Opportunities for Energy Efficiency Improvement in a Renewable Fuels Process

Master's Thesis within the Sustainable Energy Systems programme

LISA LECEROF

Department of Energy and Environment
Division of Heat and Power Technology
CHALMERS UNIVERSITY OF TECHNOLOGY
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MASTER'S THESIS

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LISA LECEROF

SUPERVISOR(S):

Maria Arvidsson

Annette Wendt

EXAMINER

Simon Harvey

Department of Energy and Environment
Division of Heat and Power Technology
CHALMERS UNIVERSITY OF TECHNOLOGY

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Department of Energy and Environment
Division of Heat and Power Technology
Chalmers University of Technology
SE-412 96 Göteborg
Sweden
Telephone: + 46 (0)31-772 1000

Cover:
A capture of rapeseed flowers growing in a field, which is one of the raw materials used for one kind of renewable fuel production. Source: *Baum im Feld* by Petr Kratochvil

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ABSTRACT

A larger part of the cooling and heating demand in a renewable fuels plant is accomplished by using air-cooling and steam in several alternating steps. This study examine what opportunities there are for energy efficiency improvements on site, in order to reduce the electricity and fuel consumption in the plant.

The evaluation was done by using pinch tools on three different case studies: theoretical case based on technical specifications, operational case based on averaged measurements from the process and the adjusted operational case with changed temperature targets respectively.

The investigated cases had total heating demands of 18.7 MW, 20.0 MW and 19.2 MW and total cooling demands of 20.4 MW, 18.9 MW and 17.7 MW respectively.

Results from the pinch analysis of the theoretical case indicated that there were three ways to improve energy efficiency in the plant: by removal of pinch rule violations, by optimizing the temperature level at which utility was supplied or increasing methanol condensation temperatures in an integrated HEN (i.e. raising saturation pressure).

Actual temperatures and flows in the operational case were analysed. By changing target temperatures the plant could save 2 123 kW of energy, both from heating and electricity. Assuming the operation would be adjusted accordingly, the adjusted operational case was created. On this case the full pinch analysis with retrofit suggestions was performed, as well as investigating the full potential of optimizing hot utility levels and condensing methanol at a higher temperature.

By performing a retrofit of the existing HEN, 1 517 kW can be heat integrated. This would save the plant 5.54 MSEK annually in reduced utility costs.

By heating the two flows entering the reactors with MP steam before using HP steam, 5 100 kW could be saved. This shift of hot utility level would save the plant 3.74 MSEK annually in reduced steam costs.

By building two new flashing steps at 8 bar, which would supply condensing methanol at 135ºC, approximately 2 194 kW would become available. If integrated in such a way that it could replace the corresponding load in MP steam, 15.5 MSEK in eliminated steam costs would be saved annually.

It was recommended that options for flashing methanol at intermediate pressures, combined with a new retrofit investigation, would be the best alternative to investigated further.

Key words: Energy efficiency, Pinch analysis, Retrofit, Utility optimization

Contents

Preface

In this study, opportunities for energy efficiency improvements in a renewable fuels plant have been investigated using pinch analysis tools. The study was carried out from September 2013 to January 2014.

This study was carried out with Maria Arvidsson at Chalmers University of Technology and Annette Wendt as supervisors. Without their valuable input this investigation would probably have halted halfway through. My co-workers at the site are highly appreciated for their encouragement, assisting me with data and patiently answering my many questions about general site information. I want to express my gratitude to Staffan Wellander, my invaluable source of knowledge in the English language, for patiently correcting my grammatical errors. Last but not least, I would like to thank my examiner, Professor Simon Harvey, for his active involvement in this project.

Göteborg January 2014

Lisa Lecerof

Notations

Abbreviations:

MeOH	Methanol
HEN	Heat exchanger network
HX	Heat exchanger
HP	High pressure (42 bar)
MP	Medium pressure (8 bar)
LP	Low Pressure (3 bar)
LP_{cond}	Low pressure condensate (liquid at its saturation point)
Vap.	Vapour
Cond.	Condensate
Liq.	Liquid
CW	Cooling water
CCW	Chilled cooling water
MER	Maximum energy recovery
CC	Composite curve
ΔT_{min}	Minimum temperature difference
Q_{Cmin}	Minimum cooling demand
Q_{Hmin}	Minimum heating demand
GCC	Grand composite curve
SEK	Swedish kronor
MSEK	Million Swedish kronor

Symbols:

T_S	Starting temperature
T_T	Target temperature
C_P	Heat capacity
$\dot{m}$	Mass flow
H_S	Enthalpy at starting conditions
H_T	Enthalpy at target conditions
Q	Load
U	Overall heat transfer coefficient
A_{hx}	Heat exchanger area
ΔT_{lm}	Logarithmic mean temperature difference
C	Cost

Dictionary:

Hot utility	Heat supplied from a heating medium, for example steam
Cold utility	Cooling supplied from a cooling medium, for example cooling water
A-U	Annotation for process streams
HXxx	Annotation for an existing heat exchanger in place on a process stream

1 Introduction

As global warming and greenhouse gases have become a hot topic on the political agenda, incentives for transforming our current infrastructure have been created. This is for example done through financial support funding and tax relief for projects that promote sustainable solutions.

Companies can increase profits from supplying more environmentally friendly products and sustainability has become part of the business strategy. Among the various alternative technologies that can contribute to a more sustainable society, the switching of fuel feedstock away from fossil crude oil in the transport sector plays an important part. Several different fuel and powertrain systems compete on the market today. For vehicles that emit less CO_2 the main alternative fuels consist of biogas from organic residues, methanol (MeOH) from crops and biodiesel produced from vegetable oils (Statens energimyndighet, 2013). There is a continuous debate among politicians, scientists and lobbyists about which option is the greatest reducer of CO_2 so that the best future technology is promoted. In this debate the calculated emissions from feedstock to end use product are often used as an argument. The calculated numbers can influence customers' choice and what subsidies companies might get to support their production. This in turn makes their product even more likely to gain market shares early on.

In the production chain from raw material to market, there are several instances where emissions could be reduced in different ways. One is to decrease the energy losses in the product-refining step.

1.1 Background

This thesis investigates energy usage in a process plant producing renewable fuels suitable for Nordic climate conditions. The plant owner is the largest supplier on the Scandinavian market, and the unit is one of Europe's most modern. It was put into operation during 2007 and has since then been improved with respect to product quality and fuel sustainability criteria set up by the EU. The unit uses a solid-state catalyst, which is a new technology that differs from the conventional liquid base catalysed processes. This cuts down on the number of separation steps and results in higher purity of the final products (Nage, Kulkarni, Kulkarni, & Topare, 2012).

The factory is built from a bought license, and is a completely new concept. In the original design, little effort was made to minimize the energy usage. Most of the cooling and heating demand in the plant is currently accomplished using cooling water and boiler steam. In order to reduce the steam consumption, and thereby reduce boiler fuel demand, it is of interest to examine how the heat exchanger network (HEN) can be improved. Therefore it is important to establish energy saving targets. Energy savings imply that as much excess process heat as possible should be re-used internally to thereby save both external heating and cooling (hot- and cold utility). Since process cooling is accomplished using air cooler fans, reduced cold utility usage reduces the electric power consumption of the plant. Increased process heat recovery thus contributes to reducing the environmental impact of the plant product.

1.2 Purpose and objective

The purpose of this thesis work is to identify opportunities for energy efficiency improvements in an existing process to save fuel (through steam usage reductions) and electricity. A rough estimation of the energy savings target is performed using the original process concept. Thereafter a more thorough study is performed based on process data collected from the plant. The energy saving targets is established using pinch analysis tools and pinch rule violations are identified. Retrofit opportunities are suggested, followed by a preliminary economic evaluation of the potential in utility cost savings. The thesis also investigates other energy efficiency measures that involve process modifications or changes to the temperature level of process steam supply.

1.3 Scope

The investigation is limited to the plant, energy efficiency opportunities that can be accomplished by heat exchanging with other processes at the same site are not considered. This is because the production area is situated far away relative to other production facilities in operation on the same site. Possible heat integration would thus require new long and expensive piping. The analysis does not account for equipment performance losses due to operation at off-design conditions after retrofit.

2 The Plant

At the plant, a range of products is manufactured and the site is divided into nine different production areas. Many of these are interconnected, and sometimes the products are refined from one another. The investigated plant shares the same system for steam distribution and cooling water as the other areas, but is completely independent in the sense of material integration.

2.1 Process description

The plant has two reactors with intermediate product separation steps in between them, which increases the conversion of the reactants. The supplied feedstock consists of rapeseed oil and methanol. A general process flow sheet is presented in Figure 2.1.

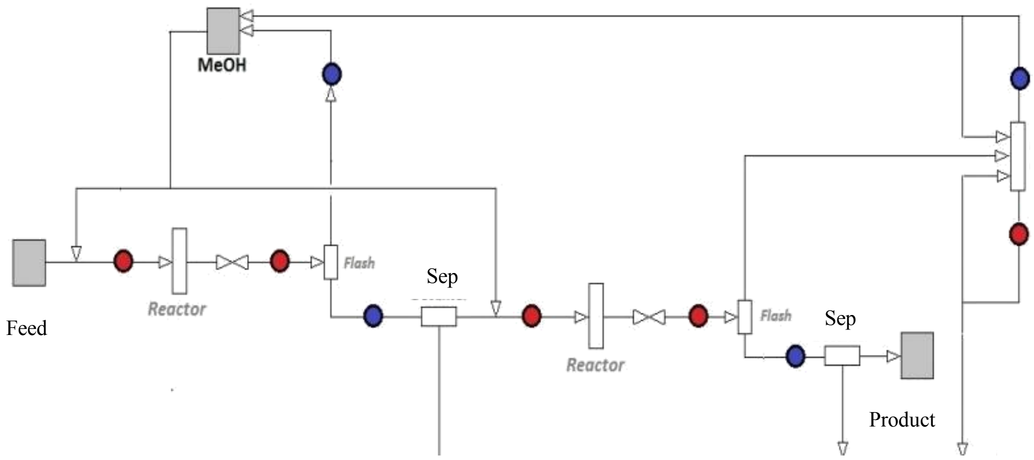

Figure 2.1 General flow sheet of the process. Cold streams requiring heating are indicated with red dots. Hot streams requiring cooling are indicated with blue dots.

The first reactor feed consists of a mixture of a renewable feedstock and methanol which is preheated and then introduced into the reactor. Methanol and tri-glycerides react in the catalyst beds, forming ester together with glycerol. The reaction is illustrated in Figure 2.2.

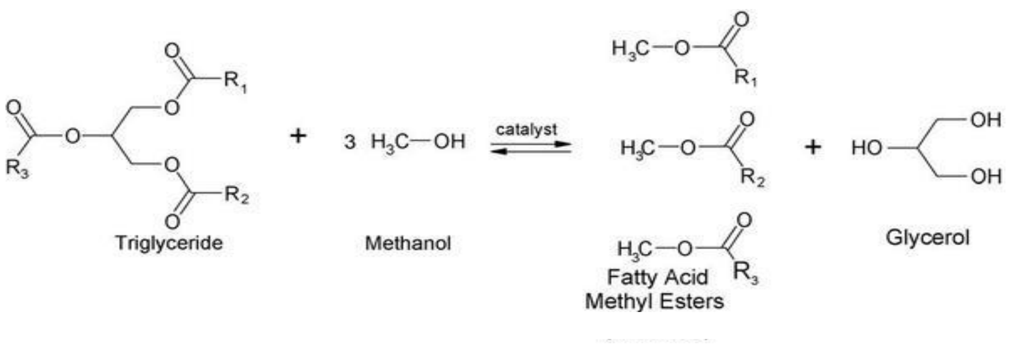

Figure 2.2 Tri-glycerides react with methanol in the reactors (Tan & Lee, 2011).

In order to achieve a high conversion, and to minimize unwanted side reactions, methanol is supplied in stoichiometric excess. To drive the reaction further, it is desirable to remove the glycerol product in an intermediate separation step. The excess methanol present after the first reactor needs to be flashed away, requiring further heating. Glycerol is then removed from the ester at decreased temperatures. Before the second reactor, recycled methanol is again added to the mixture, which is then heated. The same separation operations are then repeated in further process steps so that strict product purity specifications are reached.

There are also necessary purification steps for methanol and glycerol. To ensure that no glycerol or by-products follow with methanol to the storage tank and vice versa, some of the flash streams enter a distillation column. Methanol ready for reuse is received in the top distillate. The bottom glycerol fraction is, however, enriched in water from unwanted side reactions. Therefore the glycerol and water is further processed.

This results in two purified product streams with methyl esters and glycerol.

2.2 The energy system

The main flows relevant for continuous steady-state operation of the process are included in the analysis. Streams used for start-up, sample cooling and some streams working in batch mode are omitted. When referring to a stream, the capital letters A-U are used. There are red dots on cold (blue) streams in demand of heating, and blue dots on hot (red) streams in demand of cooling.

Note that the configuration does not depict the current heat exchanger network (HEN), only where the heating- and cooling demand of process parts. Table 2.1 describes the actual HEN and utility heaters and coolers in place.

Internal heat exchangers have been split up in Figure 2.1 to represent cooling and heating demands in separate places on the concerned streams. In Table 2.1 the stream then has two heat exchanger (HX) notations on it. The HX unit number is indicated with bold typing in Table 2.1 and appears twice, once in the hot- and once in the cold stream section. The locations where heat integration takes place are also illustrated in detail in Figure 2.3.

Table 2.1 A reference table to Figure 2.1, listing the process streams (A-U) with heating (cold streams, written in blue) and cooling demands (hot streams, written in red) in the existing HEN, process location and current medium to satisfy the demand. Bold-faced heat exchangers indicate internal heat exchange and hence appear twice. LP_{cond} - low pressure condensate; LP – low pressure steam; MP-medium pressure steam; HP – high pressure steam; CCW – chilled cooling water; CW – cooling water; Air – air coolers

Stream	HX	Medium
A	HX1	$LP_{cond.}$
B	**HX2**	**O**
C	HX3	HP
D	HX4	LP
E	HX5	**N**
F	HX6 HX7	LP HP
G	HX8	MP
H	HX9	MP
I	HX10	MP
J	HX11	MP
K	HX12	**S**
L	HX13	MP

M	HX14	Air
N	**HX5** **HX15**	**E** CW
O	**HX2** **HX16**	**B** CW
P	HX17	CW
Q	HX18	CCW
R	HX19	Air
S	**HX12**	**K**
T	HX20	CCW
U	HX21	CCW

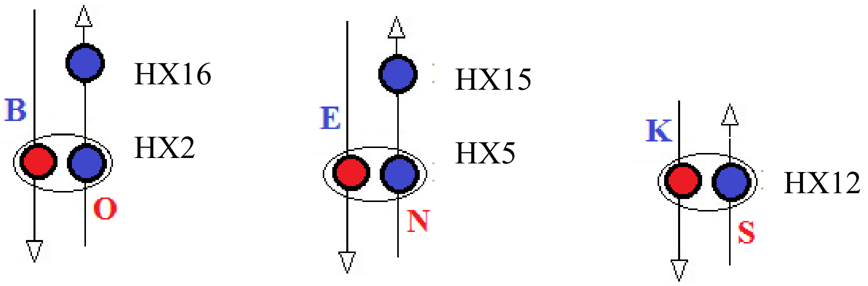

Figure 2.3 An illustration of what the HEN looks like where the net is currently heat integrated. The HX combinations are also indicated in Table 2.1.

The current HEN consists of 21 units - ten heaters, eight coolers and three heat exchangers transferring heat between process streams. In Table 2.2 below, the different types of exchangers are summarized in more detail. Among the heating units one of them uses low pressure (LP) condensate at 3 bar, two use LP steam at 3 bar, five use medium pressure (MP) steam at 8 bar and the two heaters just before the reactors use high pressure (HP) steam at 42 bar. Additionally, three units heat exchange internal process streams with each other. In all, the plant is estimated to require 20 MW of both cooling and heating. The internal heat recovery in the plant reaches no more than 2 MW; the rest is satisfied by supplying hot and cold utility.

Table 2.2 Summary of HX, utility types and their approximate temperature levels available in the plant. The steam levels are assumed to be at their saturation point, condensing at constant temperatures. Return temperature of LP condensate at 100 ºC is a process requirement. Cond. – condensate; No. – number; Temp. – temperature;

No. of Heaters	**Heating medium**	**Temp.** In/Out (ºC)
2	HP steam, 42 bar	253
5	MP steam, 8 bar	170
2	LP steam, 3 bar	133
1	LP_{cond}, 3 bar	133/100
No. of Coolers	**Cooling medium**	**Temp.** (ºC)
2	Air	Ambient
3	CW	23/33
3	CCW	0/5

All exchangers in place are of the shell- and tube type except from HX5 (see Table 2.1), which is a plate heat exchanger.

