device to the heat exchanger — where hot stream $i$ exchanges heat with cold stream $j$ at stage $k$ ($L\_id\_ex_{i,j,k}$) — will be represented by the variable $L\_id\_ex_{i,j,k}$ and calculated by:

$$L\_id\_ex_{i,j,k,a} \geq id_{i,k,a} - ex_{i,j,k,a} - \mathcal{M}\left(1 - \sigma_{i,j,k}\right) \tag{4.31}$$

$$L\_id\_ex_{i,j,k,a} \geq ex_{i,j,k,a} - id_{i,k,a} - \mathcal{M}\left(1 - \sigma_{i,j,k}\right) \tag{4.32}$$

$$L\_id\_ex_{i,j,k,a} \geq 0 \tag{4.33}$$

$$i \in \mathrm{H}, \quad j \in \mathrm{C}, \quad k \in \mathrm{K} \quad a \in \mathrm{A}$$

$$L\_id\_ex_{i,j,k} = \sum_{a \in \mathrm{A}} L\_id\_ex_{i,j,k,a} \tag{4.34}$$

$$i \in \mathrm{H}, \quad j \in \mathrm{C}, \quad k \in \mathrm{K}$$

Similarly, the length of the pipe branch that transports hot stream $i$ from the heat exchanger — where hot stream $i$ exchanges heat with cold stream $j$ at stage $k$ — to the outlet device ($L\_ex\_od_{i,j,k}$) will be represented by the variable $L\_ex\_od_{i,j,k}$ and calculated by:

$$L\_ex\_od_{i,j,k,a} \geq ex_{i,j,k,a} - od_{i,k,a} - \mathcal{M}\left(1 - \sigma_{i,j,k}\right) \tag{4.35}$$

$$L\_ex\_od_{i,j,k,a} \geq od_{i,k,a} - ex_{i,j,k,a} - \mathcal{M}\left(1 - \sigma_{i,j,k}\right) \tag{4.36}$$

$$L\_ex\_od_{i,j,k,a} \geq 0 \tag{4.37}$$

$$i \in \mathrm{H}, \quad j \in \mathrm{C}, \quad k \in \mathrm{K} \quad a \in a$$

$$L\_ex\_od_{i,j,k} = \sum_{a \in \mathrm{A}} L\_ex\_od_{i,j,k,a} \tag{4.38}$$

$$i \in \mathrm{H}, \quad j \in \mathrm{C}, \quad k \in \mathrm{K}$$

The length of the pipe branch from the inlet device to the heat exchanger and, from the heat exchanger to the outlet device can be summed:

$$L\_branch_{i,j,k} = L\_id\_ex_{i,j,k} + L\_ex\_od_{i,j,k} \tag{4.39}$$

$$i \in \mathrm{H}, \quad j \in \mathrm{C}, \quad k \in \mathrm{K}$$

It is worth discussing the role of the term $\mathcal{M}\left(1 - \sigma_{i,j,k}\right)$ in the inequalities of the equations (4.31) and (4.32) and equations (4.35) and (4.36). If hot stream $i$ is matched with cold stream $j$ for heat exchange at stage $k$ ($\sigma_{i,j,k} = 1$), then the term $\mathcal{M}\left(1 - \sigma_{i,j,k}\right)$ is cancelled out. As a consequence, the right-hand side of the inequalities becomes the differences between the coordinates of the equipment being connected.

If hot stream $i$ is not matched with cold stream $j$ for heat exchange at stage $k$ ($\sigma_{i,j,k} = 0$), then the right-hand side of the inequalities is controlled by the large constant $\mathcal{M}$. In this way, the variable length of the pipe branch on the left-hand side of the inequalities can assume any non-negative value. This variable will be minimised in the objective function, as a consequence, it will assume the minimum value possible, which is zero.

If hot stream $i$ is not matched with any cold stream for heat exchange, then a null length will be assigned to all pipe branches. This situation is depicted in Part **(d)** of Figure 4.14. In this case, we need another pipe to transport hot stream $i$ directly from the inlet to the outlet device of the stage. The length of pipe to do that will be represented by the variable $L\_id\_od_{i,k}$ and calculated by:

$$L\_id\_od_{i,k,a} \geq id_{i,k,a} - od_{i,k,a} - \mathcal{M} \sum_{j \in \mathrm{C}} \sigma_{i,j,k} \tag{4.40}$$

$$L\_id\_od_{i,k,a} \geq od_{i,k,a} - id_{i,k,a} - \mathcal{M} \sum_{j \in \mathrm{C}} \sigma_{i,j,k} \tag{4.41}$$

$$L\_id\_od_{i,k,a} \geq 0 \tag{4.42}$$

$$i \in \mathrm{H}, \quad k \in \mathrm{K} \quad a \in \mathrm{A}$$

$$L\_id\_od_{i,k} = \sum_{a \in \mathrm{A}} L\_id\_od_{i,k,a} \tag{4.43}$$

$$i \in \mathrm{H}, \quad k \in \mathrm{K}$$

In the equations (4.40) and (4.41), the term:

$$\mathcal{M} \sum_{j \in \mathrm{C}} \sigma_{i,j,k}$$

functions to make zero the length of the pipe that connects the inlet and outlet devices — the variable $L\_id\_od_{i,k,a}$ on the left-hand side — in case hot stream $i$ is matched with any cold stream for heat exchange.

The Part **(c)** of Figure 4.14 shows that hot stream $i$ is transported from the outlet device of stage $k = 1$ to the inlet device of stage $k = 2$. We can generalise the transport of hot stream $i$ from one stage to another by means of the indices $k$ and $k + 1$. The length of pipe to connect the outlet device of stage $k$ to the inlet device of stage $k + 1$ will be represented by the variable $L\_od\_id_{i,k}$ and calculated by:

$$L\_od\_id_{i,k,a} \geq od_{i,k,a} - id_{i,k+1,a} \tag{4.44}$$

$$L\_od\_id_{i,k,a} \geq id_{i,k+1,a} - od_{i,k,a} \tag{4.45}$$

$$i \in \mathrm{H}, \quad k \in \mathrm{K}, \quad a \in a$$

$$L\_od\_id_{i,k} = \sum_{a \in \mathrm{A}} L\_od\_id_{i,k,a} \tag{4.46}$$

$$i \in \mathrm{H}, \quad k \in \mathrm{K}$$

The index $k$ in the equations (4.44) to (4.46) must assume a value less than the maximum number of stages in the superstructure. The reason why is that the index $k$ in the variable $id_{i,k+1,a}$ is summed with the number one $(k + 1)$.

To leave stage $k = 2$, hot stream $i$ is transported from the outlet device to the terminal device (Part **(e)** of Figure 4.14). The stage $k = 2$ is the last stage of the superstructure. Thus, we can say that hot stream $i$ is transported from the outlet device of the last stage specified by $k = n_k$ to the terminal device. The length of pipe to connect the outlet device of stage $k = n_k$ to the terminal device will be represented by the variable $L\_od\_td_{i,a}$ and calculated by:

$$L\_od\_td_{i,a} \geq od_{i,k,a} - td_{i,a} \tag{4.47}$$

$$L\_od\_td_{i,a} \geq td_{i,a} - od_{i,k,a} \tag{4.48}$$

$$i \in \mathrm{H}, \quad k \in \mathrm{K}, \quad a \in \mathrm{A}$$

$$L\_od\_td_i = \sum_{a \in \mathrm{A}} L\_od\_td_{i,a} \tag{4.49}$$

$$i \in \mathrm{H}, \quad k \in \mathrm{K}$$

It is worth remembering that the terminal device is a cooler if hot stream $i$ needs to reject an extra heat load, otherwise it is a simple joint between two pipe segments. For the cases that the terminal device is a cooler, the length of the pipe that transports the cold utility will not be calculated, as assumed in the first chapter of this thesis, Subsection 1.4.5.

After passing through the terminal device, hot stream $i$ is transported to its endpoint (Part **(f)** of Figure 4.14). The length of pipe to connect the terminal device to the endpoint will be represented by the variable $L\_td\_EP_i$ and calculated by:

$$L\_td\_EP_{i,a} \geq td_{i,a} - EP_{i,a} \tag{4.50}$$

$$L\_td\_EP_{i,a} \geq EP_{i,a} - td_{i,a} \tag{4.51}$$

$$i \in \mathrm{H}, \quad a \in \mathrm{A}$$

$$L\_td\_EP_i = \sum_{a \in \mathrm{A}} L\_td\_EP_{i,a} \tag{4.52}$$

$$i \in \mathrm{H}$$

The summation of the lengths of all the pipes that transport hot stream $i$ will give us the length of pipe for the transport of hot stream $i$ ($L\_total_i$):

$$\begin{aligned} L\_total_i = L\_SP\_id_i + \sum_{j \in \mathrm{C}} \sum_{k \in \mathrm{K}} L\_branch_{i,j,k} + \sum_{k \in \mathrm{K}} L\_id\_od_{i,k} \\ + \sum_{k \in \mathrm{K}} L\_od\_id_{i,k} + L\_od\_td_i + L\_td\_EP_i \end{aligned} \tag{4.53}$$

$$i \in \mathrm{H}$$

The dependent variables are summarised in Table 4.3.

**Table 4.3:** A summary of the pipes that transport hot stream $i$ and the variables to denote their lengths.

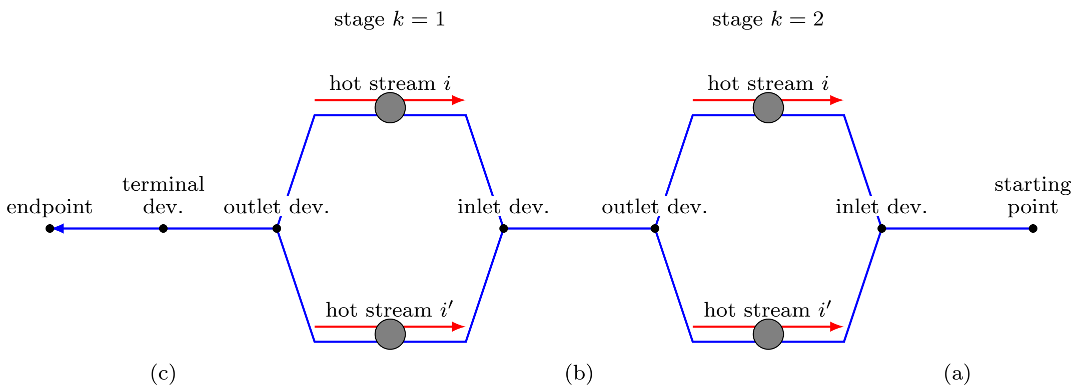
| pipe | variable for the length of pipe |
|---|---|
| from the starting point to the inlet device of the first stage | $L\_SP\_id_i$ |
| from the inlet device to the heat exchanger | $L\_id\_ex_{i,j,k}$ |
| from the heat exchanger to the outlet device | $L\_ex\_od_{i,j,k}$ |
| from the inlet device to the outlet device inside the same stage | $L\_id\_od_{i,k}$ |
| from the outlet device of a stage to the inlet device of the next stage | $L\_od\_id_{i,k}$ |
| from the outlet device of the last stage to the terminal device | $L\_od\_td_i$ |
| from the terminal device to the endpoint | $L\_td\_EP_i$ |
| the total length of pipe for the transport of hot stream $i$ | $L\_total_i$ |

### Dependent variables and constraints: cold stream $j$

To calculate the length of pipe for the transport of cold stream $j$, it is necessary to make two modifications in the equations (4.30) to (4.54). The first one is to replace the index $i$ with the index $j$. The second one is on the stages $k$. The order of the stages is a piece of information used to specify the inlet device that is connected to a starting point, the outlet device that is connected to an endpoint and the outlet and inlet devices of adjacent stages.

The stages of the superstructure used in this work are numbered from left to right in ascending order. The hot streams run from left to right following the order of the stages. In an opposite manner, the cold streams run from right to left and pass through the stages in descending order. In this way, the inlet device that is connected to the starting point of cold stream $j$ is positioned in the last stage of the superstructure (Part **(a)** of Figure 4.15).

**Figure 4.15:** The topology of cold stream $j$ running from left to right in a two-stage superstructure.




To calculate the length of pipe from the starting point of cold stream $j$ to the inlet device of the last stage, it is necessary to specify that the index $k$ in the equations (4.28) to (4.30) assume the value $k = n_k$. As cold stream $j$ runs through the stages in descending order, there is a pipe that transports cold stream $j$ from the outlet device of stage $k$ to the inlet device of stage $k - 1$ (Part **(b)** of Figure 4.15).

The length of that pipe can be calculated by replacing the index $k + 1$ in the coordinates of the inlet device shown in the equations (4.44) and (4.45) with the index $k - 1$. Notice that the index $k$ must assume values greater than one to avoid the index $k - 1$ being equal to zero because there is no stage number zero in this mathematical model.

The outlet device that is connected to the endpoint of cold stream $j$ is positioned in the first stage of the superstructure (Part **(c)** of Figure 4.15). There is an intermediate point between the outlet device of the first stage and the endpoint of cold stream $j$, which is the terminal device. To calculate the length of pipe from the outlet device of the first stage to the terminal device of cold stream $j$, it is necessary to specify that the index $k$ in equations (4.47) to (4.49) assumes the value stage $k = 1$.

By making the above-described modifications, one achieves the following equation to calculate the length of pipe for the transport of cold stream $j$:

$$\begin{aligned} L\_total_j = L\_SP\_id_j + \sum_{i \in \mathrm{H}} \sum_{k \in \mathrm{K}} L\_branch_{j,i,k} + \sum_{k \in \mathrm{K}} L\_id\_od_{j,k} \\ + \sum_{k \in \mathrm{K}} L\_od\_id_{j,k} + L\_od\_td_j + L\_td\_EP_j \end{aligned} \tag{4.54}$$

$$j \in \mathrm{C}$$

**Objective function**

The objective function will be defined as the minimisation of the length of the pipes that transport all the hot streams and all the cold streams:

$$\min \sum_{i \in \mathrm{H}} L\_total_i + \sum_{j \in \mathrm{C}} L\_total_j \tag{4.55}$$

## 4.4 Example: simultaneous design of heat exchanger network and heat exchanger plant layout

In this example, we will demonstrate the simultaneous design of HEN and heat exchanger plant layout in a stepwise fashion. In order to do that, let us consider a process with the hot streams H1 and H2 and the cold streams C1 and C2 adapted from Shenoy (1995). The process flowsheet diagram is shown in Figure 4.16. Stream C1 is carried through the pipe rack ($\mathbf{PR_1}$) and needs to be heated to enter the reactor ($\mathbf{R_1}$). The product of the reactor (Stream H1) needs to be cooled before entering the distillation column ($\mathbf{DC_1}$). Stream C2 is withdrawn from the top of the distillation column, then it is heated and transported to the pipe rack. The bottom product of the distillation column (Stream H2) is cooled and transported to the pipe rack as well.

The supply and target temperatures ($T_{supply}$ and $T_{target}$) of the streams and their heat capacity flow rate ($\dot{m}cp$) are shown in Table 4.4. According to Shenoy (1995), the optimal annual costs of utilities and heat exchangers results from a hot utility consumption of 360 kW. The corresponding a driving force for heat transfer is 13 °C. The pinch point on the hot and cold sides are 125 °C and 112 °C, respectively.

Based on the information presented above, we will firstly design only the HEN of the process by solving of the mathematical model of the superstructure. Next, we will design the HEN and heat exchanger plant layout in a simultaneous manner.

**Figure 4.16:** The process flowsheet diagram with the streams H1, H2, C1 and C2 adapted from Shenoy (1995).

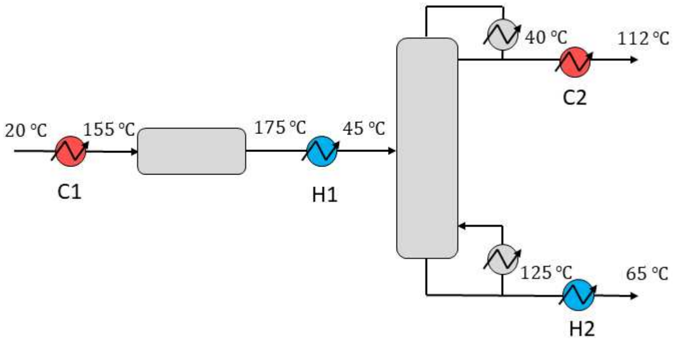



**Table 4.4:** The properties of the streams H1, H2, C1 and C2 for the HEN design problem.

| stream | $T_{supply}$ (°C) | $T_{target}$ (°C) | $\dot{m}cp$ (kW) |
|---|---|---|---|
| H1 | 175 | 45 | 10 |
| H2 | 125 | 65 | 40 |
| C1 | 20 | 155 | 20 |
| C2 | 40 | 112 | 15 |

To demonstrate the capabilities of our methodology, a number of restrictions will be applied to design the HEN and heat exchanger plant layout. All the computations will be carried out by using the General Algebraic Modelling System (GAMS) software version 40.4.0 with CPLEX solver version 22.1.0.

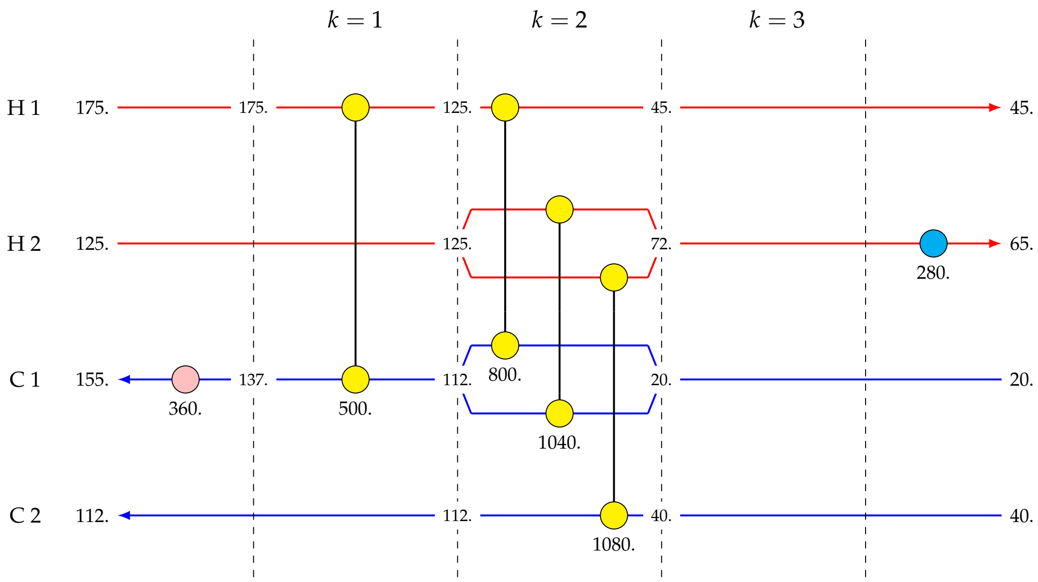
### 4.4.1 HEN design through superstructure

To design the HEN, we will fix the hot utility consumption to 360 kW and the minimum driving force for heat transfer of 13 °C. The objective function will be defined as the minimisation of the number of heat exchangers. The number of stages of the superstructure will be defined so that a stream can exchange heat in series with as many streams of the other type as there are. The streams should be able to exchange heat below and above the pinch point. There are only the streams H1 and C1 below the pinch point. All the two hot and two cold streams exchange heat above the pinch point. The number of stages in the superstructure will be one for below the pinch point plus two for above the pinch point, which totals three stages.

### Results and discussion

The HEN designed by solving the mathematical model of the superstructure consists of six heat exchangers. Stream C1 exchanges heat in parallel with the two hot streams. Then, Stream C1 exchanges heat in series with Stream H1 and the hot utility, respectively. Similarly, Stream H2 exchanges heat in parallel with the two cold streams and continues to exchange heat with cold utility. Stream C2 reaches its target temperature by exchanging heat with Stream H1 only. In summary, there are four process-to-process heat exchangers, one cooler and one heater. The HEN topology is depicted in Figure 4.17.

**Figure 4.17:** The HEN topology obtained from through superstructure by minimising the number of heat exchangers. The hot utility consumption was fixed to 360 kW. The minimum driving force for heat transfer fixed to 13 °C. The heat flow rate is shown below in each heat exchanger.

