CHALMERS

Process integration study for increased energy efficiency of a PVC plant

Master's Thesis within the Sustainable Energy Systems programme

ÅSA LINDQVIST

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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Cover:
INEOS' PVC production site in Stenungsund.

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ABSTRACT

Increasing energy prices combined with the need to reduce greenhouse gas emissions makes energy efficiency important both for economical and environmental reasons. In the energy-intensive industry, such as the process industry, there is a substantial potential for energy savings. INEOS ChlorVinyls, a major chlor-alkali producer, has a PVC production site in Stenungsund, on the West coast of Sweden. The site is a large consumer of both electricity and different fuels, and it is therefore relevant to perform a systematic energy efficiency study of the site.

This master's thesis has investigated the possibilities for increasing the energy efficiency of INEOS' PVC production site by increased internal heat recovery. The production of PVC is divided into three sub-processes. In this study focus has been on the second sub-process, which is the VCM production plant. Since the site is to be retrofitted in a near future, and the production capacity will be extended, the study was made for a future scenario with increased production. To identify options for increasing the energy efficiency of the site, pinch analysis has been used. Pinch analysis is a systematic method for identifying opportunities for improving the integration of processes in order to decrease the amount of external heating and cooling needed.

The results show that theoretically 7.8 MW of steam and the same amount of cooling water could be saved by internal heat exchange in the future VCM plant. This corresponds to 37% of the steam use. The pinch violations in the heat exchanger network were identified, and different retrofits are proposed, aiming to eliminate the five largest pinch violations in the system. Both direct heat transfer and the use of a heat transfer media such as water have been investigated. The suggested retrofits lead to steam savings between 4.5 and 6.5 MW, and an economic evaluation shows that the annual savings could reach 20 MSEK with a short pay-back period of about one year. Both in the existing and future VCM-plant, there is a large excess of low-grade heat below the pinch, which could be used for heating at other parts of the site. This would enable additional steam savings of the same size as in the proposed retrofits for increased internal heat transfer.

Key words: PVC plant, VCM plant, Energy efficiency, Pinch analysis

Studie av processintegration för ökad energieffektivitet i en PVC-anläggning
Examensarbete inom masterprogrammet *Sustainable Energy Systems*
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SAMMANFATTNING

Stigande energipriser tillsammans med behovet av att minska utsläppen av växthusgaser gör energieffektivisering mycket viktigt, både av ekonomiska och miljömässiga skäl. Inom den energiintensiva industrin, exempelvis processindustrin, finns en avsevärd potential för energibesparing. INEOS ChlorVinyls, en stor kloralkaliproducent, har en PVC-produktionsanläggning i Stenungsund på Sveriges västkust. Anläggningen förbrukar stora mängder av både el och olika bränslen, och det finns därför anledning att utföra en systematisk energieffektivitetsanalys av anläggningen.

Detta examensarbete har undersökt möjligheterna att öka energieffektiviteten för INEOS anläggning genom ökad intern värmeåtervinning. Produktionen av PVC är uppdelad i tre delsteg. Denna studie har fokuserat på VCM-fabriken, som är produktionens andra steg. Eftersom delar av anläggningen kommer att byggas ut inom en snar framtid, och produktionskapaciteten kommer att utökas, är studien gjord för ett framtida scenario med ökad produktion. För att hitta möjliga energieffektiviseringsåtgärder har pinchanalys använts. Pinchanalys är en systematisk metod för att hitta möjligheter för förbättrad processintegration, med syfte att minska mängden extern värme och kyla.

Resultaten visar att 7.8 MW ånga och lika mycket kylvatten teoretiskt kan sparas genom ökad intern värmeväxling i den framtida VCM-fabriken. Det motsvarar 37 % av ångbehovet. Pinchbrotten i värmeväxlarnätverket har identifierats, och olika förbättringsförslag har undersökts, med målet att eliminera de fem största pinchbrotten i systemet. Både direktvärmeväxling och användning av ett värmeöverföringsmedium, såsom vatten, har undersökts. De föreslagna åtgärderna ger ångbesparingar på mellan 4.5 och 6.5 MW, och en ekonomisk utvärdering visade att de årliga besparingarna kunde uppgå till 20 miljoner kronor för åtgärder med en kort återbetalningstid på ca ett år. I VCM-fabriken finns ett stort värmeöverskott under pinchtemperaturen, vilket skulle kunna användas i andra delar av anläggningen. Det skulle kunna möjliggöra ytterligare ångbesparingar av samma storleksordning som de föreslagna åtgärderna för värmeväxling.

Nyckelord: PVC, VCM, energieffektivitet, pinchanalys

Contents

Preface

In this Master's Thesis, opportunities for increased energy efficiency in a PVC production plant located in Stenungsund on the West coast of Sweden have been studied using pinch analysis. The project has been carried out in cooperation between Chalmers University of Technology, Division of Heat and Power Technology, and INEOS ChlorVinyls in Stenungsund. The aim of the thesis was to investigate possibilities for increased energy efficiency at INEOS' PVC production site in Stenungsund.

I would like to thank my supervisor Roman Hackl and my examiner Simon Harvey at Chalmers, Division of Heat and Power Technology, together with my supervisor Kent Olsson at INEOS for all their help and time spent on this project. I also would like to thank all employees at INEOS who have helped me during my time in Stenungsund.

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Notations

Abbreviations:

°C	Degree Celsius
CC	Composite Curves
CW	Cooling water
EDC	Dichloroethane
GCC	Grand Composite Curve
GJ	GigaJoule
GWh	GigaWatt hour
H_2	Hydrogen
HCl	Hydrochloric acid
HP	High Pressure
kW	KiloWatt
LP	Low Pressure
MER	Maximum Energy Recovery
MJ	MegaJoule
MP	Medium Pressure
MSEK	Million Swedish crowns
MW	MegaWatt
MWh	MegaWatt hour
NaCl	Sodium chloride
NaOH	Sodium hydroxide
NRTL	Non-random two-liquid model
SEK	Swedish crowns
VCM	Vinyl chloride
PVC	Polyvinyl chloride

Symbols:

A_{hx}	Heat exchanger area
C_e	Cost of equipment
C	Inside battery limits cost
C_{FC}	Total capital cost
ΔT_{lm}	Logarithmic mean temperature difference
ΔT_{min}	Minimum temperature difference
Q	Load
U	Overall heat transfer coefficient

1 Introduction

The increasing energy prices worldwide and the threat of escalating global warming makes energy efficiency a matter of utter importance. The world's growing population and the increased energy intensity causes the global energy use to rise rapidly (US Energy Information Administration, 2010). Together with the debates concerning oil scarcity and greenhouse gas emissions it is obvious that energy efficiency actions must be taken, both for environmental and economical reasons.

INEOS ChlorVinyls is a major chlor-alkali producer and the largest manufacturer of PVC in Europe (INEOS ChlorVinyls). One of their manufacturing sites is located in a large chemical cluster in Stenungsund, on the West coast of Sweden. The INEOS site in Stenungsund has about 300 employees and produces mainly PVC (INEOS ChlorVinyls Sweden). The site was built during the 1960s, when energy prices were low and energy efficiency was consequently a lower priority. Today, there are other conditions and there are economic incentives to be more restrictive with the use of energy. It is therefore relevant to perform a systematic energy efficiency study of the site. The site is a large energy consumer. In 2009, about 500 GWh of electricity and 420 GWh of fuels were consumed (INEOS ChlorVinyls, 2009). For comparison, an average Swedish house has a total energy demand of 24000 kWh annually (Swedish Energy Agency, 2011).

Not only INEOS but the entire chemical cluster in Stenungsund has a large energy usage (Hackl, Andersson, & Harvey, 2010). An ongoing PhD research project (Hackl, Andersson, & Harvey, 2010) is investigating opportunities for increased process integration between the different companies in the cluster in order to save energy. This master's thesis will contribute to this research by providing detailed knowledge about the energy situation and process integration opportunities at the INEOS site.

The INEOS site for PVC production consists of three different parts; a chlorine production plant, a VCM (vinyl chloride) production plant and finally a PVC production plant. This thesis will mainly focus on the VCM production plant, since it is the most steam consuming part. The possibilities to integrate the VCM plant with other parts of the site will be investigated as well.

The company is planning to build a new chlorine production plant, due to the fact that the current production technology uses mercury and will be prohibited in a few years (Swedish Ministry of the Environment, 2010). The new plant will have increased capacity and thus the production in the other plants also will be affected. The targeted time plan is to have the new plant installed and running in 2015.

Pinch analysis can be used to investigate opportunities for integration of parts of the VCM plant in order to save energy. Since the site is to be retrofitted in the near future, the study is performed for the case of increased production after commissioning of the new chlorine production plant.

1.1 Purpose and objective

The purpose of this master's thesis is to identify options for increasing the energy efficiency of the INEOS PVC production site in Stenungsund by increased internal heat exchange and thereby reduced demand for external heating and cooling. Since a part of the plant is to be replaced and the production capacity will be extended in the near future, the study is performed for a hypothetical plant with a production capacity corresponding to that planned after the retrofit.

The objective of this master's thesis is to determine minimum cooling and heating demand of the plant and to identify practical energy efficiency measures which may be implemented in order to reduce the steam demand. This is done using pinch analysis to investigate the possibilities for increased internal heat exchange. The energy efficiency measures are evaluated economically and the consequences for CO_2 emissions are briefly discussed. The thesis is focusing on the VCM production plant, but possibilities for integration with units in the PVC plant are investigated as well.

The results of the project may be used further in the ongoing research project of the chemical cluster situated in Stenungsund (Hackl, Andersson, & Harvey, 2010).

2 Process description

The INEOS ChlorVinyls site in Stenungsund produces about 200 ktons of PVC annually (INEOS ChlorVinyls, 2009). The raw materials purchased are mainly salt and ethylene and main products sold are different kinds of PVC and sodium hydroxide (NaOH).

Table 2.1 Consumption and production at the INEOS site in Stenungsund (INEOS ChlorVinyls, 2009)

	Consumed [kton/yr]	*Produced [kton/yr]*
NaCl	200	
Ethylene	80	
NaOH		130
PVC		200

The site is a large energy consumer, and is participating in the Programme for Improving Energy Efficiency in Energy Intensive Industries (PFE), which is run by the Swedish Energy Agency. The energy use and CO_2 emissions for the site are presented in Table 2.2. Note that the CO_2 emissions are on-site emissions only, i.e. emissions associated with generation of the large amounts of electricity used at the site are not included.

Table 2.2 Energy use and carbon dioxide emissions at the site in 2009 (INEOS ChlorVinyls, 2009)

	Annual use [GWh]	*Annual emissions [kton]*
Electricity	500	
Fuel gas and methane	280	
Hydrogen	90	
Liquid byproducts	15	
Oil	30	
Carbon dioxide	60	

The fuel gas is a mixture of combustible gases, containing mostly hydrogen and methane, purchased from a nearby site. The detailed composition can be found in Appendix 5. Fuel oil is used as a back-up fuel in case the nearby site cannot supply the total fuel demand.

The very large electricity consumption is to large extent due to the electrolysis in the first part of the process. In order to reduce the emissions of carbon dioxide at the global level, it is therefore important to reduce the use of electricity, but this is beyond the scope of this Master's Thesis.

2.1 Process overview

The PVC production process can be divided into three separate parts as shown in the figure below; the chlorine production plant, the VCM plant and the PVC plant. This Master's thesis has focused on the second part, the VCM production plant, which will be described more in detail in Chapter 2.2.

A general overview of the process is shown in Figure 2.1.