2.2.1 Steam distribution network

Heat is produced in three furnaces that deliver steam to the whole plant. Three fuels are burned in a mixture: methane-rich fuel gas supplied by a pipeline from a neighbouring industrial plant, internally produced waste streams enriched in combustible products and finally, natural gas purchased from the gas grid

The steam distribution network on the site is quite complex and is illustrated in Figure 2.4. There are only pipelines for distributing MP and HP steam to the plant, i.e. LP steam must be produced locally. MP steam is generated from internal flashing of HP condensate so that imported MP steam can be kept at a minimum. The MP condensate is then gathered and flashed into LP steam and LP condensate. There are also possibilities to throttle MP steam to 3 bar if the internal flashing is not sufficient. After the LP condensate is gathered and utilized in the pre-heater, HX1, it is sent back to the feed water tank at 100ºC.

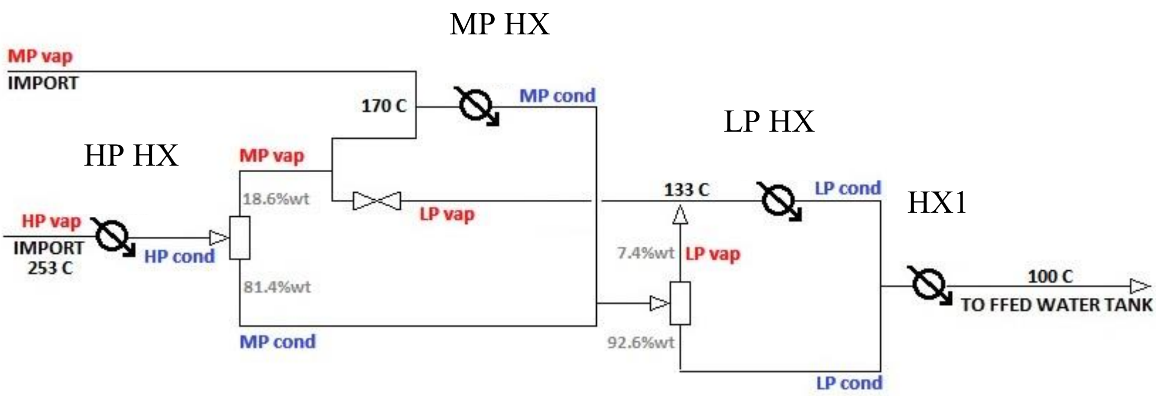

Figure 2.4 The steam distribution network and steam consumers in the plant. The associated streams connected to the HX can be found in Table 2.1. The vapour phase is abbreviated "vap" and condensate with "cond".

2.2.2 Cooling system

To manage the cooling demand on site, eight different units in three different cooling categories are used. Two air cooled units used explicitly on the plant, three coolers using ordinary cooling water (CW) from a nearby lake and three units that need chilled cooling water (CCW) due to the low dew point of methanol, as it gets purified in the recycling system at lower pressures.

Large fans manage the air-cooling taking place on stream M and R in Table 2.1. Each heat exchanger package consists of two fans operating in pairs. When the first has reached its full capacity the additional demand is taken care of by starting the second one. The reflux system on the distillation column uses one fan in on/off mode. The second one has the ability to operate with different frequencies depending on the additional load, to reach a specified outlet temperature. The pair of air fans that take care of the flash vapour fractions in stream M can only work in on/off mode. The total installed engine capacities in the air-cooling units are 184 kW.

The CW is common to the whole site and passes forced draft air-cooling towers on its way back to the crude water tank. Its supply temperature to the site is 22ºC and is returned at 33ºC approximately.

CCW consists of a water-glycol mixture to avoid freezing. It is produced in a closed ammonia cycle according to the one shown in Figure 2.5. Heat for evaporation is taken from the water stream at an inlet temperature of 5ºC. The cooled water leaves at 0ºC, and is supplied to all production areas on site. Compressed ammonia vapour is then condensed at high pressure by indirectly rejecting heat to the surroundings. After passing an expansion valve that relieves the pressure, it is evaporated again by heat from the CCW stream, and the cycle is closed.

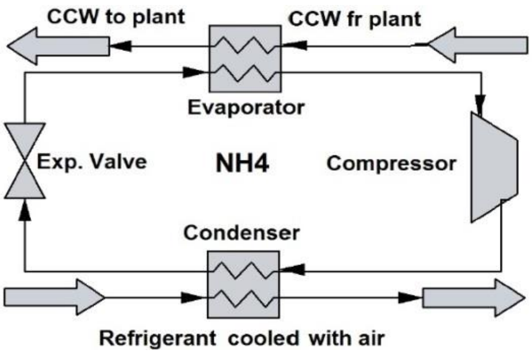

Figure 2.5 Flow sheet of the ammonia (NH_4) compression cycle producing CCW. Exp. – expansion;

3 Methodology

Pinch analysis tools are used to establish energy saving targets. This chapter presents a brief overview of the basic concepts for carrying out such an analysis, starting with the different cases to be analysed and the necessary data collecting procedures.

3.1 Case presentation

The plant has an extensive and well-documented heating and cooling system. Few changes have been made to the original design when it comes to equipment, so the available data of loads and temperatures are used without further verification in an initial theoretical case study (hereafter referred to as the theoretical case, corresponding to operation of the process according to original concept).

However, current operation of the process differs with respect to the original concept, thus the theoretical case data will be revised with respect to measured averaged flows and temperatures in the process to better correspond to actual operating data. Mass balance consistency is assured by checking if the averaged flows add up, and is adjusted where necessary. The same applies to critical points where energy balances are inconsistent and temperatures need to be changed. The process with adjusted averaged measurements will hereafter referred to as the operational case.

By comparing the operational case with theoretical data, areas where set points might be too high or too low can be looked further into. Assuming measures to reset these points are taken to save energy, an adjusted operational case is established (hereafter referred to as adjusted operational case). Here options for increasing internal heat recovery will be investigated so that further energy savings can be made.

In Figure 3.1 the road map through the three different cases are described, and the reader is encouraged to keep them in mind when the results are presented and discussed in the following chapters. Below, the headline text in italics describes assumptions behind the data collected. Then a short description of the work to be done follows and finally, what results that can be expected by doing so are presented.

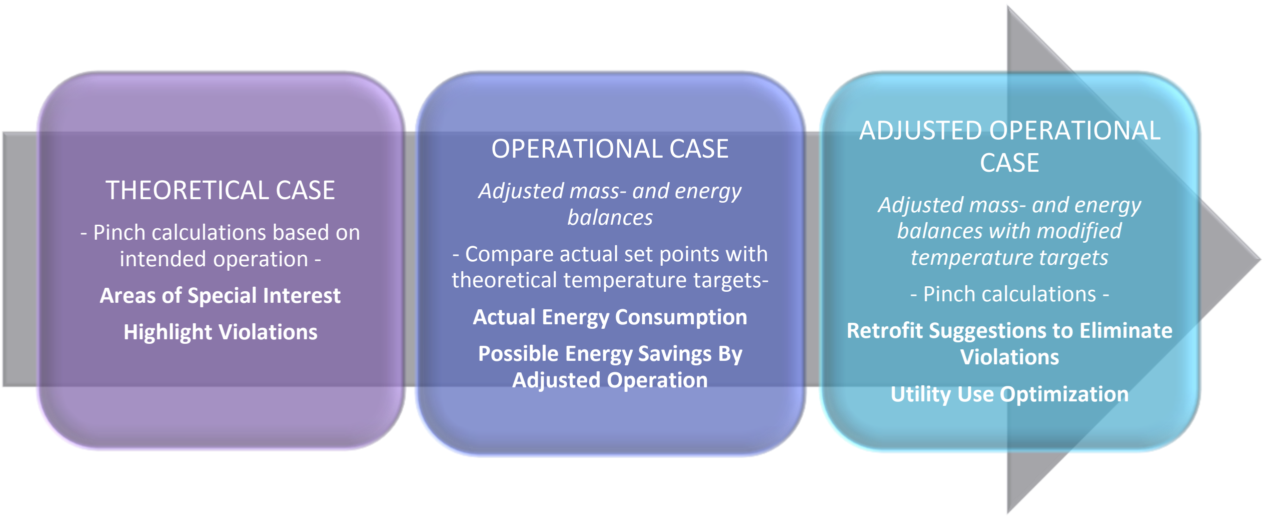

Figure 3.1 In the figure the different cases and their purpose are presented as a "report road map". Headlines refer to each case, text marked in italics describe the underlying basis for calculation, then a short description of the work that will be performed for each case is given and the expected results are marked in bold face.

3.1.1 Theoretical case

As mentioned above this case consists of theoretical data corresponding to the original process concept, which is assumed to have been generated using engineering calculation tools including mass and energy balances. This data is used for a rapid assessment of the potential for improved process energy efficiency, thus little effort is made in this work to verify the data quality.

3.1.2 Operational case

Averaged measured values of flows and temperatures in the process are used in this case, with physical data from the theoretical case as a basis of comparison. Flows are adjusted so that mass balances are verified. Gas flow meters are judged to be less liable than pure liquid flow meters. However, liquid flows that have a risk of flashing in the measurement point are also regarded as unreliable. Where flows are missing and operation has been changed so that flow data from the theoretical case is not likely to be valid anymore, assumptions of new mass balances will be backed up with simulations using the simulation software Aspen Plus. Heat capacities from the theoretical case will be kept as they are, or scaled to match new/deviating temperature levels. Digital temperature meters are viewed as the most reliable measurements in the process. In a few points manual temperature readings needs to be performed. Their reliability is less then digital readings and will consequently be more prone to adjustment for closing the energy balances.

Current operation differs from the original concept in several aspects, requiring process data adjustments as described below:

- The distillation column in Figure 2.1 runs at a temperature that is 16% lower than in theory, and with 67% of the theoretical reflux mass flow. This consequently affects the target temperature of stream I in the column reboiler, HX10 and mass flow in the distillate condenser, stream R in HX19, respectively.
- Methanol from the flash system in stream M and distillate in stream R is divided into gas cooling, condensing and liquid cooling parts to represent the load distribution in HX14 and HX19 according to standard pinch analysis procedures.
- The feed tank in Figure 2.1 is heated, which increases the inlet temperature of stream A in HX1 by 100%.
- Pre-heating of feed in stream A cannot exceed 100ºC due to a risk of water evaporation; therefore the operational target temperature in HX1 is set to be 95% of its theoretical value.
- Reactant flows (stream C and F in Figure 2.1) are not heated as far as their theoretical target, so in the operational case the target temperature in HX3 and HX7 is lowered by 4.8% compared to the theoretical value.
- The target temperature before the second separation (stream O in Figure 2.1) is raised to increase performance in the following separation operation. Therefore the outlet temperature on stream O, exiting from HX16, is raised by 50% compared to the theoretical value.

The HX numbers, in connection with the streams referred to in the text, are tabulated in Table 2.1.

3.1.3 Adjusted operational case

Deviations from the theoretical case listed in the previous section have been implemented for practical reasons. Some deviations are required in order to get the site to work properly, and others because the suggested technical solutions were modified as the facility was built. However, there are a number of instances in the operational case where there are no obvious explanations for deviation from the original concept. These points have been looked into further to see whether only a change in target temperatures can save energy, without making any significant investments. The adjusted operational case is then based on the assumption that the new targets are implemented, and then investigated further to see how even more energy savings can be made by increasing the degree of internal heat recovery.

3.2 Data collection

In the analysis, stream characteristics such as start and final temperatures, $T\ (°C)$, specific heat capacity, $C_P\ (kJ/(kg \cdot K))$, and mass flow, $\dot{m}\ (kg/s)$, are collected in order to determine the required process heating- and cooling loads on the plant. These quantities will naturally vary with production rate and stream composition, but when performing the analysis they are set to fixed values. Therefore the system boundaries and data collection should be carefully selected to represent the desired operation before proceeding (Kemp, 2006). The choice will affect the outcome of the analysis, and consequently conclusions that can be drawn from it.

The plant is evaluated for a production rate corresponding to its full capacity. The system boundary is drawn around the plant, since it is stand-alone with regard to process streams, and only shares utility supply with the site as a whole. Theoretical data (theoretical case) is compared with measured averaged flow and temperatures in the process (operational case) matching this condition. Some targets are changed, since it is sometimes not possible to run a site exactly like it was theoretically intended to in the initial design stages. Theoretical values for heat capacities are kept as they are. The simulations, upon which they are based, are as close as one can get without redoing them all over again with changed operating conditions, or running extensive labs. Where data is not sufficient, simulations in Aspen Plus and Chemcad are performed to fill in the gaps.

3.2.1 Aspen Plus and Chemcad simulations

Simulations are used in the operational case to establish reliable data at locations where mass balances do not add up, or when data deviates from the theoretical case data to a large extent. Locations where the theoretical case data cannot be assumed to be valid in the operational case as well is where temperatures might be much higher or lower. This in turn affects the outcome of running a unit operation. In areas where mass balances do not add up, the flow meters might be subjected to turbulence or flashing across the measurement point, which strongly affects the measurement. The unit operations simulated were flash systems in Figure 2.1.

The esters are a combination of different methyl esters, and the composition originally depends on the type of fatty acids present in the feedstock (Chemstations, Inc., 2012). The presence of more polar molecules, such as methanol and glycerol, makes it likely that substances interact with each other. Therefore it is appropriate to use the NRTL (non-random two-liquid) property model in the simulations (Chemstations, Inc., 2012).

This equation of state model makes use of activity coefficients between binary pairs to account for how much they deviate from ideal gases. The coefficients are evaluated for all possible pair-combinations of the substances in the flow composition.

Aspen Plus provides a larger range of different methyl esters, and the effects of elevated temperatures were simulated in this software. The ester flow composition was approximated as 63$\%_{wt}$ methyl oleate, 30$\%_{wt}$ linoleic acid (instead of methyl linoleate), and 7$\%_{wt}$ methyl palmitate (Chemstations, Inc., 2012). In addition, methanol and glycerol were used in proportions representative for the feed composition in the flash system.

Chemcad only has the properties of one main ester component in its database (methyl oleate (Chemstations, Inc., 2012)). It was used when simulating effects of flashing methanol at elevated pressures. The different software was used for two different purposes since their availability was limited at the time of simulation.

3.3 Principles behind pinch analysis

Pinch analysis is based on the thermodynamic features of process streams and was developed in Manchester during the 70's (Linnhoff & Flower, 1978). It investigates how much heat energy that can theoretically be recovered within the system, based on the first- and second law of thermodynamics.

A stream is heated or cooled between a starting temperature, T_S, and a target temperature, T_T. For $T_S < T_T$ the stream is defined as cold, since it needs heat to reach the target, and correspondingly for $T_S > T_T$, the stream is defined as hot (Linnhoff & Flower, 1978). If flow and heat capacities are known, the heat load, Q, for each stream can be calculated according to equation 3.1 below. If the heat loads between the different temperature-ranges are plotted, a cold and a hot composite curve (CC) is formed respectively, as illustrated in Figure 3.2.

$$Q = \dot{m} \cdot C_P \cdot (T_S - T_T) = \dot{m} \cdot (H_S - H_T) \qquad [kW] \qquad (3.1)$$

When heat is exchanged between two streams, the hot stream needs to have a higher temperature than the cold one at all times. If the streams reach the same temperature, there will be no further driving force for heat transfer. This is a thermodynamic restriction. Therefore the two CCs should not intersect. By sliding the curves relative to one another, the global minimal allowable temperature difference (ΔT_{min}) in heat exchangers is varied, which affects the amount of internal heat recovery potential and minimum utility requirements. A large temperature difference gives a smaller heat exchanger area since heat transfer is more effective, but also allows for less recovery potential. The opposite applies for small values (Harvey, 2011). From the curve the internal heat recovery potential is shown as the part where the hot and cold curves overlaps each other ($Q_{recover}$). Where no overlap exists, heating or cooling from external utilities is needed, corresponding to the minimum cooling and heating demands, Q_{Cmin} and Q_{Hmin}, respectively.