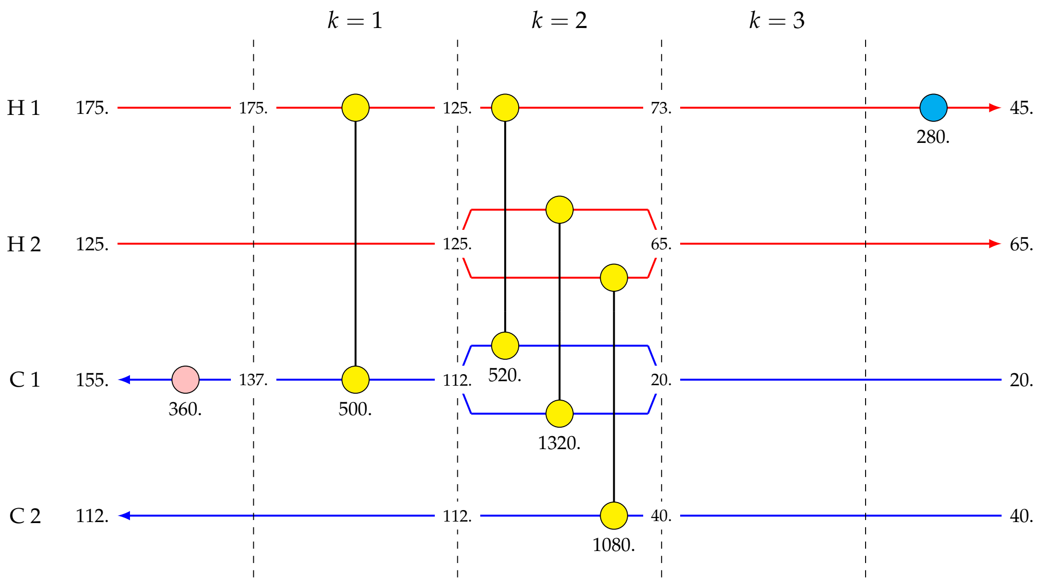



### restriction on the HEN topology: Stream H2 does not exchange heat with the cold utility

Now, we will apply a number of constraints to the mathematical model of the superstructure for HEN. Those constraints will explore different HEN topologies for the same hot utility consumption fixed of 360 kW and minimum driving force for heat transfer. Let us start off by forbidding Stream H2 to exchange heat with the cold utility. The HEN obtained from this prohibition has the same topology of the previous one with the exception of the cooler, which was replaced from Stream H2 to H1.

The heat flow rate in the heat exchangers was redistributed. Stream H2 could reach its target temperature by increasing the heat flow rate with Stream C1. As a consequence, the heat flow rate between Stream H1 and Stream C1 decreased. This required Stream H1 to reject the heat surplus to the cold utility. The HEN topology is depicted in Figure 4.18.

**Figure 4.18:** HEN topology obtained by restricting Stream H2 to not exchanging heat with the cold utility.

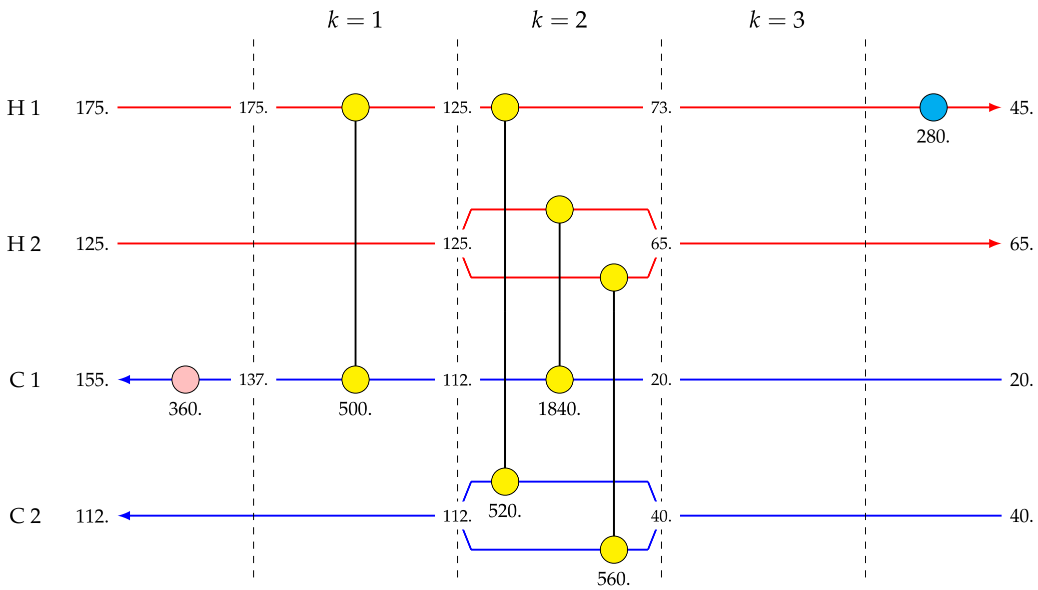



**restriction on the HEN topology: Stream C1 cannot exchange heat in parallel**

The two HEN topologies obtained thus far have two heat exchangers arranged in parallel on Stream C1. We will restrict Stream C1 to exchanging heat in series only. This restriction results in a still not seen HEN topology. Stream C1 exchanges heat in series with Stream H2, Stream H1 and the hot utility, respectively. There is one less heat exchanger installed on Stream C1 compared to the other HEN presented in figures 4.17 and 4.18.

The new feature of the HEN is that the other cold stream, Stream C2, exchanges heat in parallel with the two hot streams. This can also be understood from the view point of the hot streams: Stream H1 exchanges heat with both the streams C1 and C2 rather than doing this with Stream C1 twice. Since one heat exchanger was removed from Stream C1 but a new one was added to Stream C2, the HEN remains with six heat exchangers (Figure 4.19).

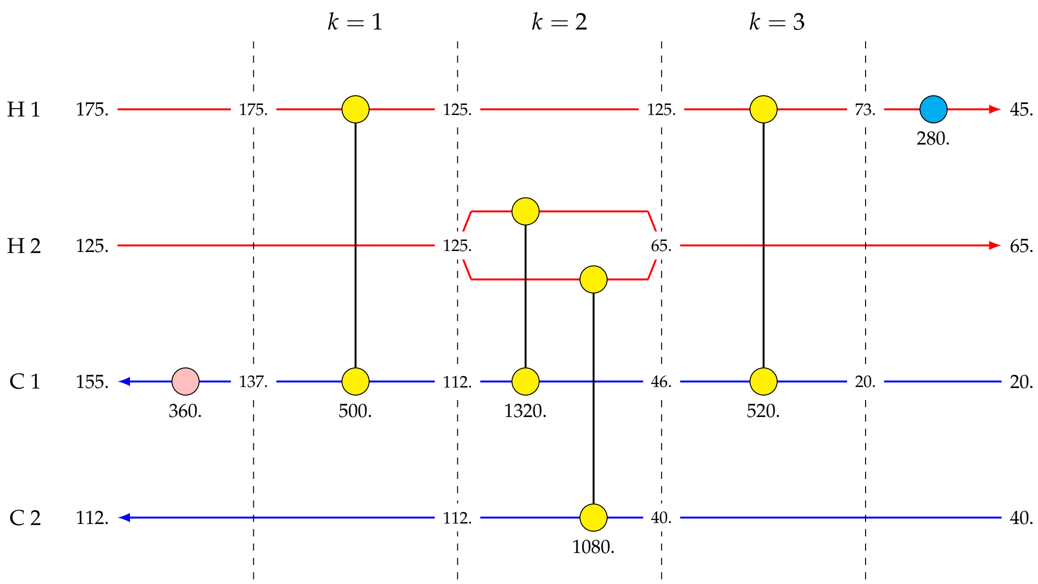
**Figure 4.19:** HEN topology obtained by restricting the cold stream C1 to not exchange heat in parallel.




### restriction on the HEN topology: the cold streams C1 and C2 cannot exchange heat in parallel

Now, we will restrict both cold streams C1 and C2 to not exchanging heat in parallel. The HEN obtained from this restriction is the same in terms of streams matched and heat flow rate as those shown in Figure 4.18. The only topological difference is that Stream C1 first exchanges heat with Stream H1 and secondly with Stream H2 rather than them both in parallel (Figure 4.20).

**Figure 4.20:** HEN topology obtained by restricting the cold streams C1 and C2 to not exchanging heat in parallel.



**restriction on the HEN topology: summary**

In summary, all of the HENs obtained here have six heat exchangers, which are: four process-to-process heat exchangers, one cooler and one heater. The cooler can be installed either on Stream H1 or Stream H2. As long as the minimum drive force for heat transfer is 13.0 °C, to reach the energy target of 360 kW, Stream H2 should exchange heat with at least two cold streams.

We went through four HEN topologies. The Topology T212 was obtained without applying any constraint on the mathematical model of the superstructure for HEN. The Topology T211 was obtained by restricting a single hot stream to not exchanging heat with the cold utility. To obtain the Topology T221, a single cold stream was restricted to not exchanging heat in parallel. The Topology T201 resulted from the restriction for all the cold streams to not exchanging heat in parallel. The HEN topologies and the restrictions that they resulted from are summarised in the tables 4.5 and 4.6, respectively.

**Table 4.5:** A summary of the HEN topologies obtained by minimising the number of heat exchangers for a hot utility consumption fixed to 360 kW and minimum driving force for heat transfer fixed to 13 °C.

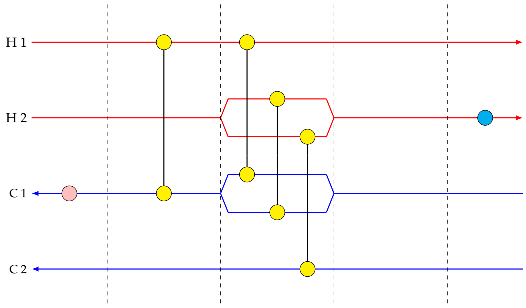
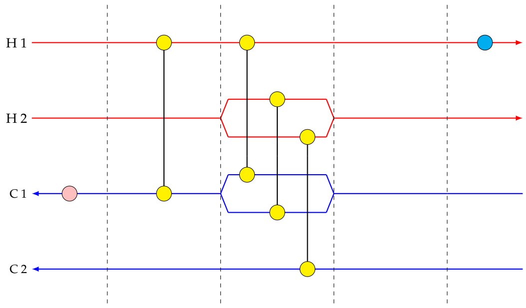
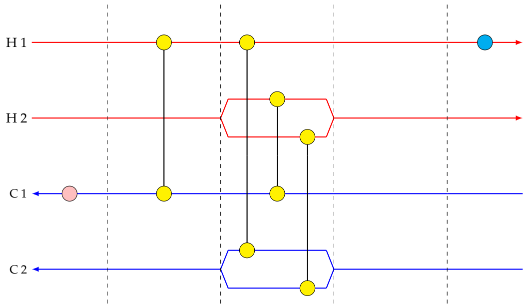
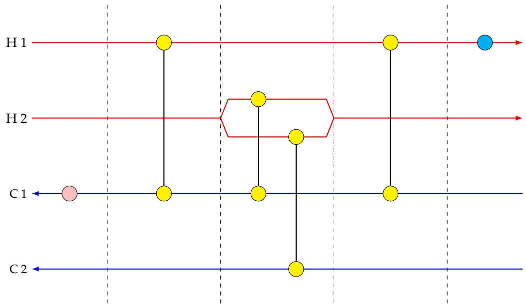
| Name | Topology |
|---|---|
| Topology T212 |  |
| Topology T211 |  |
| Topology T221 |  |
| Topology T201 |  |

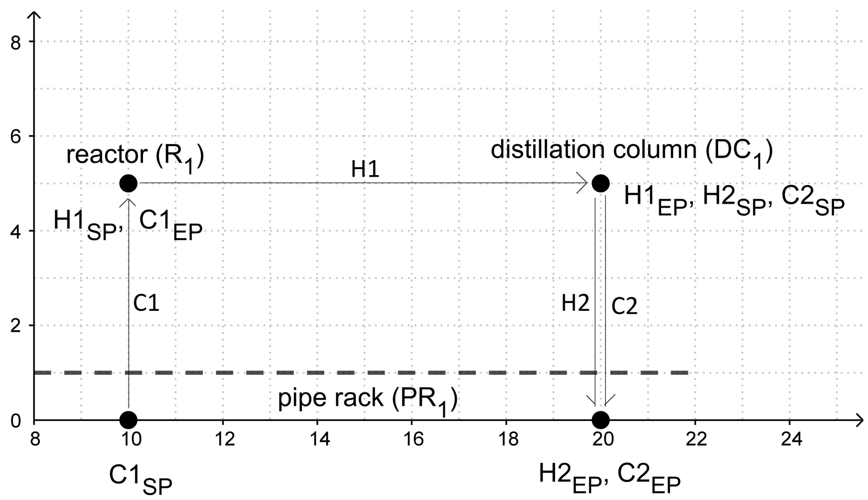
**Table 4.6:** The restrictions applied to obtain different HEN topologies by minimising the number of heat exchangers.

| name of the topology | restriction applied |
|---|---|
| Topology T212 | None. |
| Topology T211 | A specific hot stream is not allowed to exchange heat with the cold utility. |
| Topology T221 | A specific cold stream is not allowed to exchange heat in parallel. |
| Topology T201 | All the cold streams are not allowed to exchange heat in parallel. |

### 4.4.2 Simultaneous design of HEN and heat exchanger plant layout

Let us now assume a plant for the process introduced in this section. The layout of the plant is depicted in Figure 4.21. As shown in this figure, the reactor and distillation column are represented as points and the pipe rack as a dashed line. The coordinates of the starting (*SP*) and endpoints (*EP*) of the streams in the plant are shown in Table 4.7.

**Figure 4.21:** The plant of the process with the equipment and the starting and endpoints of the streams H1, H2, C1 and C2. The arrows illustrate a connection between the starting and endpoint of each stream.




A HEN will be designed for the process with not more than six heat exchangers. The location of heat exchangers in the plant will be calculated so as to minimise the total length of pipe required for the transport of the streams. The hot utility

**Table 4.7:** The coordinates of the starting and endpoints for the application of the superstructure for heat exchanger plant layout.

| stream | starting point ($SP$) | | | endpoint ($EP$) | | |
|---|---|---|---|---|---|---|
| | $x$ | $y$ | equipment | $x$ | $y$ | equipment |
| H1 | 10 | 5 | $\mathbf{R_1}$ | 20 | 5 | $\mathbf{DC_1}$ |
| H2 | 20 | 5 | $\mathbf{DC_1}$ | 20 | 0 | $\mathbf{PR_1}$ |
| C1 | 10 | 0 | $\mathbf{PR_1}$ | 10 | 5 | $\mathbf{R_1}$ |
| C2 | 20 | 5 | $\mathbf{DC_1}$ | 20 | 0 | $\mathbf{PR_1}$ |

consumption is fixed to 360 kW. The minimum driving force for heat transfer is 13 °C.

### Results and discussion

The HEN topology obtained by minimising the total length of pipe is the same as the previously obtained Topology T211, which is depicted again in Figure 4.22. The heat exchangers have their coordinates in the plant shown right above them in Figure 4.22. The coordinates of the heat exchangers are also shown in Table 4.8. The total length of pipe for the transport of the streams is 45 lu. To obtain HEN Topology T211 in the previous subsection, it was necessary to forbid the heat exchange between Stream H2 and the cold utility. This time, we obtained HEN Topology T211 without applying this restriction.

**Figure 4.22:** Topology T211 obtained by minimising the total length of pipe.

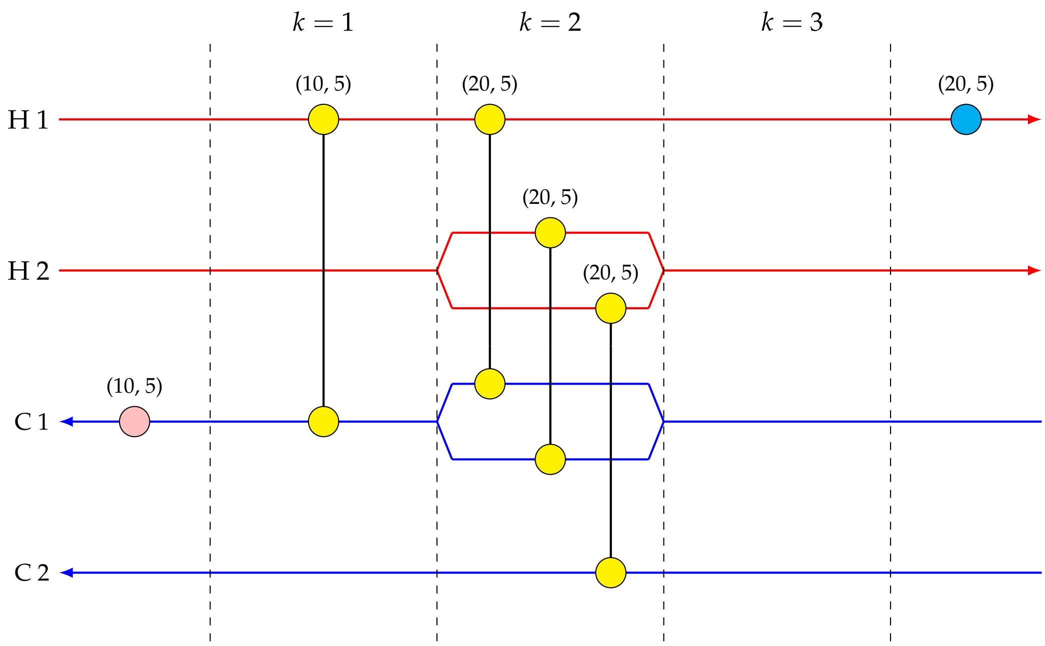


**Table 4.8:** The coordinates of the heat exchangers in the plant of the process obtained from Topology T211. The terms "CUT" and "HUT" stand for 'cold utility' and 'hot utility', respectively.

| heat exchanger | coordinate $x$ | coordinate $y$ |
|---|---|---|
| H1/C1 (stage $k = 1$) | 10.0 | 5.0 |
| H1/C1 (stage $k = 2$) | 20.0 | 5.0 |
| H2/C1 | 20.0 | 5.0 |
| H2/C2 | 20.0 | 5.0 |
| H1/CUT | 20.0 | 5.0 |
| HUT/C1 | 10.0 | 5.0 |

The location of heat exchangers in the plant can be analysed in terms of the length of additional pipe[1]. No additional pipe is required for the transport of both hot streams H1 and H2. For Stream C1 to be transported from the pipe rack, passing through the heat exchangers in parallel, to the reactor, it is required 20 lu of additional pipe. Stream C2 can be transported from the distillation column, passing through a heat exchanger, to the pipe rack without requiring additional pipe. The length of pipe for the entire path of the stream and the length of additional pipe are compared in Table 4.9.

**Table 4.9:** The length of pipe for the entire path of the stream and the length of additional pipe obtained from Topology T211.

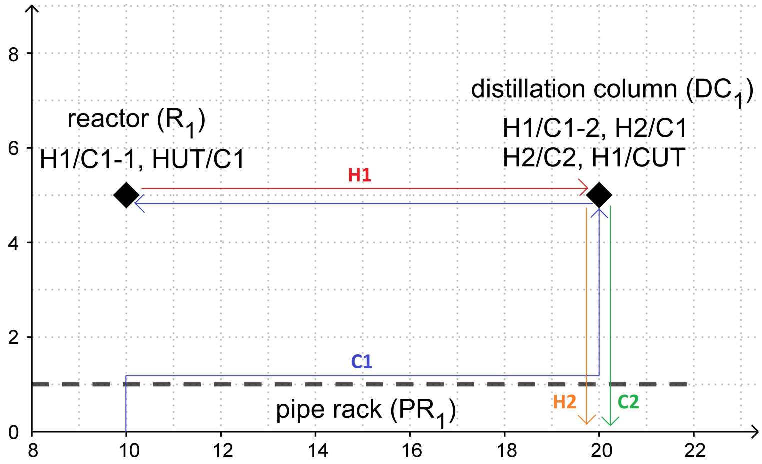
| stream | length of pipe (lu) | |
|---|---|---|
| | computed | additional |
| H1 | 10.0 | 0.0 |
| H2 | 5.0 | 0.0 |
| C1 | 25.0 | 20.0 |
| C2 | 5.0 | 0.0 |
| total | 45.0 | 20.0 |

The plant layout with the heat exchangers at their coordinates is depicted in Figure 4.23 by small black lozenges. This figure also shows possible pipe routes for the transport of the streams. The pipe routes were made by us and were not calculated from the presented mathematical models. The streams H1, H2 and C2 share the distillation column at the coordinates (20; 5) by means of their starting or endpoints. The length of the pipes that transport these streams was minimised by placing their heat exchangers at the same coordinates as the distillation column. The streams H2 and C1 do not pass through the same coordinates in the plant of

[1]The length of additional pipe was defined in Section 2.1.

the process. Thus, exchanging heat between them will always require additional pipe.

**Figure 4.23:** The plant of the process with the heat exchangers obtained from Topology T211.




The pipes that transport the streams H2 and C1 pass through heat exchangers in parallel. The length of these pipes is affected not location by the location of the heat exchangers but also the location of the splitter and mixer devices. The farther the heat exchangers are from the splitter and mixer devices, the longer are the pipes. The two heat exchangers in parallel installed in the pipe that transports Stream H2 are located at the same coordinates (20; 5) as the splitter and mixer devices. In this way, the splitter and mixer devices do not require additional pipe for the transport of Stream H2. The same occurs for Stream C1 with its heat exchangers, splitter and mixer devices.

### Restrictions on the HEN topology

In Subsection 4.4.1, we applied a number of restrictions presented in Table 4.6 to obtain different HEN topologies. We will apply those restrictions again to evaluate the change of pipe length with respect the HEN topology. In addition, we aim at knowing whether there will be HEN topologies different from those previously obtained in Table 4.5.

### Results and discussion

The HEN topologies obtained by applying the restrictions are the same as those previously obtained: Topology T212, Topology T201 and Topology T221. All of these HEN topologies lead to the same objective value, and therefore, require the same total length of pipe of 45 lu. Thus, obtaining Topology T211 (Figure 4.22) rather than Topology T212, Topology T201 and Topology T221 was just a matter of coincidence. These HEN topologies and the restrictions applied to obtain them will be discussed as this section continues.

### Restriction: Stream H1 cannot exchange heat with the cold utility

Topology T212 was obtained by forbidding Stream H1 to exchange heat with the cold utility (Figure 4.24). The coordinates of the heat exchangers in the plant are shown in Table 4.10. The total pipe length (45 lu) and length of pipe individually (Table 4.9) are the same as those obtained from HEN Topology T211.