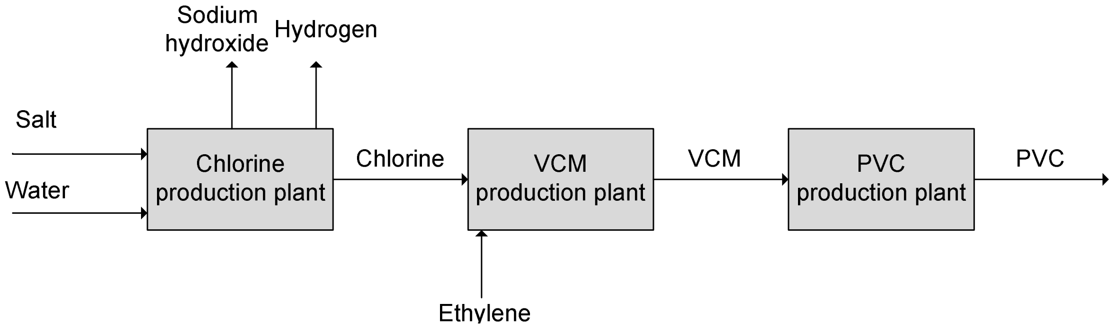

Figure 2.1 General overview of the production plants

Today, no heat is transferred between the different parts of the site, and the steam demand in the three plants is covered by a steam boiler. In the VCM plant however, there is generation of steam which is used internally.

The following figures describing the process are based on a process description published in the company's environmental report (INEOS ChlorVinyls, 2009).

2.1.1 The chlorine plant

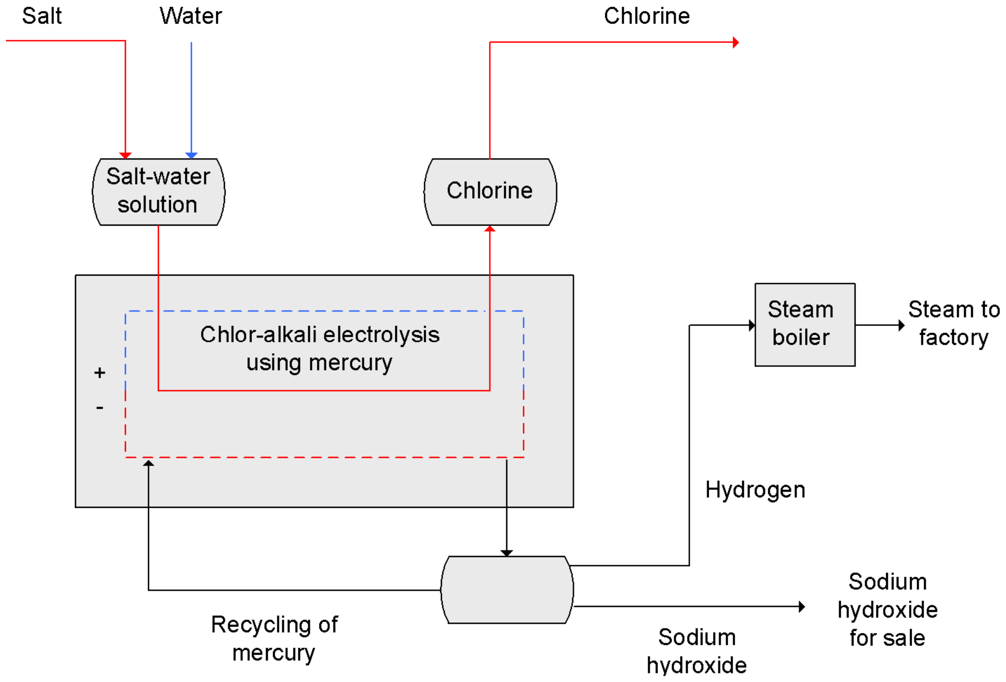

Figure 2.2 Present chlorine production

Chemical reaction: $2NaCl + 2H_2O \rightarrow 2NaOH + H_2 + Cl_2$

Today, electrolysis is used to produce chlorine from common salt as seen in Figure 2.2. Hydrogen gas and sodium hydroxide are produced as well. The current production technology uses mercury and needs to be replaced in a few years. The planned chlorine production plant will use membrane technology and will have a larger production capacity than the existing plant.

Since chlorine is currently produced by electrolysis, electricity is the dominating energy source in the chlorine plant. The chlorine production plant consumes about 380 GWh of electricity and 8 GWh of steam annually (INEOS ChlorVinyls, 2008).

The energy demand in the planned chlorine production plant using membrane technology will differ from the current demand. The membrane process uses electrolysis as well, but the anolyte and the catolyte are separated by a cation-exchange membrane, that selectively transmits sodium ions (Schmittinger et al, 2006). The membrane process is more energy efficient. Therefore the electricity demand of the membrane process will be lower than for the current process. However, since the production capacity will be increased the total electricity use will be larger. The NaOH produced using the membrane technology will be more diluted than today's' NaOH (Schmittinger et al, 2006). The sodium hydroxide needs to be concentrated before it can be sold, and therefore heat will be needed, resulting in a larger steam demand in the new chlorine production plant.

2.1.2 The VCM plant

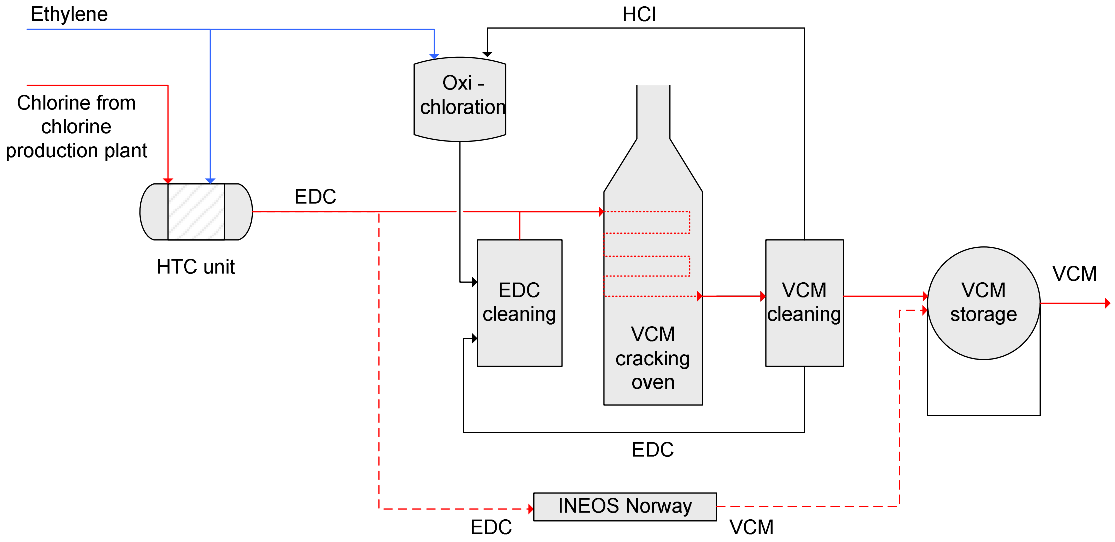

Figure 2.3 VCM plant

Chemical reaction: $C_2H_4 + Cl_2 \rightarrow C_2H_4Cl_2 \rightarrow C_2H_3Cl+HCl$

The second part of the process, the VCM production plant, is divided into different steps. First, the chlorine produced in the chlorine plant reacts with ethylene and forms EDC (dichloroethane). The EDC is further reacted to VCM (Vinyl Chloride) in a cracking oven. Because of the limited capacity of the VCM production parts, some EDC is exported to another INEOS site in Norway, from which VCM also is imported. Today, about three times more VCM is imported than EDC exported.

The EDC production part of the process has a higher production capacity than that utilized today, and will operate at a higher production level than the current when the new chlorine production plant is built. The VCM production unit, however, is operating at its maximum capacity today and the VCM production will remain unchanged. All additional produced EDC will be exported, but the import of VCM will not change.

The VCM production plant has a large energy demand. 240 GWh of steam are used annually, of which about 150 GWh are generated in the process. Another 110 GWh of fuels are used for combustion in the cracking oven (INEOS ChlorVinyls, 2008).

2.1.3 The PVC plant

In the last part of the production the VCM is polymerized into PVC. The process is shown in Figure 2.4. As in the EDC production plant, not all installed capacity is currently utilized and the production will increase after the installation of the new chlorine production plant.

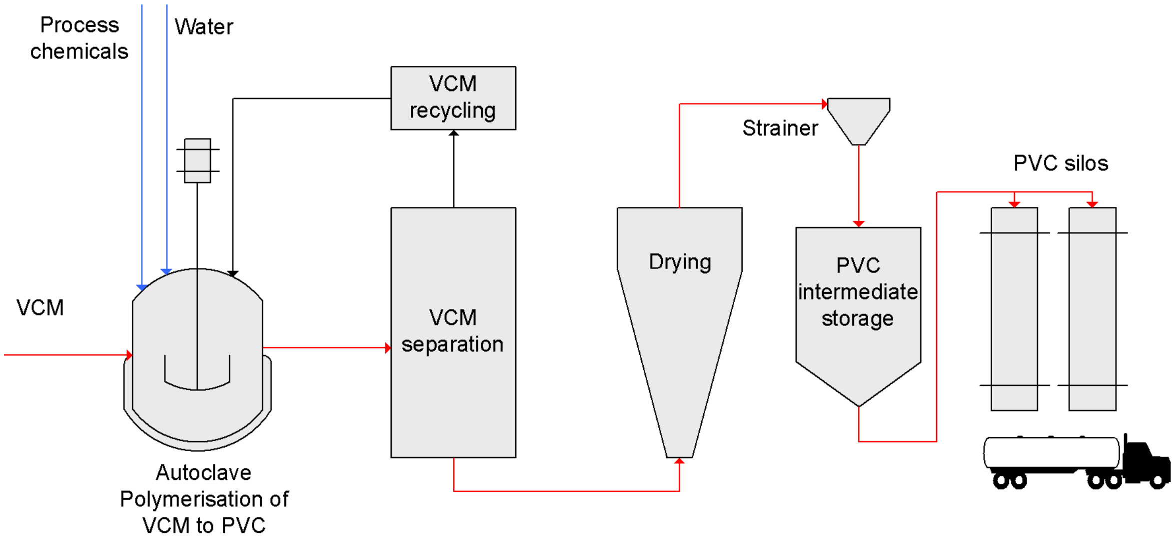

Figure 2.4 PVC plant

Chemical reaction: n·$C_2H_3Cl \rightarrow (C_2H_3Cl)_n$

The VCM is polymerized into PVC in autoclaves, and thereafter dried. The polymerisation takes place in several batch reactors, which makes the drying utilities able to operate continuously. There are four parallel production systems, producing two different kinds of PVC, known as s-PVC and p-PVC.[1]

The PVC production plant consumes about 90 GWh of steam and 80 GWh of fuels for combustion each year (INEOS ChlorVinyls, 2008).

2.2 Detailed description of the VCM plant

In the VCM plant, VCM is formed from the chlorine produced in the chlorine plant and purchased ethylene. As seen in Figure 2.3, the VCM plant can be divided into four different units, which are connected to each other. The process contains several heating and cooling units. The heating and cooling operations achieved in the heat exchangers marked in yellow were included in the study.

The calculated current loads for the heat exchangers are presented. For input data and assumptions for calculations of heat exchanger loads, see Appendix 4.

2.2.1 The HTC unit

In the first part, the HTC unit, ethylene and chlorine react to form EDC (dichloroethane). The main product is EDC, which is transported to the VCM-unit or exported. The second largest product is a by-product called moist EDC, which goes to the EDC-cleaning unit.

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[1] s-PVC (suspension PVC) is mostly used for harder products, like pipes, whereas p-PVC (paste PVC) is used for softer products, like floors (INEOS ChlorVinyls - Extern miljöredovisning enligt EMAS, 2009). The size of the s-PVC particles is about 0.1 mm and the p-PVC particles are even smaller (INEOS ChlorVinyls –PVC från salt till återvinning).