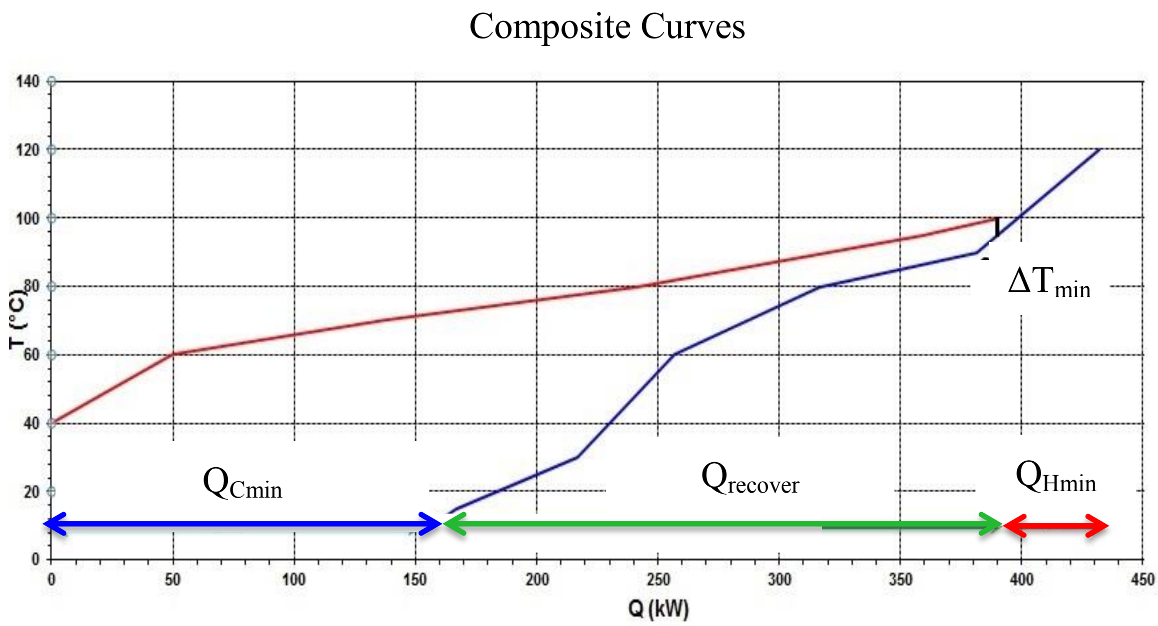

Figure 3.2 Principle sketch showing the hot (red line) and a cold (blue line) composite curves. The total energy available in a temperature range is summed up for hot and cold streams respectively. Where the curves overlap there is a potential to recover heat by transferring it from the hot streams to the cold ones, $Q_{recover}$. The remaining loads have to be supplied by hot (Q_{Hmin}) and cold (Q_{Cmin}) utility.

3.3.1 Energy targeting via the grand composite curve (GCC)

Another graph that is typically employed in pinch analysis is the grand composite curve (GCC). Adding $\frac{\Delta T_{min}}{2}$ to T_S and T_T for cold streams, and subtracting it from hot streams, ensures that ΔT_{min} is maintained when summing up and cascading the energy availability/deficit of the hot and cold streams in the temperature intervals generating one single curve. When temperatures are shifted in this way it is said that we work with "shifted temperatures", also referred to as "interval temperatures" (Kemp, 2006). Plotting the interval temperatures on the y-axis, and the corresponding enthalpy change on the x-axis, generates the GCC, shown in Figure 3.3.

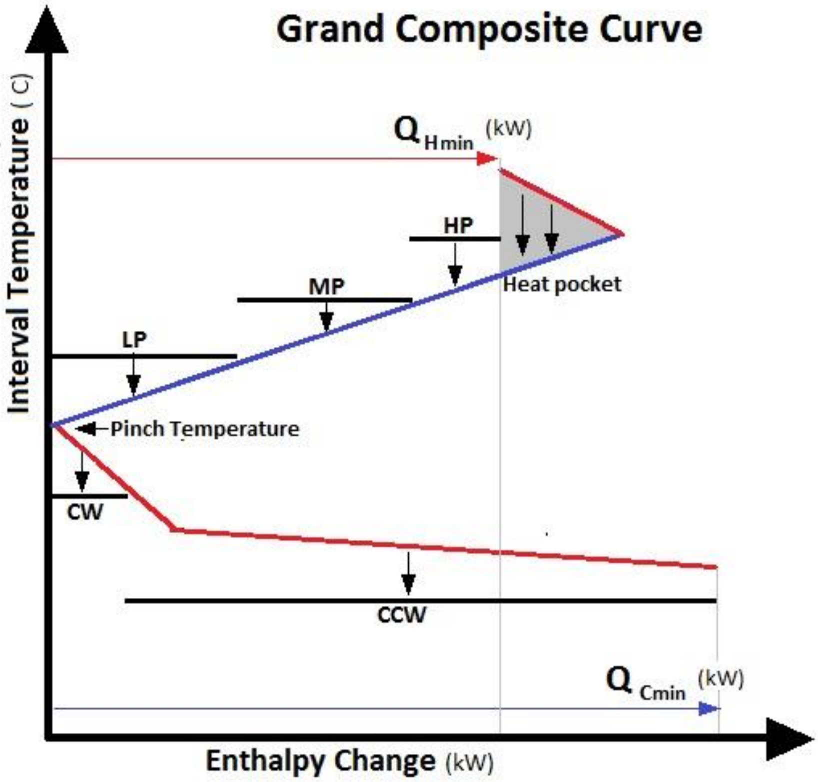

Figure 3.3 Principle sketch of a GCC. The energy availability/deficit of the streams in each temperature interval are summed up and cascaded down to underlying temperature intervals in the process.

In a GCC, maximum energy recovery (MER) is reached both above and below the pinch temperature, and the energy targets (Q_{Hmin} and Q_{Cmin}) can be read from the curve, as illustrated in Figure 3.3. The pinch temperature is where enthalpy change is zero and the curve intersects the y-axis.

In a GCC there is a heat deficit above the pinch. When all the heat available from higher temperatures in hot streams has been internally transferred to cold streams, there will still be a need for additional heating. Hence, in order to accomplish a system with maximum heat recovery no cooling should be done here. Since all heat available in streams above the pinch has been recovered in a GCC, the remaining heating demand is the minimum (Q_{Hmin}). The opposite applies below the pinch: minimum cooling demand (Q_{Cmin}) can be seen, and here no heat should be supplied since an overall heat surplus prevails. If heat transfer across the pinch should connect these two distinct regions above and below the pinch, the result will be an increased demand for cooling and unnecessary heating. The extra amounts of energy supplied by external heating above the pinch will be cascaded down in the region that already has a surplus of heat, thus requiring additional external cooling.

The reasoning above can be summarized in three golden rules that should be followed when designing a HEN: First, do not use cold utility above the pinch. Second, do not use hot utility below the pinch and third, do not transfer heat across the pinch. Breaking these rules is termed "pinch rule violations", and if they are not followed the final design will consume more energy than the minimum requirement (Kemp, 2006).

The required temperature at which the hot and cold utility should be supplied to satisfy Q_{Hmin} and Q_{Cmin} can also be targeted from the GCC. By introducing the utility temperature level, in interval temperatures, the load can be extended until it intersects the

GCC. By maximizing the cheaper utility levels first, costs for heating and cooling can be kept to their theoretical minimum.

3.3.2 Heat exchanger network (HEN) design and retrofit

An existing HEN can be analysed in order to identify the locations where the pinch rules are violated, as well as the corresponding heat flows. The network can thereafter be modified (retrofitted) with the objective of reducing or eliminating these pinch rule violations, leading to energy efficiency improvements for the process as a whole. In practice, retrofit projects usually aim at identifying the most efficient changes (in terms of energy recovered per invested unit of surface area, and ultimately by money earned per invested unit of surface area) to reduce the utility usage and hence rarely lead to MER (Maximum Energy Recovery) for the modified process. Limitations such as requirements for return on capital investments, geographical distances or large pressure differences between streams of interest, as well as thermodynamic limitations needs to be taken into consideration, just to mention a few.

In a retrofit it is important to consider the topography of existing HX units to reduce the capital investment costs. Old HX:s should be utilized in the new retrofit network as much as possible. This could reduce the investment cost for the new installed heat transfer area necessary, and the number of new units could be kept low. The equipment size and necessary heat transfer area in each exchanger is estimated by calculating A_{hx} from equation 3.2. The loads will be known through the pinch analysis and the stream match in the retrofit determines the resulting temperatures. By assigning an overall heat transfer coefficient for a new stream match in the HX unit from Figure 8.1, U $(kW/(m^2 \cdot K))$, the area can be estimated by using equation 3.2.

$$Q = U \cdot A_{hx} \cdot \Delta T_{lm} \qquad [kW] \tag{3.2}$$

Here Q (kW) is the total amount of energy transferred across the HX area, A_{hx} (m^2), with streams having a logarithmic mean temperature difference of ΔT_{lm}.

The HEN design should start with the streams that are closest to the pinch. This is due to their low degree of freedom, and makes sure that there will not be an impossible match of streams left in the end (Kemp, 2006). When matching two streams one should aim at ticking off at least one of the loads so that the number of new units is kept at its minimum. A pinch match is when both streams start/end at the same interval temperature. In this case it is extra important that the streams do not intersect within the heat exchanger. To ensure that temperatures will not violate ΔT_{min} in a pinch match, the following should apply to both streams (Franck, Persson Elmeroth, Vamling, & Harvey, 2011):

Below the pinch: $\dot{m}Cp_{cold} \leq \dot{m}Cp_{hot}$

Above the pinch: $\dot{m}Cp_{cold} \geq \dot{m}Cp_{hot}$

These guidelines are valid for network design in general and applies to retrofits as well.

3.4 Economic evaluation of utility cost savings

A very rough economic evaluation of utility cost savings is carried out for each retrofit suggestion. The annual savings made due to the improvements in the HEN will consist of a reduced utility consumption. In the steam distribution network of the renewable fuel production site the different steam levels (HP, MP, LP, and LP_{cond}) are connected by internal flashing and only HP and MP steam can be imported, see Figure 2.4. Accordingly, the price of the steam of the different pressure levels can be estimated from the price of HP steam (C_{HP}). In order to give a quick estimation of the utility cost savings, the cost for MP steam (generated by internal flashing of HP condensate) is estimated by the change of specific enthalpy of evaporation. As can be seen in Figure 2.4, some of the MP steam is created without extra cost by flashing HP condensate. The MP steam load above this margin is imported, and the cost (C_{MP}) is calculated according to equation 3.3. This is a very rough estimation and a more dynamic analysis of the complex steam network is required for more accurate results of the utility cost savings.

$$\Delta H_{vap_{HP}} = 1698\ kJ/kg$$

$$\Delta H_{vap_{MP}} = 2048\ kJ/kg$$

$$C_{HP} = 0.24\ SEK/kg\ = \frac{0.24\ \cdot 3600}{1.698} =\ 509\ SEK/MWh$$

$$1\ kg\ \cdot \Delta H_{vap_{HP}} = x\ kg\ \cdot \Delta H_{vap_{MP}} \tag{3.3}$$

$$\rightarrow x = \frac{\Delta H_{vap_{HP}}}{\Delta H_{vap_{MP}}} = \frac{1698}{2048} = 0.83\ kg$$

$$C_{MP} = 0.83 \cdot 0.24 = 0.1992\ SEK/kg = 422\ SEK/MWh$$

LP steam is only generated internally on the site. Loads above what is available from flashing MP condensate will have the same price as MP steam, since that is where it originates according to Figure 2.4.

The savings from reducing electricity consumption in the fan engines, if the demand for air cooling is decreased on the process stream side, is estimated based on the assumption that the two are proportional according to 23.2.

$$Electricity\ cost\ reduction =$$

$$= \frac{Load\ reuction\ in\ HX14\ \ or\ HX19\ on\ process\ side\ (kW)}{Total\ process\ load\ in\ HX14\ or\ HX19\ (kW)} \cdot$$

$$\cdot\ C_{tot_{electricity\ used\ by\ fan\ engines\ in\ HX14\ or\ HX19}}\ (SEK/yr) \qquad \left[\frac{SEK}{yr}\right] \tag{3.4}$$

The price for CW is based on the assumption that the temperature difference over each HX in the plant corresponds to the temperature difference over the cooling tower. It also assumes that all CW has the same price as crude water and a density of 1000 kg/m^3.

$$C_{crude\ water} = 0.45\ SEK/m^3$$
$$\delta_{water} = 1000\ kg/m^3$$
$$Cp_{water} = 4.18\ kJ/(kg \cdot K)$$

$$\rightarrow C_{CW} = 0.45 \cdot 10^{-3}\ SEK/kg$$

$$\Delta H_{CW} = 4.18 \cdot (33 - 23) = 41.8\ kJ/kg$$

$$\rightarrow CW\ cost\ reduction =$$

$$= \frac{Load\ reduction\ of\ CW\ on\ process\ side\ (kW)}{41.8\ \left(\frac{kJ}{kg}\right)} \cdot 0.45 \cdot 10^{-3}\ \left(\frac{SEK}{kg}\right) \cdot t_{yr}$$

Further utility price calculations are reported in Appendix 1.

4 Results and Discussion

Results from the pinch analysis of the theoretical case showed that there are a number of areas of interest for improving heat recovery within the process. Thereafter, pinch analysis was performed for the operational case. Here it was noted that changing target temperatures for certain streams to the values indicated in the theoretical case could save energy. Assuming that operation is adjusted accordingly, the adjusted operational case was created. Pinch analysis was performed for the adjusted operational case, and specific retrofit suggestions were made in order to achieve energy savings.

4.1 Case data

All the pinch data used in the investigated cases (theoretical case, operational case and adjusted operational case) are presented in Table 4.1. The table should be used as a reference throughout the reading of the results and will also be referred back to in the text. To track the location of each exchanger Table 2.1 should be consulted. The changes in target temperatures in the operational and adjusted operational case are explained in more detail in Section 4.4.

The total heating (18.7 MW, 20.0 MW and 19.2 MW) and total cooling demands (20.4 MW, 18.9 MW and 17.7 MW) for the investigated cases (theoretical case, operational case and the improved operational case respectively) are summarized in Table 4.1.

Table 4.1. Pinch data used for the investigated cases. Numbers within parenthesis are calculated based on the resulting load on the cold streamside. MeOH – methanol; [a] divided into vapour, condensing, and liquid. [b] divided into vapour and condensing. [c] Divided into condensing and liquid.

		Theoretical Case			**Operational Case**			**Adjusted Op. Case**		
Stream	**HX**	**Load** (kW)	**T_{start}** (°C)	**T_{target}** (°C)	**Load** (kW)	**T_{start}** (°C)	**T_{target}** (°C)	**Load** (kW)	**T_{target}** (°C)	**Load** (kW)
A	HX1	20	100	893	40	95	606	40	67	297
B	**HX2**	55.5	79.3	748	26	70	1316	50	67	527
MIX A+B	MeOH	79.3	99.2	333	70	80	163	67	67	0
C	HX3	99.2	210	3860	80	200	3992	67	200	4401
D	HX4	99.6	99.7	1030	96	102	1207	96	102	1207
E	**HX5**	50	73.6	340	60	78	251	50	67	237
F	HX6 HX7	77 100	100 210	703 4035	74 97	97 200	780 3830	67 97	97 200	1043 3810
G	HX8	79.4	117.6	2050	75	132	3322	75	118	3162
H	HX9	107.7	135	420	132	155	489	118	135	411
I	HX10	136	147.1	4300	108	125	3788	108	125	3788
J	HX11	124	124.1	30	124	125	30	124	125	30
K	**HX12**	48	99	131	48	99	131	48	99	131
L	HX13	121	148	111	90	148	171	90	148	171

Total Heating Demand		**18 652 kW**			**20 075 kW**			**19 216 kW**		
MIX A+B	Feed	100	99.2	9.2	94	80	163	67	67	0
					112	77	98	112	77	98
M[a]	HX14	90	50	7907	77	70	7926	77	70	7926
					70	20	1029	70	50	412
N	**HX5** HX15	93 74	74 50	340 400	102 (87)	(87) 60	251 448	102 (88)	(88) 50	237 628
O	**HX2** HX16	135 79	79 40	748 510	155 -	60 -	1316 -	135 97	97 60	527 512
P	HX17	55	15	810	55	15	513	55	15	513
Q[b]	HX18	135	10	231	155	20	19	135	20	16
					20	17	227	20	17	227
R[c]	HX19	77	65	9210	77	65	5965	77	65	5965
		-	-	-	65	25	753	65	50	384
S	**HX12**	149	70	131	149	70	131	149	70	131
T	HX20	35	15	40	35	15	40	35	15	40
U	HX21	104.6	10	81	90	10	51	90	10	51
Total Cooling Demand		**20 409 kW**			**18 930 kW**			**17 667 kW**		

4.2 Theoretical case

4.2.1 Selecting ΔT_{min}

Pinch calculations were performed with the theoretical case data presented in Table 4.1 with different values of ΔT_{min} to investigate possible threshold effects. The results on potential energy recovery within the system from varying the global ΔT_{min} are presented in Figure 4.1. As can be seen, the penalty in “lost” energy recovery potential is relatively constant between each 5ºC-increase of ΔT_{min}, and so the energy penalty was not very significant. Increasing ΔT_{min} obviously affects the required HX area, however, these requirements were not quantified in this project. It was decided to proceed with a minimum difference of 10ºC, since this value corresponds to that implemented in a number of heat exchangers in the existing HEN. Due to time limitations, no other values of ΔT_{min} were investigated in this study.