**Figure 4.24:** Topology T212 obtained by minimising the total length of pipe and forbidding Stream H1 to exchange heat with the cold utility.

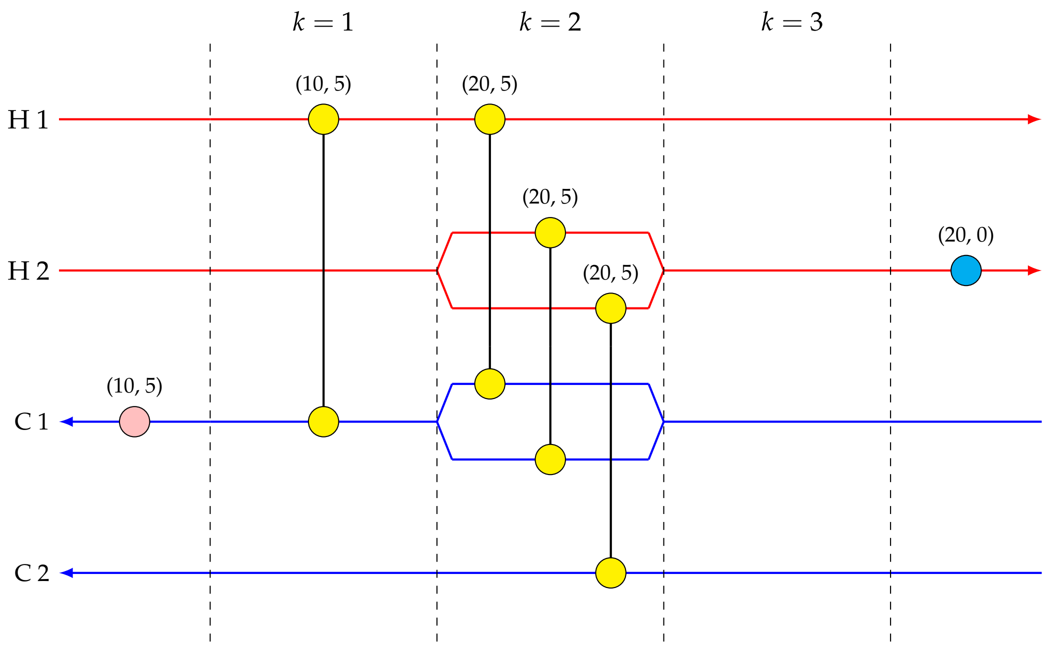



**Table 4.10:** The coordinates of the heat exchangers in the plant obtained from Topology T212.

| heat exchanger | coordinate $x$ | coordinate $y$ |
|---|---|---|
| H1/C1 (stage $k = 1$) | 10.0 | 5.0 |
| H1/C1 (stage $k = 2$) | 20.0 | 5.0 |
| H2/C1 | 20.0 | 5.0 |
| H2/C2 | 20.0 | 5.0 |
| H2/CUT | 20.0 | 0.0 |
| HUT/C1 | 10.0 | 5.0 |

This results was expected because the only difference between Topology T211 and Topology T212 is that the cooler is replaced from Stream H1 to Stream H2. The mathematical model developed by us does not optimise the length of pipe for the transport of utility streams. In this manner, as long as the coolers and heaters can be placed anywhere in the plant of the process, they will be placed so as to not increase the length of pipe for the transport of the process streams.

### Restriction: the cold streams C1 and C2 cannot exchange heat in parallel

Restricting the cold streams C1 and C2 to not exchanging heat in parallel gives Topology T201 (Figure 4.25). The coordinates of the heat exchangers in the plant of the process are shown in Table 4.11. For Stream C1 to be transported it is required additional pipe 20 lu long. The other streams can be transported without requiring additional pipe. The length of pipe for the entire path of the streams and the length of additional pipe are compared in Table 4.12.

**Figure 4.25:** Topology T201 obtained by minimising the total length of pipe and restricting the cold streams C1 and C2 to not exchanging heat in parallel.

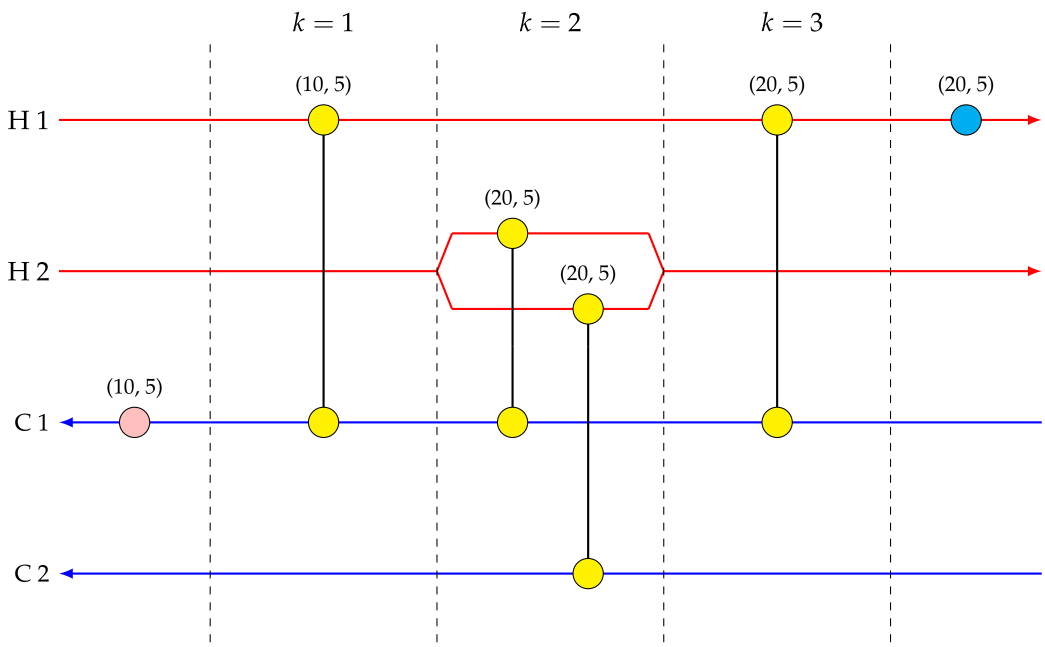



**Table 4.11:** The coordinates of the heat exchangers in the plant of the process obtained from Topology T201.

| heat exchanger | coordinate $x$ | $y$ |
|---|---|---|
| H1/C1 (stage $k=1$) | 20.0 | 5.0 |
| H1/C1 (stage $k=3$) | 20.0 | 5.0 |
| H2/C1 | 20.0 | 5.0 |
| H2/C2 | 20.0 | 5.0 |
| H1/CUT | 20.0 | 5.0 |
| HUT/C1 | 10.0 | 5.0 |

**Table 4.12:** The length of pipe for the entire path of the streams and the length of additional pipe obtained from Topology T201.

| stream | length of pipe (lu) computed | additional |
|---|---|---|
| H1 | 10.0 | 0.0 |
| H2 | 5.0 | 0.0 |
| C1 | 25.0 | 20.0 |
| C2 | 5.0 | 0.0 |
| total | 45.0 | 20.0 |

### Restriction: stream C1 cannot exchange heat in parallel

The restriction for Stream C1 to not exchange heat in parallel led to Topology T221. This is the only topology where Stream H1 exchanges heat with Stream C2 (Figure 4.26). The coordinates of the heat exchangers in the plant are shown in Table 4.13. The heat exchanger for the streams H2 and C1 is located at the same coordinates (10; 5) as the reactor. This location allows Stream C1 to be transported without requiring additional pipe whereas requires Stream H2 additional pipe 20 lu long. The length of pipe for the entire path of the streams and the length of additional pipe are shown in Table 4.14.

**Figure 4.26:** Topology T221 obtained by minimising the total length of pipe and restricting the cold stream C1 to not exchange heat in parallel.

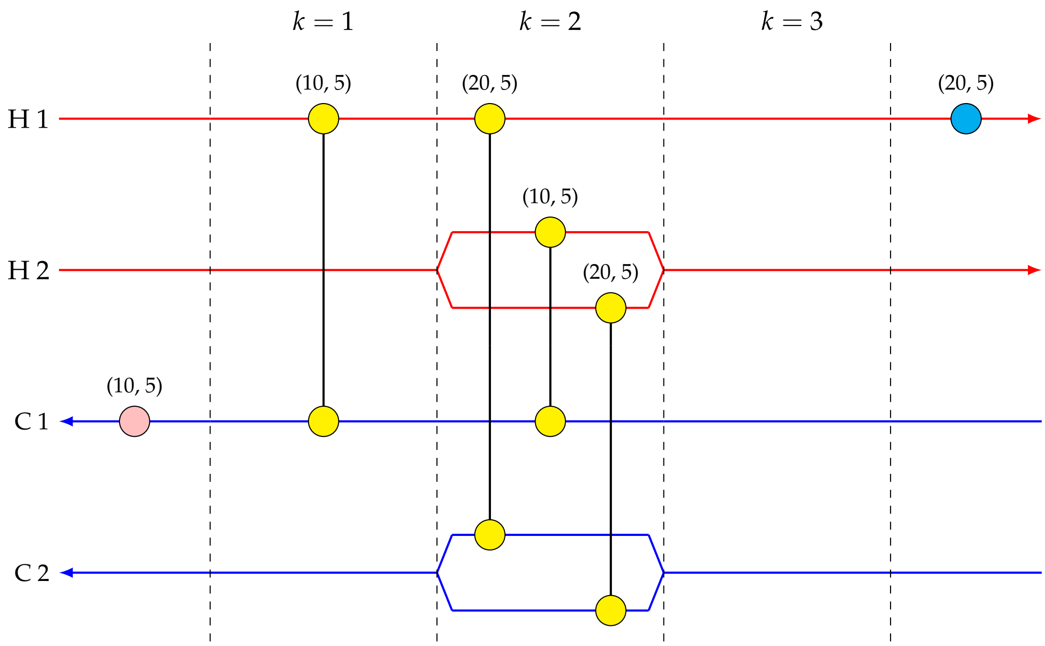


**Table 4.13:** The coordinates of the heat exchangers in the plant obtained from Topology T221.

| heat exchanger | coordinate $x$ | $y$ |
|---|---|---|
| H1/C1 | 10.0 | 5.0 |
| H1/C2 | 20.0 | 5.0 |
| H2/C1 | 10.0 | 5.0 |
| H2/C2 | 20.0 | 5.0 |
| H1/CUT | 20.0 | 5.0 |
| HUT/C1 | 10.0 | 5.0 |

**Table 4.14:** The length of pipe computed and the length of additional pipe obtained from Topology T221.

| stream | length of pipe (lu) computed | additional |
|---|---|---|
| H1 | 10.0 | 0.0 |
| H2 | 25.0 | 20.0 |
| C1 | 5.0 | 0.0 |
| C2 | 5.0 | 0.0 |
| total | 45.0 | 20.0 |

## 4.5 Heat exchanger placement in specific zones of the plant: a superstructure version

Thus far, we have presented the mathematical model of the superstructure for heat exchanger plant layout. The location of heat exchangers in the plant is calculated so as to minimise the total length of pipe required for the transport of the streams. There are not any constraints that restrict the number of locations where the heat exchangers can be placed.

Previous in this thesis (Section 3.4), we explained that, from the practical viewpoint, some locations may not be appropriate for heat exchanger placement. Then, we developed a mathematical model that allows the user to specify the zones of the plant where heat exchanger placement is allowed. At that time, our focus was on the **sequential** design of HEN and heat exchanger plant layout. In this section, we will adapt the mathematical model previously presented to the current situation: **simultaneous** heat exchanger network and plant layout.

#### Mathematical model: process-to-process heat exchangers

To represent a zone where the heat exchangers can be placed, we will continue using the index $z$ of the set Z. The physical size of the zones will continue to be specified by means of boundaries on the rectangular axes. The lower and upper boundaries of zone $z$ on the axis $a$ are represented by the parameters $low_{z,a}$ and $upp_{z,a}$, respectively.

To denote whether the heat exchanger for hot stream $i$ and cold stream $j$ at stage $k$ is placed in zone $z$, we will use the binary variable $\psi_{i,j,k,z}$, as follows:

$$\psi_{i,j,k,z} = \begin{cases} 1, & \text{if the heat exchanger for hot stream } i \text{ and cold stream} \\ & j \text{ at stage } k \text{ is placed in the zone } z; \\ 0, & \text{otherwise.} \end{cases} \tag{4.56}$$

$$i \in \mathrm{H}, \quad j \in \mathrm{C}, \quad k \in \mathrm{K}, \quad z \in \mathrm{Z}$$

If hot stream $i$ and cold stream $j$ exchange heat with each other at stage $k$ ($\sigma_{i,j,k} = 1$), then their heat exchanger has to be placed inside one of the specified zones:

$$\sum_{z \in \mathrm{Z}} \psi_{i,j,k,z} = \sigma_{i,j,k} \tag{4.57}$$

$$i \in \mathrm{H}, \quad j \in \mathrm{C}, \quad k \in \mathrm{K}$$

For the heat exchanger to be inside the zone it was specified to, its coordinate on the axis $a$ has to be encompassed within the boundaries of the zone:

$$ex_{i,j,k,a} \geq \sum_{z \in \mathrm{Z}} low_{z,a}\, \psi_{i,j,k,z} \tag{4.58}$$

$$ex_{i,j,k,a} \leq \sum_{z \in \mathrm{Z}} upp_{z,a}\, \psi_{i,j,k,z} \tag{4.59}$$

$$i \in \mathrm{H}, \quad j \in \mathrm{C}, \quad a \in \mathrm{A}$$

**Mathematical model: coolers and heaters**

The placement of coolers and heaters in the plant can also be restricted to specific zones. This can be done by replacing the variables in equations (4.56) to (4.59) as described as follows for the cooler of hot stream $i$:

- To denote whether hot stream $i$ exchanges heat in the cooler, the binary variable $\sigma_{i,j,k}$ should be replaced with the binary variable $\sigma\_td_i$.

- To denote whether the cooler is placed inside zone $z$, the binary variable $\psi_{i,j,k,z}$ should be replaced with $\psi\_td_{i,z}$.

- To limit the coordinate of the cooler to the boundaries of zone $z$, the variable $ex_{i,j,k,z}$ should be replaced with the variable $td_{i,a}$.

To define zones in the plant where the heater of cold stream $j$ can be placed, the procedure showed above can be used by replacing the index $i$ with the index $j$.

**Minimum and maximum number of heat exchangers placed inside a single zone**

We can specify a minimum and a maximum number of heat exchangers placed in a single zone by means of the parameters, $\lambda_z$ and $\kappa_z$, respectively:

$$\sum_{i \in \mathrm{H}} \sum_{j \in \mathrm{C}} \sum_{k \in \mathrm{K}} \psi_{i,j,k,z} + \sum_{i \in \mathrm{H}} \psi\_td_{i,z} + \sum_{j \in \mathrm{C}} \psi\_td_{j,z} \geq \lambda_z \tag{4.60}$$

$$\sum_{i \in \mathrm{H}} \sum_{j \in \mathrm{C}} \sum_{k \in \mathrm{K}} \psi_{i,j,k,z} + \sum_{i \in \mathrm{H}} \psi\_td_{i,z} + \sum_{j \in \mathrm{C}} \psi\_td_{j,z} \leq \kappa_z \tag{4.61}$$

$$z \in \mathrm{Z}$$

## 4.6 Example: the use of zones for simultaneous design of heat exchanger network and plant layout

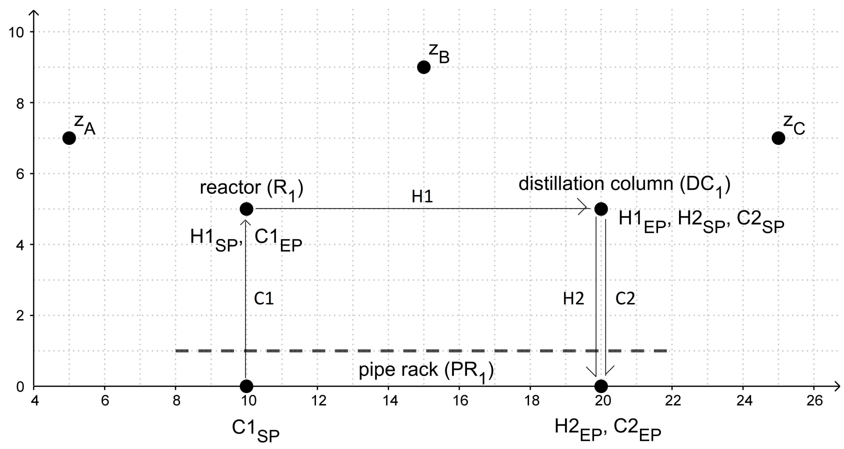
In this section, we will restrict the number of locations where the heat exchangers can be placed in the plant. In order to do that, the HEN and heat exchanger plant layout will be designed in a simultaneous manner for the problem stated in Section 4.4. The objective function will be defined as the minimisation of the total length of pipe for the transport of the streams.

To design the HEN, the hot utility consumption is fixed to 360 kW. The minimum driving force for heat transfer is 13 °C. The HEN is restricted to having a maximum of six heat exchangers. To design the heat exchanger plant layout, the heat exchangers will be restricted to being placed only in the zones depicted in Figure 4.27. The coordinates of these zones are shown in Table 4.15.

**Table 4.15:** The coordinates of the zones where it is allowed to place the heat exchangers.

| zone | coordinate $x$ | coordinate $y$ |
|---|---|---|
| A | 5.0 | 7.0 |
| B | 15.0 | 9.0 |
| C | 25.0 | 7.0 |

**Figure 4.27:** The layout of the plant with the zones $z_A$, $z_B$ and $z_C$ where the heat exchangers can be placed. The arrows are used for helping the reader to identify the starting and endpoints of the streams.




Four cases will be studied. In Case $\mathrm{ZR}_1$, there will be no restriction on the maximum number of heat exchangers per zone. In the cases $\mathrm{ZR}_2$, $\mathrm{ZR}_3$ and $\mathrm{ZR}_4$ a maximum of two heat exchangers will be allowed per zone. Additional restrictions will be applied to the cases $\mathrm{ZR}_3$ and $\mathrm{ZR}_4$. In Case $\mathrm{ZR}_3$, the coolers and heaters can only be placed in Zone $z_B$. In Case $\mathrm{ZR}_4$, the streams H2 and C1 are allowed to exchange heat only in Zone $z_B$. A summarised description of the cases is shown in Table 4.16.

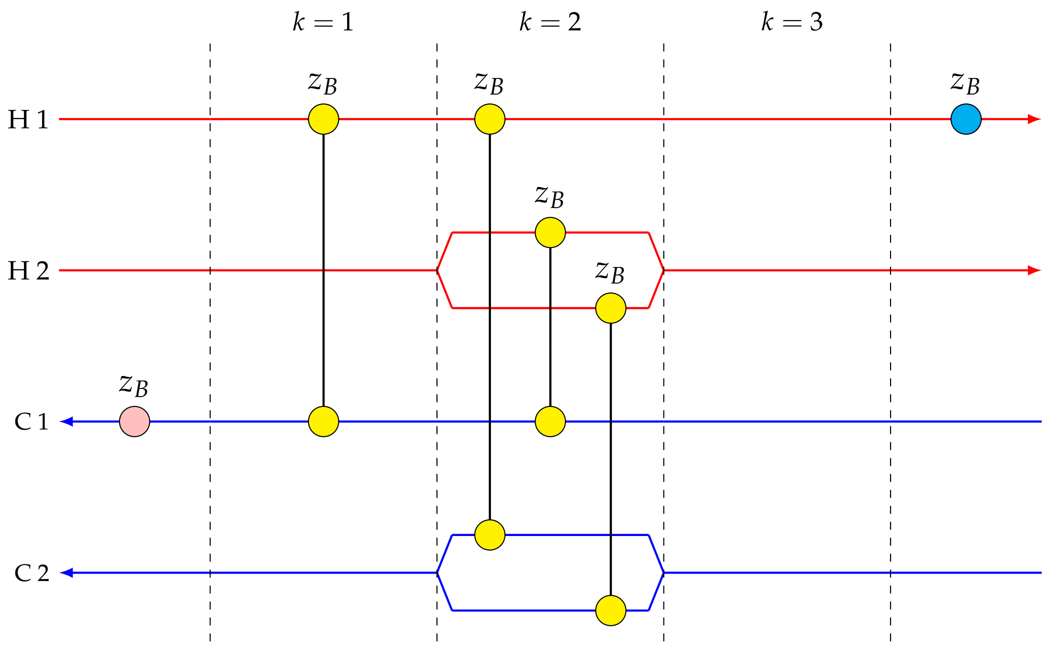
**Table 4.16:** The cases for heat exchanger plant layout design.

| case | description |
|---|---|
| $\mathrm{ZR}_1$ | There is no restriction on the maximum number of heat exchangers per zone. |
| $\mathrm{ZR}_2$ | A maximum of two heat exchangers can be placed inside a single zone of the process plant. |
| $\mathrm{ZR}_3$ | In addition to $\mathrm{ZR}_1$, coolers and heaters are restricted to being placed in $z_B$. |
| $\mathrm{ZR}_4$ | In addition to $\mathrm{ZR}_1$, any heat exchanger for the streams H2 and C1 is restricted to not being placed in $z_B$. |

### Results and discussion: Case $\mathrm{ZR}_1$

The topology of the HEN designed in Case $\mathrm{ZR}_1$ is the same as that (Topology T221) obtained in previous part of this example. Topology T221 is depicted here again in Figure 4.28. In the plant, all heat exchangers are located in Zone $z_B$ at the coordinates (15, 9). The total length of pipe for the transport of the streams is 87 lu.