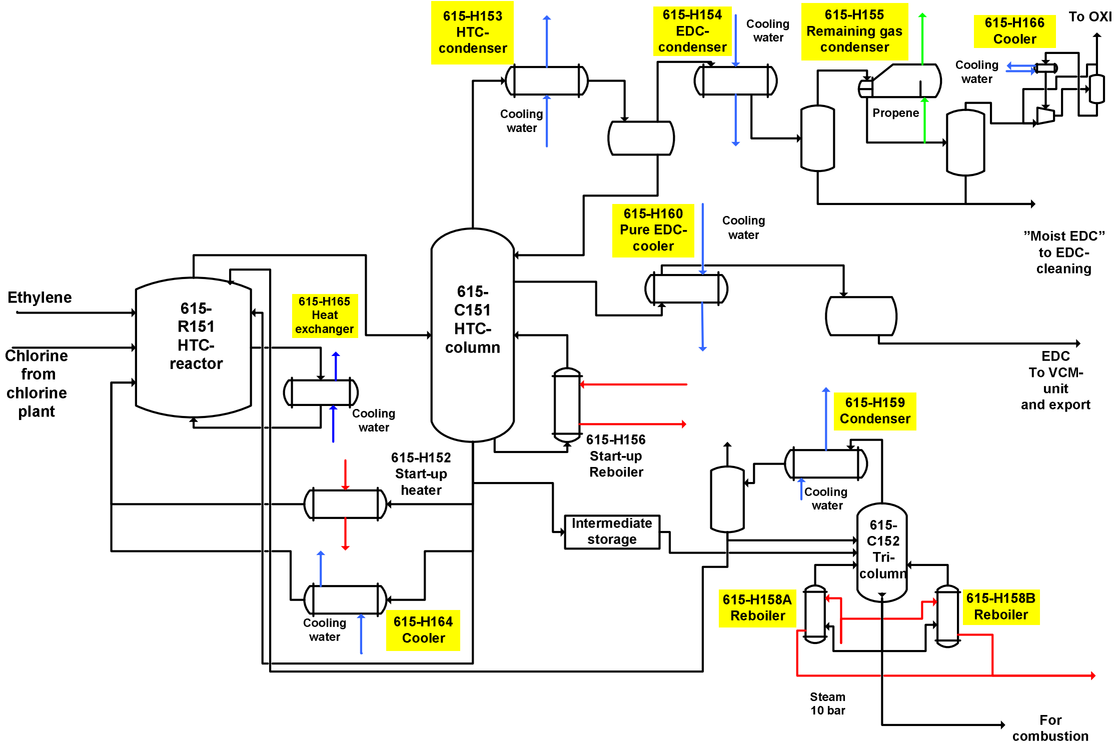


Figure 2.5 The HTC-unit

The main energy demand is for cooling in the HTC-unit. The only heating needed is in the tricolumn. In the existing process, 8 heat exchangers are in use, although the reboilers and the condenser to the tricolumn are only used part of the time. H164 and H165 are not in use today, but will probably be in use when the production increases. The heat exchanger with the largest duty is the HTC-condenser at the top of the HTC-column.

Table 2.3 Calculated loads of the heat exchangers in the HTC unit, for current operating conditions

	Description:	*Load [kW]:*	*Utility:*
H153	HTC-condenser	9500	Cooling water
H154	EDC-condenser	150	Cooling water
H155	Remaining gas condenser	5	Propene
H158AB	Reboiler Tri-column HTC	170	10 bar steam
H159	Condenser Tri-column HTC	160	Cooling water
H160	Pure EDC-cooler	360	Cooling water
H166	Cooler	30	Cooling water

In H155 the process fluid is cooled down to -16°C, and thus refrigerant is required instead of cooling water.

2.2.2 The VCM unit

In the VCM unit, the EDC produced in the other parts reacts and forms VCM (Vinyl chloride). VCM is the desired product of the VCM plant. In this unit, hydrochloric acid is formed as well, and is transported to the OXI unit (see chapter 2.2.3). Moist EDC is left as a by-product, and is transported to the EDC-cleaning part.

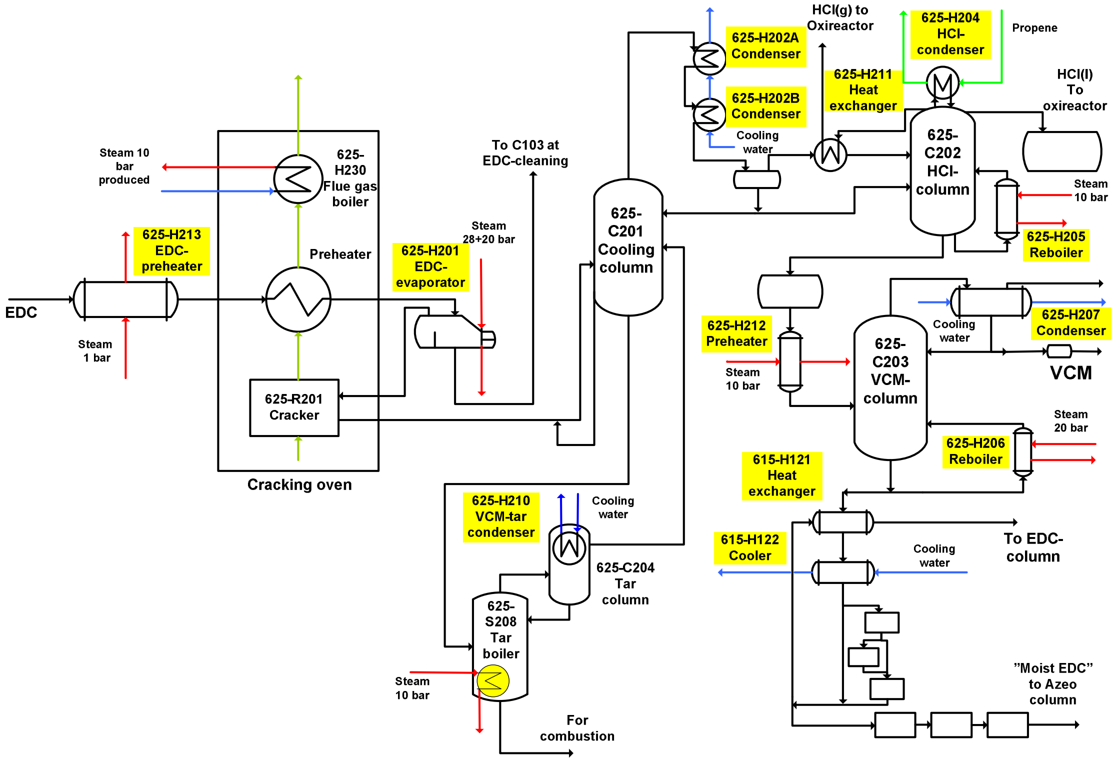


Figure 2.6 The VCM unit

In the VCM unit, 14 different heat exchangers are studied. The largest cooling demand occurs in the condensers on the cooling column after the cracker. The cracker consumes a lot of fuel but was not included in this study, since it is not possible to replace the fuel with increased heat exchanging. The heat in the flue gases was not included in the main analysis, but is discussed in a later section (Chapter 4.7), where increased energy utilisation from the flue gases was investigated.

Table 2.4 Calculated loads of the heat exchangers in the VCM unit, for current operating conditions

	Description:	*Load [kW]:*	*Utility:*
H121	From VCM-column/To EDC column	300	
H122	Cooler before R102	390	Cooling water
H201	EDC-evaporator	3400	20 bar steam
H202A	Condenser cooling column	7000	Cooling water
H202B	Condenser 2 cooling column	2100	Cooling water
H204	HCl condenser	1600	Propene
H205	Reboiler HCl-column	1600	10 bar steam
H206	Reboiler VCM column	2200	20 bar steam
H207	Condenser VCM column	2900	Cooling water
H210	VCM tar condenser	110	Cooling water
H211	HCl preheater/cooler inlet HCl column	110	
H212	Preheater ingoing stream to VCM-column	460	10 bar steam
H213	EDC-preheater	2000	1 bar steam
S208	VCM tar boiler	180	10 bar steam

The only two internal heat exchangers in the VCM plant are located in the VCM unit. The HCl-condenser, H204, operates at a temperature of -32°C and uses propene as a cold utility.

2.2.3 The OXI unit

In the OXI unit, the hydrochloric acid formed in the VCM unit is mixed with ethylene and oxygen and forms EDC. The EDC is further transported to the EDC-cleaning. The excess ethylene leaving the first three reactors is mixed with chlorine and forms EDC in the fourth reactor. The reaction in all four Oxi-reactors is exothermic, and the heat generated is used for production of steam, which is used in the VCM plant. As with the heat in the flue gases, the steam generated in the Oxi-reactors was not included in the main analysis, but is discussed in a later section (Chapter 4.7), where the demand for external hot utility was investigated.

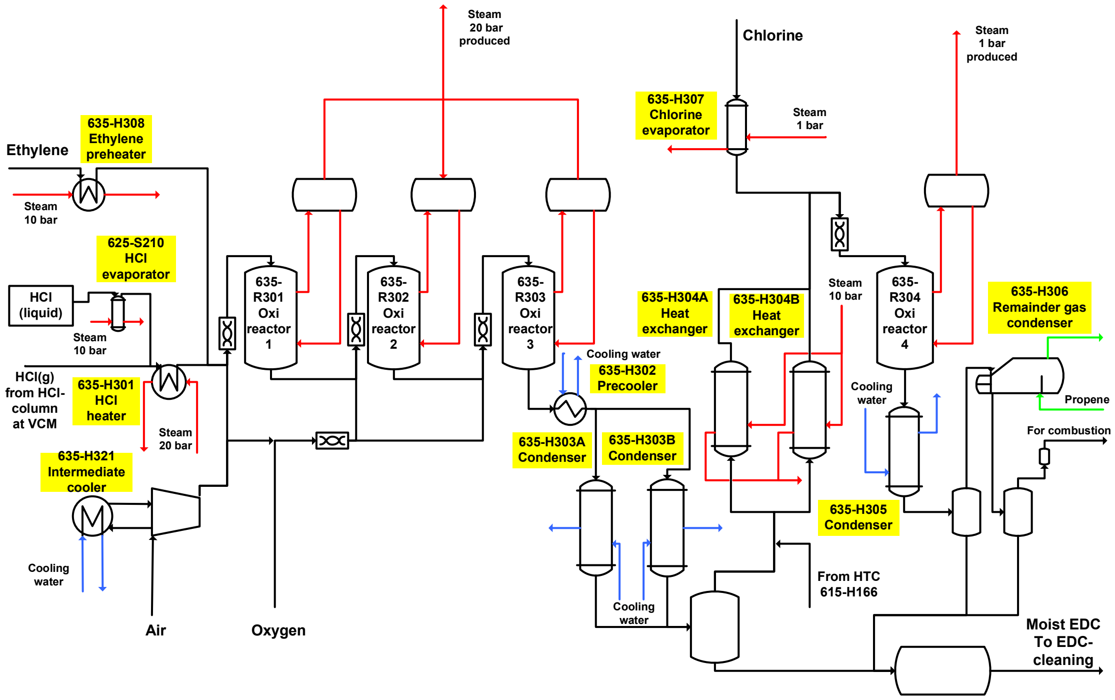


Figure 2.7 The OXI unit

In the OXI unit there are 12 heaters or coolers. The unit has a higher demand for cooling than for heating, because of the condensation after the exothermic reactors.