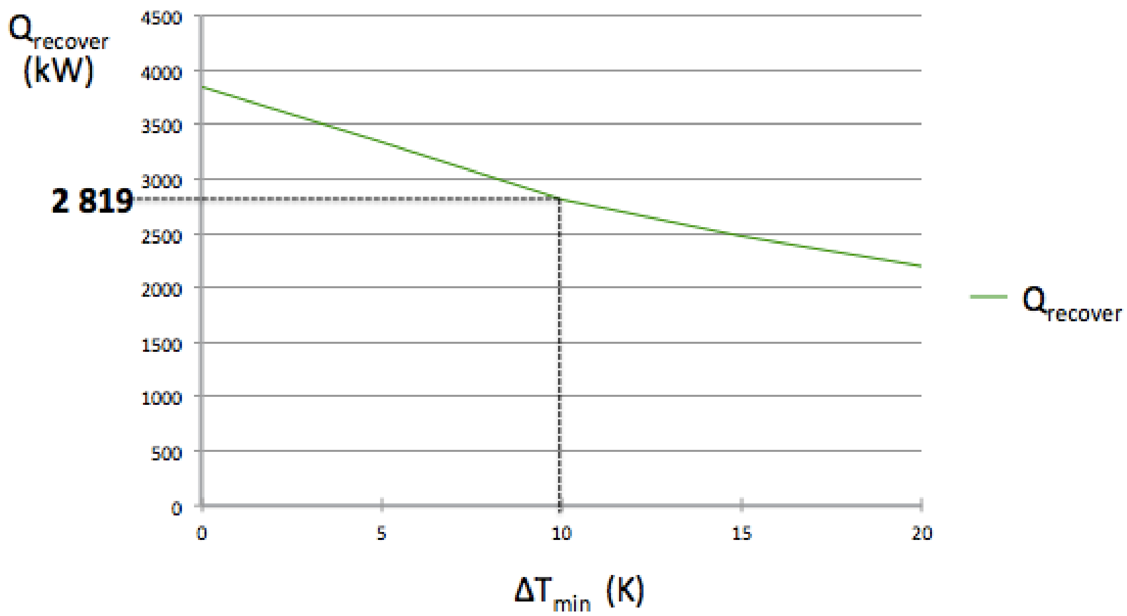

Figure 4.1. Impact of ΔT_{min} on the internal heat recovery potential. The indicated value of recovery potential corresponds to a ΔT_{min} of 10°C

4.2.2 Pinch analysis and initial conclusions

In the theoretical case the total demand for process stream heating is 18 652 kW, and the need for cooling is 20 409 kW according to the data tabulated in Table 4.1. Summing up the HX:s that require some type of utility in the current network HEN (shown in Table 2.1) indicates a hot utility consumption of 17 432 kW and a cold utility consumption of 19 189 kW. Hence, 1 220 kW are recovered through internal heat recovery.

From the data presented for the theoretical case in Table 4.1 the CC was plotted, and is shown in Figure 4.2. The minimum heating demand is 15 833 kW and the minimum cooling demand reach 17 590 kW. The recovery potential is 2 819 kW, hence it is possible to increase the internal heat recovery by 1 599 kW in the theoretical case HEN.

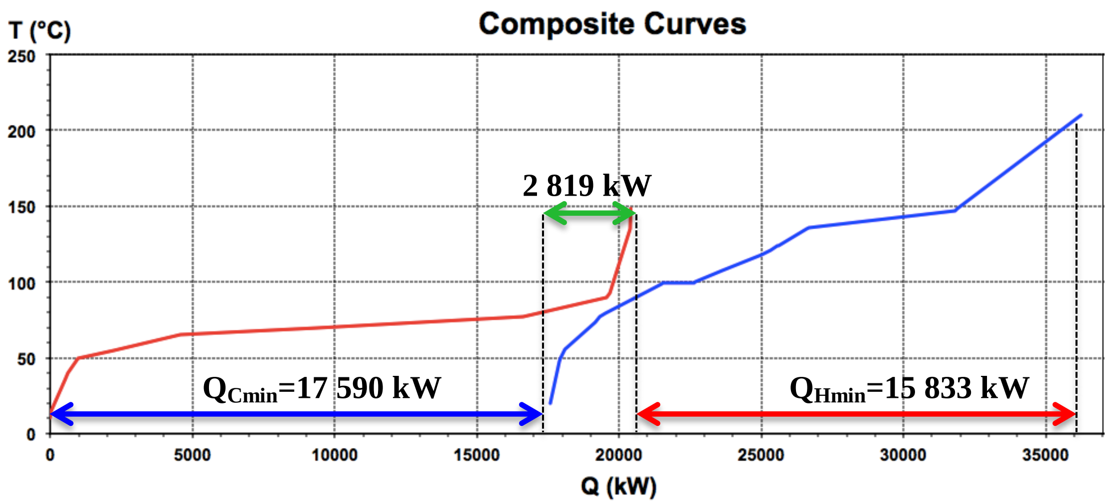

Figure 4.2 The figure shows the hot (red) and cold (blue) CC in real temperatures. ΔT_{min} is 10K.

The pinch violations identified in the theoretical case with a ΔT_{min} of 10°C are shown in Table 4.2.

Table 4.2 Summarization of the pinch rule violations in the current HEN in Table 2.1 using the HX data of the theoretical case in Table 4.1.

Heating Below Pinch		**Transfer Across Pinch**		**Cooling Above Pinch**	
Location	**Quantity** (kW)	**Location**	**Quantity** (kW)	**Location**	**Quantity** (kW)
HX1:	670	HX2:	596	HX18	83
HX6:	107	HX12:	50	HX21	13
HX8:	32	HX5:	48	-	-
Total	808	**Total**	694	**Total**	96

In all nine pinch violations were identified in the current HEN, comprising a total of 1 599 kW. The largest violations are related to feed and methanol preheating in HX1, HX2 and HX6, just before the two reactors in Figure 2.1. It is also interesting to note that all of the existing internal heat exchangers illustrated in Figure 2.3 transfer heat across the pinch. Cooling above the pinch was less of a problem in the theoretical case. Extra attention was given to the heat exchangers highlighted in these results when moving forth to the proceeding cases.

Areas of special interest for energy efficiency improvement

Looking at the hot stream CC (red line) in Figure 4.2, the flat section is mainly a contribution from the condensation of methanol at approximately 1.5 bar (which in the theoretical case is specified to take place between 90-65ºC). The condensation of methanol comprises the largest cooling demand and consists of several streams, mainly originating from the flash system and reflux from the distillation column in stream M and R respectively.

If methanol could be condensed at a higher temperature, i.e. a higher saturation pressure, part of the flat section on the hot CC would shift upwards thus allowing the two CC curves to shift and increase their overlap. Consequently there could be a chance to significantly increase the heat recovery potential within the system. The easiest location to perform this adjustment is preferably in the flash system with methanol going to HX14 in stream M, since flashing involves a single separation step. Increasing the pressure in the column to integrate methanol condensing in HX19, will affect all of the separation steps taking place on each tray, altering product purity and column operation.

The pinch temperature location did cast some light on the mixing points. The mixing points were non-isothermal and so one stream heats the other as they blend. The two largest mixing points are where methanol is combined with feed before the two reactors, see Figure 2.1. Quantifying the first one gave inconsistent results, as seen in the two loads labelled “Mix A+B” on the hot and cold side in Table 4.1. Since their temperatures were in the region around the pinch, and might transfer heat across it, the new pre-heating targets were set to match the cold pinch temperature to then blend isothermally instead.

4.3 Operational case

Data from process measurements was analysed to establish energy targets for the operational case. This required adjustments of mass flow and temperature values. The resulting temperatures and loads are presented in Table 4.1.

4.3.1 Pinch calculations

In the operational case the plant required 20 075 kW of heat and 18 930 kW of cooling. With the current HEN the plant consumes 18 215 kW of hot utility and 17 070 kW of cold utility, and consequently 1 862 kW are recovered internally. Pinch calculations resulted in a minimum heating demand of 16 509 kW and a minimum cooling demand of 15 363 kW, so an addition of 1 704 kW could be recovered internally.

The pinch temperature is 67ºC for cold streams and 77ºC for hot streams, compared to 80ºC and 90ºC respectively in the theoretical case. The reason for the significant decrease of pinch temperature was because the methanol in stream M and R was restricted to the temperature span for gas cooling, condensation and liquid cooling according to common pinch analysis procedures (Kemp, 2006). In the theoretical case the condensation was assumed to take place throughout the whole interval between 90-50ºC, stated in Table 4.1. Consequently the pinch temperature was lowered to the probable interval for phase change, since the largest cooling load corresponds to methanol condensation.

4.4 Comparison of theoretical and operational cases

Some general points for revision appeared when comparing how the site was planned to run in the theoretical case with the target set points in the operational case. The changes discussed in the following sections constitute the new set of data making up the adjusted operational case in Table 4.1.

4.4.1 Increased methanol tank temperature

The methanol tank temperature is considerably lower, 25°C instead of 55°C. The lower temperature could stem from the lack of controllability of the two air-coolers in HX14, which is either on or off as stated in Section 2.2.2. By introducing frequency controllers that vary the fan speed in order to meet a specified value of outlet temperature, the tank temperature could be higher. This would consequently lower electricity consumption in the fans. The benefit by doing this is not only a reduced cooling demand but also reduced heating, since immediately after the storage tank the temperature of stream B is raised again in HX2. The difference in methanol vapour pressure between the two temperatures is 0.46 bar (Goodwin, 1987). The tank and the security vent are designed to withstand pressures up to 4.5 bar, and in operation tank pressure reaches approximately 2.5 bar. Therefore the security vent to the flare should manage to keep the gases within the tank instead of releasing them to the flare.

4.4.2 Sub-cooling of reflux in the distillation column

The sub-cooling of methanol also applies to the distillation column reflux in HX19, which in the operational case also has a measured return temperature of 25°C. The reason for this might be to reach the intended methanol tank temperature. Changing this target is of a somewhat more complex nature then adjusting the methanol tank temperature. However, to match the new target in the tank the reflux temperature was increased in the adjusted operational case, as stated in Table 4.1. Depending on the column regulation system it could however mean that more reflux needs to be returned to the column to maintain purity specifications in the distillate and bottom fraction. By raising the target temperature in HX19, a higher mass flow could compensate for the lower degree of cooling. This is not regarded in the adjusted operational case.

4.4.3 Increased temperatures in the flash system

In the flash system the target values for the outlet temperatures from HX8 and HX9 have been raised by approx. 25°C, compared to the theoretical case as can be seen when comparing the data in Table 4.1. A probable reason for the elevated temperatures is to get a lower methanol content in the final product, see Figure 2.1. However, simulations in Aspen Plus imply that by raising temperatures this much, there is a risk of increasing the concentration of single bonded methyl esters in the flash tank top fraction. One possible option to prevent this effect is to raise the flash pressure as well. In that way the ester would stay in the bottom flow, and the leaving methanol gets a higher pressure, hence would condensate at a higher temperature. This would turn part of the cooling load in HX14 into potentially useful heat if the network were to be changed, as discussed in Section 0. A second option is to decrease temperatures back to levels set in the theoretical case and save MP steam, which is the option performed in this study for the adjusted operational case.

4.4.4 Non-isothermal mixing

In order to eliminate unnecessary risks of pinch violations in the larger mixing points, the pre-heating should only proceed to the cold pinch point and then blend. By doing this the pre-heat loads can be satisfied by internal heat exchange in the retrofit, before they blend, and then enter to a unit using hot utility above pinch. This would minimize the number of units needed to straighten up the violation. It decreases the loads in the pre-heaters but increases the load in the final heating steps HX3 and HX7 using HP steam. The adjustments would be according to the pinch rules, so that one HX unit can take care of the whole pre-heat load by internal heat recovery in a retrofit. However, it might not be as beneficial to only change these temperatures without making additional changes to the HEN.

4.4.5 Energy savings associated with adjustments in temperature

Each suggestion for change of operation temperature in the operational case above was associated with a change in energy use. These can be seen in Table 4.3. The new temperature targets created the adjusted operational case. If all the adjustments suggested in this chapter were to be implemented, the plant could directly reduce its energy demand with 2 123 kW. The operational case and the adjusted operational case rely on the same mass balance, so that the loads are comparable and stem only from the effects of changed temperature targets.

Table 4.3. The table shows how the load in each exchanger involved change when the temperature adjustments to the operational case discussed above are implemented to become the adjusted operational case.

		Operational Case			Adjusted operational Case			Change in Load	Reason behind changed target
Stream	**HX**	T_{start} (ºC)	T_{target} (ºC)	Load (kW)	T_{start} (ºC)	T_{target} (ºC)	Load (kW)	ΔQ (kW)	
A	HX1	40	95	606	40	67	297	-308	4.4.4 Non-isothermal mixing
B	HX2	26	70	1316	50	67	527	-790	4.4.4 Non-isothermal mixing 4.4.1 Increased methanol tank temperature
MIX A+B	MeOH	70	80	163	67	67	0	-163	4.4.4 Non-isothermal mixing
C	HX3	80	200	3992	67	200	4402	+410	4.4.4 Non-isothermal mixing
E	HX5	60	78	251	50	67	237	-14	4.4.4 Non-isothermal mixing
F	HX6 HX7	74	200	4610	67	200	4853	+243	4.4.4 Non-isothermal mixing
G	HX8	75	132	3322	75	118	3162	-160	4.4.3 Decreased temperatures in the flash system
H	HX9	132	155	489	118	135	411	-78	4.4.3 Decreased temperatures in the flash system
Change in Heating Demand:								**-860**	
MIX A+B	Feed	94	80	163	67	67	0	-163	4.4.4 Non-isothermal mixing
M	HX14 liq	70	20	1029	70	50	412	-617	4.4.1 Increased methanol tank temperature
N	HX5 HX15	102	60	699	102	50	865	+166	Achieve better separation
O	HX2 HX16	155	60	1316	135	60	1039	-277	4.4.3 Decreased temperatures in the flash system
Q	HX18 gas	155	20	19	135	20	16	-3	4.4.3 Decreased temperatures in the flash system
R	HX19 liq	65	25	753	65	50	384	-369	4.4.2 No sub-cooling of reflux (4.4.1 Increased methanol tank temperature)
Change in Cooling Demand:								**-1263**	

Energy savings associated with adjustments in temperature might not automatically be a financial benefit with the current HEN, but it gives an insight on where energy saving potentials due to changes in operation can be found if changes were to be done.

4.5 Adjusted operational case

4.5.1 Pinch analysis

In the adjusted operational case the total heating demand is 19 216 kW and the cooling demand is 17 667 kW. Summing up the utility consumption in the current HEN, using the data of the adjusted operational case, gave 18 321 kW of consumed hot utility and 16 772 kW of cold utility, and consequently 895 kW is recovered internally.

Figure 4.3 illustrates the resulting CC of the adjusted operational case. The pinch interval temperature is still 72ºC. This is because the saturation pressure of the condensing methanol is the same in the operational and improved operational cases. The minimum hot utility demand is 16 652 kW and the minimum cold utility demand is 15 103 kW, and it would be possible to recover 2 564 kW through internal HX.