**Figure 4.28:** Topology T221 obtained for Case $ZR_1$. The name of the zone where a heat exchanger is placed is shown above the heat exchanger.




### Results and discussion: the cases $ZR_2$, $ZR_3$ and $ZR_4$

The HEN obtained in the cases $ZR_2$, $ZR_3$ and $ZR_4$ has the same topology as those obtained in the previous part of this example. The topology of the HEN obtained in Case $ZR_2$ is Topology T212. As shown in Figure 4.29, Stream C1 exchanges heat in parallel with the hot streams H1 and H2 in Zone $z_B$. Then, Stream C1 exchanges heat in series with Stream H1 and the hot utility in Zone $z_A$, respectively.

Stream H2 exchanges heat in parallel with the cold streams C1 and C2 in different zones. The heat exchanger for the streams H2 and C2 is located in Zone $z_C$. Stream H2 also exchanges heat with the cold utility in Zone $z_C$. The layout of the plant with the heat exchangers placed in the zones is depicted in Figure 4.30. The total length of pipe is 118 lu. In comparison with Case $ZR_1$, the total length of pipe increased by 31 lu.

**Figure 4.29:** Topology T212 obtained in Case $ZR_2$.

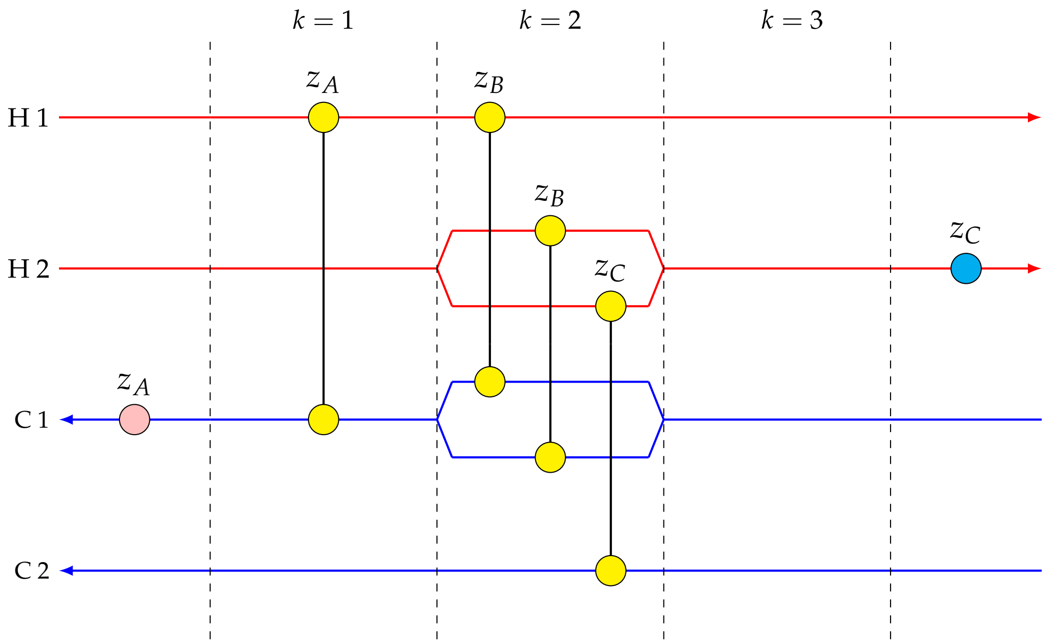
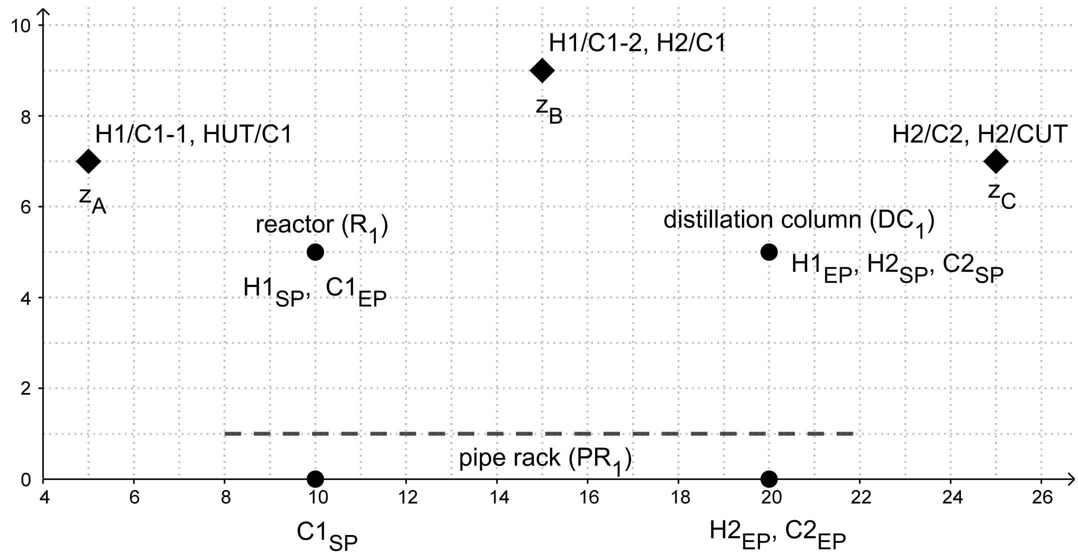
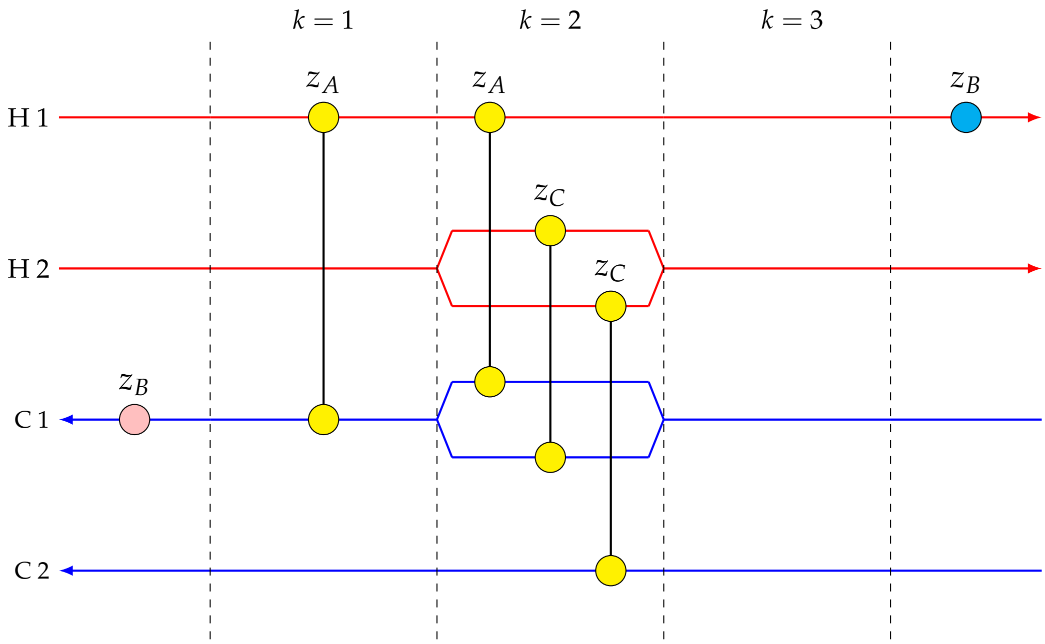



**Figure 4.30:** Case $ZR_1$. The layout of the plant with the heat exchangers placed in the zones.



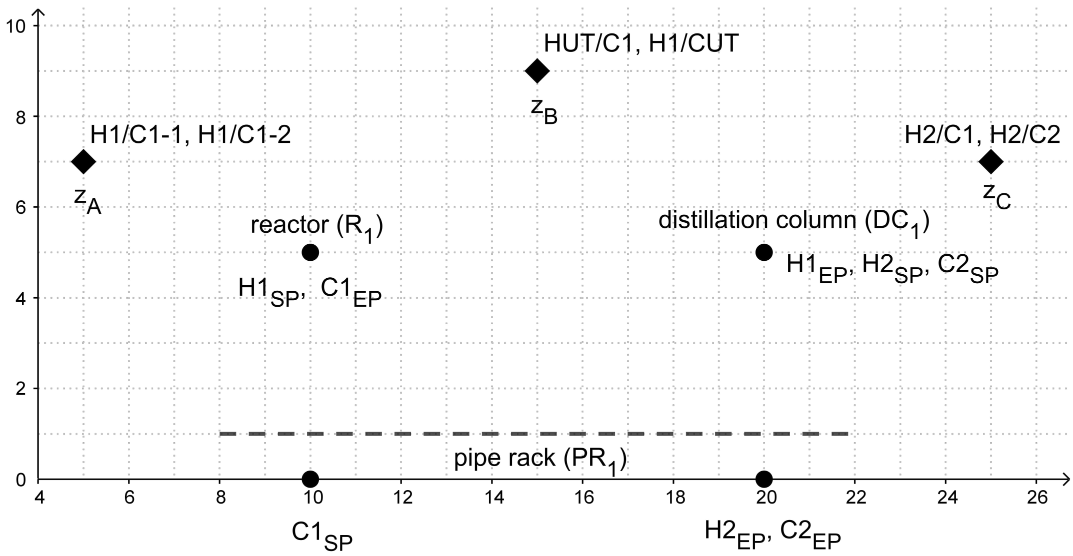

As shown in Figure 4.30, the heater (HUT/C1) and the cooler (H2/CUT) are located in the zones $z_A$ and $z_C$, respectively. In Case $ZR_3$, we evaluated how the HEN topology and the total length of pipe change by restricting the coolers and heaters to being placed only in zone $z_B$. The HEN designed in Case $ZR_3$ has Topology T211 (Figure 4.31).

**Figure 4.31:** Topology T211 obtained in Case $ZR_3$.




As shown in Figure 4.31, Stream C1 exchanges heat in parallel with the hot streams H1 and H2 in the zones $z_A$ and $z_C$, respectively. Then, Stream C1 exchanges heat one more time with Stream H1 in zone $z_A$ and is transported to Zone $z_B$, where it exchanges heat with the hot utility. Stream H2 exchanges heat in parallel with the cold streams C1 and C2 in Zone $z_C$. The layout of the plant is depicted in Figure 4.32. The total length of pipe is 134 lu. The total length of pipe required in Case $ZR_3$ is 47 lu longer than that required in Case $ZR_1$.

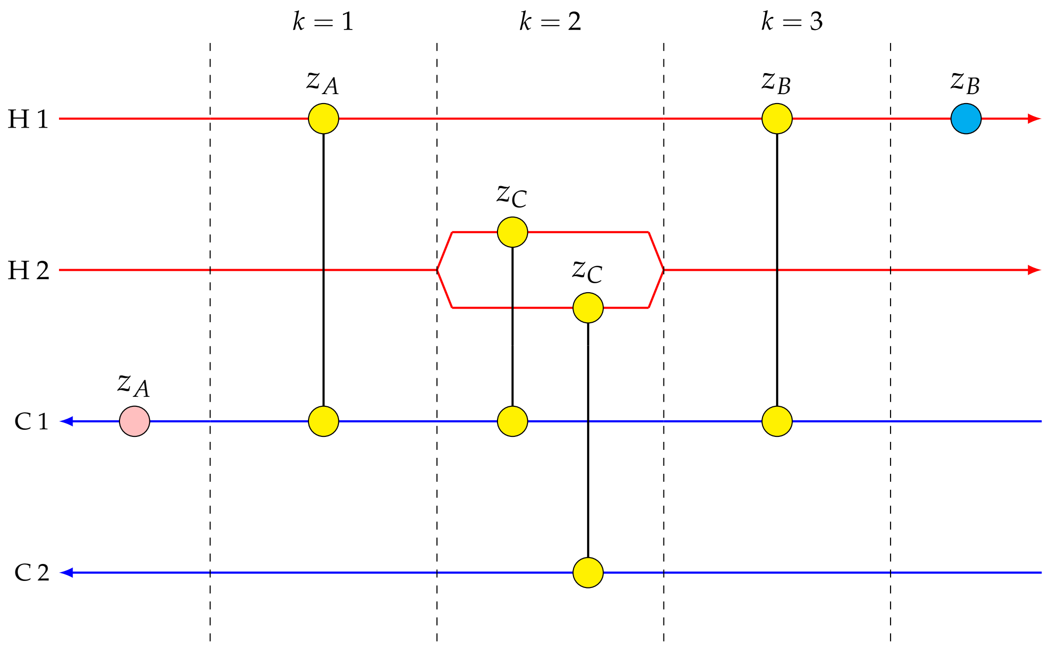
**Figure 4.32:** Case $ZR_3$. The layout of the plant of with heat exchangers placed in the zones.




In the previous part of this example, the layout of the plant Figure 4.21 showed that Stream C1 enters the reactor whereas Stream H2 leaves the distillation column. Besides that, the streams C1 and H2 are transported through different parts of the pipe rack. These streams have no place of the plant in common. As a consequence, exchanging heat between the streams C1 and H2 always requires additional pipe.

The best option to minimise the length of additional pipe is to place the heat exchanger in Zone $z_B$. This can be confirmed in the layout of the plant shown in the cases $\mathrm{ZR}_1$, $\mathrm{ZR}_2$ and $\mathrm{ZR}_3$. In Case $\mathrm{ZR}_4$, a situation was considered that restricts the heat exchanger for the streams H2 and C1 to not being placed in Zone $z_B$. The designed HEN has Topology T201 (Figure 4.33).

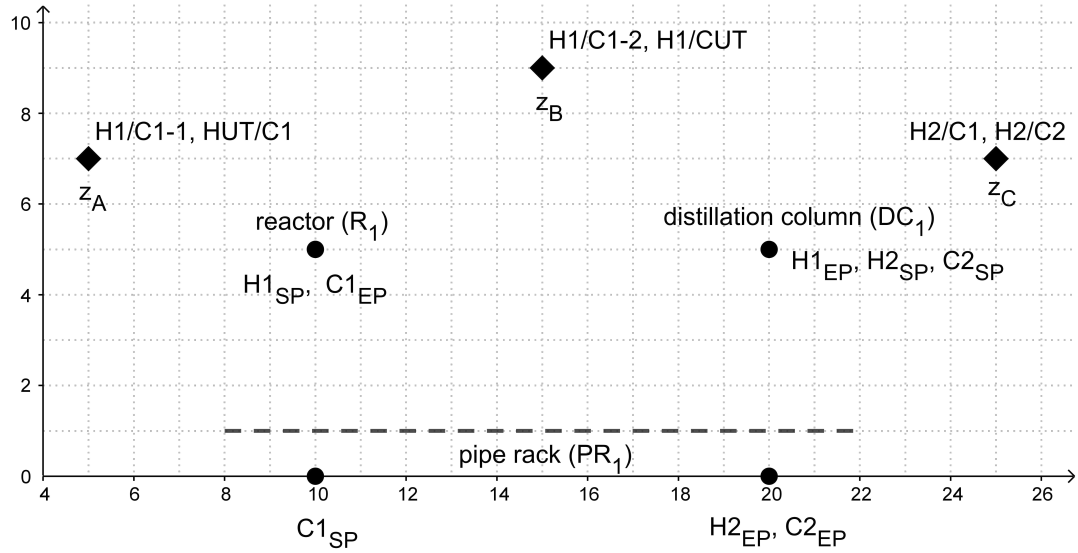
**Figure 4.33:** Topology T201 obtained in Case $\mathrm{ZR}_3$.




As shown in Figure 4.33, the heat exchanger for the streams H2 and C1 is located in Zone $z_C$. Stream C1 firstly exchanges heat with Stream H1 in Zone $z_B$. Then, Stream C1 is transported to Zone $z_C$ where it exchanges heat with Stream H2. Lastly, Stream C1 exchanges heat with Stream H1 again and with the hot utility in Zone $z_A$, respectively.

Stream H2 exchanges heat in parallel with the two cold streams C1 and C2 in the same zone. The layout of the plant with the heat exchangers in the zones is depicted in Figure 4.34. Compared to the Case $\mathrm{ZR}_1$, the total length of pipe increased from 87 lu to 119 lu.

**Figure 4.34:** Case $ZR_3$. The layout of the plant with the heat exchanger placed in the zones.




### summary: placing the heat exchangers in zones

In summary, to design the HEN and heat exchanger plant layout, we restricted the heat exchangers to being placed in the zones $z_A$, $z_B$ and $z_C$. Four case were studied, which yielded four topologies for the HEN: Topology T221, Topology T212, Topology T211 and Topology T201. These HEN topologies were also obtained in the previous part of this example (Section 4.4). There is a tendency for the solver to place all the heat exchangers in Zone $z_B$. We tackled this tendency by restricting a maximum of two heat exchangers per zone. In addition, we restricted the location of specific heat exchangers with respect the zones they could be placed.

The total length of pipe varied with the heat exchanger plant layout. In Case $ZR_1$, all the heat exchangers were placed in Zone $z_B$, the total length of pipe was 87 lu. In Case $ZR_2$, by restricting a maximum of two heat exchangers per zone, the total length of pipe increased to 118 lu. The longest total length of pipe was required in Case $ZR_3$. In this case, we restricted the heater and the cooler to being located in Zone $z_B$, the total length of pipe was 134 lu. In Case $ZR_4$, by restricting the heat exchanger for the streams H2 and C1 to not being placed in Zone $z_B$, the total length of pipe was 119 lu. The zones where the heat exchangers were placed, HEN topologies and total length of pipe obtained in the cases $ZR_1$ to $ZR_4$ are summarised in Table 4.17.

**Table 4.17:** The HEN topology, heat exchanger per zone and total length of pipe obtained in the cases $ZR_1$ to $ZR_4$.

| case restriction | topology | zone $z_A$ | zone $z_B$ | zone $z_C$ | total length of pipe (lu) |
|---|---|---|---|---|---|
| $ZR_1$ | 221 | none | all | none | 87.0 |
| $ZR_2$ | 212 | H1/C1-1, HUT/C1 | H1/C1-2, H2/C1 | H2/C2, H2/CUT | 118.0 |
| $ZR_3$ | 211 | H1/C1-1, H1/C1-2 | H1/CUT, HUT/C1 | H2/C1, H2/C2 | 134.0 |
| $ZR_4$ | 201 | H1/C1-1, HUT/C1 | H1/C1-2, H1/CUT | H2/C1, H2/C2 | 119.0 |

## 4.7 Conclusion

In this chapter, a method was developed to design the HEN and heat exchanger plant layout simultaneously. The concept of superstructure is the basis of this method. The HEN is designed through the superstructure proposed by Yee and Grossmann (1990). The superstructure for heat exchanger plant layout was completely developed by us. Both superstructures were formulated as mathematical optimisation models. The objective of the optimisation is to minimise the total length of pipe for transporting the streams.

The superstructure for HEN consists of a mixed-integer linear programming model. The continuous variables are the heat flow rate, temperature of the streams and utility consumption. An upper boundary can be set on the utility consumption. The integer variables denote whether two streams exchange heat with each other. The objective function is defined as the minimisation of the number of heat exchangers. The superstructure was used to design the HEN of the bio-ethanol-to-gasoline conversion process (SERTH; LESTINA, 2014).

The superstructure for heat exchanger plant layout was firstly presented in a conceptual manner. The concept is that the length of pipe is calculated with the heat exchangers at all the possible coordinates. The optimal set of coordinates is the one that minimises the total length of pipe. In the mathematical model, the same integer variables of the superstructure for HEN are used. This characteristic is the key to the simultaneous design.

To implement the mathematical models, the General Algebraic Modelling System (GAMS) was used. The HEN and heat exchanger plant layout were simultaneously designed for a four-stream problem (SHENOY, 1995). The mathematical models were solved to optimality in less than one second. It was an expected result because of the small number of streams and simplicity of the plant layout.

Minimising the length of pipe drives the design towards lower costs of pipe and pumping. However, the true minimisation of the costs of pipe and pumping can only be achieved if they are explicitly included in the objective function. Besides that, the length of pipe is not consistent in terms of units with other relevant costs, such as costs for utility consumption and heat exchangers. As a consequence, the length of pipe cannot be summed together with these kinds of costs in the objective function. The purpose of the next chapter is to extend the present method to allow the minimisation of the costs of pipe and pumping.

# Chapter 5

# Simultaneous design of HEN and heat exchanger plant layout: costs of pipe and pumping

A traditional strategy for the design of HENs is to minimise the costs of utilities and heat exchangers. In the previous chapter, we showed that these costs can be indirectly minimised by setting upper boundaries on the utility consumption and number of heat exchangers. In the current chapter, we will present the equations to calculate the costs of utilities and heat exchangers. Then, we will demonstrate the direct minimisation of the costs of utilities and heat exchangers in the design of HENs.

The traditional strategy for the design of HENs does not include the cost of pumping to overcome the energy loss in pipes and heat exchangers. The cost of pipes to transport the streams is not included either. In doing so, the costs of pipe and pumping cannot be balanced against the costs of utilities and heat exchangers. The consequence is that the impact of the costs of pipe and pumping on the HEN topology cannot be evaluated.