Table 2.5 Calculated loads of the heat exchangers in the OXI unit, for current operating conditions

	Description:	*Load [kW]:*	*Utility:*
H301	HCl heater	340	10 bar steam
H302	Precooler	2600	Cooling water
H303AB	Condenser after precooler H302	1600	Cooling water
H304AB	Heater before R304	250	10 bar steam
H305	Condenser after R304	380	Cooling water
H306	Remainder gas condenser	240	Propene
H307	Chlorine evaporator	110	1 bar steam
H308	Ethylene preheater	240	10 bar steam
H321	Intermediate air cooler	310	Cooling water
S210	HCl- evaporator	60	10 bar steam

Since the process fluid in H306 is to be cooled down to -21°C, propene is used.

2.2.4 The EDC-cleaning unit

In the EDC-cleaning, the moist EDC produced in the other three units is refined to EDC, which is then sent to the VCM unit for cracking. The EDC-cleaning consists of six columns, and hence there are six reboilers and five condensers. The heat demand and the cooling demand are similar, the largest load being located on the EDC-column.

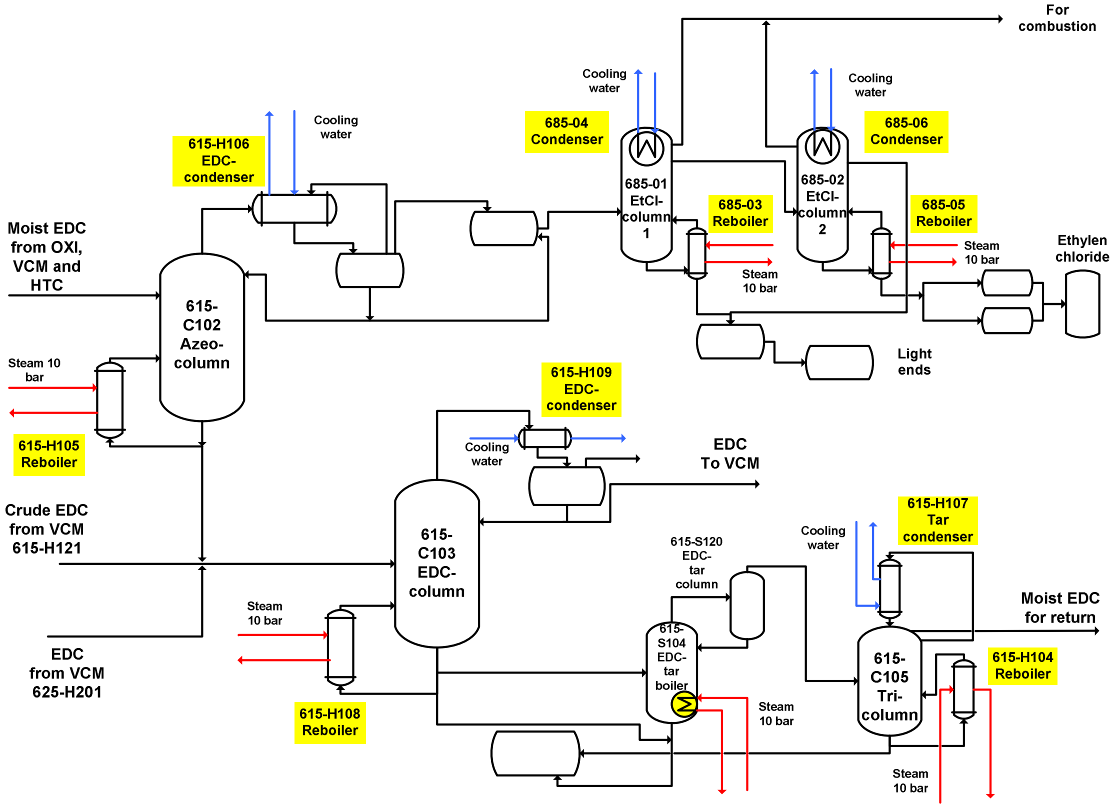


Figure 2.8 The EDC-cleaning unit

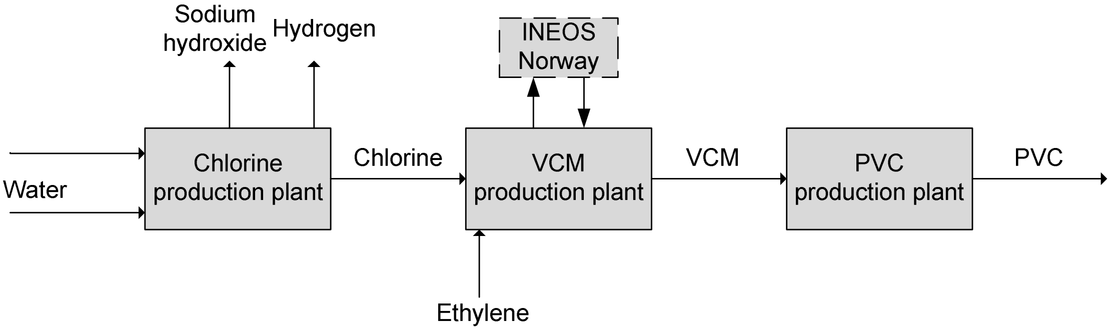
The reboiler at the EDC-column has the largest steam demand in the VCM-plant.

Table 2.6 Calculated loads of the heat exchangers in the EDC-cleaning unit, for current operating conditions

	Description:	*Load [kW]:*	*Utility:*
685-03	Reboiler EtCl-column 1	70	10 bar steam
685-05	Condenser EtCl column1	60	Cooling water
685-04	Reboiler EtCl-column 2	20	10 bar steam
685-06	Condenser EtCl-column 2	20	Cooling water
H104	Reboiler Tri-column EDC	70	10 bar steam
H105	Reboiler Azeo column	3900	10 bar steam
H106	Condenser Azeo column	2300	Cooling water
H107	Tar condenser (Condenser Tri-column EDC)	80	Cooling water
H108	Reboiler EDC-column	5800	10 bar steam
H109	Condenser EDC-column	6600	Cooling water
S104	EDC tar boiler	80	10 bar steam

2.3 Assumptions about the future VCM plant

After construction of the new chlorine plant, the production capacity is assumed to be increased by 100%. This means that the flow into the VCM production plant will be doubled. The EDC production part, the HTC unit, has a higher production capacity than utilized today and the production will have a twofold increase. However, the VCM production is limited, and the VCM unit is operating at its maximum today. The abundance of EDC will be exported to a site in Norway, from which VCM will be imported. Both the production and the import of VCM will remain unchanged, and hence there will be no change in the PVC production plant.



The flow rates into the HTC reactor will be doubled, and consequently the heat and cooling demands for the HTC unit will be doubled. However, the reflux from the HTC-condenser on top of the HTC-column is limited, and cannot exceed 120 m^3/h, in order to avoid flooding of the column. This means that some of the cooling demand has to be covered by other coolers. As seen in Figure 2.6, the heat exchangers that first will be taken into use will be the already existing heat exchangers H164 and H165, but there will probably be a need for investment in an additional cooler (Hnew) as well.

Today the tri-column is only used one third of the time. When the production is extended, it is assumed that the tri-column will be used continuously but on a level corresponding to 2/3 of the level today. The production of EDC is assumed to be doubled, and hence the load of the pure EDC-cooler, 615-H160, is also assumed to be doubled. The situation at the top of the HTC-column is more complicated. As mentioned, the reflux is limited, and cannot increase by more than 38%. However, the production of light products is assumed to be doubled, like the production of all other products. This means that a new steady-state is assumed to be reached, with a slightly higher content of volatile compounds in the stream leaving the top of the HTC-column. Since the HTC-condenser is substantially larger than the following coolers, it is possible for the reflux to increase by only 38%, and the mass flow through the following coolers to increase with 100%. The increased flow in the other heat exchangers will barely increase the load of the HTC-condenser, which increases by about 38%. The existing heat exchangers H164 and H165 were modelled using the software “Aspen-designer heat and rating” to approximate their cooling capacity. As earlier mentioned the additional cooling demand is assumed to be covered by a new cooler.

The production capacity of the VCM unit is limited and it is operating at its maximal capacity today, thus the production will remain unchanged. Since the production in the OXI unit is based on the level of hydrochloric acid produced in the VCM unit, it will remain unchanged as well. However, before the fourth reactor, there is a stream coming from the HTC-unit. This one is small compared to the other stream, and the total increase is assumed to be negligible.

The EDC-cleaning unit has incoming streams both from the HTC-unit, the VCM-unit and the OXI-unit. Both the VCM unit and the OXI-unit are assumed to operate at the same level as today. A brief comparison of the different flow rates indicates that if the flow rate from the HTC-unit is increased by a factor of two, and the flow rate from the VCM-unit and the OXI-unit remains constant, the inlet flow rate to the Azeocolumn will increase by approximately 6%. The inlet flow to the EDC-column will increase by 4%. Based on this, the production is assumed to remain unchanged in all units but the HTC unit.

Table 2.7 Summary of assumptions of the future production in the VCM plant

Unit:		*Description:*	*Future conditions:*
VCM			**No change in production**
OXI			**No change in production**
EDC			**No change in production**
HTC			
	H153	HTC-condenser	Maximal load before the column floods
	H154	EDC-condenser	Doubled load
	H155	Remaining gas condenser	Doubled load
	H158AB	Reboiler Tri-column HTC	2/3 of today's load
	H159	Condenser Tri-column HTC	2/3 of today's load
	H160	Pure EDC-cooler	Doubled load
	H166	Cooler	Doubled load
	H164	Cooler	Started
	H165	Cooler	Started
	Hnew	Cooler	Remaining load

3 Methodology

Pinch analysis was used to identify possible energy savings in the VCM plant. Process data was extracted from the process control system and from local measurement points. Simulations in Aspen plus were necessary to approximate flow rates and compositions not measured at the site, and to determine heat exchanger loads. Different measures for increasing the energy efficiency of the plant were investigated. Economic evaluations of the measures as well as an evaluation of reduced carbon dioxide emissions were performed.

3.1 Data extraction

Data was gathered using different sources. At the site, the process operation software “Aspen process explorer” is used, making it possible to access process data. Process data was taken as a mean value for a representative time period. All data and the time period used are listed in Appendix 2. Since the value over a time period can be seen, this is the most trustworthy data. Some data not included in the database was taken from local measurement points, such as pressure indicators, being more uncertain since such data is not measured and recorded continuously. Additional temperatures needed are measured at the pipes in the site, assuming no temperature difference on the inside and the outside of the pipe in the case of steel pipes. In some cases with no possibilities for measurements, data was taken from the site’s process descriptions and equipment design data.

For a plan of the process, including measurement points, please see Appendix 1.

At certain points in the process extractions of the streams are analyzed. Their composition is reported in a program called LabMaster. The points where the compositions are measured are shown in the process plan in Appendix 1. For the case where the composition of the stream is unknown, it is primarily calculated, secondarily simulated in Aspen, and if nothing else is possible the composition is taken from an old Aspen plus simulation of the whole VCM plant, performed in 2001. For process data and assumptions for each heat exchanger, please see Appendix 4.

3.2 Aspen simulations

Aspen plus was used to calculate the heat transferred in the heat exchangers. In some of the cases, where the temperatures in and out of the heater/cooler and the pressure, flow rate and composition were known, the simulations were used only to determine the load of the heat exchangers. However, in some other cases, process simulations had to be used to determine stream data not measured in the process.

The property method used is NRTL (Non-random two-liquid model). The NRTL model can describe vapour-liquid equilibrium for non-ideal solutions, according to the built-in help function in Aspen plus. The NRTL equation uses activity coefficients and has advantages compared to simpler models when handling non ideal-mixtures (Prausnitz, Lichtenthaler, & Azevedo, 1999). One exception is the two coolers after the first three Oxi-reactors in the OXI-part, which is modelled separately. In this case, hydrochloric acid is formed and dissolved into water, for which an electrolyte package had to be activated and the property method ENRTL-RK was used.