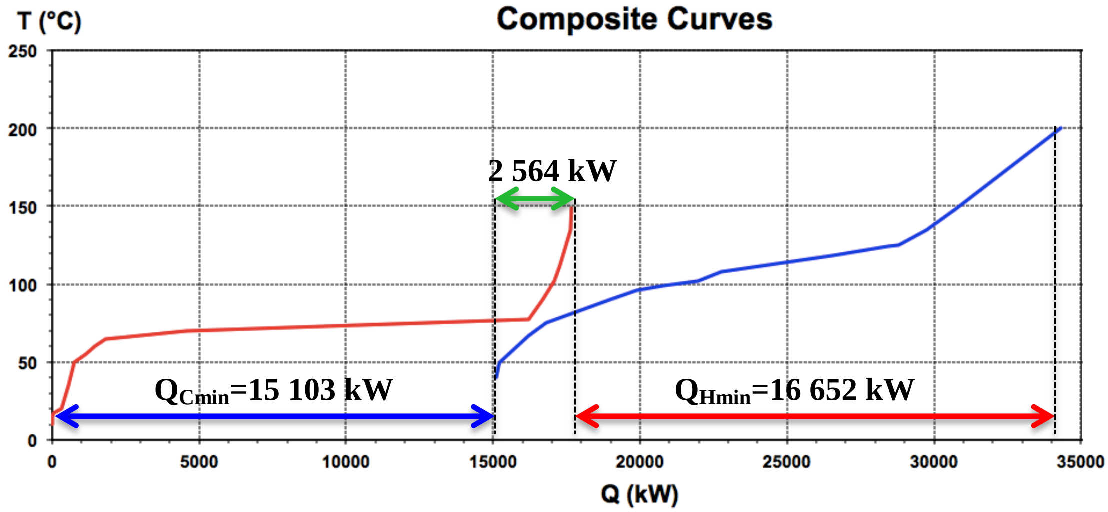

Figure 4.3 Hot (red line, streams to be cooled) and cold (blue line, streams to be heated) CC for the adjusted operational case. ΔT_{min} is 10K.

The pinch violations identified in the adjusted operational case with a ΔT_{min} of 10°C are shown in Table 4.4 below. They total pinch violation load is 1 669 kW. The network in place on the violating streams is illustrated in Figure 4.5.

Table 4.4 Pinch violations in the current HEN, if the process were to be run according to the adjusted operational case. To refer back to associated streams see Table 2.1.

Heating Below Pinch		**Transfer Across Pinch**		**Cooling Above Pinch**	
Location	**Quantity** (kW)	**Location**	**Quantity** (kW)	**Location**	**Quantity** (kW)
HX1:	297	HX2:	527	HX16	277
		HX5:	237	HX15	179
		HX12:	46	HX18	8
				HX14	98
Total	297	**Total**	810	**Total**	562

Pre-heating of feed in HX1 constitutes a smaller pinch violation than shown in Table 4.2 for the theoretical case, since the fact that the storage tank is heated now has been taken into consideration. Since the pinch temperature is lower, cooling above the pinch has merged as a greater problem than indicated in the pre-study of the theoretical case. In the theoretical case the problem was only restricted to the internal heat exchange in HX5 and HX2, which both transferred heat across the pinch. With a lower pinch temperature the following coolers HX15 and HX16 (see Figure 2.3), which ensures that streams O and N reach their target temperature, ends up partly operating on the wrong side of the pinch as well. Out of their original load shown in Table 4.1, 28.5% and 54% now constitute pinch violations, respectively. The other two places where cooling take place above the pinch is when the methanol flash streams M and Q are cooled down to condensation temperature (in HX14 and HX18, see Table 4.1). Since these violations are so small they will not be dealt with further when retrofits are considered in Section 4.7. The same applies to streams K and S in HX12 illustrated in Figure 2.3. Consequently the retrofit networks will be 152 kW away from reaching MER, and 1 517 kW will be recovered in each retrofit. This corresponds to 91% of the total pinch violations in the adjusted operational case HEN.

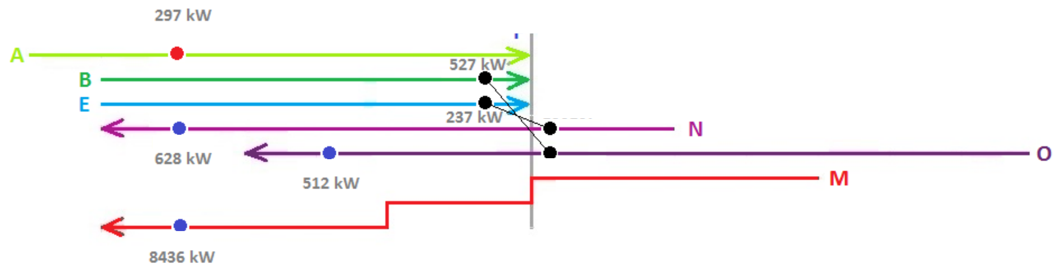

Figure 4.4 The HX that constitutes a violation and their loads are illustrated. The size of the violation each HX comprise can be seen in Table 4.4. In this figure HX18 and HX12 are excluded.

Areas of special interest for energy efficiency improvement (continued)

Looking in Figure 4.3, the distinct flat section of the hot composite curve remains. This section is the methanol condensation, and the observations made in the theoretical case apply for the adjusted operational case as well. Elevating part of the condensation load to a higher temperature shows a potential to internally recover some of the condensation heat above the pinch. The methanol condensation pressure plays a key role in how much heat that is possible to recover in the process. Since the cold composite curve is less steep the benefits of a higher condensation pressure is likely to become even greater here than in the theoretical case. Therefore a quick assessment of the potential of recovering more heat of condensation will be made in Section 4.9, but a more detailed investigation is outside the scope of this thesis work.

4.6 New stream system boundary for retrofit consideration

Energy efficiency in the HEN for the adjusted operational case is increased by the removal of utility usage or stream matches constituting the violation. Instead the streams are re-matched, so that the heat they contain is properly utilized by other available streams above and below the pinch with a heating or cooling demand. The different criteria used when deciding whether to include or exclude a stream in a retrofit have been mentioned earlier in Section 3.3.2

All HX units located in inaccessible areas (and their connected streams) were omitted from the retrofit study. Two exchangers, HX13 and HX20 are situated high above ground, along the distillation column. Considering that they are not any direct targets for pinch violation and have low loads, the streams involved in the HX units (L and T) will be omitted from the system. Stream J and stream U are also excluded, due to their low load reported in Table 4.1.

Another stream that preferably should not be heat integrated is the reflux in the distillation column condenser (stream R, exchanger HX19). The column is mainly regulated through adjusting the reflux, so the available load varies, and maintaining a high degree of freedom is important for column controllability.

The streams that will not be considered in the retrofit of the adjusted operational case are summarized in Table 4.5, and their location is shown in Figure 2.1. Note that the table also includes the smaller pinch violations discussed in the previous section.

Table 4.5 The streams excluded for consideration in a retrofit and their respective load.

Stream	HX	Reason for Exclusion	Load (kW)	Stream	HX	Reason for Exclusion	Load (kW)
J	HX11	Small Load	30	Q	HX18	Small Violation	16
K	**HX12**	Small Violation	131				227
L	HX13	Inaccessible	171	S	**HX12**	Small Violation	131
				T	HX20	Inaccessible	40
				U	HX21	Small Load	51
				R	HX19	Controllability	(6 346)
Total			332	Total			465

The omitted units (excluding HX19 which is omitted due to controllability issues) account for 1.7% of the total heating demand and 2.6% of the total cooling demand in the adjusted operational case. Thus, this will not have a major impact on the final conclusions of the retrofit study.

4.7 Retrofit study

The streams left considered for retrofit are shown as coloured arrows in Figure 4.5. Streams that constitute pinch violations and that were considered for re-matching are shown with individual colours. The arrows run between the start and target temperature for the respective stream, with the stream name indicated by the starting temperature. For reference data see Table 4.1.

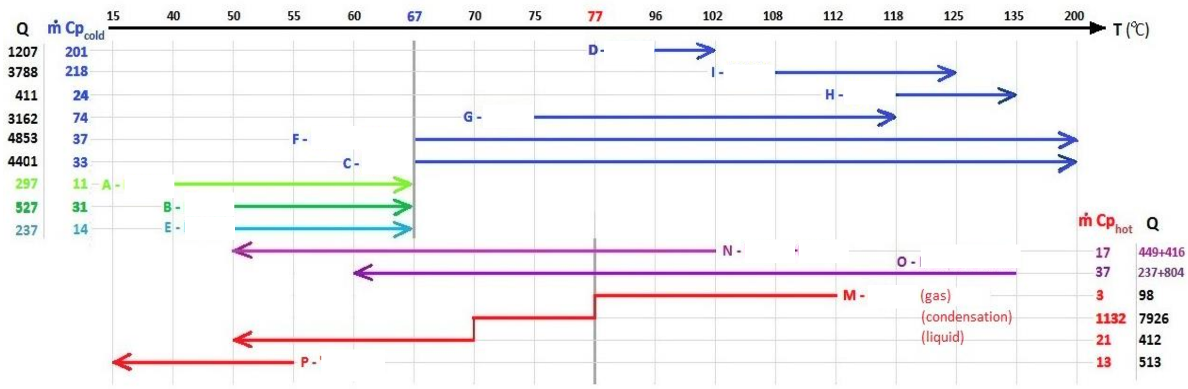

Figure 4.5 The figure illustrates the streams considered for retrofit. The loads (kW) and mCp-values (kW/K) are included in the margins, to the left for cold streams and to the right for hot streams. Individually coloured arrows comprise the pinch violations that were decided to consider for a retrofit in the adjusted operational case. The arrows in red and blue represent the other process streams available for a re-match.

The new matches were combined according to the pinch rules explained in Section 3.3.2. In order to break the pinch violation with a hot stream above pinch, heating a cold stream below the pinch, cold streams were integrated with hot streams below or above pinch. Matches identified as possible were combined into several retrofit suggestions. The different retrofit suggestions were then evaluated based on: targeting for minimizing the total HX area, or to use already existing units to a larger extent (i.e. minimizing the new HX area), or targeting for maximized utility cost savings. Since the interesting pinch violations for retrofit consideration already was determined in Section 4.5.1 all retrofit suggestions will save the same amount of energy (1 517 kW).

In order to estimate the new overall heat transfer coefficient (U) for each new stream combination the diagram shown in Appendix 3 was used. This was required in order to target the required HX area by using equation 3.2.

4.7.1 Below the pinch point

Cold streams that needed to be heated to the pinch point by internal heat recovery include streams A, B and E. Matching any of them with the hot stream P is not possible since this stream must be cooled to such a low temperature that none of the cold streams were sufficient. Correspondingly, stream P is too cold to heat any of the pinch streams A, B and E to their pinch point target temperature. Because neither of the streams could be ticked off, stream P was excluded from the retrofit study at this stage.

Stream M is the largest of the three remaining hot streams (M, N and O) with 7 926 kW of heat available. Streams N and O have 449 kW and 236 kW of heat available below the pinch, respectively according to the margin in Figure 4.5. The “tick-off rule” implies that N could be matched with A or E but not B. Steam O is only barely sufficient to heat E. Note that this match (O-E) resembles the one already in place in HX5. However, in the retrofit the hot stream O enters at the hot pinch temperature instead, to avoid the previous pinch violation. Stream M is the only option for pre-heating methanol in stream B so that it can be ticked off. The small load of 98 kW comprising the initial gas cooling of stream M down to condensation would remain a pinch violation though (all according to Table 4.4 for the violation in HX14), since the cooling would start above the hot pinch temperature. However, stream M could potentially supply heat to all three streams without a significant change in temperature, since it is a condensing stream. This high condensation load constitutes a large potential for heat integration, and it was therefore retained as a candidate hot stream.

All possible matches discussed above are in accordance with the “$\dot{m}Cp$ rule” (see Section 3.3.2), which is important since all these matches are located at the pinch, i.e. all of the streams in consideration require heating and cooling to the pinch temperature. An overview of the possible matches with resulting new total areas is presented in Table 4.6.

Table 4.6. When the “tick- off rule” and “mCp rule” were followed, the number of possible matches narrowed down to the ones presented in this table. The streams have the same colour as given in Figure 4.6 for simple identification.

	Below the Pinch					
Match	**Cold Stream**	**Hot Stream**	**Q** (kW)	**ΔT_{lm}** (K)	**U** (kW/m^2/K)	**A_{hx}** (m^2)
1	A	N	297	14.1	0.273	77
	B	M	527	16.9	0.490	63
	E	O	237	9.9	0.450	53
2	A	M	297	20.5	0.304	48
	B	M	527	16.7	0.490	63
	E	N	237	11.3	0.450	47
3	A	M	297	20.5	0.304	48
	B	M	527	16.7	0.490	63
	E	M	237	16.4	0.518	28

4.7.2 Above the pinch point

The only streams in Figure 4.5 that are cold enough to cool the hot streams N and O, entering decanter 1 and 2, down to the hot pinch temperature were streams C and F, fed to reactors one and two respectively, as illustrated in Figure 2.1. The reactor streams contain enough heat to tick off streams N and O, and their $\dot{m}Cp$-values are such that ΔT_{min} is maintained throughout the exchange.

Table 4.7 The hot streams N and O require new matches in order for them to be cooled above the pinch. In the table the possible combinations and their estimated required heat exchanger area are shown. Their respective stream representation can be seen in Figure 4.6.

	Above the Pinch					
Match	**Cold Stream**	**Hot Stream**	**Q** (kW)	**ΔT_{lm}** (°C)	**U** (kW/m^2/K)	**A_{hx}** (m^2)
4	C	N	416	15.4	0.377	72
	F	O	804	23.6	0.450	76
5	F	N	416	15.8	0.450	60
	C	O	804	22.8	0.377	94

4.7.3 Targeting for minimum HX area

The matches above and below the pinch were combined into six different retrofit suggestions in all, presented in Table 4.8. The new HX areas in Table 4.6 and Table 4.7 are rough estimations of the new heat transfer area required, given the approximations used to estimate the U-value in the new stream match. Retrofit suggestion 3 (combining match 3 and 4) shows the smallest totally required heat exchanger area.

Table 4.8. Summary of the estimated heat exchanger areas required for the possible stream matches above and below pinch. Retrofit 3 (combining match 3 and 4) would give the retrofit with the smallest heat exchanger area and is highlighted in green.

Retrofit	**1**	**2**	**3**	**4**	**5**	**6**
Match	**1+4**	**2+4**	**3+4**	**1+5**	**2+5**	**3+5**
A_{hx} (m^2)	341	306	287	347	312	293

4.7.4 Opportunities for re-using existing HX units

The discussion in this chapter was based on the assumption that existing HX:s can retain their original location with respect to one of the streams being exchanged in the existing HEN, and that a new stream can be exchanged with this stream without further restrictions (as illustrated in Figure 4.6-Figure 4.9). The opportunity for re-using an existing HX is noted where applicable.

Table 4.9 The table shows what streams that are heat exchanged in existing HX:s involved in the HEN. In the next column the retrofit options for each stream involved are shown. These are the new streams that could run through the existing HX after the retrofit.

Existing HX	**Stream Match** In Existing HEN	**Stream Match** / In Retrofit (nr.)			
HX1	A / LP_{cond}	A / N A / M		(1, 4) (2, 5)	
HX5	N / E	N / A (N / E) N / C N / F	(1) (2) (1,2,3) (4,5,6)	E / O E / N E / M	(1, 4) (2, 5) (3, 6)
HX2	O / B	O / E O / C O / F	(1) (1,2,3) (4,5,6)	B / M	(1, 2, 3, 4, 5, 6)
HX15	N / CW	Same			
HX16	O / CW	Same			

Retrofit suggestion 1

Retrofit 1 combines matches 1 and 4, shown in Table 4.6 and Table 4.7. The complete network is illustrated in Figure 4.6.

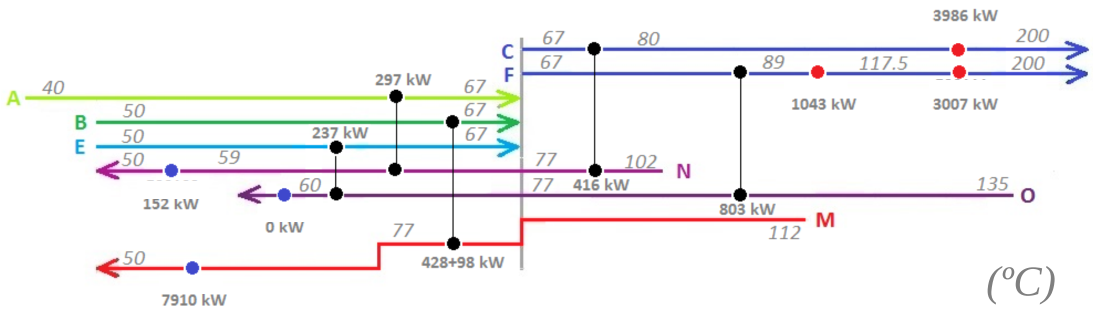

Figure 4.6 New heat exchangers and their respective loads for retrofit 1. The new inlet and outlet temperature of streams are indicated with grey italic numbers in °C. The new loads in each HX are indicated with bold grey typing in kW.

Retrofit 1 in Figure 4.6 still includes stream N in cooler HX15 like before (see Table 2.1). The new cooling load is 152 kW, compared to the previous load of 628 kW (see Table 4.1). Therefore it will probably be able to handle the new retrofit adjustments, even though the inlet temperature of stream N will be changed. The new load in HX16 is zero, and so this exchanger would be kept as backup for stream O in this retrofit.