In the current chapter, we will develop the mathematical model to calculate the costs of pipe and pumping. This mathematical model will work as an extension of that for designing the heat exchanger plant layout, which was developed in the previous chapter. We will demonstrate that by designing the HEN and heat exchanger plant layout simultaneously, the costs of pipe and pumping are balanced against the costs of utilities and heat exchangers. Thus, we are able to evaluate the impact of the costs of pipe and pumping on the HEN topology.

## 5.1 Direct approach for minimum utility and heat exchanger cost

The theme of this section is the design of HENs. The mathematical model of the superstructure for HEN allows the minimisation of costs of utilities and heat exchangers. These costs can be minimised in an indirect or direct manner. The current section presents an approach for the **direct** minimisation of the costs of utilities and heat exchangers. An indirect approach was presented in the previous chapter, Subsection 4.1.3.

In the **direct** approach, the definitive HEN results from the optimal trade-off from the costs of utilities and heat exchangers. A non-optimal trade-off may be obtained if the solver is trapped in a local optimum or the solution of the mathematical model is infeasible. The cost of utilities can be calculated from the utility consumption ($\dot{q}\_cooler_i$ and $\dot{q}\_heater_j$) and by assuming an annual price per unit of power:

$$ac\_cold\text{-}utility = pcu \sum_{i \in \mathrm{H}} \dot{q}\_cooler_i \tag{5.1}$$

$$ac\_hot\text{-}utility = phu \sum_{j \in \mathrm{C}} \dot{q}\_heater_j \tag{5.2}$$

where the parameters $phu$ and $pcu$ are the annual prices per unit of power of the hot utility and cold utility, respectively (ULRICH; VASUDEVAN, 2006).

The cost of heat exchangers will be calculated from their heat transfer areas plus a fixed charge. Throughout the design of the superstructure, hot stream $i$ and cold stream $j$ may or may not match for heat exchange at stage $k$. If they do ($\sigma_{i,j,k} = 1$), then the cost of the heat exchanger will be calculated, otherwise ($\sigma_{i,j,k}$) its cost will be equal to zero:

$$C\_heat\text{-}exchanger_{i,j,k} = \mathbf{a}\ \sigma_{i,j,k} + \mathbf{b}\ \left(A_{i,j,k}\right)^{\mathbf{c}} \tag{5.3}$$

$$i \in \mathrm{H}, \quad j \in \mathrm{C}, \quad k \in \mathrm{K}$$

where **a**, **b** and **c** are parameters related to the type, material and pressure rating of the heat exchanger (SINNOT; TOWLER, 2020; SMITH, 2005).

Previously in the first chapter of this thesis (Subsection 1.4.5), we assumed that all the heat exchangers operate with pure countercurrent flow. This assumption allows us to calculate the heat transfer area between hot stream $i$ and cold stream $j$ at stage $k$ ($A_{i,j,k}$) using the log mean temperature difference:

$$A_{i,j,k} = \frac{\dot{q}_{i,j,k}}{U_{i,j}\ LMTD_{i,j,k}} \tag{5.4}$$

$$i \in \mathrm{H}, \quad j \in \mathrm{C}, \quad k \in \mathrm{K}$$

where the overall heat transfer coefficient ($U_{i,j}$) will be calculated by:

$$\frac{1}{U_{i,j}} = \frac{1}{h_i} + \frac{1}{h_j} \tag{5.5}$$

$$i \in \mathrm{H}, \quad j \in \mathrm{C}$$

The log mean temperature difference ($LMTD_{i,j,k}$) can be calculated from the temperature locations. The temperature locations $k$ and $k+1$ represent the temperatures on the left and right ends of the heat exchanger, respectively. To denote the difference between the temperatures of hot stream $i$ and cold stream $j$ at the temperature location $k$, we will define the variable $dt_{i,j,k}$.

The log mean temperature difference in its standard form is undefined when the temperature difference on the left and right ends of the heat exchanger have the same value. To avoid this problem, we will calculate the log mean temperature difference by means of the Paterson (1984) approximation:

$$LMTD_{i,j,k} = \frac{2}{3}\sqrt{dt_{i,j,k}\, dt_{i,j,k+1}} + \frac{dt_{i,j,k} + dt_{i,j,k+1}}{6} \tag{5.6}$$

$$i \in \mathrm{H}, \quad j \in \mathrm{C}, \quad k \in \mathrm{K}$$

The greater the temperature difference is, the smaller the heat transfer area is. To minimise the heat transfer area and, therefore the cost of heat exchangers, the solver will attempt to make the temperature difference as great as possible. In this way, we can calculate difference between the temperatures of hot stream $i$ and cold stream $j$ on the left and right ends of the heat exchanger by using an inequality of the form *less than or equal to*:

$$dt_{i,j,k} \leq t_{i,k} - t_{j,k} + \Gamma\,\left(1 - \sigma_{i,j,k}\right) \tag{5.7}$$

$$dt_{i,j,k+1} \leq t_{i,k+1} - t_{j,k+1} + \Gamma\,\left(1 - \sigma_{i,j,k}\right) \tag{5.8}$$

$$i \in \mathrm{H}, \quad j \in \mathrm{C}, \quad k \in \mathrm{K}$$

The right-hand side of the inequalities (5.7) and (5.8) is dependent on the match between hot stream $i$ and cold stream $j$ at stage $k$. If these streams are matched for heat exchange ($\sigma_{i,j,k} = 1$), then the term $\Gamma\,\left(1 - \sigma_{i,j,k}\right)$ is cancelled out. In this way, the right-hand side of the inequalities becomes the actual difference between the temperatures of the streams. If hot stream $i$ and cold stream $j$ are not matched for heat exchange at stage $k$ ($\sigma_{i,j,k} = 0$), then the large constant $\Gamma$ controls the right-hand side of the inequalities. As a result, the variable on the left-hand side does not affect the temperature of the streams.

**The cost of coolers and heaters**

To calculate the cost of coolers and heaters, a formulation analogous to that presented in the equations (5.3) to (5.8) can be used. The inlet and outlet temperatures of the cold utility stream will be represented by the parameters $T^{in}_{cold\text{-}utility}$ and $T^{out}_{cold\text{-}utility}$, respectively. The cost of a cooler for hot stream $i$ will be calculated by:

$$C\_cooler_i = \mathbf{a}\ \sigma\_td_i + \mathbf{b}\ (A\_cooler_i)^{\mathbf{c}} \qquad i \in \mathrm{H} \tag{5.9}$$

$$A\_cooler_i = \frac{\dot{q}\_cooler_i}{U\_cooler_i\ LMTD\_cooler_i} \qquad i \in \mathrm{H} \tag{5.10}$$

$$U\_cooler_i = \frac{1}{h_i} + \frac{1}{h_{cold\text{-}utility}} \qquad i \in \mathrm{H} \tag{5.11}$$

$$LMTD\_cooler_i = \frac{2}{3}\sqrt{dtcu_{i,k=n_k+1}\left(T\_target_i - T^{in}_{cold\text{-}utility}\right)} + \frac{dtcu_{i,k=n_k+1} + \left(T\_target_i - T^{in}_{cold\text{-}utility}\right)}{6} \qquad i \in \mathrm{H}, \quad k \in \mathrm{K} \tag{5.12}$$

$$dtcu_{i,k=n_k+1} \leq t_{i,k=n_k+1} - T^{out}_{cold\text{-}utility} + \Gamma\ (1 - \sigma\_td_i) \qquad i \in \mathrm{H}, \quad k \in \mathrm{K} \tag{5.13}$$

To calculate the cost of a heater, the inlet and outlet temperatures of the hot utility stream will be represented by the parameters $T^{in}_{hot\text{-}utility}$ and $T^{out}_{hot\text{-}utility}$, respectively. The cost of a heater for cold stream $j$ will be calculated by:

$$C\_heater_j = \mathbf{a}\ \sigma\_td_j + \mathbf{b}\ \left(A\_heater_j\right)^{\mathbf{c}} \qquad j \in \mathrm{C} \tag{5.14}$$

$$A\_heater_j = \frac{\dot{q}\_heater_j}{U\_heater_j\ LMTD\_heater_j} \qquad j \in \mathrm{C} \tag{5.15}$$

$$U\_heater_j = \frac{1}{h_j} + \frac{1}{h_{hot\text{-}utility}} \tag{5.16}$$
$$j \in \mathrm{C}$$

$$LMTD\_heater_j = \frac{2}{3}\sqrt{\left(T^{in}_{hot\text{-}utility} - T\_target_j\right)\ dthu_{j,k=1}} + \frac{\left(T^{out}_{hot\text{-}utility} - T\_target_j\right) + dthu_{j,k=1}}{6} \tag{5.17}$$
$$j \in \mathrm{C}, \quad k \in \mathrm{K}$$

$$dthu_{j,k=1} \leq T^{out}_{hot\text{-}utility} - t_{j,k=1} + \Gamma\ \left(1 - \sigma\_td_j\right) \tag{5.18}$$
$$j \in \mathrm{C}, \quad k \in \mathrm{K}$$

The cost of heat exchangers the cost of coolers and the cost of heaters can be expressed on an annual basis. The reader is encouraged to refer to other sources to learn about cost annualisation, such as Sinnot and Towler (2020), Chapter 6, Section 6.7.6. In this thesis, to express the cost of heat exchangers, coolers and heaters on an annual basis, they will be multiplied by the constant $(af)$.

The objective function can be defined as the minimisation of the annual costs of utility consumption, heat exchangers, coolers and heater:

$$\begin{aligned}\min pcu \sum_{i\in \mathrm{H}} \dot{q}\_cooler_i + phu \sum_{j\in \mathrm{C}} \dot{q}\_heater_j \\ + af \sum_{i\in \mathrm{H}} \sum_{j\in \mathrm{C}} \sum_{k\in \mathrm{K}} C\_heat\text{-}exchanger_{i,j,k} \\ + af \sum_{i\in \mathrm{H}} C\_cooler_i + af \sum_{j\in \mathrm{C}} C\_heater_j\end{aligned} \tag{5.19}$$

## 5.2 Example: minimisation of the costs of utilities and heat exchangers

In this section, a HEN will be designed for the bio-ethanol-to-gasoline conversion process. This process was introduced in the previous chapter, Section 4.2. To design the HEN, the mathematical model of the superstructure for HEN will be solved (Section 4.1). The number of stages of the superstructure will be equal to five. The objective function will be defined as the minimisation of the costs of utilities and heat exchangers. These costs will be calculated by solving the mathematical model developed in Section 5.1. The mathematical models of the superstructure for HEN and for the costs of utilities and heat exchangers will be solved simultaneously.

The parameters to calculate the costs of utilities and heat exchangers will be taken from Souza et al. (2016) and are shown in Table 5.1. To calculate the heat transfer area, we will assume that the heat transfer coefficient of all the streams is $200\,\mathrm{W\,m^{-2}\,K}$. The computations will be carried out using the General Algebraic Modelling System (GAMS) version 40.4.0 with the solver DICOPT.

**Table 5.1:** The data to estimate the annual costs of utilities and heat exchangers.

| constant | value | description |
|---|---|---|
| $pcu$ | 10 | annual price per unit of power ($/kW year) |
| $phu$ | 120 | |
| $af$ | 0.264 | annualisation factor |
| **a** | $30 \times 10^3$ | cost law parameters |
| **b** | 750 | |
| **c** | 0.8 | |

To facilitate the search for the optimal solution, we will reduce the size of the feasible region by setting limits on the parameters of the problem (WILLIAMS, 2013). The mathematical models will be solved over a range of upper limits on the hot utility consumption. To start the solving process, we will define the upper limit on the hot utility consumption as equal to the sum of the heat load of the cold streams.

The sum of the heat load of the cold streams C1 (682.4 kW), C2 (1170.7 kW) and C3 (74.7 kW) is equal to 1927.8 kW. Next, we will define the upper limit on the hot utility consumption based on the first solution obtained for the mathematical models. This procedure will be repeated to define the other next upper limits on the hot utility consumption. In doing so, we will be decreasing the upper limit on the hot utility consumption and solving the mathematical models successively. We will stop solving the mathematical models when the costs of utilities and heat exchangers does not decrease anymore.

Based on previous experiments, we discovered that the solution of the mathematical models is strongly sensible to the minimum driving force for heat transfer. The smaller minimum driving force for heat transfer is, the easier is for the solver to find a feasible solution. In the solution found, the temperature difference in the heat exchangers does not necessarily is equal to the minimum driving force for heat transfer.

For example, by setting a value of 0.1 °C, we may obtain temperature differences in the heat exchangers that make sense in practice, for example 8.0 °C. In contrast, by setting a minimum driving force for heat transfer that makes sense in practice, we may obtain an infeasible solution. Those results suggest us to solve the mathematical models over a range of values for the minimum driving force for heat transfer. For each upper limit on the hot utility consumption, the minimum driving force for heat transfer will be varied from 8.1 °C to 0.1 °C through uneven steps. That being said, let us carry on the computations.

### Results and discussion

Three values of costs of utilities and heat exchangers were found. By varying the minimum driving force for heat transfer from 8.1 °C to 3.0 °C, the costs of utilities and heat exchangers is approximately $\$1.95 \times 10^5\,\mathrm{year}^{-1}$. The corresponding hot utility consumption is 717.0 kW. For the minimum driving force for heat transfer of 0.8 °C, the costs of utilities and heat exchangers is $\$1.32 \times 10^5\,\mathrm{year}^{-1}$. In this case, the hot utility consumption is 126.2 kW. When we allowed the minimum driving force for heat transfer to be as small as 0.1 °C, the costs of utilities and heat exchangers did not change significantly. However, the hot utility consumption decreased to 74.7 kW.

### Results and discussion: reduction of the upper limit on the hot utility consumption

The mathematical model was solved again for an upper limit on hot utility consumption of 126.2 kW. The minimum driving force for heat transfer was defined 5.5 °C. The result is the lowest costs of utilities and heat exchangers found thus far: $\$1.08 \times 10^5\,\mathrm{year}^{-1}$. The corresponding hot utility consumption is 74.7 kW.

To verify if there was an even lower costs of utilities and heat exchangers, the mathematical model was solved one more time. The upper limit on the hot utility consumption was 74.7 kW. In addition, the number of heat exchangers was restricted to being less than or equal eight. The results show that for a minimum driving force for heat transfer of 0.5 °C, the costs of utilities and heat exchangers is $\$1.06 \times 10^5\,\mathrm{year}^{-1}$. The hot utility consumption found is 32.7 kW.

The mathematical model was solved for the last time. The upper limit on the hot utility consumption was 32.7 kW. The restriction for number of heat exchangers was kept. The result is that the costs of utilities and heat exchangers did not decrease anymore over the range of minimum driving force for heat transfer. The hot utility consumption remained 32.7 kW.

### Summary of the results

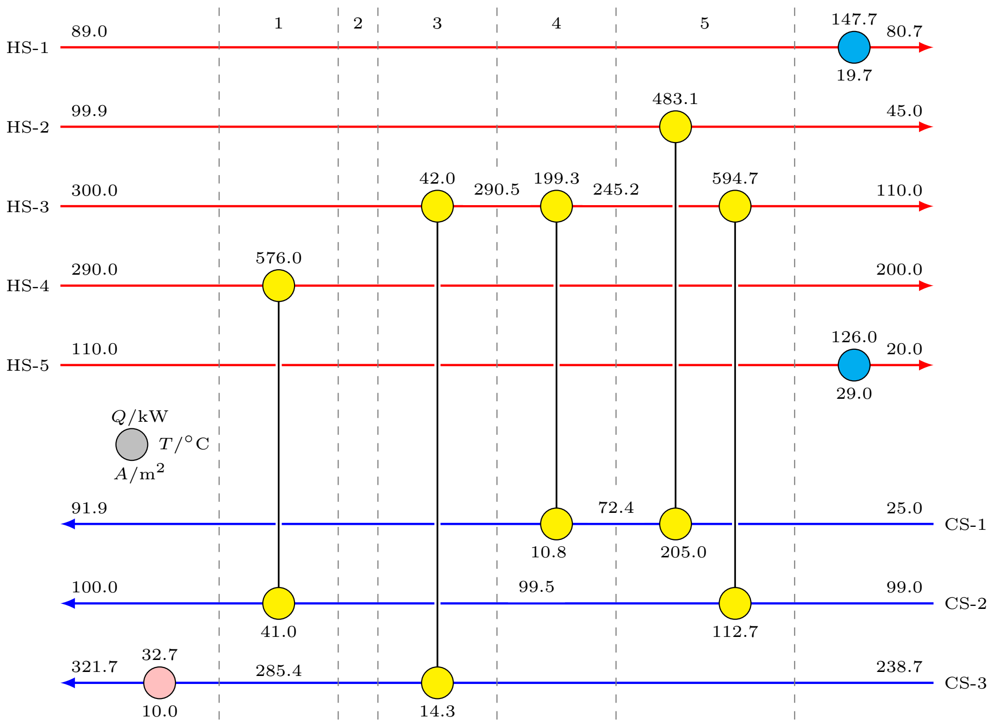
The superstructure mathematical model for HEN was solved a number of times. The objective was to minimise the costs of utilities and heat exchangers. To reduce the size of the feasible region, the upper limit of hot utility consumption was decreased in a stepwise fashion. In addition, the minimum driving force for heat transfer was varied through uneven steps from 8.1 °C to 0.1 °C.

The results showed that by varying the minimum driving force for heat transfer, the utility consumption decreased for the same objective value. It was also shown that for the same utility consumption there may exist more than one objective value. The results are shown in Section 5.2. The topology of the HEN for the lowest costs of utilities and heat exchangers ($\$1.06 \times 10^5\,\text{year}^{-1}$) is depicted in Figure 5.1. The total heat transfer area is around of $442.3\,\text{m}^2$. The mathematical model was solved in an average time of 3 s.

**Table 5.2:** The results for the costs of utilities and heat exchangers and the hot utility consumption.

| input | | output | |
|---|---|---|---|
| min. driving force for heat transfer (°C) | upper limit on hot utility consump. (kW) | cost of utilities and heat exchangers ($10^5\,\$\,\text{year}^{-1}$) | actual hot utility consump. (kW) |
| 3.0 | 1927.8 | 1.95 | 717.0 |
| 0.8 | 1927.8 | 1.32 | 126.2 |
| 5.5 | 126.2 | 1.08 | 74.7 |
| 0.5 | 74.7 | 1.06 | 32.7 |

**Figure 5.1:** The topology of the HEN obtained by minimising the costs of utilities and heat exchangers simultaneously.



## 5.3 Energy loss and cost of pumping

As the fluids flow through pipes, energy is dissipated due to viscous effects and friction with the pipe walls. Energy is also dissipated in the flow through valves, bends, flanges and other fittings mainly because of flow separation (FOX et al., 2020). The energy loss of the fluids is overcome by means of pumps. This section presents a mathematical model to calculate the energy loss of the fluids as they are transported through pipes and heat exchangers. The energy loss of the fluids will be used to calculate the power required for the pumps and the cost to operate them[1].

The assumptions to develop the mathematical model that calculates energy loss were made in the first chapter of this thesis, Subsection 1.4.5. In addition, we will make the following assumptions:

- the stream flows at a constant velocity;
- the stream does not undergo a phase change;
- the effect of the temperature change on the stream properties is negligible.

To develop this mathematical model, we will use various variables and indices. They were defined in the mathematical models of the superstructure for HEN (Section 4.1) and plant layout (Section 4.3). We think that the large number of variables with different labels and indices may be confusing. To facilitate the understanding of this content, we will firstly introduce the equations in a simple form.

### Pipes

The energy loss per unit of mass of the fluid as it flows through pipes will be calculated by means of Section 5.3 (FOX et al., 2020):

$$ELP = f \frac{L}{D} \frac{\overline{V}^2}{2}$$

where $f$ is the friction factor, $\overline{V}$ is the average fluid velocity at a section along the pipe, $D$ is the pipe internal diameter and $L$ is the length of pipe. The pipe internal diameter can be calculated from the stream mass flow rate, a constant velocity and the stream physical properties:

$$D = \sqrt{\frac{4\,\dot{m}}{\rho\,\overline{V}\,\pi}}$$

[1]The cost to purchase pumping equipment will not be considered.

where $\dot{m}$ is the mass flow rate and $\rho$ is the specific mass. With this information, the Reynolds number can be calculated and the friction factor obtained from the Moody chart (FOX et al., 2020).

**Heat exchangers**

The energy loss per unit of mass from the flow through a heat exchanger will be calculated using the method developed by Polley, Panjeh Shahi, et al. (1990). That method will be summarised as follows:

- estimating pressure drop of heat exchangers requires the knowledge of the complete heat exchanger design;
- Polley, Panjeh Shahi, et al. (1990) proposed a rapid design algorithm to obtain the basic geometric parameters of the heat exchanger for a required heat flow rate and maximum allowed pressure drop;
- a similar approach was proposed by Polley and Shahi (1991) to calculate consistent heat transfer area requirements in retrofit of HENs with pressure drop constraints;
- these methods are based on a simple relationship between heat exchanger pressure drop, heat transfer area and heat transfer coefficients:

$$\Delta P = K\, h^{\epsilon}\, A$$

where:

- $K$ is a constant related to some fluid properties, such as thermal conductivity and volumetric flow rate, and the geometric parameters of the heat exchanger, such as tube size;
- $\epsilon$ is a constant equal to 5.1 for the fluid in the shell-side and 3.5 for the fluid in the tube-side of the heat exchanger;
- $h$ is the heat transfer coefficient;
- $A$ is the heat transfer area;

The equation presented above is obtained in unit of pressure. In the mathematical model for calculating heat exchanger pressure drop, we will modify that equation to obtain the energy loss per unit of mass. To do that, we will divide the right-hand side of that equation by the stream specific mass ($\rho$).