The composition and the flow rate leaving the Oxi-reactors had to be calculated manually, since there were no measurement points at all. For calculations of the flow rate and composition leaving the Oxi-reactors, please see Appendix 3.

In Appendix 4, there is a summary of the results for the different heat exchangers, and of what input data and assumptions that were used. Because of uncertainties in the data and assumptions, there will be uncertainties in the results. The uncertainties most affecting the results are discussed in Chapter 4.8.

3.3 Pinch analysis

Pinch analysis is a systematic method for identifying opportunities for improving the integration of processes in order to decrease the amount of external heating and cooling needed. This is achieved by increasing the share of heating and cooling that is done by internal heat exchanging. Using pinch analysis, the minimum heating and cooling demand and the maximum potential for heat exchanging can be determined. Thereafter, improvements of the current network are investigated.

A material process flow which must be heated or cooled is defined as a stream. A stream which needs to be heated is defined as a cold stream, and a stream in need of cooling is defined as a hot stream, regardless of their absolute temperature. The heat loads for all hot streams over a temperature range is added, forming the hot composite curve. In the same way a cold composite curve can be formed (Kemp, 2007).

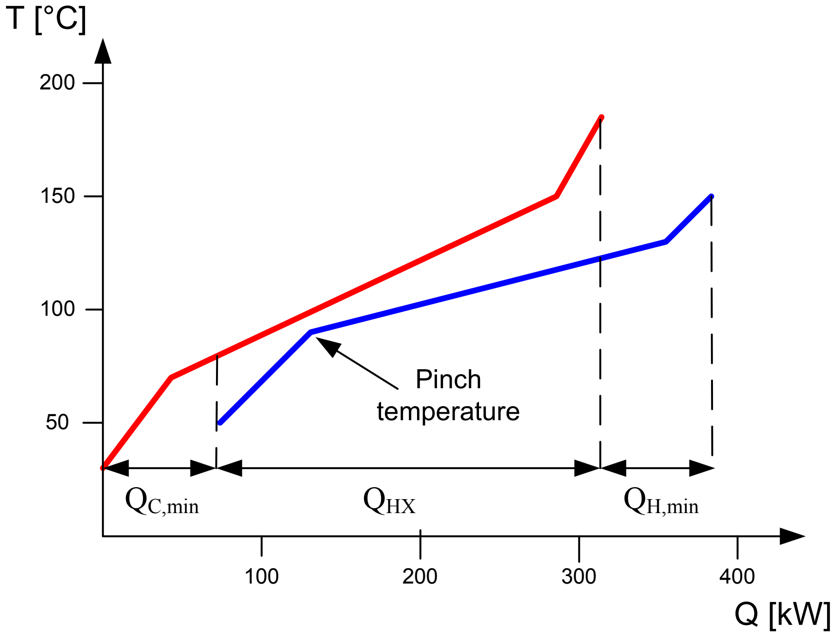


Figure 3.1 Composite curves

The red (uppermost) line in Figure 3.1 represents the hot composite curve, whereas the blue (bottommost) line represents the cold composite curve. For a chosen minimum allowable temperature difference (ΔT_{min}) in the heat exchangers, the pinch point can be determined. The pinch point is the point where the minimum temperature difference occurs, and limits how much heat that it is possible to recover internally within the process (Q_{HX}). Also the minimum heating and cooling demands ($Q_{H,min}$ and $Q_{C,min}$) can be calculated from the composite curves. The minimum heating and cooling demands are the needs for heating and cooling if maximal internal heat exchange is applied.

The composite curve temperature profiles can be plotted in so called shifted or interval temperatures, which are obtained by adding $\Delta T_{min}/2$ to cold streams and subtracting the same from hot streams (Kemp, 2007). This means that if the cold and hot composite curves intersect, the temperature difference is equal to the chosen minimum temperature difference.

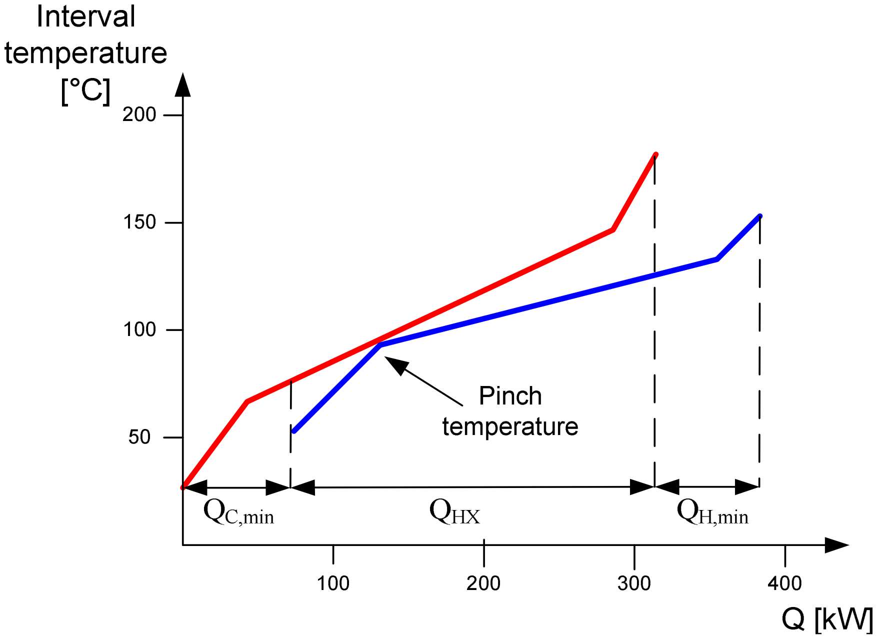


Figure 3.2 Composite curves with interval temperatures

In a similar way the grand composite curve (GCC) is made, where all streams (both hot and cold) in the temperature range is added, resulting in only one curve. Interval temperatures should be used. A positive slope of the curve indicates a heat demand, and a negative slope indicates excess heat and a cooling demand.

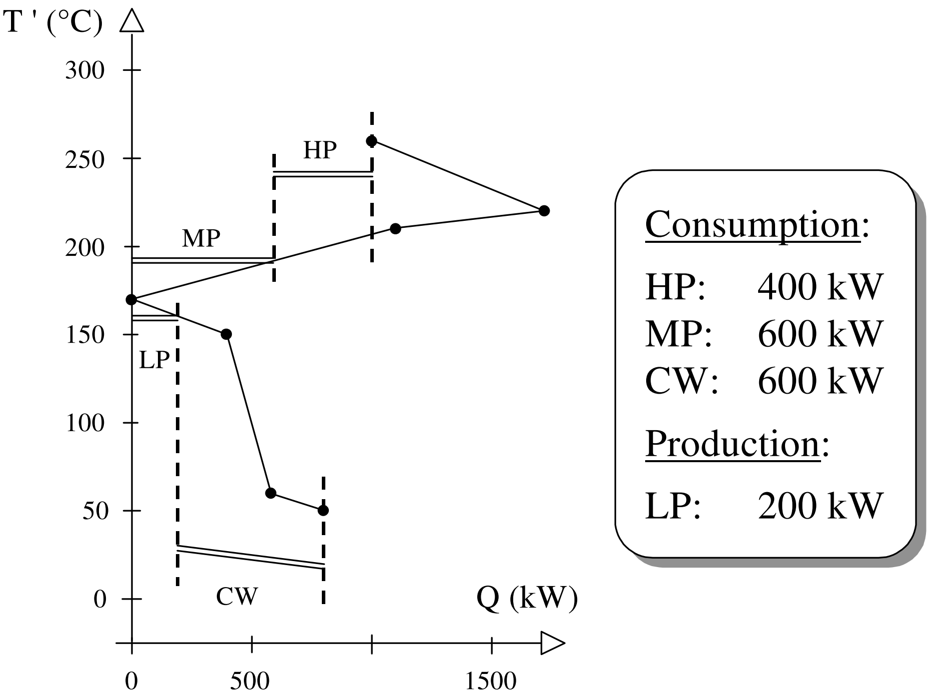


Figure 3.3 Example of a grand composite curve (Harvey, S et al., 2009)

As in the composite curves, the minimum heating and cooling demands and the pinch temperature can be seen. To the right of the dotted line in the upper part of Figure 3.3, a so-called pocket can be seen. In the pockets, no external heating or cooling is needed. The GCC can be used to conduct a graphical analysis of the utility levels needed and how the corresponding utility loads should be set to achieve minimum utility costs. Also the potential for producing steam or hot water for district heating, or for installation of heat pumps, can be determined.

Above the pinch, only external heating is needed. In the same way there is only need for external cooling below the pinch. When designing a heat exchanger network for obtaining minimum heating and cooling, the design should start at the pinch point and thereafter work outwards. (Kemp, 2007)

There are three “golden rules” when designing a network for maximum heat recovery. Above the pinch there should be no cooling, below the pinch there should be no heating, and no heat should be transferred across the pinch (Kemp, 2007). Violations of these rules result in a higher use of energy than needed for a maximum heat recovery network.

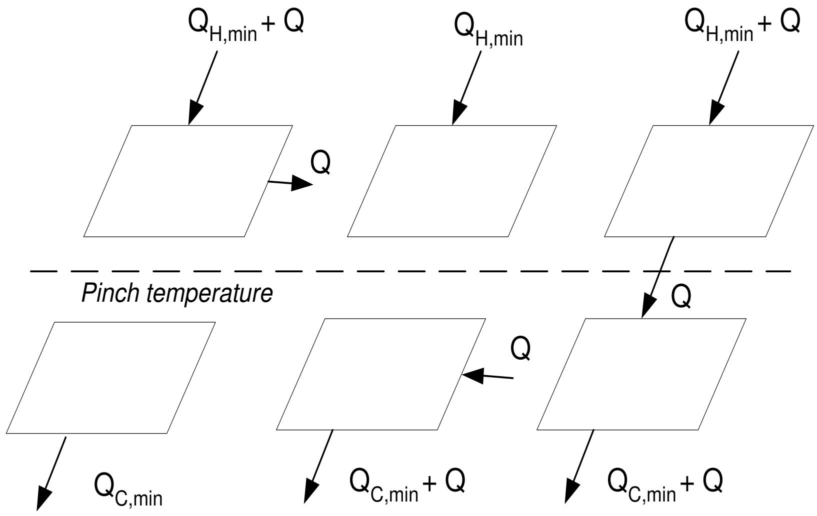


Figure 3.4 Illustration of violations of the three golden rules (Harvey, S et al., 2009)

When studying a site in order to save energy, it is important to identify pinch violations and remove them to improve heat efficiency. Pinch violations can be removed by modifying the temperature range in which the heat exchangers operate, either by installing a new heat exchanger or by changing the order of the existing ones.

A pinch analysis is performed as follows:
The first step is to identify which streams that should be included, and thereafter extract stream data. Data extraction is often the most time consuming part of a pinch analysis. The data needed is start and target temperature, and the heat content or demand of the stream. When the data extraction is done, an appropriate value of ΔT_{min} is to be chosen. The minimum temperature difference can be set as a global value, or as individual values for each stream. A small minimum temperature difference will allow more heat to be transfered, but will also increase the heat exchanger areas needed. The pinch analysis determines the pinch temperature as well as the minimum process heating and cooling demands. Comparison with the current demands allows the potential for improvements to be determined. Thereafter, the existing network of heaters, coolers and heat exchangers can be examined to identify pinch violations. Pinch violations leads to a higher energy demand than the minumim, as shown in Figure 3.4. When the pinch violations in the existing system are identified, ways to eliminate them should be examined. This could be done by modifying existing heat exchangers, or adding new ones.

For the pinch analysis, the Excel add-in “Pro-pi” was used.