Probably none of the old exchangers are appropriate to operate with the new matches above the pinch, since this involves high-pressure streams C and F. Therefore new equipment would have to be built in the retrofit network.

Retrofit suggestion 2

In retrofit 2, match number 2 and 4 is combined. The complete network is illustrated in Figure 4.7.

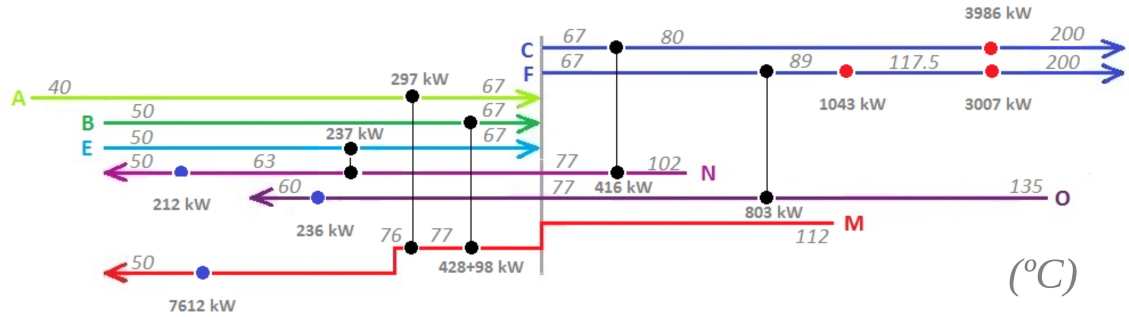

Figure 4.7 New heat exchangers and their respective loads for retrofit 2. The new inlet and outlet temperature of streams are indicated with grey italic numbers in °C. The new loads in each HX are indicated with bold grey typing in kW.

Note that the previous HEN with stream N and E in HX5 is preserved (see Table 2.1). The only difference is that stream N now enters at a lower temperature level, which decreases ΔT_{lm} in comparison to the existing operation conditions in Table 4.9. Therefore the old equipment could be used, but would probably need to be extended. However, since this is a plate HX, adding more plates to the existing ones could quite easily be done.

HX15 now exchanges 212 kW with stream N, and HX16 needs to remove 236 kW from stream O according to the loads indicated in Figure 4.8. Both these loads are lower than what has been handled in the previous network (see Table 4.1), and complementing with extra units would probably not be necessary in retrofit 2.

Retrofit suggestion 3

In this retrofit match 3 and 4 are combined. The new network is illustrated in Figure 4.8. The only difference from retrofit 2 is that stream E is matched with stream M (condensing methanol).

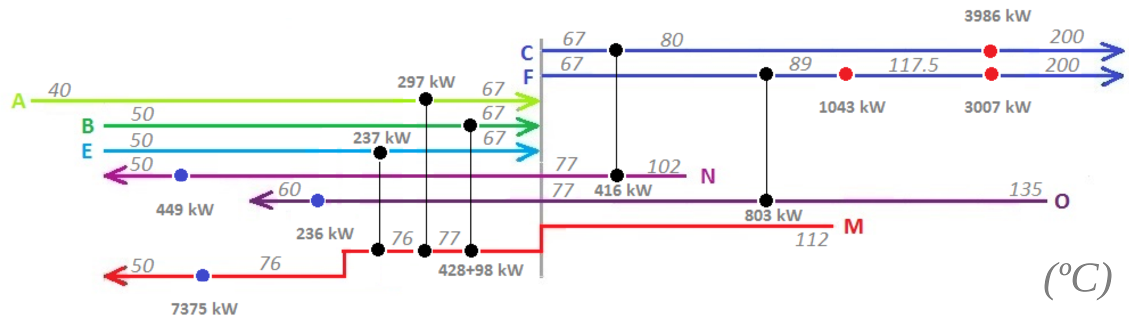

Figure 4.8 New heat exchangers and their respective loads for retrofit 3. The new inlet and outlet temperature of streams are indicated with grey italic numbers in °C. The new loads in each HX are indicated with bold grey typing in kW.

Retrofit suggestions 4, 5 and 6

The retrofits include the match between 1+5, 2+5 and 3+5. The difference in these networks is streams N and O. They are swapped and exchange heat with streams F and C, respectively, see Figure 4.9 and Appendix 4 for detailed networks.

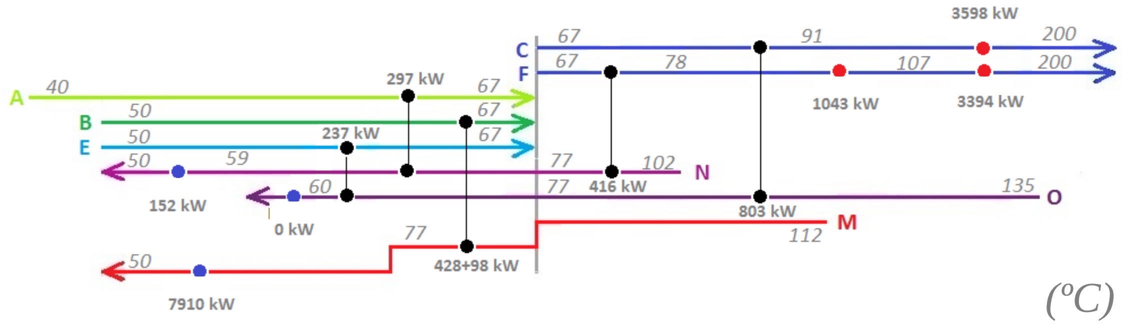

Figure 4.9 New heat exchangers and their respective loads for retrofit 4. The new inlet and outlet temperature of streams are also indicated with grey italic numbers

4.8 Hot utility level optimization

As mentioned in the GCC-theory in Section 3.3.1, there is an opportunity to save hot utility costs, by shifting the utility load pressure levels to the lowest possible. However the GCC is somewhat misleading to use for this purpose, since it depicts the minimum heating and cooling demand of the streams assuming MER between the hot and the cold streams already are achieved. Since the utility level optimization aimed at comparing how the loads were distributed on the streams using hot utility only in the current HEN, these streams were combined in a CC, see the blue curve in Figure 4.10 and Figure 4.11.

The hot utilities at their current temperature levels (see Table 2.2) and loads in the adjusted operational case (see Table 4.1) were additionally included as red lines in Figure 4.10 and Figure 4.11. The orange lines represent the hot utilities at optimized temperature levels (i.e. the lowest required). It should be noted that the results in this section are based on rough estimations of the effects of change in the steam distribution system and a more thorough analysis is required for more accurate results.

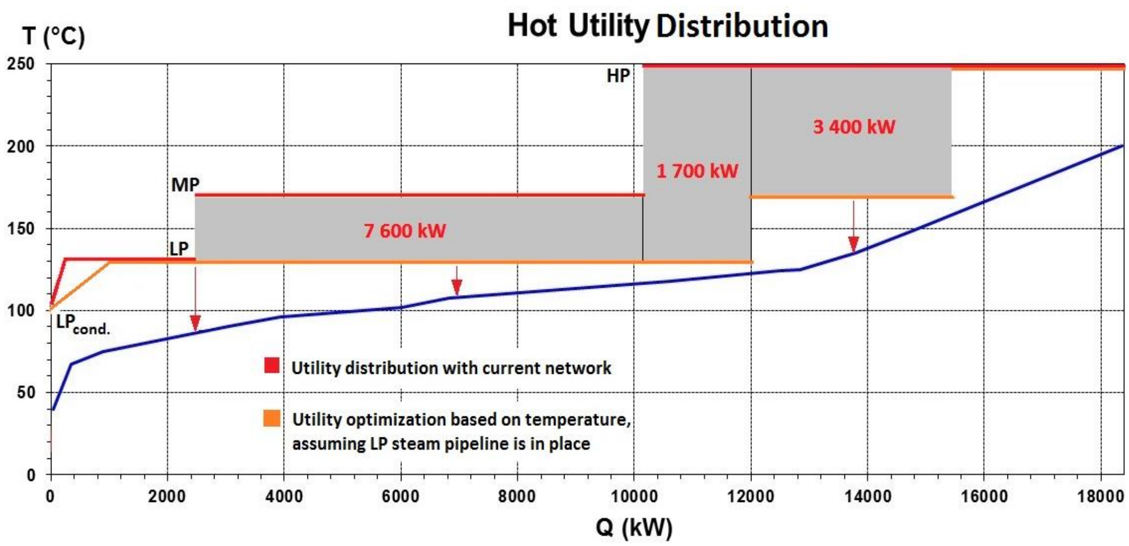

Figure 4.10 The figure illustrates how much of the distributed steam loads that could be supplied at lower temperature levels if LP steam were readily available on the site. ΔT_{min} = 10ºC

As illustrated in Figure 4.10, 3 400 kW of HP steam load could be supplied with MP steam, corresponding to lowering the steam mass flow with 0.34 kg/s (based on enthalpy difference, see Table 4.13). The potential for utilizing LP utility steam in the process is so large that it also could be possible to shift 1 700 kW of the HP steam load down LP steam, corresponding to a lowered steam mass flow of 0.22 kg/s. The largest utility shift is accomplished by using 7 600 kW of LP steam instead of MP steam. However, this only corresponds to a steam flow reduction of 0.2 kg/s.

To accomplish these results there would have to be an LP steam pipeline from the common utility system to the plant in place, but, as illustrated in Figure 2.4, there is not. Also, significant investments in HX units adapted for LP steam heating would be required to supply the hot utility according to optimum. This would in principle involve one investment on every continuous stream on site, and is probably not a realistic option to consider.

The lack of LP steam supply to the plant was taken into consideration in Figure 4.11, which only maximized the LP condensate and LP steam available from flashing MP condensate[1]. The MP steam load was then maximized until it reached a ΔT_{min} of 10ºC with the process streams in the blue curve. Lastly, the remaining load had to be covered by HP steam.

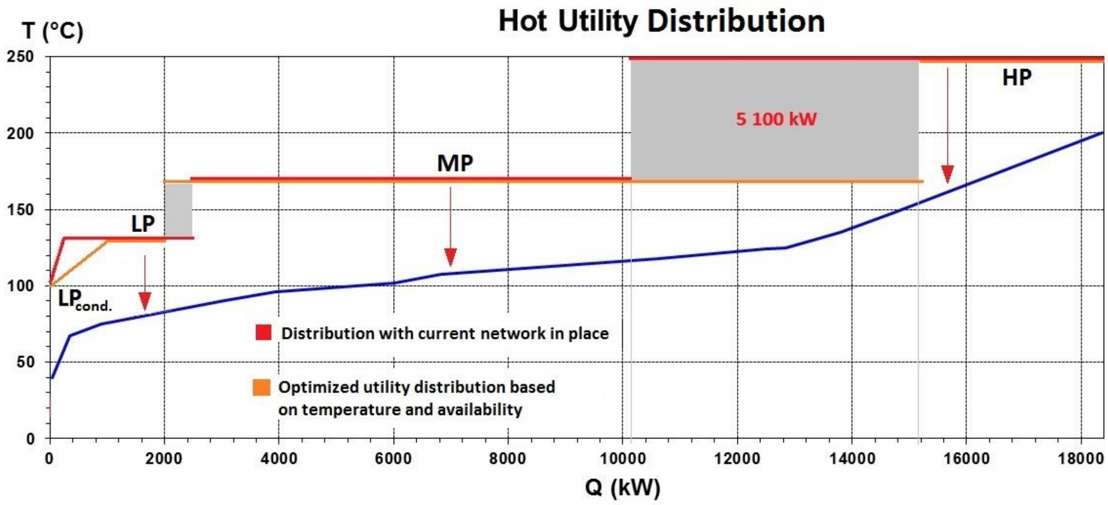

Figure 4.11. The curve illustrates how much of the HP steam load that could be supplied by MP steam instead, based on temperature levels of the process streams.

According to the graph in Figure 4.11, a total of 5 100 kW HP steam could ideally be replaced with MP steam corresponding to a decrease in steam mass flow of 0.51 kg/s, as reported in Table 4.13. Installing two MP steam heaters on stream C and F in Figure 2.1 would accomplish this shift, since HP steam is only supplied to HX3 and HX7 before the reactors, see Table 2.1. This option is therefore more attractive than the previous one, from an investment point of view. A summation of the results from the hot utility level optimization is tabulated in Table 4.12 and Table 4.13.

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[1] Note that the LP and LP_{cond} levels represent what is available with the consumption of HP and MP steam in the adjusted operational case, the results from the calculations are shown in Appendix 1. Iterations will be needed to decide what actually is available in the optimized system.

Table 4.10 A summation of how much hot utility that is consumed by the current HEN in the adjusted operational case, and how much the HEN could consume if it was optimized according to the orange lines in Figure 4.11 and Figure 4.12. The loads in the optimized networks are targeted graphically and are approximated values.

Hot Utility	Ad. Op. Case HEN Load (kW)	Optimized Load With LP Steam (kW)	Optimized Load Without LP Steam (kW)
HP	8 211	3 200	3 200
MP	7 562	3 200	13 200
LP	2 250	11 000	1 000
LP_{cond}	297	1 000	1 000

Table 4.11 Summary of the graphical findings from performing utility load optimization. The steam level shifts indicated above the bold line in this table is associated to Figure 4.11 and would be possible if there were LP steam supplied to the plant. The steam level shift below the bold line in this table is based on what is possible with the current steam network shown in Figure 2.4. Calculations are based on the load reduction, targeted graphically, and the difference in enthalpy of condensation between the steam levels.

Steam Level Shift	Load (kW)	Steam Flow Reduction (kg/s)
HP→MP	3 400	$3\,400 * (1/2048 - 1/1698) = 0.34$
HP→LP	1 700	$1\,700 * (1/2164 - 1/1698) = 0.22$
MP→LP	7 600	$7\,600 * (1/2164 - 1/2048) = 0.20$
HP→MP	5 100	$5\,100 * (1/2048 - 1/1698) = 0.51$

4.9 Energy savings associated with modifying the condensation pressure of methanol

The influence of an elevated condensation temperature of methanol in the flash systems has also been discussed as a potential energy saver earlier in the text. The process flow is throttled from a high to a low pressure level immediately after the two reactors. This is done so that a maximum amount of methanol evaporates in the flash step.

Instead of throttling methanol directly down to the low pressure level in the flash, it could be flashed to an intermediate pressure level first. At higher pressures the saturation temperature (i.e. the condensation temperature) is raised. This increases the potential for the heat of condensation to be at useful temperature level (i.e. to heat a cold stream). In order to investigate the potential to introduce an intermediate flash step, simulations of the flash tank were performed in Chemcad. The results from this flash simulation are shown in Appendix 7.

The methanol was flashed after reactor 1 and 2 (see Figure 2.1). The simulation was performed at a pressure of 8 bar corresponding to a condensation temperature of 135ºC. The vapour stream mass flow became 2.2 kg/s from each flash. The streams released 4 400 kW of condensation heat, which now could be utilized in the process.

By introducing two intermediate flash steps, the downstream mass flow rates were additionally changed. After the most obvious flows downstream were adjusted accordingly in the adjusted operational case, the process had a total heating demand of 17 747 kW and a total cooling demand of 17 020 kW.

Results from pinch calculations can be seen below in Figure 4.12. The CC shows that the adjusted operational case with two intermediate flashing steps has a minimum heating demand of 11 097 kW and a minimum cooling demand of 10 370 kW. The potential heat recovery could then be increased to cover 6 650 kW in all, where 4 400 kW is supplied by condensing methanol. How this potential could be recovered in practice was not investigated further in this study since it would require a new pinch analysis and retrofit study on the new stream system.

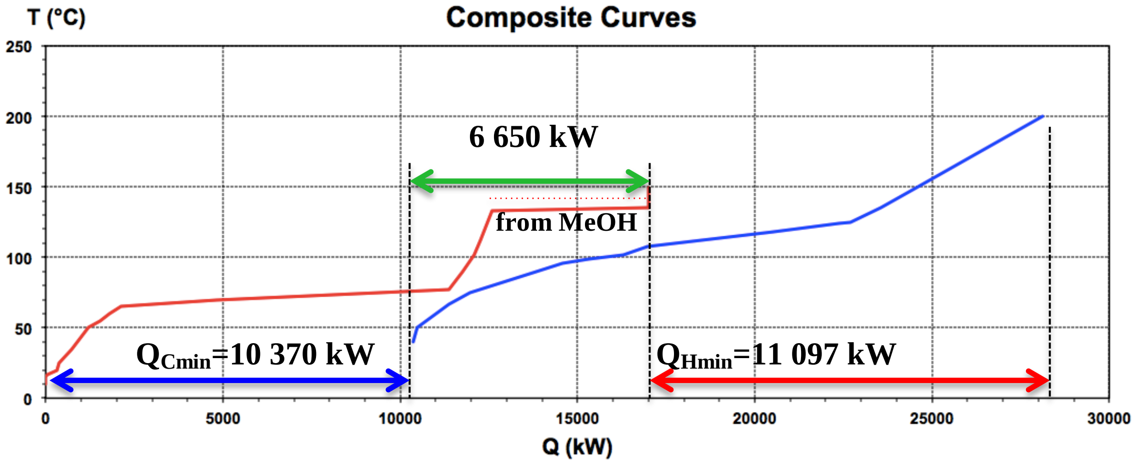

Figure 4.12 The figure illustrates how part of the hot CC is elevated by the methanol that condensates at 135ºC. It enables a higher degree of internal heat exchange to take place.