**Mathematical model**

The mathematical models of the superstructure for HEN (Section 4.1) and plant layout (Section 4.3) defined the pipes and heat exchangers that we want to calculate the energy loss from. The mass flow rate through the **pipes not connected to heat exchangers** is that provided at the beginning of the problem. The average flow velocity will be taken from a typical value. The internal pipe diameter for hot stream $i$ to flow at the average flow velocity will be calculated by:

$$D_i = \sqrt{\frac{4\,\dot{m}_i}{\rho_i\,\overline{V}_i\,\pi}} \qquad (5.20)$$

$$i \in \mathrm{H}$$

The length of pipe is a variable obtained by solving the mathematical model of the superstructure for heat exchanger plant layout. The energy loss per unit mass of hot stream $i$ as it flows through the **pipes not connected to heat exchangers** will be calculated by:

- from the starting point to the inlet device of the first stage (Figure 5.2-**a**):

$$ELP\_SP\_fid_i = f_i\,\frac{L\_SP\_id_i}{D_i}\,\frac{\overline{V}_i^2}{2} \qquad (5.21)$$

$$i \in \mathrm{H}, \quad k \in \mathrm{K}$$

- from the outlet device of stage $k$ to the inlet device of stage $k$+1 (Figure 5.2-**b**):

$$ELP\_od\_id_{i,k} = f_i\,\frac{L\_od\_id_{i,k}}{D_i}\,\frac{\overline{V}_i^2}{2} \qquad (5.22)$$

$$i \in \mathrm{H}, \quad k \in \mathrm{K}$$

- from the inlet to the outlet device when hot stream $i$ does not exchange heat with cold stream $j$ at stage $k+1$ (Figure 5.2-**c**):

$$ELP\_id\_od_{i,k} = f_i\,\frac{L\_id\_od_{i,k}}{D_i}\,\frac{\overline{V}_i^2}{2} \qquad (5.23)$$

$$i \in \mathrm{H}, \quad k \in \mathrm{K}$$

- from the outlet device of the last stage to the terminal device (Figure 5.2-**d**):

$$ELP\_od\_td_i = f_i\,\frac{L\_od\_td_i}{D_i}\,\frac{\overline{V}_i^2}{2} \qquad (5.24)$$

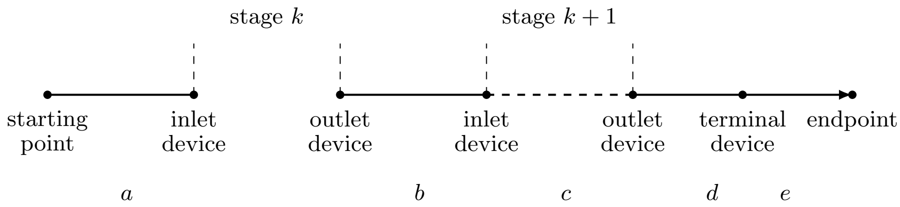
$$i \in \mathrm{H}, \quad k \in \mathrm{K}$$

- from the terminal device to the endpoint (Figure 5.2-**e**):

$$ELP\_td\_EP_i = f_i \, \frac{L\_td\_EP_i}{D_i} \, \frac{\overline{V}_i^2}{2} \tag{5.25}$$

$$i \in \mathrm{H}, \quad k \in \mathrm{K}$$

**Figure 5.2:** The pipes not connected to heat exchangers.




The mass flow rate through the **pipes connected to heat exchangers** will be calculated from a heat balance around the heat exchanger. The mass flow rate of hot stream $i$ as it flows to exchange heat with cold stream $j$ at stage $k$ is given by:

$$\dot{m}_{i,j,k} = \frac{1}{c_{\mathrm{P}i}} \, \frac{\dot{q}_{i,j,k}}{t_{i,k} - t_{i,k+1}} \tag{5.26}$$

$$i \in \mathrm{H}, \quad j \in \mathrm{C}, \quad k \in \mathrm{K}$$

where $c_{\mathrm{P}i}$ is the specific heat capacity. To obtain the average flow velocity, we will assume that the fluid flows at the same velocity as that through the pipes not connected to heat exchangers ($\overline{V}_i$). The internal diameter of the pipe for hot stream $i$ to flow at the average flow velocity will be calculated by:

$$D\_branch_{i,j,k} = \sqrt{\frac{4\,\dot{m}_{i,j,k}}{\rho_i \, \overline{V}_i \, \pi}} \tag{5.27}$$

$$i \in \mathrm{H}, \quad j \in \mathrm{C}, \quad k \in \mathrm{K}$$

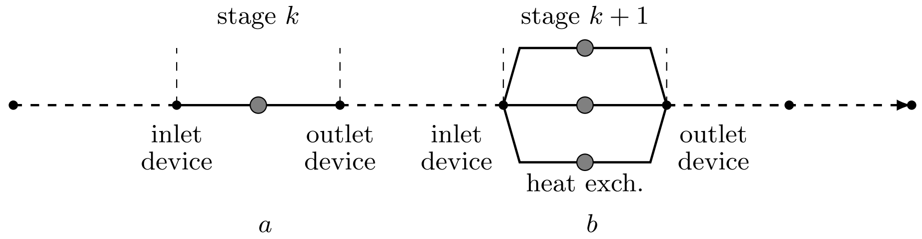
The **pipes connected to heat exchangers** transport hot stream $i$ through the segments depicted in Figure 5.3-**a**:

- from the inlet device to the heat exchangers;
- from the heat exchangers to the outlet device of the stage.

The energy loss per unit mass of hot stream $i$ as it flows through a pipe to exchange heat with cold stream $j$ at stage $k$ will be calculated by:

$$ELP\_branch_{i,j,k} = f_i \, \frac{L\_branch_{i,j,k}}{D\_branch_{i,j,k}} \, \frac{\overline{V}_i^2}{2} \tag{5.28}$$

**Figure 5.3:** The pipes connected to heat exchangers.




$$i \in \mathrm{H}, \quad j \in \mathrm{C}, \quad k \in \mathrm{K}$$

where the friction factor ($f_i$) will be assumed the same as that calculated for the pipes not connected to heat exchangers. In a heat exchanger, the energy loss per unit of mass of hot stream $i$ as it exchanges heat with cold stream $j$ at stage $k$ will be calculated (POLLEY; PANJEH SHAHI, et al., 1990) by:

$$ELEX_{i,j,k} = \frac{K_i\, h_i^{\epsilon}\, A_{i,j,k}}{\rho_i} \tag{5.29}$$

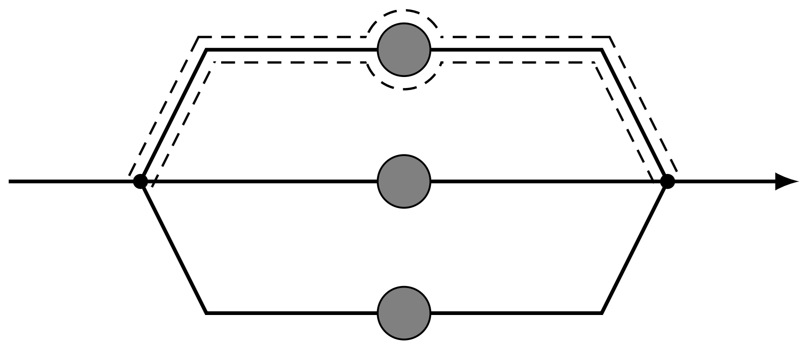
$$i \in \mathrm{H}, \quad j \in \mathrm{C}, \quad k \in \mathrm{K}$$

where the heat transfer area $A_{i,j,k}$ is obtained by solving the mathematical model of the superstructure for HEN. If hot stream $i$ exchanges heat in parallel with various cold streams $j$ at stage $k$, then the pipe will be split into pipe branches (Figure 5.3-**b**). By summing the equations (5.28) and (5.29), we can obtain the energy loss per unit mass in each pipe-branch-to-heat-exchanger system (Figure 5.4):

$$ELS_{i,j,k} = ELP\_branch_{i,j,k} + ELEX_{i,j,k} \tag{5.30}$$

$$i \in \mathrm{H}, \quad j \in \mathrm{C}, \quad k \in \mathrm{K}$$

**Figure 5.4:** A pipe-branch-to-heat-exchanger system enclosed by a dashed line.



If we assume that all the pipe branches at stage $k$ share the same pump, then the pump should have power enough for hot stream $i$ to flow through the pipe-branch-to-heat-exchanger system where the energy loss is the highest. To obtain the highest energy loss per unit mass at stage $k$, we will define the variable $EL\_highest_{i,k}$ and restrict it to being greater than or equal to the energy loss per unit mass in all pipe-branch-to-heat-exchanger systems at stage $k$:

$$\begin{aligned} EL\_highest_{i,k} &\geq ELS_{i,j=1,k} \\ EL\_highest_{i,k} &\geq ELS_{i,j=2,k} \\ &\vdots \\ EL\_highest_{i,k} &\geq ELS_{i,j=n_j,k} \end{aligned} \tag{5.31}$$

$$i \in \mathrm{H}, \quad j \in \mathrm{C}, \quad k \in \mathrm{K}$$

The variable $EL\_highest_{i,k}$ will be minimised in the objective function. As a result, the variable $EL\_highest_{i,k}$ will assume the minimum value that satisfies Equation (5.31). This value is the highest energy loss per unit of mass in all the pipe-branch-to-heat-exchanger systems at stage $k$.

The idea that the energy loss per unit mass is different over the pipe-branch-to-heat-exchanger systems makes sense from the solely mathematical point of view. In practice, the mass flow rate is distributed over the pipe-branch-to-heat-exchanger systems so as to equalise the energy loss per unit mass. This concept is analogous to the resistance of electric circuits in parallel.

Obtaining different values of energy loss per unit mass is a consequence from an assumption made to develop the mathematical model of the superstructure for HEN in the previous chapter, Section 4.1 on page 88. It was assumed that the outlet temperature in the heat exchangers is the same through all the pipe branches. To satisfy this assumption, the mass flow rate through each pipe branch should have a certain value regardless of the energy loss per unit mass.

The solution output from our mathematical model may be inconsistent for practical purposes. On the one hand, it imposes a mass flow rate through the pipe-branch-to-heat-exchanger systems. On the other hand, it designs different energy loss per unit mass through the pipe-branch-to-heat-exchanger systems. To correct the inconsistency, all the pipe-branch-to-heat-exchanger systems should have the highest energy loss per unit mass. In other words, the energy loss per unit mass of the pipe-branch-to-heat-exchanger systems should increase up to the highest value. The increase in energy loss per unit mass can be achieved by installing a device, such as a valve.

### The total energy loss per unit of mass

The **pipes not connected to heat exchangers** contribute to the energy loss per unit of mass from:

- the starting point to the inlet device of the first stage (Equation 5.21);

- the outlet device of stage $k$ to the inlet device of $k+1$ (Equation 5.22) summed over all stages $k$ ;
- the outlet device of the last stage to the terminal device (Equation 5.24);
- the terminal device to the endpoint (Equation 5.25);
- the inlet to the outlet device (Equation 5.23) summed over all stages $k$.

The **pipes connected to heat exchangers** contribute to the energy loss per unit of mass from:

- the pipe-branch-to-heat-exchanger system (Equation 5.31) summed over all stages $k$.

The total energy loss per unit of mass of hot stream $i$ results from the summation of the energy losses from all the pipes not connected and connected to heat exchangers:

$$\begin{aligned} EL_i = ELP\_SP\_fid_i + \sum_{k\in \mathrm{K}} ELP\_od\_id_{i,k} + ELP\_od\_td_i \\ + ELP\_td\_EP_i + \sum_{k\in \mathrm{K}} ELP\_id\_od_{i,k} + \sum_{k\in \mathrm{K}} EL\_highest_{i,k} \end{aligned} \tag{5.32}$$

$$i \in \mathrm{H}$$

The energy losses per unit of mass of cold stream $j$ can be calculated by replacing the index $i$ with the index $j$ in the equations (5.20) to (5.32). In doing so, the total energy loss per unit mass of cold stream $j$ will be given by:

$$\begin{aligned} EL_j = ELP\_SP\_fid_j + \sum_{k\in \mathrm{K}} ELP\_od\_id_{j,k} + ELP\_od\_td_j \\ + ELP\_td\_EP_j + \sum_{k\in \mathrm{K}} ELP\_id\_od_{j,k} + \sum_{k\in \mathrm{K}} EL\_highest_{j,k} \end{aligned} \tag{5.33}$$

$$j \in \mathrm{C}$$

### Annual cost of pumping

The pumping power required to transport hot stream $i$ and cold stream $j$ from their starting point, passing through the heat exchangers, to their endpoint can be determined by multiplying the mass flow rate by the total energy loss per unit of mass:

$$\begin{aligned} pp_i = \dot{m}_i \, EL_i \\ pp_j = \dot{m}_j \, EL_j \end{aligned} \tag{5.34}$$

$$i \in \mathrm{H}, \quad j \in \mathrm{C}$$

The cost of pumping for the transport hot stream $i$ and cold stream $j$ will be given by:

$$\begin{aligned} cost\_pumping_i &= awl\, ec\, \frac{1}{\eta}\, pp_i \\ cost\_pumping_j &= awl\, ec\, \frac{1}{\eta}\, pp_j \\ i \in \mathrm{H}, &\quad j \in \mathrm{C} \end{aligned} \tag{5.35}$$

where $awl$ is the amount of hours the pump works per year, $ec$ is the electricity price per kWh and $\eta$ is the pump efficiency. Notice that the cost of pumping is given on an annual basis. The total cost of pumping will be calculated by:

$$cost\_pumping\_total = \sum_{i \in \mathrm{H}} cost\_pumping_i + \sum_{j \in \mathrm{C}} cost\_pumping_j \tag{5.36}$$

## 5.4 Cost of pipe

The typical methods for estimating the cost of pipe require the layout, material and cost of fittings. The methods vary in accuracy from an order-of-magnitude estimate to a more accurate estimate, such as the Material and Labour Takeoff Method (PETERS et al., 2003). The cost to purchase pipes can be estimated from vendors' quotations and historical data of other projects. There are rules of thumb, diameter and velocities recommended for designing the pipe system with the lowest cost (SINNOT; TOWLER, 2020).

In this chapter, we will develop a formulation to calculate the cost of pipe for the transport of the streams. The streams are transported by pipes from their starting points, passing through the heat exchangers, to their endpoints. The length of the pipes is obtained by solving the mathematical model presented in the previous chapter, Equation (4.54). To represent the length of pipe for the transport of hot stream $i$ and cold stream $j$, we defined the variables $L\_total_i$ and $L\_total_j$, respectively.

The cost of pipe will include contingency pipe for bypassing physical obstacles, absorb thermal expansions and contractions. We will assume that the length of contingency pipe is equal to a percentage ($\upsilon\_cont$) of the length of pipe for the transport of the streams:

$$\begin{aligned} L\_cont_i &= \upsilon\_cont_i\, L\_total_i \\ L\_cont_j &= \upsilon\_cont_j\, L\_total_j \\ i \in \mathrm{H}, &\quad j \in \mathrm{C} \end{aligned} \tag{5.37}$$

We will include the cost of fittings in the cost of pipe. Fittings are pieces of equipment installed in the pipes, such as valves, tees, elbows and flanges. To calculate the cost of fittings, we will make an assumption that the cost of fittings is equivalent to the cost of a pipe of a certain length. We are aware that the concept of equivalent length of pipe is usually applied to pressure drop instead of cost. The consequence of extending the concept of equivalent length of pipe to cost is that an inaccurate result may be obtained. We will also assume that the equivalent length of pipe for fittings is equal to a percentage ($\upsilon\_fit$) of the length of pipe for the transport of the streams:

$$\begin{gathered} L\_fit_i = \upsilon\_fit_i \, L\_total_i \\ L\_fit_j = \upsilon\_fit_j \, L\_total_j \\ i \in \mathrm{H}, \quad j \in \mathrm{C} \end{gathered} \tag{5.38}$$

To calculate the cost of pipe, we will assume that the cost of a pipe is proportional to its length. The cost of pipe for the transport of the streams along with the costs of contingency pipe and fittings will be calculated by:

$$\begin{gathered} cost\_pipe_i = \beta_i \left( L\_total_i + L\_cont_i + L\_fit_i \right) \\ cost\_pipe_j = \beta_j \left( L\_total_j + L\_cont_j + L\_fit_j \right) \\ i \in \mathrm{H}, \quad j \in \mathrm{C} \end{gathered} \tag{5.39}$$

where the parameter is $\phi$ is the cost of pipe per unit length. The cost of pipe can be obtained on an annual basis by multiplying Equation (5.39) by the annualisation factor $af$. The total cost of pipe will be calculated by:

$$\begin{gathered} cost\_pipe\_total = \sum_{i \in \mathrm{H}} cost\_pipe_i + \sum_{j \in \mathrm{C}} cost\_pipe_j \\ i \in \mathrm{H}, \quad j \in \mathrm{C} \end{gathered} \tag{5.40}$$

## 5.5 Example: the impact of piping and pumping power cost in the network topology

A strategy commonly adopted in the academic literature for designing of HENs is to fix the consumption of utilities and minimise the cost of heat exchangers. In that strategy, other costs are indirectly neglected under the assumption that they cannot modify the HEN topology. In this section, we will show a case in which the costs of pipe and pumping modify the HEN topology. In order to do that, a HEN will be designed for a simple process so as to minimise the cost of heat exchangers. Then, the HEN will be redesigned including the minimisation of the costs of pipe and pumping.

### HEN design: minimisation of the cost of heat exchangers

Let us consider a process with one hot stream (H1) and three cold streams (C1, C2 and C3). The hot stream is to be cooled from 200 °C to 150 °C by rejecting 1500 kW. The cold streams are to be heated from 100 °C to 150 °C by absorbing 500 kW each. The film heat transfer coefficient of all the streams is assumed to be 200 W m$^{-2}$ K. The stream data is summarised in Table 5.3.

**Table 5.3:** The supply temperature, target temperature, heat load and heat transfer coefficient of the streams H1, C1, C2 and C3

| stream | $T_{supply}$ (°C) | $T_{target}$ | $\dot{q}$ (kW) | $h$ (W m$^{-2}$ K) |
|---|---|---|---|---|
| H1 | 200.0 | 150.0 | 1500.0 | |
| C2 | | | | 200.0 |
| C3 | 100.0 | 150.0 | 500.0 | |
| C4 | | | | |

To satisfy the heat balance of the streams, a HEN will be designed by solving the mathematical model of the superstructure (Section 4.1). We will define the superstructure with three stages and a minimum driving force for heat transfer of 1.0 °C. The objective function will be defined as the minimisation of the cost of heat exchangers.

The equations to calculate the cost of heat exchangers were presented in Section 5.1. The annualisation factor and the parameters **a**, **b** and **c** required to calculate the cost of heat exchangers will be taken from the previously shown Section 5.2. The computations will be carried out using the General Algebraic Modelling System (GAMS) version 40.4.0 with the solver DICOPT.

### Results and discussion: minimisation of the cost of heat exchangers

In the designed HEN, the hot stream exchanges heat in parallel with the three cold streams. The topology of the HEN is depicted Figure 5.5. The heat flow rate between the hot stream and each cold stream is 500 kW. The area of each heat exchanger is 100 m$^2$. The annual cost of each heat exchanger is \$15.8 × 10$^3$ year$^{-1}$, which totals \$47.4 × 10$^3$ year$^{-1}$.

The arrangement of the heat exchangers in parallel or in series affects the temperature difference between the streams. The greater is the number of heat exchangers arranged in parallel, the smaller is the total heat transfer area and, consequently, is the cost of heat exchangers. By arranging two heat exchangers in parallel followed by another heat exchanger (Figure 5.6), the total heat transfer area increases by

$37\,\mathrm{m}^2$. Similarly, by arranging the three heat exchangers in series (Figure 5.7), the total heat transfer area increases by $45\,\mathrm{m}^2$.