3.4 CO_2 emissions evaluation

Today, both hydrogen produced at the site and fuel gas bought from a nearby company are used as boiler fuels for steam production. The fuel assumed to decrease in usage when the heat demand decreases is the purchased fuel gas, a mixture of mainly hydrogen and methane. In Table 3.1, the main properties of the fuel are shown. The detailed composition of the gas can be found in Appendix 5.

Table 3.1 Average values purchased fuel, February 2011

	Composition [Mol %]	*LHV [MJ/kg]*	*CO_2 Emission factor [ton/TJ fuel]*
Hydrogen	46.6		
Methane	49.2		
Mixture		56.4	44.3

Assuming a boiler efficiency of 0.9 gives:

$$\mathrm{CO_2 emission\ factor}\left[\frac{\mathrm{ton}}{\mathrm{TJ_{steam}}}\right] = \frac{\mathrm{CO_2 emission\ factor}\left[\frac{\mathrm{ton}}{\mathrm{TJ_{fuel}}}\right]}{0{,}9} = 49.194 \qquad (3.1)$$

This corresponds to emissions of 180 kg CO_2/MWh_{steam}. A reduction of the steam use in the process by 1 MW results in an annual steam saving of 30.326 TJ (8.42 GWh) assuming an annual downtime of 2 weeks (running time 8424 h/yr). The decrease in CO_2 emissions is calculated as follows:

$$\mathrm{Annual\ reduction\ in\ CO_2 emissions}\left[\frac{\mathrm{ton\ CO_2}}{\mathrm{MW_{steam\ reduced}}\cdot\mathrm{yr}}\right] =$$

$$= 49.194\left[\frac{\mathrm{ton\ CO_2}}{\mathrm{TJ_{steam}}}\right]\cdot 30.326\left[\frac{\mathrm{TJ_{steam}}}{\mathrm{MW_{steam\ reduced}}\cdot\mathrm{yr}}\right] = 1492\left[\frac{\mathrm{ton\ CO_2}}{\mathrm{MW_{steam\ reduced}}\cdot\mathrm{yr}}\right] \qquad (3.2)$$

This gives an annual decrease in CO_2 emissions of 1.5 kton per MW steam reduced. For comparison, the annual carbon dioxide emissions from the whole site in Stenungsund are about 60 kton/yr (INEOS ChlorVinyls, 2009).

The site is participating in the European Emissions Trading Scheme (EU ETS). As from 2013, the distribution of emission certificates will decrease, and instead of being based on the emissions from the site, they will be based on the most carbon dioxide efficient sites in Europe (Swedish Environmental Protecting Agency, 2011). As a result, the economic value of reduced carbon dioxide emissions can be estimated from the value of emission certificates. During 2009 and 2010, the price of emission certificates was around 15€/ ton CO_2 (Swedish Energy Agency, 2010). Together with the emissions calculated in equation 3.1 and 3.2, this gives that the value of the avoided carbon dioxide emission for 1 MW of reduced steam is 22000€/yr, which corresponds to 205000 SEK/yr[2]. This can also be expressed as 24 SEK/ MWh_{steam}.

The price is likely to rise in the future, when the amount of certificates issued for free is decreased.

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[2] The exchange rate used for SEK/€ is an average value for the time period May 2010 to May 2011 and is set to 9.17 SEK/€ (E24, 2011).

3.5 Economic evaluation

The retrofit proposals were evaluated with respect to economic performance. This was done by using estimated investment costs, and estimated annual utility cost savings and savings because of reduced carbon dioxide emissions to determine payback period for investments in heat recovery equipment.

3.5.1 Investment costs

The investment costs were estimated using the factorial method of cost estimation (Sinnot & Towler, 2009).

First, the area needed for the heat exchanger was calculated from:

$$Q = UA_{hx}\Delta T_{lm}, \tag{3.3}$$

Where Q is the heat transferred in the heat exchanger, U is the overall heat transfer coefficient (please see Appendix 6), and ΔT_{lm}, the mean logarithmic temperature, is calculated from:

$$\Delta T_{lm} = \frac{(T_{hot,out} - T_{cold,in}) - (T_{hot,in} - T_{cold,out})}{ln\left(\frac{(T_{hot,out} - T_{cold,in})}{(T_{hot,in} - T_{cold,out})}\right)} \tag{3.4}$$

The heat exchanger is assumed to be aU-tube shell and tube heat exchanger, and the cost of a carbon steel heat exchanger is approximated with (Sinnot & Towler, 2009):

$$C_{e,CS,2007} = 24000 + 46 \cdot A_{hx}^{1,2} \quad , \qquad [\$] \tag{3.5}$$

where A_{hx} is the heat transfer area in the heat exchanger in [m^2], and $C_{e,CS,2007}$is the equipment cost for a carbon steel heat exchanger on a US Gulf Coast basis, Jan 2007. (CE index 509.7). In November 2010 the CE index was 556.7 (Chemical engineering, 2011), giving the equipment cost:

$$C_{e,CS} = C_{e,CS,2007} \cdot \frac{\text{CE index 2010}}{\text{CE index2007}} \qquad [\$] \tag{3.6}$$

The cost of the heat exchanger, including installation costs, is calculated from (Sinnot & Towler, 2009):

$$C = C_{e,CS} \cdot \left(\left(1 + f_p\right) \cdot f_m + \left(f_{er} \cdot f_{el} \cdot f_i \cdot f_c \cdot f_s \cdot f_l\right)\right), \tag{3.7}$$

where C is the inside battery limits (ISBL) cost, and $C_{e,CS}$ is the cost of the equipment in carbon steel. Together with Eq. 3.6, this gives the inside battery limit cost:

$$C = C_{e,CS,2007} \cdot \frac{\text{CE index 2010}}{\text{CE index2007}} \cdot \left(\left(1 + f_p\right) \cdot f_m + \left(f_{er} + f_{el} + f_i + f_c + f_s + f_l\right)\right) \tag{3.8}$$

The constants used are found in

Table 3.2 and Table 3.3.

Table 3.2 Typical factors for estimation of project fixed capital cost (Sinnot & Towler, 2009)

Constant	*Explanation*	*Value of constant (Process type: fluids)*
f_{er}	Equipment erection	0.3
f_p	Piping	0.8
f_i	Instrumentation and control	0.3
f_{el}	Electrical	0.2
f_c	Civil	0.3
f_s	Structures and buildings	0.2
f_l	Lagging and paint	0.1

The material constants used are found in Table 3.3. The material used in the calculation will be the one used in the existing heater/cooler on the stream. The process fluids are often corrosive, and expensive materials are needed in the heat exchangers. In most of the existing heat exchangers, Monel is used. However, in the heat exchanger after the Oxi-reactors, H302, tantalum is used. Since no material factor for tantalum is found, it is approximated with Inconel.

Table 3.3 Material cost factors (Sinnot & Towler, 2009)

Material	f_m
Carbon steel	1
304 Stainless steel	1.3
Monel	1.65
Nickel and Inconel	1.7

The total fixed capital cost includes as well design and engineering cost and other offsite costs. The cost factors used are showed in Table 3.4.

Table 3.4 Outside battery limits costs (Sinnot & Towler, 2009)

Constant	*Explanation*	*Value of constant (Process type: fluids)*
OS	Offsite costs	0.3
D&E	Design and engineering	0.3
X	Contingency	0.1

The total fixed capital cost can be calculated by equation (3.5)(Sinnot & Towler, 2009).

$$C_{FC} = C \cdot (1 + OS) \cdot (1 + D\&E + X) \qquad [\$] \tag{3.9}$$

The exchange rate used for USD/SEK is an average value for the time period May 2010 to May 2011 and is set to 6.92 SEK/USD (E24, 2011) .

3.5.2 Annual savings

As in the evaluation of emissions in Chapter 3.4, the decreased use of fuel is assumed to lead to reduced purchase of fuel gas.

The annual savings are calculated from the reduced use of fuel. The price of the fuel is 95 SEK/GJ$_{\mathrm{fuel}}$ (342SEK/MWh$_{\mathrm{fuel}}$), and the efficiency of the steam boilers is assumed to be 0.9. In the calculations of annual savings, the plant is assumed to have an annual downtime of two weeks.

$$\text{Fuel savings } \left[\frac{\text{SEK}}{\text{yr}}\right] = Q_{Steam\ reduction}[\text{MW}] \cdot 8424 \left[\frac{\text{h}}{\text{yr}}\right] \cdot 3{,}6 \left[\frac{\text{GJ}}{\text{MWh}}\right] \cdot \frac{95 \left[\frac{\text{SEK}}{\text{GJ}_{\text{fuel}}}\right]}{\eta_{boiler}} \tag{3.10}$$

This gives reduced fuel costs of 380 SEK/MWh$_{\mathrm{steam}}$ saved, and annual fuel savings of 3.2 MSEK per MW steam saved.

In Chapter 3.4, the savings resulting from decreased carbon dioxide emissions were calculated to 24 SEK/ MWh$_{\mathrm{steam}}$ reduced.

This gives that the total savings are about 400 SEK/MWh$_{\mathrm{steam}}$ saved, and the annual savings are 3.4 MSEK per MW steam reduced in the process.

3.5.3 Payback time

The payback time is calculated as the time to recover the investment cost based on the annual utility cost savings and savings because of reduced carbon dioxide emissions.

$$\text{Payback time [yr]} = \frac{\text{Investment cost [MSEK]}}{\text{Annual savings } \left[\frac{\text{MSEK}}{\text{yr}}\right]} \tag{3.10}$$

No interest rate was considered, since the pay-back period turned out to be short.

4 Results

4.1 System definition and process heating and cooling requirements

About 50 heat exchangers were included in the study. Most of them are heaters and coolers using heating or cooling utilities, only two are internal stream heat exchangers. Steam at 1, 10, 20 and 28 barg is used for heating, and cooling water and propene are used for cooling. Heat exchanges only used for start-up were not included in the study.

This thesis has focused on the future production, but an analysis of today's plant was made as well, and is presented in Appendix 8. The current heat demand for the process is 20.9 MW, and the cooling demand is 38.6 MW. The potential for internal heat exchanging is 7.8 MW. All the results shown in this chapter are for the future production level.

The future plant will have a larger cooling demand, because of the increased cooling demand in the HTC unit. The heat demand remains almost unchanged. If no energy efficiency measures are implemented when extending the production, the plant will have a heat demand of 20.8 MW and a cooling demand of 48.9 MW.

The estimated loads of the heat exchangers in the future system are presented in Table 4.1. A more detailed table, with for example condensation temperatures, is found in Appendix 9.