The pinch temperature is still 66ºC for cold streams and 77ºC for hot streams, as in the operational- and adjusted operational case without intermediate flashing. This is because the methanol reflux in HX19 remains the same. Changing column pressure to utilize condensation heat here as well would alter the whole distillation unit operation and is not recommended without detailed further investigations.

4.10 Utility cost savings

Retrofit

The retrofits presented in Section 4.7 are composed of different combinations of match 1-5 in Table 4.7 and Table 4.8. Each stream match is associated with a utility use reduction and a new HX area investment. The details on how these have been calculated can be seen in Appendix 5.

Table 4.12 A summation of what utility cost reductions that could be expected after a retrofit.

Retrofit	**1**	**2**	**3**	**4**	**5**	**6**
Match	**1+4**	**2+4**	**3+4**	**1+5**	**2+5**	**3+5**
Utility Saving (MSEK/yr)	5.54	5.46	5.39	5.54	5.46	5.39

Retrofit 1 is the option that potentially would cut down on utility costs the most. However there is not a great difference between the different options financially since the match above pinch is the greater part of the total saving.

Utility level optimization

In this case no energy is saved. The load switch makes it possible to reduce the mass flow of imported steam to the site by supplying heat at a lower temperature/pressure level with a larger value of ΔH_{vap}. The potential comprise in lowering 5 100 kW of HP steam down to MP steam. This would require an investment in MP steam HX units on stream C and F. The reduction of HP steam import would generate a utility cost reduction of 3.74 MSEK annually, according to calculations in Appendix 6.

Modifying condensation pressure of MeOH

This option changes the whole stream system in the plant, and to be able to investigate the full investment cost, a new retrofit would be required. The possible retrofit options will be restricted to the new condensation temperature of methanol and there is probably a better optimum then the one tested in this study. However, the load available in the resulting streams is enough to cover the MP steam load in HX15 with a maintained ΔT_{min} of 10ºC in the new HX/condenser. If this replacement were to be made it would save 15.5 MSEK annually according to calculations in Appendix 7.

5 Conclusions

To conclude, this study shows that there are potentials to reduce the energy consumption in three different ways. Each suggestion could be considered on its own, or combined to save even more energy in the plant. In Table 5.1a-c below the most important findings from this thesis work are summarized.

Table 5.1a Key figures from the analysis of the streams in each case description and the results from the pinch calculations respectively with a ΔT_{min} of 10K.

	Theoretical Case Load (kW)	**Operational Case Load** (kW)	**Adjusted Operational Case Load** (kW)
FROM PROCESS DEFINITION			
HP	7 895	7 822	8 211
MP	6 911	7 800	7 562
LP	1 733	1 987	2 250
LP_{cond}	893	606	297
Total hot utility use	**17 432**	**18 215**	**18 320**
Total heating demand	**18 652**	**20 076**	**19 215**
Air	17 117	15 771	14 785
CW	1 720	961	1 653
CCW	352	337	334
Total cold utility use	**19 189**	**17 069**	**16 772**
Total cooling demand	**20 409**	**18 930**	**17 667**
Recovered internally	**1 220**	**1 861**	**895**
FROM PINCH CALCULATIONS			
$\mathbf{Q_{Hmin}}$	**15 833**	**16 509**	**16 652**
$\mathbf{Q_{Cmin}}$	**17 590**	**15 363**	**15 103**
$\mathbf{Q_{Recover}}$	**2 819**	**3 567**	**2 564**
Pinch Violations	**1 599**	**1 704**	**1 699**

Table 5.1b Here the worst pinch rule violations are shown, identified in the pinch analysis of the adjusted operational case. The pinch temperature were 67°C for cold streams and 77°C for hot streams (ΔT_{min}=10°C).

Heating Below Pinch		**Transfer Across Pinch**		**Cooling Above Pinch**	
Location	**Quantity** (kW)	**Location**	**Quantity** (kW)	**Location**	**Quantity** (kW)
HX1:	297	HX2:	527	HX16	277
		HX5:	237	HX15	179

Table 5.2c This table sum up the different investment options identified throughout this investigation, and their associated potential to reduce utility costs within the plant.

Energy Efficiency Measure	**Potential Utility Load Reduction** (kW)	**Utility Cost Reduction** (MSEK/yr)
Retrofit Suggestion	1 517	5.46
Utility Optimization	5 100*	3.74
MeOH Condensation	4 400	15.5**

*This load is not reduced, but corresponds to a steam level shift from HP- down to MP steam.
** Assuming the entire load could be integrated to replace MP steam

5.1 Pinch study

The largest pinch violation at the plant occurs in HX2, where the finished product (at 151°C, i.e. above the pinch temperature) is used to pre-heat methanol from the tank (at 26°C, i.e. below the pinch temperature). In the adjusted operational case, 527 kW of heat is transferred across and 277 kW cooled away above the pinch. In the operational case the ester stream is much warmer, making this violation even larger. If this violation were to be removed, the hot ester in stream O could be of better use somewhere else. Its temperature is hot enough to heat a stream above the pinch and therefore steam costs could be cut by 808 kW. This study showed that the best way to utilize the heat in stream O is to integrate it with stream C going to reactor 1.

Heat recovery from methanol condensation in stream M could supply the heat necessary for pre-heating of liquid methanol and feed in exchanger HX2 and HX1 respectively. The differences in utility cost savings between the matches below the pinch point are most probably a consequence from overestimating the CW price in comparison to the electricity reduction. The pinch temperature will most probable always be found at the condensation temperatures of methanol, as it is the largest cooling load in the whole process, and is the limiting factor for heat recovery. This would prevent it from transferring heat across the pinch when/if heat integrated. LP condensate, used in HX1, might not be worth so much money in itself but, by utilizing the energy it contains somewhere above the pinch instead, there are “opportunity costs” to investigate. The potential of the hot stream O, on the other side of HX2, have already been concluded to be of great value elsewhere in the process.

There is not much difference between retrofit 1-3 and 4-6 from the initial screening performed in this investigation. Therefore all options were subjected for further discussion in the report, since it was too early to exclude one in front of the other at this stage. It can be concluded that a retrofit in the plant would recover 1 517 kW of heating and cooling, and have similar effects on the utility cost savings.

5.2 Adjusted operation

If the violation in HX2 were to be removed, the suggestion of raising the methanol tank temperature would take effect. As of today, this change would probably only transfer the reduced load in HX14 to the CW in HX16 on the ester side of the process.

The same applies to the mixing points. As long as the pre-heating with hot utility below the pinch remains, the streams should be heated as far as possible before they blend. The violation will not increase because of it, as long as both streams have target temperatures above the cold pinch, and HP utility steam in the following heaters can be saved.

The only recommendation in Section 4.4 that immediately would save energy and have financial motivations is to decrease flash temperatures in HX8 and HX9. In turn, this also reduces the risk of receiving ester in the methanol top flow.

5.3 Improved load distribution between utility levels

The amount of heat supplied at the HP steam temperature level is large, approximately 45% of the total hot utility consumption in the plant, and only performed in two units, HX3 and HX7. Investing in two MP steam heating steps on streams C and F would show an immediate reduction in HP steam import, but on the counter side increase the MP steam import. Steam still has to be imported to the site, instead of completely eliminating the demand by integrating it with another process stream as the other suggestions in this report imply. The financial benefits of only swapping the temperature levels are not as large as to actually perform heat integration, which increase the energy efficiency and consequently eliminate utility loads. The change of utility level investment is minimally invasive on the process but only helps with the symptoms of high energy consumption, not the disease, which is cured by increasing the degree of heat integration on the site.

It is also worth noting that the steam distribution network has not been completely investigated, only within the system boundary of the plant. How steam levels are created originally in the boiler houses will be crucial in order to determine what effects in fuel consumption a swap of utility distribution levels will have.

5.4 Methanol availability at a higher condensation pressure

The pinch analysis and retrofit suggestions in this thesis work focused on what it possible to adjust and achieve in system as it is today. If the largest heat source on the plant, which is condensing methanol, could be made available at higher temperature levels the system would change. The preliminary results in the CC indicate that retrofit options in this system would increase the potential energy integration considerably. Making methanol vapour, which condenses at a temperature corresponding to MP steam, available, could replace entire steam loads.

If intermediate heating steps with MP steam were to be investigated, as discussed in Section 5.3 above, a large MP steam consumer would be created (stream C and F going to the reactors). Combined with the intermediate flashing, condensing methanol could replace the new MP load created in the utility optimization. The system would go from using HP steam to condensing methanol in order to cut the HP steam load.

A large consumer of MP steam in the current HEN is the distillation column re-boiler, HX10 in Table 4.1. The results in Section 4.9 indicate that the entire load could be replaced by recovering heat from condensation of methanol from the intermediate flashing steps (4 400 kW). This would imply that one large retrofit integration could be made, instead of several small ones as was considered in this retrofit study, to integrate a larger amount of energy. Another benefit by integrating the hypothetical methanol flow with the distillation column re-boiler, instead of installing two intermediate heaters on steam C and F, is that the column runs at a much lower pressure then the reactor streams. This would probably cut down on the equipment investment cost.

6 Future Work

Recommendations for future work involve quantifying how large the pinch violation in HX2 is in the operational case, and perform a more accurate economic evaluation. It is the worst offender in the system (transferring 527 kW across and cooling away 277 kW above the pinch) and diminishes suggested process optimizations, which could save energy - essentially for free. The future investigation should take its starting point in the operational case in this study. However, the data used when calculating the load in HX2 needs to be validated, since the resulting figures showed significant differences between the methanol- and the ester side in the initial data collection of the operational case. To get a consistent energy balance both sides was adjusted. The resulting load finally suggested that no cooling occurs in HX16, which is unlikely.

A future study should investigate the potential for flashing methanol at intermediate pressures more thoroughly. Preferably a curve indicating what methanol top flow to expect at different temperature and pressure levels should be done. The corresponding energy content in the top flow condensation should also be included, as well as the chain effects on mass flow rate changes, affecting downstream heating and cooling demands. For this, more extensive simulations would be required. When a new stream system has been established, a retrofit should be done all over again. There are probably more beneficial retrofits to identify with this system change in place, in addition to those covered in this report.

If a less invasive path should be chosen, where utility level optimization is the more attractive option, future works should include a full survey of the steam net for the entire site. This is necessary to conclude whether a steam shift would result in less fuel consumption in the boiler houses or only result in more HP steam throttled down to MP steam levels.

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8 Appendix

Appendix 1 – Utility price calculations

These calculations were not possible to perform without using some of the results from the study. The share of vapour received from flashing was based on isenthalpic conversion of saturated liquid at high pressure into a two-phase vapour-liquid system at a lower pressure at its saturation point.

$C_{HP} = 0.24\, SEK/kg$

$\Delta H_{vap_{HP}} = 1698\, kJ/kg$

$\Delta H_{vap_{MP}} = 2048\, kJ/kg$

$\Delta H_{vap_{LP}} = 2164 kJ/kg$

$$1\, kg\ \cdot \Delta H_{vap_{HP}} = x\, kg\ \cdot \Delta H_{vap_{MP}}$$

$$\rightarrow x = \frac{\Delta H_{vap_{HP}}}{\Delta H_{vap_{MP}}} = \frac{1698}{2048} = 0.83\, kg$$

$$C_{MP} = 0.83 \cdot 0.24 = 0.1992\, SEK/kg$$

Since steam is created internally by flashing condensate, these amounts are generated without any extra cost. When the consumption exceeds these margins there is a price on LP and MP steam according to the calculation above. Below, the total amount of kW available for free is calculated:

HP steam is always imported, to supply heat to the reactor flows:

$$\dot{m}_{HP_{cond}} = \frac{Q_{HX3} + Q_{HX7}}{1698} \left[\frac{kg}{s}\right]$$

In the system MP steam is also generated from flashing. The amount provided for free from HP condensate is:

$$\dot{m}_{MP_{vap_{free}}} = 0.186 \cdot \dot{m}_{HP_{cond}} \left[\frac{kg}{s}\right]$$

$$\rightarrow Q_{MP_{free}} = \dot{m}_{MP_{vap_{free}}} \cdot 2048\ [kW]$$

Not all of the HP condensate turns into MP steam. The remaining 81.4$\%_{wt}$ becomes MP condensate. The MP steam consumers also make a contribution to how much LP steam that will be created through flashing:

$$\dot{m}_{MP_{cond}} = \frac{Q_{HX8} + Q_{HX9} + Q_{HX10} + Q_{HX11} + Q_{HX13}}{2048} + 0.814 \cdot \dot{m}_{HP_{cond}} \left[\frac{kg}{s}\right]$$

$$\rightarrow Q_{LP_{vap_{free}}} = 0.074 \cdot \dot{m}_{MP_{cond}} \cdot 2164\ [kW]$$

In the end, all steam imported to the plant will end up like LP condensate. It is assumed to be saturated and then have a restriction to not get colder then 100ºC. In the end this is a by-product from the total steam consumption, and so this hot utility have a maximum limitation:

$$Cp_{cond} = 4.18 \frac{kJ}{kg \cdot K}$$
$$Q_{LP_{cond,max}} = \dot{m}_{MP_{cond}} \cdot 4.18 \cdot (133 - 100) \ \ [kW]$$

Table 8.1. A summation of the availability of free MP steam from flashing HP condensate, and also how much LP steam and LP condensate that can be generated from all of the previous stem consumers. It is assumed that the original loads of HP and MP steam remain the same in each case.

Steam Level	Theoretical Case	Operational Case	Adjusted Op. Case
$Q_{HP\ tot}$ (kW)	7 895	7 802	8 212
$Q_{MP\ tot}$ (kW)	6 911	7 800	7 562
$Q_{MP,free}$ (kW)	1 781	1 760	1 852
$Q_{LP,free}$ (kW)	1 146	1 209	1 222
$Q_{LPcond,\ max}$ (kW)	988	1 041	1 052

The utility price for air-cooling stems from the electricity price. Since this is an unknown area in the process, the cooling demand on the process side is estimated from the calculated consumption. It is assumed that the installed capacity in HX14 is proportional to the total site electricity cost. Then the reductions made on the process side are assumed to be proportional to how much electricity that is saved in the engines.

Since the retrofit savings only is evaluated for the adjusted operational case, this is the process demand/load used. Furthermore, the energy savings only concern the cooler HX14, and so this installed engine capacity is the only one included (meaning HX19 is not considered). With this in mind, the procedure can be seen in the equations below:

$$\sum Total\ electricity\ consumption\ in\ plant = 763\ 180\ kWh/month$$
$$t = 24 \cdot 31\ h/month$$
$$C_{el} = 0.45\ SEK/kWh$$
$$Q_{HX14} = 8\ 436\ kW$$
$$P_{intsalled_{HX14}} = 2 \cdot 37\ kW$$

$$P_{installed_{tot}} = \frac{763180}{24 \cdot 31} = 1026\ kW$$

$$\rightarrow HX14\ constitutes\ 7.2\%\ of\ the\ total\ installed\ capacity$$

$$\rightarrow C_{HX14} = 0.072 \cdot 763180 \cdot 0.45 = 24\ 770\ SEK/month$$

$$\rightarrow Utility\ cost\ reduction = \frac{Load\ reuction\ in\ HX14\ on\ process\ side\ (kW)}{8436} \cdot 24770 \cdot 12\ \left[\frac{SEK}{yr}\right]$$

The price for cooling water, CW, is based on the crude water price, and that CW in the plant have the same temperature difference as the cooling towers that gather CW from all other sites:

$C_{crudewater} = 0.45\ SEK/m^3$
$\delta_{water} = 1000\ kg/m^3$

$\rightarrow C_{CW} = 0.45 \cdot 10^{-3}\ SEK/kg$

$\Delta H_{CW} = 4.18 \cdot (33 - 23) = 41.8\ kJ/kg$

$$\rightarrow Utility\ cost\ reduction = \frac{Load\ reduction\ of\ CW\ on\ process\ side\ (kW)}{41.8} \cdot 0.45 \cdot 10^{-3} \cdot t_{yr}\ \left[\frac{SEK}{yr}\right]$$

The price for chilled cooling water, CCW, was not possible to determine. However, it only constitutes a small part of the total cooling demand so its price is assumed to be of little significance in the study.