**Figure 5.5:** The topology of the HEN with three heat exchangers in parallel.

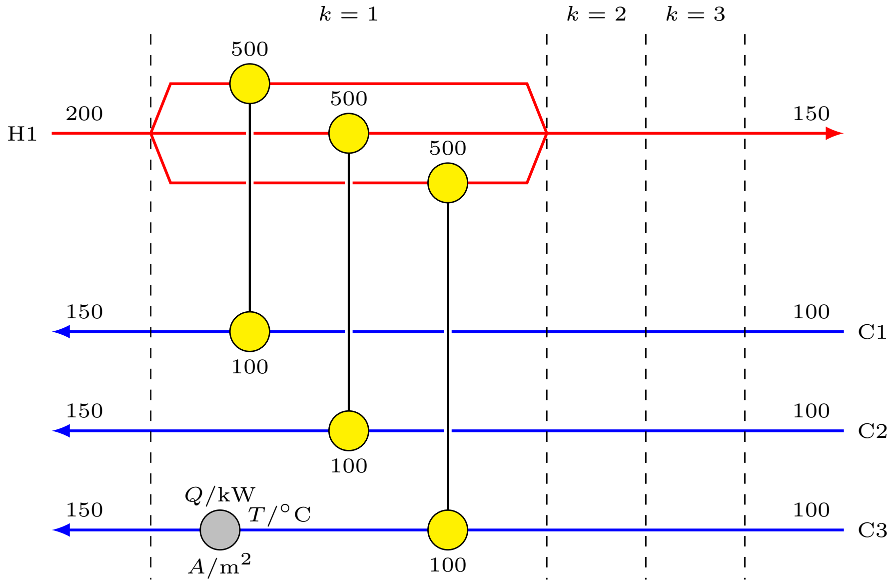



**Figure 5.6:** The topology of the HEN with two heat exchangers in parallel followed by another heat exchanger.

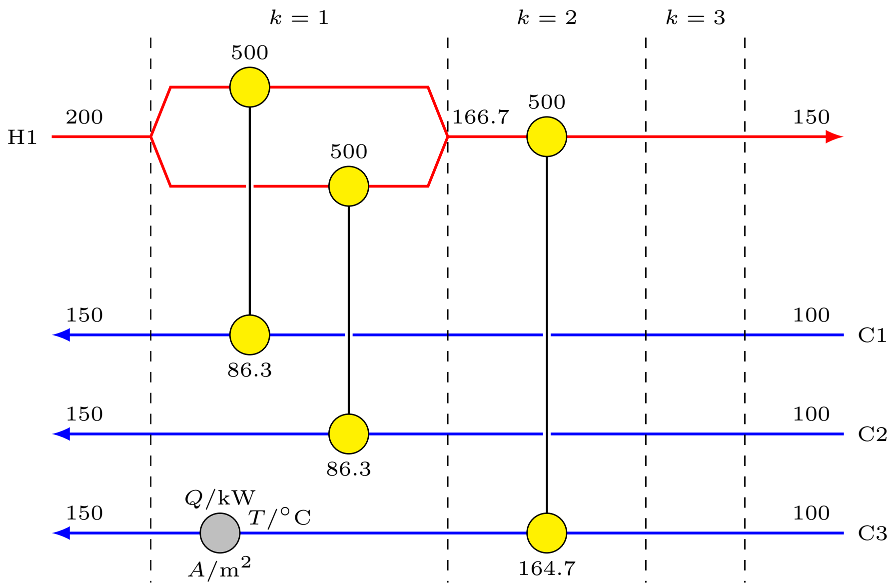
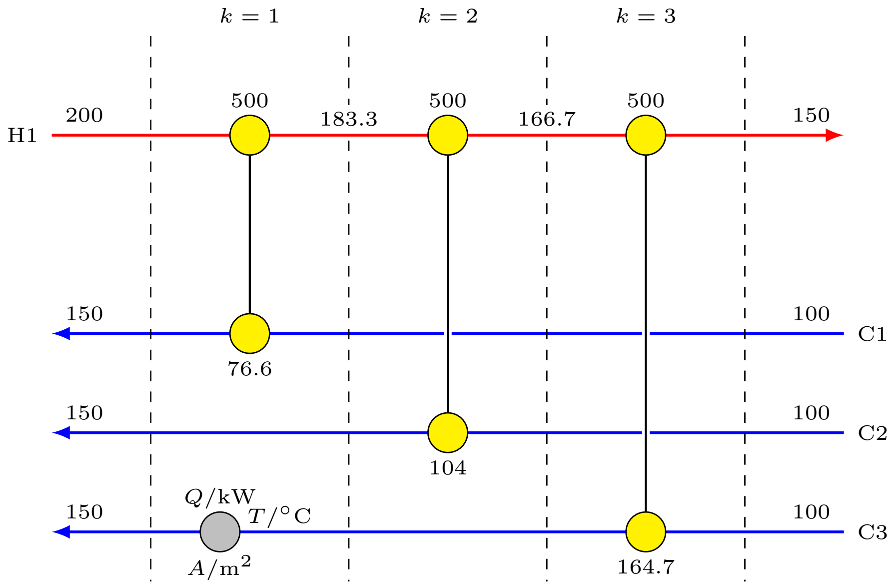


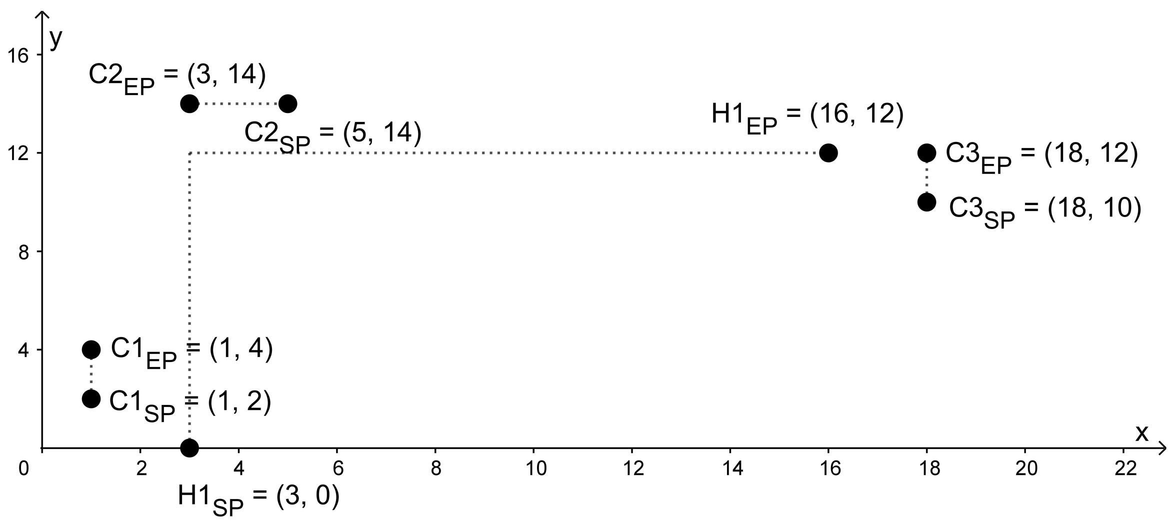
**Figure 5.7:** The topology of the HEN with three heat exchangers in series.




**HEN design: including the minimisation of the costs of pipe and pumping**

Let us now consider that the HEN for cooling and heating the streams is still to be designed. In addition to the HEN, the process requires us the layout of the starting points, endpoints and heat exchangers in the plant. The starting and endpoints have fixed coordinates in the plant, as depicted in Figure 5.8. The coordinates of the heat exchangers will be calculated so as to minimise the costs of pipe and pumping for the transport of the streams. Thus, we have a problem whose objective is to minimise the cost of heat exchangers simultaneously with the costs of pipe and pumping.

**Figure 5.8:** The plant layout and the coordinates of the starting and endpoints of the hot stream C1 and cold streams C1, C2 and C3. The dotted lines link the starting and endpoint of a stream to facilitate the visualisation of the figure.




To calculate the cost of pumping, we will assume that the pumps operate 8000 h year$^{-1}$

with an efficiency of 70 %. The price for electricity is $\$5 \times 10^{-2}\,\mathrm{kWh^{-1}}$. The power of pumping will be calculated from the energy loss in heat exchangers and pipes, as developed in Section 5.3. To calculate energy loss in heat exchangers, we will assume that the hot stream flows through the shell ($\epsilon = 5.1$) and the cold streams the tubes ($\epsilon = 3.5$) of the heat exchangers. The $K$-parameter of the hot stream is equal to $1 \times 10^{-9}$ and that of the cold streams is equal to $1 \times 10^{-6}$.

To calculate energy loss in pipes, the required data is the length of pipe, pipe internal diameter, flow velocity and friction factor. We will obtain the length of pipe by solving the mathematical model of the superstructure for heat exchanger plant layout (Section 4.3). To obtain the pipe internal diameter, let us consider that the mass flowrate of the hot stream is $15\,\mathrm{kg\,s^{-1}}$ and that of the cold streams is $4\,\mathrm{kg\,s^{-1}}$.

By dividing the mass flow rate by the assumed specific mass of the streams shown in Table 5.4, the volumetric flow rate of the hot stream is $17.6 \times 10^{-3}\,\mathrm{m^3\,s^{-1}}$ and that of the cold streams is $5.0 \times 10^{-3}\,\mathrm{m^3\,s^{-1}}$. The streams flow through pipes at an average velocity of $1.5\,\mathrm{m\,s^{-1}}$. The internal diameter of the pipes for the hot stream to flow at the average velocity is 12 cm and for the cold streams is 5 cm.

**Table 5.4:** The assumed properties of the streams H1, C1, C2 and C3.

| stream | $\rho$ $(\mathrm{kg\,m^{-3}})$ | $\mu$ $(\mathrm{N\,s\,m^{-2}})$ | $c_P$ $(\mathrm{J\,kg^{-1}\,K^{-1}})$ |
|---|---|---|---|
| H1 | 850.0 | 0.010 | 2000.0 |
| C1, C2, C3 | 800.0 | 0.002 | 2500.0 |

If a stream is split to exchange heat in parallel, the mass flow rate through the pipe branch will be calculated from a heat balance around the heat exchanger, as presented previously in Equation (5.26). The specific heat capacity to calculate the mass flow rate through the pipe branch was assumed and is shown in Table 5.4. Having the mass flow rate through the pipe branch, its internal diameter will be calculated by means of the previously presented Equation (5.27).

The friction factor will be obtained from the Moody chart. The Reynolds number of each stream can be calculated using the average velocity, the assumed specific mass and dynamic viscosity (Table 5.4). Having the Reynolds number and assuming that all the pipes are made of commercial steel with roughness of 0.025 mm, the friction factor is 0.028 for the hot stream and 0.025 for the cold streams.

The equations to calculate the cost of pipe were presented in Section 5.4. We will consider that the cost of pipe results from the contribution of the pipe for the transport of the streams, the contingency pipe and the fittings. The length of contingency pipe will be defined equal to 25.0 % of the length of pipe for the transport of the stream. The equivalent length of pipe for fittings will be defined equal to 75.0 % of the length of pipe for the transport of the stream. We will use a cost of pipe per unit length of $\$10^3\,\mathrm{lu}^{-1}$.

**Results and discussion: including the minimisation of the costs of pipe and pumping**

The total cost of pipe is mainly influenced by the length of the pipe that transports the hot stream. The length of pipe for the transport of each cold stream is the same in all the topologies depicted in the figures 5.5 to 5.7. By arranging all the heat exchangers in parallel, the total cost of pipe is the highest. The lowest cost of pipe is achieved by arranging all the heat exchangers in series. The length of pipe for the transport of the streams and the cost of pipe of the topologies of HEN are presented in tables 5.5 and 5.6.

In the topologies where the hot stream is split to exchange heat in parallel, the supply of pumping power should be enough for the hot stream to flow through the pipe-branch-to-heat-exchanger system (Figure 5.4) that has the highest energy loss. This characteristic allows the topology with all the heat exchangers arranged in parallel to be the lowest in cost of pumping. In contrast, when the pipes and heat exchangers are arranged in series, they all contribute to the energy loss. Thus, the pump should provide power for the fluid to overcome the summation of all energy losses. As a result, the topology with all the heat exchangers arranged in series has the highest cost for pumping.

By summing the cost of heat exchangers the cost of pumping and the cost of pipe, we found that the topology where all the heat exchangers are arranged in series has the lowest total cost. This way of arranging heat exchangers reduces the costs of pipe in such way that outweighs its higher costs of pumping and heat exchangers. The highest total cost is that of the topology with all the heat exchangers arranged in parallel. The total cost of the HEN in the topologies is compared in Table 5.6. This example shows that the minimisation of the costs of pipe and pumping in addition to cost of heat exchangers may change the optimal topology HEN, in the present case, from a parallel- to a in-series-arrangement.

**Table 5.5:** The length of pipe for the streams in the HEN topologies.

| network topology | pipe length (lu) hot stream | cold streams |
|---|---|---|
| all in parallel | 67.0 | 6.0 |
| two in parallel | 33.0 | 6.0 |
| all in series | 25.0 | 6.0 |

**Table 5.6:** The cost of pipe, cost of pumping and cost of heat exchangers of the HEN in the topologies.

| network topology | cost ($10^3$\$ year$^{-1}$) pipe | pumping | heat exchanger | total |
|---|---|---|---|---|
| all in parallel | 44.8 | 0.9 | 47.4 | 93.1 |
| two in parallel | 26.9 | 1.7 | 49.5 | 78.1 |
| all in series | 22.7 | 2.2 | 50.0 | 74.9 |

### A sensibility analysis: equivalent length of pipe for fittings

We will study the impact of the cost of fittings on the HEN topology. The equivalent of length of pipe for fittings will be varied. In the previous design, we defined the equivalent length of pipe for fittings equal to 75.0 % of the length of pipe for the transport of the stream. To carry out a new design of the HEN, we will use factors of 2.0, 5.0 and 10.0. Using a factor of 5.0, for example, can be understood as if the equivalent length of pipe for fittings is five times longer than the length of pipe for the transport of the stream.

The HEN was redesigned. As expected, the greater is the equivalent length of pipe for fittings, the higher is the costs of pipe and pumping. The cost of heat exchangers remains the same as those previously obtained in Table 5.6. The reason why is that the heat flow rate and temperatures were not affected by the equivalent length of pipe for fittings. For all the factors of 2.0, 5.0 and 10.0, the lowest total cost of the HEN was found in the topology with all the heat exchangers in series. The costs of pipe and pumping of the HEN in the topologies is presented in Table 5.7 over the range of equivalent length of pipe for fittings.

**Table 5.7:** The costs of pipe and pumping of the HEN in the topologies over the range of equivalent length of pipe for fittings.

| fittings factor ($v\_fit$) | network topology | costs of pipe and pumping ($10^3$\$ year$^{-1}$) |
|---|---|---|
| 2.0 | all in parallel | 73.9 |
| | two in parallel | 45.5 |
| | all in series | 39.2 |
| 5.0 | all in parallel | 141.0 |
| | two in parallel | 86.1 |
| | all in series | 73.4 |
| 10.0 | all in parallel | 254.0 |
| | two in parallel | 154.0 |
| | all in series | 131.0 |

## 5.6 Example: minimisation of the costs of utility consumption, heat exchangers, piping and pumping power

Previously in this chapter, a HEN was designed for the bio-ethanol-to-gasoline conversion process (Section 5.2). The objective function was defined as the minimisation of the costs of utilities and heat exchangers. In the current section, a HEN will be designed simultaneously with a layout of the heat exchangers in the plant. In order to do that, the mathematical models of the superstructure for HEN and heat exchanger plant layout will be solved in a simultaneous manner. The objective function will be defined as the minimisation of the costs of utility consumption and heat exchangers along with the costs of pipe and pumping.

The costs will be expressed on an annual basis. The cost of pumping will be calculated from the energy loss by solving the mathematical model developed in Section 5.3. The cost of pipe will be calculated by solving the mathematical developed in Section 5.4. The number of stages of the superstructure will be defined five. The computations will be carried out using the General Algebraic Modelling System (GAMS) version 24.0.2 with the solver DICOPT.

#### Input data for designing the HEN

To design the HEN, we will set an upper limit on the hot utility consumption. To do that, we will refer to the results previously obtained in this chapter (Section 5.2). The upper limit on the hot utility consumption will be defined from 75 kW. The minimum driving force for heat transfer will be defined 0.1 °C. To calculate the heat transfer area, we will assume that the heat transfer coefficient of all the streams is $200\,\mathrm{W\,m^{-2}\,K}$. To calculate the costs of utilities and heat exchangers, we will use the data defined previously in this chapter (Section 5.2).

#### Input data for designing the heat exchanger plant layout

To calculate the location of heat exchangers, we will assume the layout for the plant depicted in Figure 5.9. The coordinates of the starting and endpoints of the streams are shown in Table 5.8. We will restrict the heat exchangers to being placed only in the zones $z_1$, $z_2$ and $z_3$. The boundaries of the the zones $z_1$, $z_2$ and $z_3$ are shown in Table 5.9.

**Figure 5.9:** The layout assumed for the plant of the process for the conversion of bio-ethanol into gasoline.

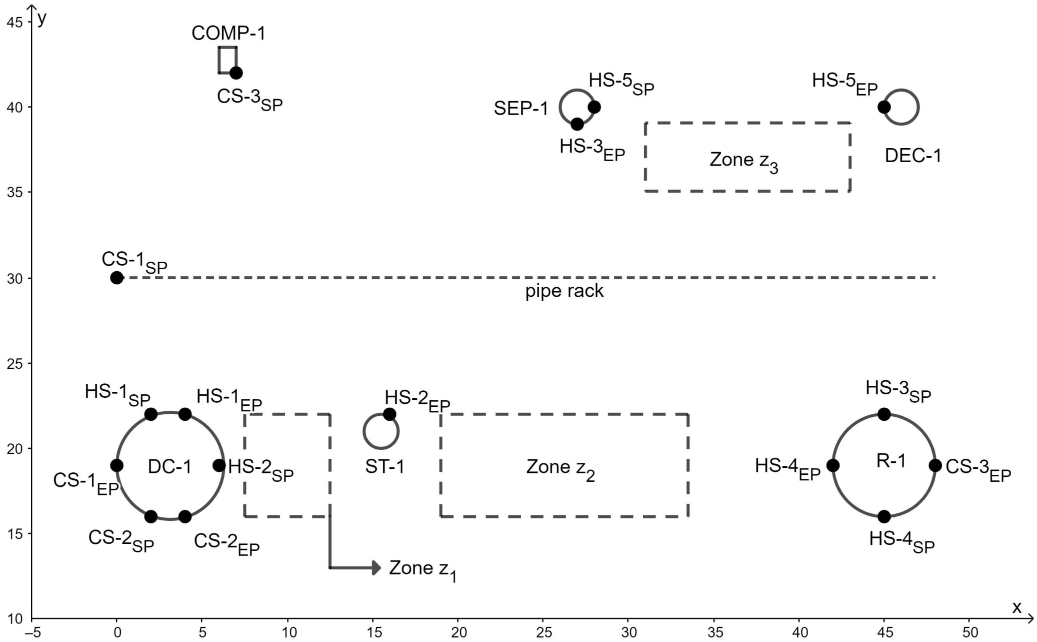



**Table 5.8:** The coordinates of the starting and endpoints for the conversion process of bio-ethanol into gasoline.

| stream | starting point ($SP$) | | | endpoint ($EP$) | | |
|---|---|---|---|---|---|---|
| | $x$ | $y$ | equipment | $x$ | $y$ | equipment |
| HS-1 | 2 | 22 | DC-1 | 4 | 22 | DC-1 |
| HS-2 | 6 | 19 | DC-1 | 16 | 22 | ST-1 |
| HS-3 | 45 | 22 | R-1 | 27 | 39 | SEP-1 |
| HS-4 | 45 | 16 | R-1 | 42 | 19 | R-1 |
| HS-5 | 28 | 40 | SEP-1 | 45 | 40 | DEC-1 |
| CS-1 | 0 | 30 | pipe rack | 0 | 19 | DC-1 |
| CS-2 | 2 | 16 | DC-1 | 4 | 16 | DC-1 |
| CS-3 | 7 | 42 | COMP-1 | 48 | 19 | R-1 |

**Table 5.9:** The lower and upper boundaries of the zones $z_1$, $z_2$ and $z_3$.

| zone | lower boundary | | upper boundary | |
|---|---|---|---|---|
| | $x$ | $y$ | $x$ | $y$ |
| $z_1$ | 7.5 | 16 | 12.5 | 22 |
| $z_2$ | 19 | 16 | 33.5 | 22 |
| $z_3$ | 31 | 35 | 43 | 39 |

In the process used in this problem there is a distillation column. To obtain the reboilers and condensers near the distillation column, we will restrict them to being placed only in Zone $z_1$. We will also restrict Zone $z_1$ to not having any other type of heat exchangers. To avoid overcrowding Zone $z_2$, we will limit the number of heat exchangers placed in it to a maximum of three. There will be no limitation on the number of heat exchangers placed in Zone $z_3$.

### Input data to calculate energy loss: stream properties

The energy loss requires the specific mass and dynamic viscosity of the streams. These properties are used to calculate the friction factor from the Reynolds number. In addition, the specific heat capacity of the streams is required to obtain the mass flow rate through the pipes connected to heat exchangers from a heat balance (Equation 5.26). We will provide these properties before the temperatures of the streams are calculated in the heat exchanges. The data available is the supply temperature, target temperature, pressure, mass flow rate and compositions of the streams (ALDRIDGE et al., 1984).

We will obtain the properties from the Aspen Hysys software. The majority of the streams are a mixture of ethanol and water. The properties of those streams will be obtained using the UNIQUAC package. The reactor product stream (Stream HS-3) and the stream removed from the separator (Stream HS-5) are a mixture of water and hydrocarbons. To obtain the properties of these two streams, we will use the Lee-Kesler-Plocker package.

The properties will be obtained at the supply and target temperatures of the streams. Then, we will use the arithmetic mean of properties to calculate the energy loss. For multi-phase streams, the properties will be taken from the phase in greater amount. To avoid miscalculating the energy loss, we will consider only the streams whose properties do not vary significantly from the supply to the target temperatures. The implications of this simplification are presented a few paragraphs below (5.6).