Table 4.1 Process heating and cooling requirements in the future VCM plant, used for pinch analysis

Unit	At:	Description	Tstart [C]	Ttarget [C]	Q [kW]	Utility
685-03	EDC	Reboiler EtCl-column 1	63.8	65	66	10 bar steam
685-04	EDC	Condenser EtCl column1	62.9	59.1	60	Cooling water
685-05	EDC	Reboiler EtCl-column 2	58.8	59	19	10 bar steam
685-06	EDC	Condenser EtCl-column 2	52.9	32.8	19	Cooling water
H104	EDC	Reboiler Tri-column EDC	133.6	145.6	74	10 bar steam
H105	EDC	Reboiler Azeo column	127.7	128.5	3858	10 bar steam
H106	EDC	Condenser Azeo column	85.6	30	2296	Cooling water
H107	EDC	Tar condenser (Condenser Tri-column EDC)	87.9	86.5	80	Cooling water
H108	EDC	Reboiler EDC-column	111	119.1	5803	10 bar steam
H109	EDC	Condenser EDC-column	89.3	35	6634	Cooling water
S104	EDC	EDC tar boiler	140	141	78	10 bar steam
H153	HTC	HTC-condenser	82.4	74.4	13065	Cooling water
H154	HTC	EDC-condenser	73.5	15.8	300	Cooling water
H155	HTC	Remaining gas condenser	15.8	-16.5	10	Propene
H166	HTC	Cooler	57.5	34	65	Cooling water
H160	HTC	Pure EDC-cooler	89.2	26.7	1011	Cooling water
H159	HTC	Condenser Tri-column HTC	87.1	13.9	104	Cooling water

H158AB	HTC	Reboiler Tri-column HTC	118	123.5	114	10 bar steam
H165	HTC	Cooler R151	115	40	1600	Cooling water
H164	HTC	Cooler HTC-column	100	40	1200	Cooling water
Hnew	HTC	New cooler HTC	100	40	3099	Cooling water
H301	OXI	HCl heater	22	179.6	342	10 bar steam
H302	OXI	Precooler	217.2	93.9	2618	Cooling water
H303AB	OXI	Condenser after precooler H302 after R303	93.9	30	1641	Cooling water
H304AB	OXI	Heater before R304	30.7	137	245	10 bar steam
H305	OXI	Condenser after R304	151.5	40	389	Cooling water
H306	OXI	Remainder gas condenser	40	-21.5	240	Propene
H307	OXI	Chlorine evaporator	3.4	69.6	109	1 bar steam
H308	OXI	Ethylene preheater	-0.4	119.9	237	10 bar steam
H321	OXI	Intermediate air cooler	160	25	307	Cooling water
S210	OXI	HCl- evaporator	-31.9	41	57,2	10 bar steam
H121-cold	VCM	From VCM-column/To EDC column	41	98	300	Internal heat exchanging
H121-hot	VCM	From VCM-column/To EDC column	157.7	118	300	Internal heat exchanging
H122	VCM	Cooler before R102	118	60	389	Cooling water
H201	VCM	EDC-evaporator	193.1	208.61	3442	20 bar steam
H202A	VCM	Condenser cooling column	132.8	66	7007,5	Cooling water
H202B	VCM	Condenser 2 cooling column	66	40	2098	Cooling water
H204	VCM	HCl condenser	-31.7	-32.3	1640	Propene
H205	VCM	Reboiler HCl-column	86.5	144.7	1559	10 bar steam
H206	VCM	Reboiler VCM column	157.7	158.1	2210	20 bar steam
H207	VCM	Condenser VCM column	41.4	17	2880	Cooling water
H210	VCM	VCM tar condenser	92.4	52.1	113	Cooling water
H211-cold	VCM	HCl preheater/cooler in HCl column	-32.3	21	110	Internal heat exchanging
H211-hot	VCM	HCl preheater/cooler in HCl column	40	13	110	Internal heat exchanging
H212	VCM	Preheater ingoing stream to VCM-column	81.8	85	457	10 bar steam
H213	VCM	EDC-preheater	26.3	118.2	1980	1 bar steam
S208	VCM	VCM tar boiler	140	141	178	10 bar steam

In Chapter 4.2 to 4.5, only the heat exchangers in the future VCM plant are included, and the analysis is based on Table 4.1. In Chapter 4.6, the possibilities to use excess heat from the VCM plant in the PVC plant are investigated. In these chapters, measures for increased energy efficiency by increased heat transfer are discussed. Finally, in Chapter 4.7, the future VCM plant is analyzed again, but this time the heat produced from the Oxi-reactors and the heat in the flue gases are included. This has been made to estimate the amounts of external steam needed, and to investigate if the heat in the flue gases could be utilized to a higher extent.

4.2 Pinch analysis of the future VCM plant

The pinch analysis is based on the stream data presented in Table 4.1.

For the pinch analysis, a global minimum temperature difference in the heat exchangers (ΔT_{min}) of 10 °C was chosen. A low minimum temperature difference increases the amount of heat possible to transfer, but also increases the areas needed for the heat exchangers. When observing the composite curves in Figure 4.1, there is no sign of possible threshold effects, and the heat demand would probably not change very much if the minimum temperature difference were to be decreased or increased.

As mentioned earlier in the chapter, the future plant will have a steam demand of 20.8 MW and a cooling demand of 48.9 MW if nothing is done to increase the energy efficiency.

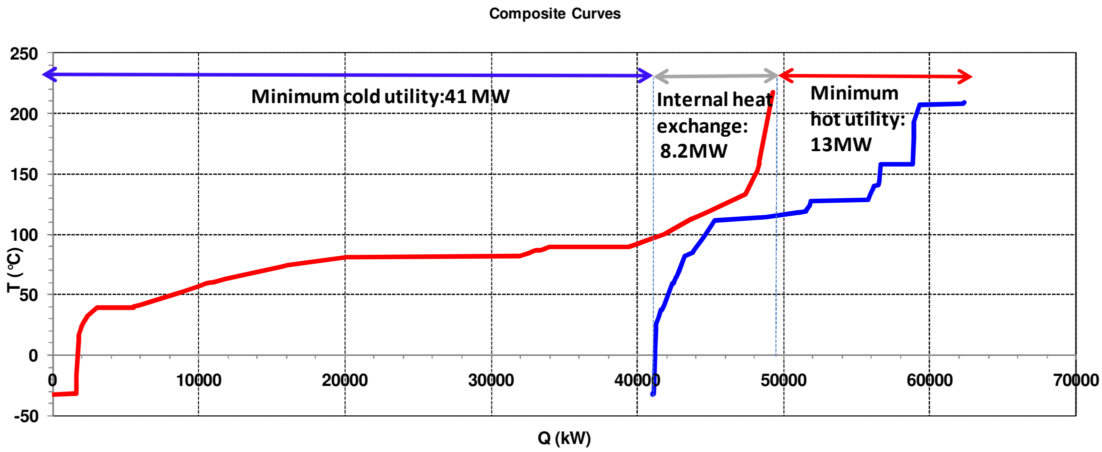

Figure 4.1 Composite curves for the future VCM plant, $\Delta T_{min}=10°C$

The future VCM plant will have a pinch temperature of 116°C (interval temperature scale), corresponding to a pinch temperature of 111°C for cold streams and 121°C for hot streams.

From the composite curves shown in Figure 4.1, it can be seen that 8.2 MW of both heating and cooling demand can be covered with internal heat exchange. Only 0.4 MW is heat exchanged today, giving an additional potential for heat exchanging of 7.8 MW. This means that the steam consumption could theoretically be reduced by 7.8 MW, corresponding to a reduction of 37% of the future demand. The very large potential for improvements can be explained by the fact that in the future plant, if no efficiency measures are implemented compared to those in the existing plant, there will be only two internal heat exchangers and about 50 heaters and coolers.

It is interesting to note that the potential for internal heat exchange for the future production level is unchanged compared to the potential for the existing plant, which was presented in the beginning of Chapter 4. This means that the additional heat in the system when the production is increased does not increase the possibilities for heat transfer. The heat demand at those temperatures could already be covered with excess heat from other hot streams. However, it increases the possibilities for transferring excess heat to other parts of the site.

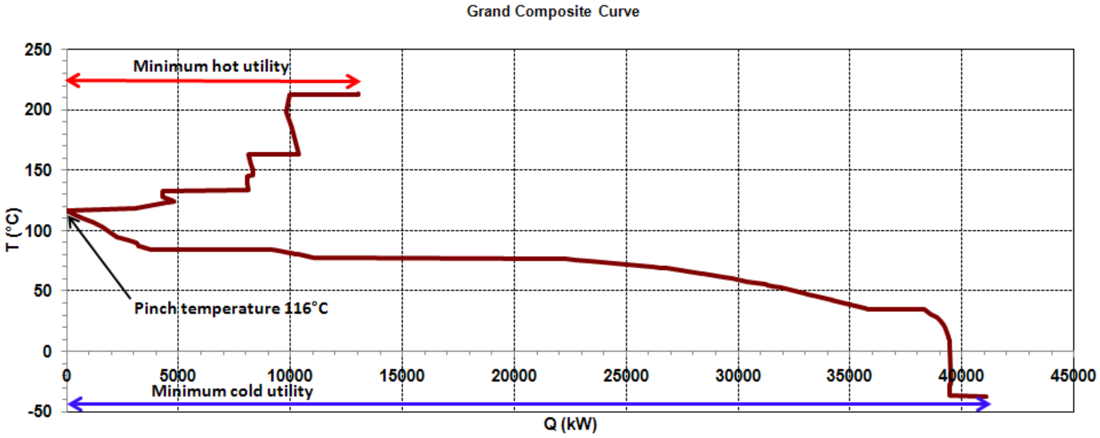

Figure 4.2 Grand composite curve future VCM plant, ΔT_{min}=10°C

As seen in the grand composite curve shown in Figure 4.2, there is a large excess of heat below the pinch temperature. This could be used either for district heat production, or for covering heat demands at other parts of the site, for example in the PVC plant. The grand composite curve (GCC) shows a surplus of 41 MW of heat, of which 32 MW is above 50°C.

The results for the heating and cooling demands are shown in Figure 4.3 and Figure 4.4.

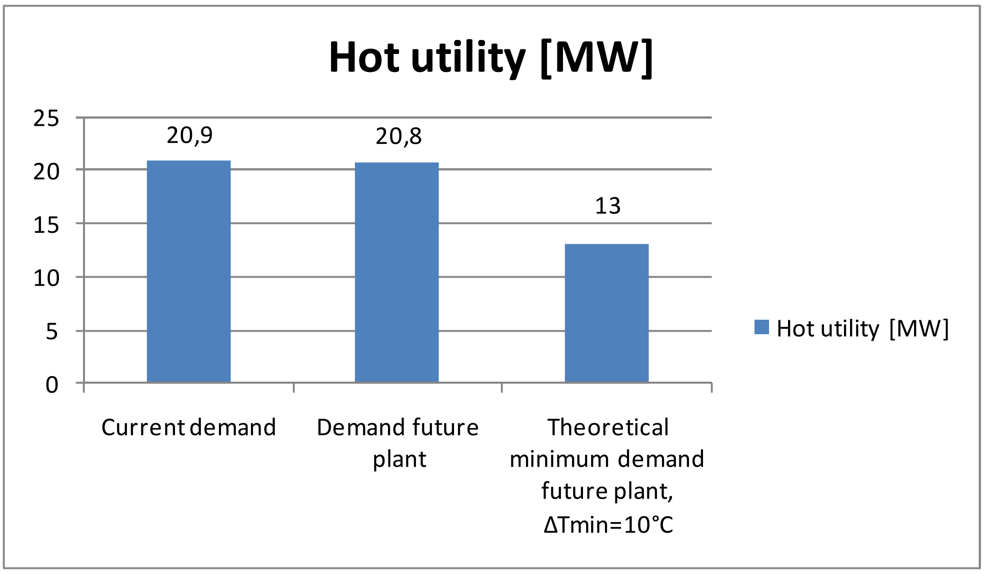

Figure 4.3 Hot utility demand in the VCM plant

According to Figure 4.3, there is a very small decrease in the heat demand for the future plant compared to the existing plant. This is no actual decrease, but is due to the fact that the reboilers of the Tri-column in the HTC-unit are supposed to be running continuously in the future, but at a level corresponding to 2/3 of the current. They are not running continuously today, but included in the pinch analysis as if they were.

Once again, the potential for energy savings of 7.8 MW should be noted, corresponding to 37% of the future demand.