Appendix 2 - Operational case process loads compared with steam side- and theoretical case loads

Since the data was available, it was investigated how well the steam side load matched the cold process side, and how the operational case hot utility loads differed from the theoretical case. There are flow meters on the steam side in all of the heaters, and so the average value was used. The results are presented in Table 8.5 where steam flow was assumed to undergo pure condensation at their saturation pressure, delivering ΔH_{vap} to the process stream. Note that these values in general are prone to error because of turbulence and other difficulties that come with measuring flow rates of gases.

Table 8.2 The hot utility demand can be calculated on either the steam side or the process side. In this table the different results are shown, and how they compare to theoretical loads. Cooling demand based on process side calculations is also included, and how they differ from the theoretical case. Numbers within parenthesis indicate that the load has been copied from another case to give comparable total utility loads.

Utility	HX	Average Steam Loads (kW)	Operational Case Process Demand (kW)	Theoretical Case Process Demand (kW)
LPcond	HX1	(606)	606	893
LP	HX4	570	1207	1030
	HX6	691	799	703
MP	HX8	2446	3322	2050
	HX9	472	489	420
	HX10	3788	(3788)	4300
	HX11	(30)	(30)	30
	HX13	131	171	111
HP	HX3	4337	3992	3860
	HX7	3450	3810	4035
Total Hot Utility		**16 521**	**18 215**	**17 432**
Air	HX14	-	9053	7907
	HX19	-	6718	9210
CW	HX15	-	448	400
	HX16	-	0	510
	HX17	-	513	(30 t/h) 810
CCW	HX18	-	246	231
	HX21	-	51	81
	HX20	-	(40)	40
Total Cold Utility			**17 070**	**19 189**

Appendix 3 – Estimation of overall heat transfer coefficient

The intervals for the overall heat transfer coefficient U was based on the diagram in Figure 8.6 (Sinnott, 1983).

The feedstock was decided to fall under the category of “Oils”. These ranges between 250-550 W/(m^2 K)

“Paraffin” was decided to resemble the esters. The value of U can range between 800-1000 W/(m^2 K).

Liquid methanol fell under the category “Brines”, since it can be used as cooling medium in many applications. Condensing methanol were more accurately put under the category “Condensation organic vapours”.

Since stream C, going to reactor 1, is a mix of methanol and feed it was assigned a combined U-value from both “Oils” and “Brines” in the same proportion. Since the second reactor stream F mainly contains esters, and only a small fraction of feed, the “Oil” component was exchanged with “Paraffin” instead.

In the end an average U value was calculated for the interval shown in Table 8.6.

Table 8.3 Here the fluid each stream has been assigned, when reading the diagram in Figure 8.6, are tabulated together with the associated value of the individual fluid coefficient range. h- fluid coefficient range;

Fluid	**Stream**	**h** (W/(m^2 K))
“Oils”	A	250-550
“Paraffin”	E N O	800-1000
“Brines”	B	775-1025
“Cond. Organic vap.”	M	825-1700
“Oils” / “Brines”	C	513-788
“Paraffin” / “Brines”	F	788-1013

The over all heat transfer coefficient noted in Table 4.6 and Table 4.7 is the average of the lowest and highest possible value of h for the stream combinations in the new exchanger. U was calculated according to the equation below:

$$U_{overall} = \frac{1}{h_{inner}} + \frac{1}{h_{outer}}$$

Where “inner” and “outer” denote the streams on each side of a shell and tube exchanger. This equation assumes that heat transfer resistance through the wall is negligible and that there is no fouling on the heat exchanger surface, which is equal on both sides of the exchanger. The average U-value used in the calculation of A_{hx} is tabulated below.

Table 8.4 The resulting estimations of the overall heat transfer coefficients based on the diagram shown in Figure 8.6

Stream Combination	A/N	B/M	E/O	A/M	E/N	E/M	C/N	F/O	F/N	C/O
$U_{avg.}$ (W/m^2/K)	273	490	450	304	450	518	377	450	450	377

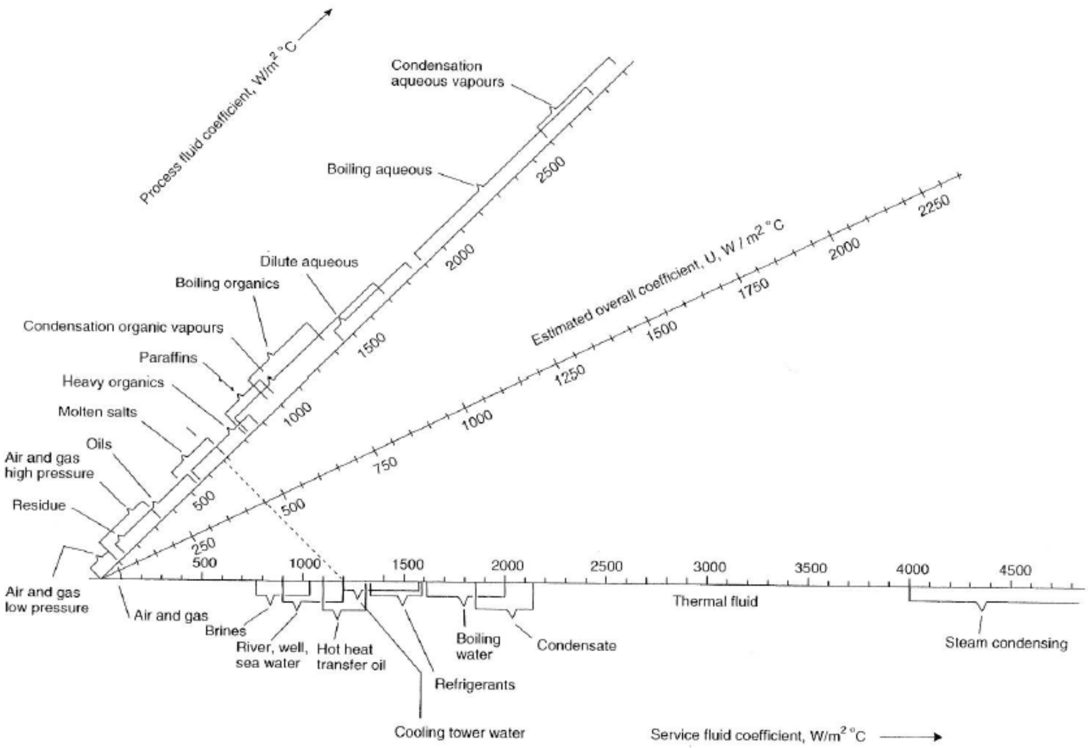

Figure 8.6 From the diagram the value of the over all heat transfer coefficient in a shell and tube HX can be read. **(Sinnott, 1983)**

Appendix 4 – Illustration of retrofit 5 and 6

Since the retrofit suggestions 4-6 are very similar to the ones made in retrofit 1-3, the resulting network for retrofit 5 and 6 are shown here in Figure 8.7 and Figure 8.8.

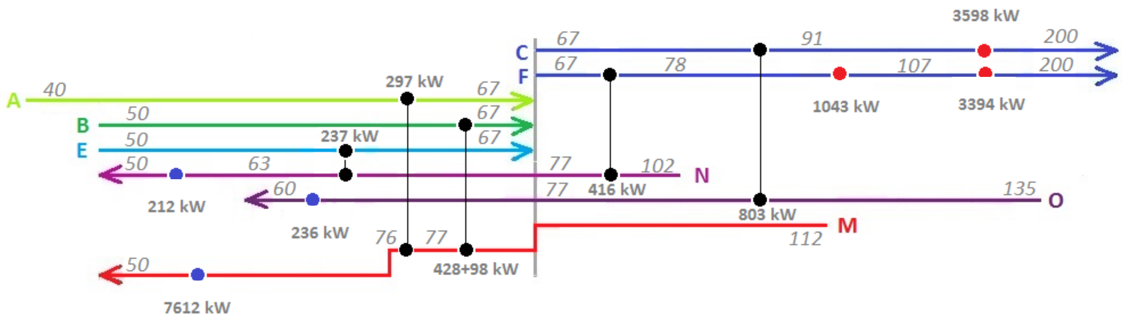

Figure 8.7 The figure illustrates the layout and loads for the network comprising retrofit 5. It is a combination between match 2 below the pinch and match 5 above the pinch. The new loads are indicated with bold grey typing given in kW. The new inlet and outlet temperatures in each exchanger are indicated with grey italic numbers given in °C

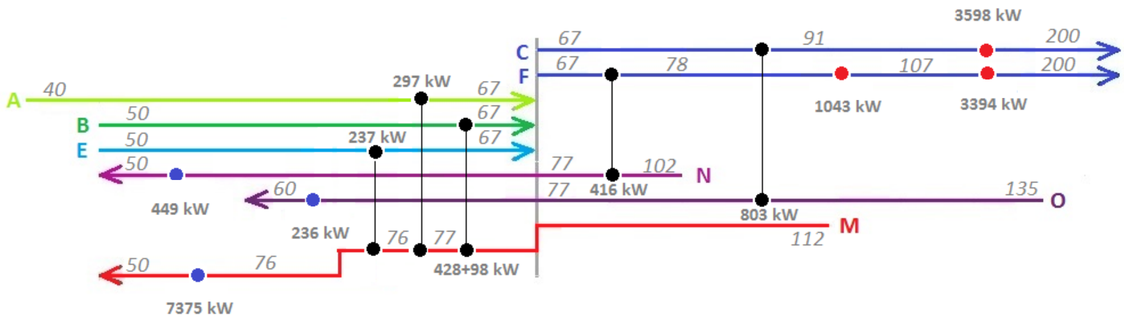

Figure 8.8 The figure illustrates the layout and loads for the network comprising retrofit 6. It is a combination between match 3 below the pinch and match 5 above the pinch. The new loads are indicated with bold grey typing given in kW. The new inlet and outlet temperatures in each exchanger are indicated with grey italic numbers given in °C

Appendix 5 – Utility cost savings through retrofit

Match 1:

B+M

Saving: 527 kW Air-cooling in HX14

$$\frac{527}{8436} \cdot 24770 \cdot 12 = 18\,570 \frac{SEK}{yr}$$

A+N

Saving: 297 kW CW in HX15.

Assume LP_{cond} is for free (could also be allocated to heat something else below pinch to increase potential savings)

$$\frac{297}{41.8} \cdot 0.45 \cdot 10^{-3} \cdot 3600 \cdot 8400 = 96\,688 \frac{SEK}{yr}$$

E+O

Saving: 237 kW CW in HX16.

$$\frac{237}{41.8} \cdot 0.45 \cdot 10^{-3} \cdot 3600 \cdot 8400 = 77\,155 \frac{SEK}{yr}$$

$$\sum Total\ Savings\ Match\ 1 = 192\,413 \frac{SEK}{yr}$$

Match 2:

B+M

Saving: $18\,570\ \frac{SEK}{yr}$

A+M

Saving: 297 kW Air-cooling in HX14. Assume LP_{cond} is for free.

$$\frac{297}{8436} \cdot 24770 \cdot 12 = 10\,465\ \frac{SEK}{yr}$$

E+N

Saving: 237 kW CW in HX15.

$$\frac{237}{41.8} \cdot 0.45 \cdot 10^{-3} \cdot 3600 \cdot 8400 = 77\,155 \frac{SEK}{yr}$$

$$\sum Total\ Savings\ Match\ 2 = 106\,190\ \frac{SEK}{yr}$$

Match 3:

B+M

Saving: $18\ 570\ \frac{SEK}{yr}$

A+M

Saving: $10\ 465\ \frac{SEK}{yr}$

E+M

Saving: 237 kW Air-cooling in HX14.

$$\frac{237}{8436} \cdot 24770 \cdot 12 = 8\ 350\ \frac{SEK}{yr}$$

$$\sum Total\ Savings\ Match\ 3 = 37\ 385\ \frac{SEK}{yr}$$

Match 4:

C+N

Saving: 416 kW HP steam in HX3, and 179 kW of CW in HX15.

$$\frac{416}{1698} \cdot 0.24 \cdot 3600 \cdot 8400 = 1.788 \cdot 10^6 \approx 1.8\ \frac{MSEK}{yr}$$

$$\frac{179}{41.8} \cdot 0.45 \cdot 10^{-3} \cdot 3600 \cdot 8400 = 58\ 273\ \frac{SEK}{yr}$$

F+O

Saving: 804 kW HP steam in HX7 (assuming the shifted load is cascaded through HX6 to finally decrease HP steam consumption only), and 277 kW CW in HX15.

$$\frac{804}{1698} \cdot 0.24 \cdot 3600 \cdot 8400 = 3.436 \cdot 10^6 \approx 3.4\ \frac{MSEK}{yr}$$

$$\frac{277}{41.8} \cdot 0.45 \cdot 10^{-3} \cdot 3600 \cdot 8400 = 90\ 177\ \frac{SEK}{yr}$$

$$\sum Total\ Savings\ Match\ 4 = 5.35\ \frac{MSEK}{yr}$$

Match 5:

C+O

Saving: 804 kW HP steam in HX3, and 277 kW of CW in HX15.

F+N

Saving: 416 kW HP steam in HX7, and 179 kW of CW in HX15

$$\sum Total\ Savings\ Match\ 5 = 5.35\ \frac{MSEK}{yr}$$

Appendix 6 – Utility cost savings through optimization

The HP steam consumption was to be minimized. The calculations of the financial benefits are shown here, assuming that the highest load possible was replaced with MP steam.

The hypothetical MP steam exchangers heated the two reactor streams to 160°C. The minimum temperature difference between MP steam and the two process streams then became 10°C.

Before reactor 1:
The load in stream C has the following heating demand if it were to be heated between ΔT:

$$\Delta T = 160 - 67 = 93°C$$

$$\rightarrow \Delta Q = 2933\ kW$$

This load is shifted from HP steam down to MP steam:

$$\sum Total\ Saving = 2933 \cdot 0.24 \cdot 3600 \cdot 8400 \cdot \left(\frac{1}{1698} - \frac{1}{2048}\right) = 2.14\ \frac{MSEK}{yr}$$

Before reactor 2:
The temperature interval for MP steam utilization becomes smaller since the pre-heater HX6 consuming LP steam is in place. The load in stream F has the following heating demand if it were to be heated between ΔT:

$$\Delta T = 160 - 97 = 63°C$$

$$\rightarrow \Delta Q = 2190\ kW$$

This load is shifted from HP steam down to MP steam:

$$\sum Total\ Saving = 2190 \cdot 0.24 \cdot 3600 \cdot 8400 \cdot \left(\frac{1}{1698} - \frac{1}{2048}\right) = 1.6\ \frac{MSEK}{yr}$$

This investment would annually save 3.74 MSEK in steam costs.

Appendix 7 – Utility cost savings from installing two intermediate flashing steps

In the study a hypothetical target temperature of $135°C$ was chosen for the methanol. The corresponding saturation pressure for the pure substance was put into the flash simulation performed in Chemcad. In order to reach the desired temperature in the actual simulation the pressure had to be somewhat decreased. The results were assumed to be applicable after the first and second reactor, before entering the flash already in place.

$T_T = 135°C$

$P_{sat,MeOH} = 10.1\ bar$ (Goodwin, 1987)

$P_{sat,Chem} = 8\ bar$

$\rightarrow \dot{m}_{top,new} = 8\ 718\ kg/h$

$\rightarrow \dot{m}_{top\ Flash\ 0} = 8\ 438\ kg/h\ \ (from\ simulation\ model)$

$(to\ be\ compared\ with\ the\ mass\ balance\ giving: 15671 - 8718 = 6953\ kg/h)$

$\Delta H_{vap_{MeOH}}(@\ 403K, sat) = 906\ kJ/kg$

$$\rightarrow\ \Delta Q_{MeOH} = \frac{8718}{3600} \cdot 906 = 2\ 194\ kW$$

Assume two of these steps were to be installed and that they could replace MP steam loads in the process:

$$\rightarrow \sum Annual\ Utility\ Saving = \frac{2 \cdot 2194}{2048} \cdot 0.24 \cdot 3600 \cdot 8400 = 15.5\ MSEK/yr$$

CHALMERS UNIVERSITY OF TECHNOLOGY
SE 412 96 Göteborg, Sweden
Phone: + 46 - (0)31 772 10 00
Web: www.chalmers.se