The output from Aspen Hysys software shows that the properties of the distillation column reflux stream (Stream HS-1) and boil-up stream (Stream CS-2) vary significantly from the supply and target temperatures because these streams undergo a phase change. The same occurs for the product of the reactor (Stream HS-3) as it changes from vapour to a mixture of two liquid phases and vapour. These streams will be excluded from the calculation of the energy loss. Table 5.10 shows the properties of the streams and the phase in greater amount. The properties of the thermal fluid of the reactor (Stream HS-4) were assumed because its composition was not presented in the referred work (ALDRIDGE et al., 1984).

[2] We are aware that this is an uncommon value of heat capacity flow rate, but it is enough for demonstration purposes.

**Table 5.10:** The specific mass, dynamic viscosity and specific heat capacity of the streams obtained from the Aspen Hysys software.

| stream | | phase in grater amount | $\rho$ ($\mathrm{kg\,m^{-3}}$) | $\mu$ ($\mathrm{mPa\,s}$) | $c_P$ ($\mathrm{kJ\,kg^{-1}\,^{\circ}C^{-1}}$) |
|---|---|---|---|---|---|
| hot stream | HS-2 | liquid | 969.9 | 0.44 | 4.2 |
| | HS-4 | | 800.0 | 0.28 | 2.7 |
| | HS-5 | | 923.4 | 0.61 | 5.1[2] |
| cold stream | CS-1 | liquid | 968.1 | 0.63 | 4.1 |
| | CS-3 | vapour | 5.3 | 0.01 | 2.2 |

### Input data to calculate energy loss: flow parameters

The energy loss requires the average flow velocity and pipe internal diameter. We will assume that the average velocity for liquid streams is $1.2\,\mathrm{m\,s^{-1}}$ and for vapour streams is $15.0\,\mathrm{m\,s^{-1}}$. The internal pipe diameter for the streams to flow at the average velocities will be calculated from the volumetric flow rate. To calculate the volumetric flow rate, we will divide the mass flow rate available in Aldridge et al. (1984) by the specific mass shown in Table 5.10. Table 5.11 shows the mass flow rate, volumetric flow rate and pipe internal diameter.

The mass flow rate of the thermal fluid of the reactor (Stream HS-3) was calculated from the ratio between its heat capacity flow rate of approximately $6.4\,\mathrm{kW\,^{\circ}C^{-1}}$ and the assumed specific heat of $2.7\,\mathrm{kJ\,kg^{-1}\,^{\circ}C^{-1}}$. By assuming that the pipes are made of commercial steel with roughness of $0.046\,\mathrm{mm}$, we will obtain the friction factor from the Moody chart (Table 5.11). To calculate the energy loss from the flow through the heat exchangers, we will assume that the hot streams pass through the shell-side ($\epsilon = 5.1$) and the cold streams the tube-side ($\epsilon = 3.5$). The value of the parameter $K$ will be assumed $1 \times 10^{-9}$ for the hot streams and $1 \times 10^{-6}$ for the cold streams.

**Table 5.11:** The mass flow rate, volumetric flow rate, pipe internal diameter and friction factor of the streams for the conversion process of bio-ethanol into gasoline.

| stream | | $\dot{m}$ ($\mathrm{kg\,s^{-1}}$) | $\dot{Q}$ ($10^{-3}\,\mathrm{m^3\,s^{-1}}$) | $D$ (cm) | $f$ |
|---|---|---|---|---|---|
| hot stream | HS-2 | 2.1 | 2.2 | 4.8 | 0.021 |
| | HS-4 | 2.4 | 3.0 | 5.6 | 0.021 |
| | HS-5 | 0.4 | 0.5 | 2.1 | 0.028 |
| cold stream | CS-1 | 2.5 | 2.6 | 5.3 | 0.021 |
| | CS-3 | 0.4 | 77.8 | 8.1 | 0.018 |

### Input data to calculate the costs of pipe and pumping

To calculate the cost of pipe, we will consider the pipe for the transport of the streams, the contingency pipe and the fittings. The contingency pipe factor will be defined 0.25. The equivalent of pipe for fittings will be defined 0.75. To define the cost of pipe per unit length, we will extrapolate the data presented by Souza et al. (2016).

Table 5.12 shows the cost of pipe per unit length for each stream. The cost of pipe for the streams HS-1, HS-3 and CS-2 will not be calculated to because their properties vary significantly from the supply to the target temperatures. To calculate the cost of pumping, we will assume that the pumps operate 8000 h year$^{-1}$ with an efficiency of 70 % and a price for electricity of \$5 $\times 10^{-2}$ kWh$^{-1}$.

**Table 5.12:** The pipe cost per unit of length of the streams for the conversion process of bio-ethanol into gasoline.

| stream | | $\phi$ (\$/lu) |
|---|---|---|
| hot stream | HS-2 | 36.4 |
| | HS-4 | 40.5 |
| | HS-5 | 22.9 |
| cold stream | CS-1 | 39.0 |
| | CS-3 | 53.5 |

### Implications of simplifying the problem

We excluded from the energy loss calculation the distillation column reflux stream (Stream HS-1), boil-up stream (Stream CS-2) and reactor product stream (Stream HS-3). A possible consequence of this simplification is that the solver can make the pipes that transport these streams too long as they do not add to the objective value. To avoid this consequence for the streams HS-1 and CS-2, we restricted any condenser and reboiler to being located in Zone $z_1$. To avoid that the pipe that transports Stream HS-3 being too long, we will restrict it to not being longer than twice the rectangular distance between the coordinates of its starting and endpoint, which results in 70.0 lu.

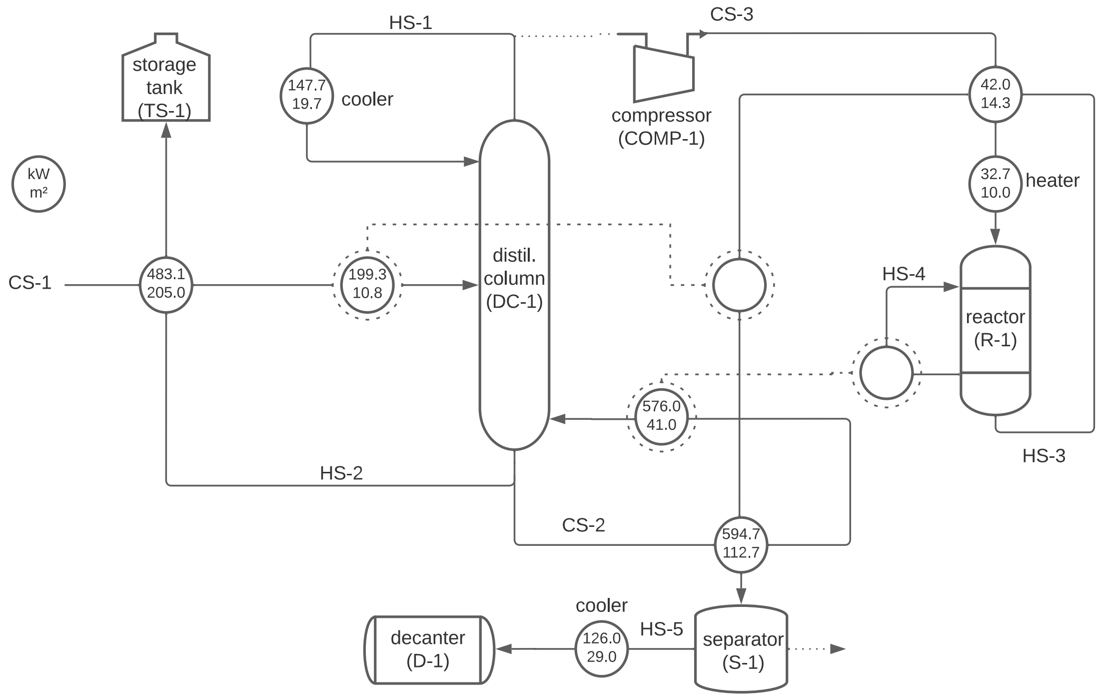
### Results and discussion: the HEN and heat exchanger plant layout

The HEN was designed by minimising the costs of utilities and heat exchangers along with the costs of pipe and pumping. The topology of the HEN is the same as that designed by minimising only the costs of utilities and heat exchangers (Figure 5.1). This result means that the costs of pipe and pumping do not affect the topology of the HEN. Thus, it can be dismissed from the design for the current problem and the assumed cost parameters.

The HEN topology is primarily affected by the cost of heat exchangers. The order of magnitude of the cost of heat exchangers is $10^4$ whereas the order of magnitude of the other costs is $10^3$. As a consequence, a variation in the number and size of heat exchangers causes a greater impact on the objective value than a variation in the length of pipe and pumping power.

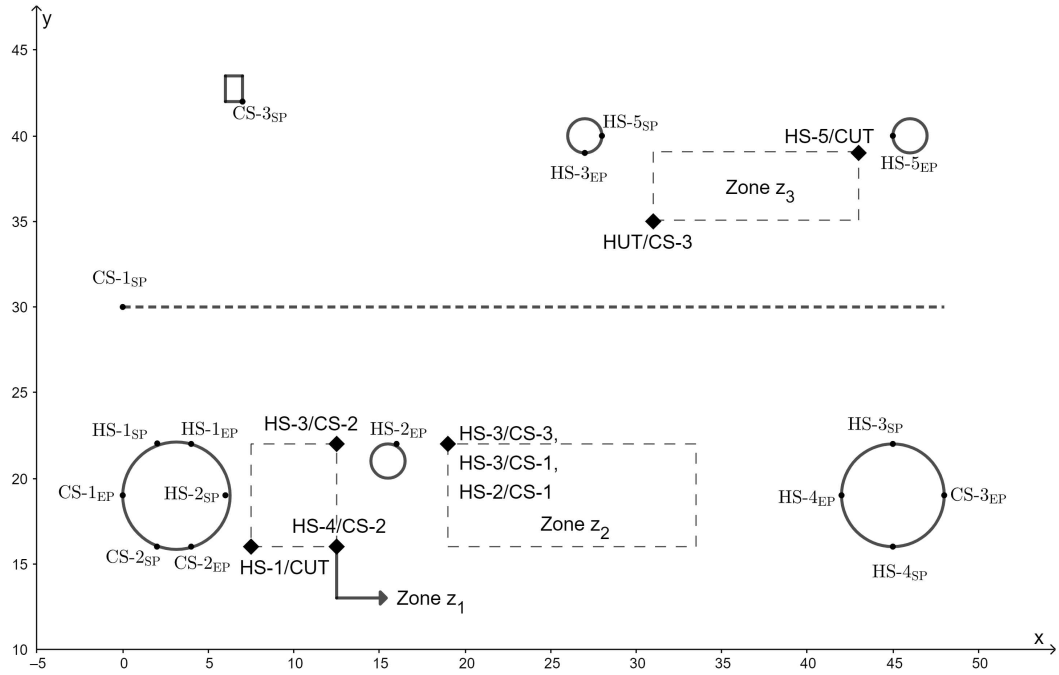
The flowsheet of the heat-integrated process is depicted in Figure 5.10. The distillation column operates with two heat-integrated reboilers and one condenser. The boil-up stream of the distillation column (Stream CS-2) exchanges heat in series with the reactor product stream (Stream HS-3) and the thermal fluid of the reactor (Stream HS-4). The reboilers could have been arranged in parallel because the boil-up stream undergoes an isothermal vaporisation.

**Figure 5.10:** The flowsheet of the process for the conversion of bio-ethanol into gasoline.




The reflux stream of the distillation column is condensed by rejecting heat to the cold utility. The two reboilers and the condenser are located in Zone $z_1$, as depicted in Figure 5.11. The top product of the distillation column is compressed (Stream CS-3) and heated by exchanging heat with the reactor product stream (Stream HS-3) and the hot utility. The two heat exchangers for the top product of the distillation column are located in Zone $z_2$.

**Figure 5.11:** The heat exchangers placed in the plant for the conversion process of bio-ethanol into gasoline.




The feed of the distillation column (Stream CS-1) is pre-heat by absorbing heat from the bottom product of the distillation column (Stream HS-2). The location of the heat exchanger to pre-heat the feed of the distillation column is Zone $z_2$. A second heat exchanger is used to pre-heat the feed of the distillation column with heat from the reactor product stream (Stream HS-3). That heat exchanger is located in Zone $z_3$.

The main product of the separator (Stream HS-5) is cooled by exchanging heat with the cold utility. The cooler is located in Zone $z_3$. In summary, there are three heat exchangers in Zone $z_1$, three heat exchangers in Zone $z_2$ and two heat exchangers in $z_3$. The coordinates of the heat exchangers along with the zones where they are located in are shown in Table 5.13.

**Table 5.13:** The coordinates of the heat exchanger in the zones of the plant of the conversion process of bio-ethanol into gasoline.

| heat exchanger | $x$ | $y$ | zone |
|---|---|---|---|
| HS-3/CS-2 | 19.0 | 22.0 | |
| HS-4/CS-2 | 12.5 | 16.0 | $z_1$ |
| HS-1/CUT | 7.5 | 16.0 | |
| HS-3/CS-3 | 19.0 | 22.0 | |
| HS-3/CS-1 | 19.0 | 19.0 | $z_2$ |
| HS-2/CS-1 | 19.0 | 22.0 | |
| HUT/CS-3 | 43.0 | 35.0 | $z_3$ |
| HS-5/CUT | 43.0 | 39.0 | |

### Results and discussion: pressure drop in pipes and heat exchangers

In this case study, we calculated the pressure drop in the pipes and heat exchangers from the energy loss. The boil-up stream (Stream CS-2), the reflux stream (Stream HS-1) and the reactor product (Stream HS-3) were excluded from the pressure drop calculations. The pressure drop in the heat exchangers was calculated on the shell-side and tube-side. The hot streams were defined to flow through the shell-side whereas the cold streams the tube-side.

The heat exchanger pressure drop of the bottom product of the distillation column is the highest (116.9 kPa). The thermal fluid of the reactor and the feed of the distillation column have the second highest heat exchanger pressure drop, which is almost five times smaller than the highest one (23.2 kPa). The heat exchanger pressure drop of the vapour stream that leaves the compressor (Stream CS-3) is smallest (1.1 kPa).

The pressure drop from the flow through the pipes varied from approximately 12.0 kPa to 32.0 kPa. The pressure drop of the main product of the separator is the highest. The feed of the distillation column has the second highest pressure drop followed by the vapour stream that leaves the compressor. The pressure drop of the bottom product of the distillation column is the smallest. Table 5.14 shows the pipe and heat exchanger pressure drop.

**Table 5.14:** The heat exchanger and pipe pressure drop of the streams. The parenthesis beside the value of heat exchanger pressure drop indicates the other stream involved in the heat exchange.

| stream | | pressure drop (kPa) | |
|---|---|---|---|
| | | heat exchanger | pipe |
| hot stream | HS-2 | 116.9 | 11.6 |
| | HS-4 | 23.4 | 15.3 |
| | HS-5 | 16.5 | 32.4 |
| cold stream | CS-1 | 23.2 (HS-2)<br>1.2 (HS-3) | 28.0 |
| | CS-3 | 1.6 (HS-3)<br>1.1 (HUT) | 20.5 |

### Results and discussion: costs of pipe and pumping

The cost of pumping to transport the vapour stream (Stream CS-3) from the compressor to the reactor is the highest ($\$1.0 \times 10^3\,\text{year}^{-1}$). The bottom product of the distillation column (Stream HS-2) requires a cost of pumping nearly seven times smaller ($\$1.6 \times 10^2\,\text{year}^{-1}$). The order of magnitude of the cost of pumping to transport the other streams is $10^1$.

The pipe to transport the vapour stream (CS-3) has the longest length and highest cost of pipe per unit length. As a result, the vapour stream has the highest pipe cost ($\$2.2 \times 10^3\,\text{year}^{-1}$). The cost of pipe to transport the feed of the distillation is the second highest ($\$1.0 \times 10^3\,\text{year}^{-1}$). The the bottom product of the distillation column and the main product of the separator requires the two smallest cost of pipe, respectively. The costs of pipe and pumping is shown in Table 5.15.

**Table 5.15:** The costs of pipe and pumping obtained for the conversion process of the bio-ethanol into gasoline.

| stream | | cost ($\$\,\text{year}^{-1}$) | |
|---|---|---|---|
| | | pipe | pumping |
| hot stream | HS-2 | 364.7 | 159.3 |
| | HS-4 | 230.0 | 12.6 |
| | HS-5 | 760.0 | 65.5 |
| cold stream | CS-1 | 100.7 | 76.9 |
| | CS-3 | 2171.4 | 1003.6 |

## 5.7 Conclusion

The first theme addressed in this chapter was the design of HENs. We presented the mathematical model to calculate the costs of utilities and heat exchangers. The cost of utilities can be calculated in a straightforward manner. The equation to calculate the cost of heat exchangers is mixed-integer nonlinear. The non-linearity is because the heat transfer area requires the variable temperature in the denominator. The cost of heat exchangers is calculated using the superstructure concept, hence the integer variables.

The remaining part of this chapter was dedicated to extending the method for designing heat exchanger plant layout. The contribution of this chapter was the mathematical model to calculate the costs of pipe and pumping. The cost of pumping takes into account the energy loss in pipes as well as heat exchangers. The energy loss in pipes is non-linear if the stream is split to exchange heat in parallel. The cost of pipe is linear and calculated from the length of pipe for the transport of the stream. Contingency pipe and fittings can be included in the calculation of the cost of pipe.

The mathematical model to calculate the costs of pipe and pumping was implemented using the General Algebraic Modelling System (GAMS). We showed that by designing the HEN and heat exchanger plant layout simultaneously, the impact of the costs of pipe and pumping on the HEN topology can be evaluated. The method we developed was demonstrated in two examples. By means of a sensibility analysis, we showed that our method can design the HEN and plant layout over a variety of scenarios.

# Chapter 6

# Overall conclusion

From a general point of view, this thesis established a relation between the concepts of heat exchanger network (HEN) and layout of chemical process plants. This relation yielded the physical location of heat exchangers in the plant, which we termed as heat exchanger plant layout. We discussed the importance of the heat exchanger plant layout for maintenance purposes, costs of pipe and pumping. A review of the academic literature revealed that the traditional design of HENs has paid limited attention to the heat exchanger plant layout. Thus, this thesis focused on the development of methods for designing the heat exchanger plant layout.

In Chapter 2, we studied how the heat exchangers placed in the plant affects the length of pipe for transporting the streams. The Shortest Length Method (SLM) was developed to calculate the shortest additional pipe length due to the heat exchange between two streams. A case study compared the SLM with a method published in the academic literature. The results showed that while the SLM provides the shortest additional pipe length due to heat exchange, the published method tends to overestimate it in the majority of the cases.

The strength of SLM is that it does not require the location of the heat exchanger to calculate the additional pipe length. Because of this characteristic, the SLM can be used to assist the HEN design by comparing the possible stream matches for heat exchange in terms of their additional pipe length. The SLM is limited to **single** heat exchanges. For **multiple** heat exchanges, such as those in series or parallel, the SLM does not provide the shortest additional pipe length.

The design of the heat exchanger plant layout was the subject of Chapter 3. We developed a method that calculates the coordinates of heat exchangers in the plant. The coordinates are calculated so as to minimise the length of pipe for transporting the streams. The length of pipe was a choice to indirectly minimise the cost of pipe. This method allows the user to set a minimum and maximum spacing between heat exchangers, restrict the number of locations available for placing heat exchangers and, prioritise the minimisation of the pipe length for specific streams.

The method developed in Chapter 3 functions sequentially after the design of the HEN is finished. The sequential order of the designs does not allow the evaluation of the potential of the heat exchanger plant layout to modify the HEN topology. The chapters 4 and 5 were dedicated to overcoming this limitation. We developed a method for the simultaneous design of the HEN and heat exchanger plant layout. This method features almost all the capabilities of the sequential one. The exception is the capability of setting a minimum and maximum spacing between heat exchangers. In addition, instead of length of pipe, the simultaneous method minimises the costs of pipe and pumping.

We presented a case study to evaluate the potential of the heat exchanger plant layout to modify the HEN topology. The case study consisted of a chemical process taken from the academic literature and a plant layout assumed by us. The result showed that the heat exchanger plant layout had no effect on the HEN topology. This result was heavily influenced by the cost parameters we input into the method. For this reason, the obtained result is applicable solely to the presented case study. We shall mention that a part of the cost parameters were taken from different sources and another part was assumed by us.

The potential of the heat exchanger plant layout to modify the HEN topology requires a wider evaluation. By saying wider, we mean an evaluation for a large set of chemical process plant with layouts of varied complexity. In addition, using consistent cost parameters. This information is scarce in the scientific literature. As a consequence, we could not carry out the wider evaluation.

Our experiments showed that the sequential method is easier to be solved than the simultaneous one. Thus, we suggest the use of the sequential method as a first step. If the costs of pipe and pumping have the same or greater order of magnitude as that of the costs of utilities and heat exchangers, then, the simultaneous method should be used, otherwise the sequential one should remain.

### Overall limitations

The set of methods developed in this thesis presents a number of limitations:

- The heat exchangers were treated as dimensionless devices. As a consequence, the nozzle direction and size of the heat exchangers were not taken into account.
- The methods do not provide the route of the pipes.
- The length of pipe is only calculated from one heat exchanger to another (including starting and endpoints). In practice, the pipes are usually passed through pipe racks as an intermediate point between two pieces of equipment.

- The cost of pumping neglected the difference of pressure from one process unit to another. For example, the increase of pressure from a separator to a reactor. In addition, the cost of pumping included only the operating cost. The cost of acquisition was neglected.
- Only bi-dimensional applications were demonstrated. However, extending the methods to tri-dimensional applications is straightforward.

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