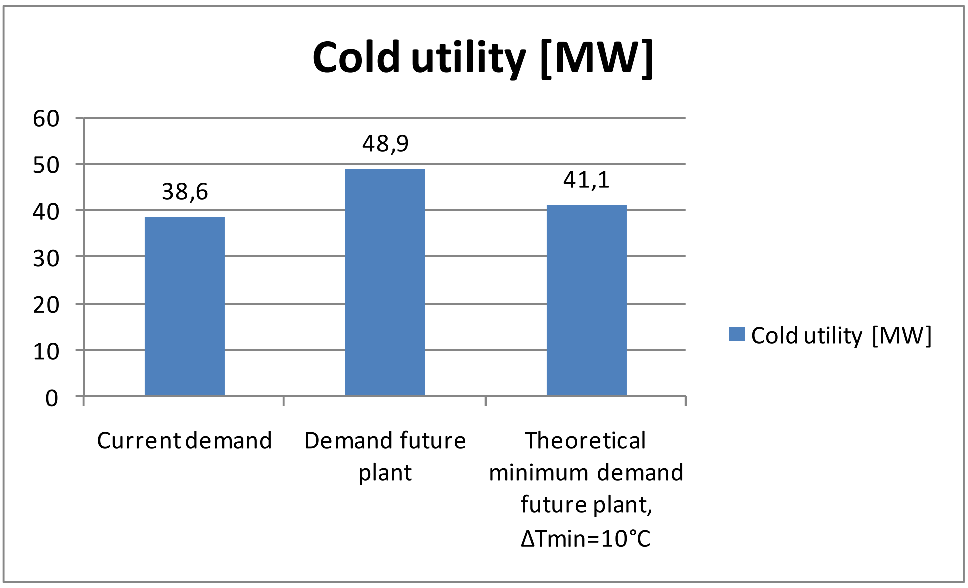

Figure 4.4 Cold utility demand in the VCM plant

From Figure 4.4, it can be seen that the cooling demand for the future plant will be higher than for the existing plant, regardless of whether increased heat exchanging in the VCM plant is implemented or not. If heat exchanging with other parts of the site is applied, the need for cooling utility might decrease more.

The theoretical potential for saving is naturally the same for hot and cold utility, since it is assuming internal heat exchange. If heat exchanging with other parts of the site is considered, it is possible that the potentials for reduction of heating and cooling demand in the VCM plant would differ, since heat is transferred across the system boundaries.

4.2.1 Possible measures for increased energy efficiency

From the composite curves and grand composite curves shown in Figure 4.1 and Figure 4.2, several options for increased energy efficiency can be identified. The potential for internal heat exchange is large, as well as the potential for exporting excess energy to other parts of the site. The grand composite curve also shows that there are possibilities for investigating heat pumps.

Internal heat exchange

As stated in Chapter 4.2, the potential for increased heat exchange is 7.8 MW. The possibility for increased heat exchange is the most thoroughly investigated of the three above mentioned options for improved energy efficiency. This is done by identifying pinch violations, which is showed in Chapter 4.3, and examining different ways of eliminating them, which is done in Chapter 4.4 and 4.5.

Export of excess heat

Another way for better utilisation of the energy is to use the excess heat from the VCM plant at other parts of the site. As seen in Figure 4.2, there is a huge heat surplus below the pinch temperature in the VCM plant. This could be used for example in the PVC plant. The PVC plant is at reasonable distance from the VCM plant, and possibilities for transferring heat from the VCM plant to the PVC plant are investigated in Chapter 4.6. This would not only reduce the demand for heat in the PVC plant, but also decrease the cooling demand of the VCM plant.

Heat pump

The heat pump option is the least studied of the mentioned options, because of the smaller potential. It is not discussed in the report other than here.

When considering a heat pump, it is important that the heat pump is placed across the pinch. This means that the heat pump should use energy from below the pinch, where there is an excess of heat, and deliver it above the pinch, where there is a deficit of heat.

The largest steam consumer in the plant is the reboiler of the EDC-column (H108). To cover parts of the demand, it could be possible to install a heat pump, using heat from the condenser. The temperature difference between the reboiler and the condenser is only about 20°C, making investigating a heat pump interesting. The EDC-column will be included in some of the proposed retrofits in Chapter 4.4. But since they do not cover all the demand of the reboiler, a heat pump could be considered regardless if an investment in increased heat exchange is made or not. The heat pump would also be placed across the pinch, causing no pinch violations.

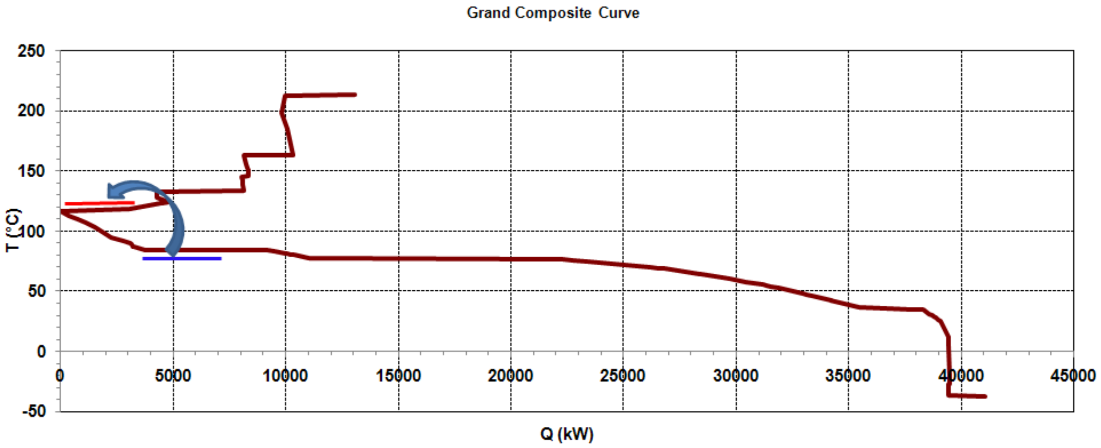

Figure 4.5 Grand composite curve for the future VCM plant, with $\Delta T_{min}=10°C$, showing a heat pump from the condenser to the reboiler of the EDC-column

A disadvantage with placing the heat pump at the EDC-column is that there could be problems with fouling.

4.3 Pinch violations in the VCM plant

For improving the energy efficiency of the plant and increase the internal heat exchange, it is important to find and eliminate the pinch violations. The analysis shows that there are a total of 15 pinch violations in the future system, giving total pinch violations of 7.8 MW. The pinch violations are shown in Appendix 11. None of the pinch violations are located in the HTC part, and it should be mentioned that the existing plant and the future plant will have the same pinch violations.

The five biggest pinch violations were investigated in detail, corresponding to about 83% of the total pinch violations. This means that if those five pinch violations are eliminated, most of the potential for heat saving in the site is achieved, and both the heating and the cooling demands could be reduced by 6.5 MW.

Table 4.2 The five largest pinch violations in the future VCM plant, ΔT_{min}=10°C

	At:	*Description:*	*Cooling above pinch [kW]:*	*Heating below pinch [kW]:*	*Heat through pinch [kW]*
H302	OXI	Precooler after Oxi-reactors	1876		
H202A	VCM	Condenser after cooling column	1680		
H205	VCM	Reboiler HCl-column		656	
H212	VCM	Preheater before VCM-column		457	
H213	VCM	EDC-preheater		1825	
Sum selected pinch violations:			3556	2938	
Total pinch violations:			3722	3787	277

Most of the pinch violations are located in the VCM and OXI units. In the OXI unit, not less than 8 out of the 10 heaters and coolers are pinch violations. However, only one of them is large enough to be considered further. The HTC unit is the only unit not having any pinch violations, since it almost only needs cooling and the temperatures are below the pinch temperature.

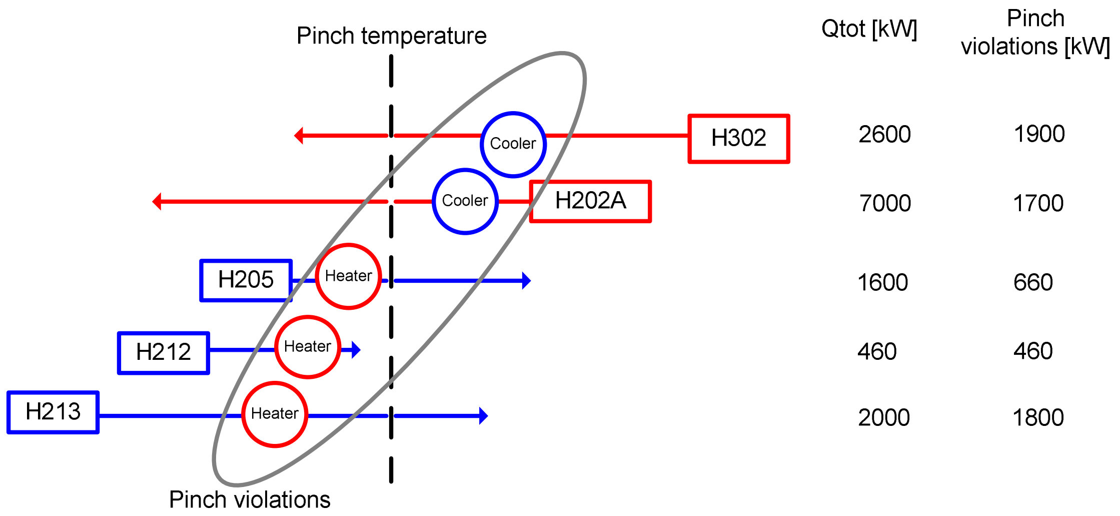

Figure 4.6 Illustration of the five largest pinch violations in the future VCM plant, ΔT_{min}=10°C

In Figure 4.6, the pinch violations are illustrated. The pinch temperature is at 116 °C (interval temperature scale).

Ways of eliminating the pinch violations will be investigated in the following chapter.

4.4 Proposed measures for increased internal heat transfer in the future VCM plant

In this chapter, the possibilities for increasing the internal heat exchange are investigated. Here, direct heat transfer is assumed. The possibilities of instead using a hot water system will be discussed in Chapter 4.5.

First, it is very important to note, that the streams are numbered by the heat exchanger that covers their heating or cooling demand today. When referring to them, it is not the actual heat exchanger that is referred to, but the stream passing it. When referring to for example H302, it is not the actual heat exchanger that is meant, but the stream today passing it, with start and target temperatures and a heat load as in the heat exchanger. When discussing heat exchanging between heat exchangers, it is thus between the streams in the current heat exchangers. For all retrofit suggestions, new heat exchangers have been considered. The possibilities of using the existing ones for new purposes are not considered. If the new heat exchangers proposed in the retrofits do not cover the total demand of the stream, the new heat exchanger is assumed to be placed on the upstream side of the existing one.

A maximum energy recovery (MER) network was created, and is shown in Appendix 12. In the MER network, 23 heat exchangers, 22 coolers and 12 heaters are used. An investment in 23 heat exchangers is unrealistic, but the MER network is made to show that it is possible to create a network making the VCM plant to need only 13 MW of steam. For comparison, the future network without any improvements will consist of only 2 heat exchangers.

When examining measures for increased internal heat exchange, focus has been on the five pinch violations selected for further investigation in Chapter 4.3. This does not necessarily mean that the above streams should be heat exchanged with each other. It should be noted that for eliminating the pinch violations in the chosen streams other streams probably have to be included. The streams that are cooled with cooling utility above the pinch should be used for heating other streams above the pinch, and the streams that are today heated with hot utility below the pinch should instead be heated with hot streams from below the pinch.

Above the pinch, there are three major cold streams at suitable temperatures which could be used for utilising the heat in the two chosen hot streams (H302 and H202A). That is the streams in reboiler H105 and reboiler H108, both located in the EDC-cleaning unit, and reboiler H205 which is located in the VCM unit. The reboiler H205 has a heat demand both below and above the pinch, and is one of the chosen pinch violations.

For heating up the chosen streams below the pinch (H205, H212 and H213), the heat below the pinch in the streams H302 and H202 could be used. Also streams from the HTC part could be used, although they are at a slightly lower temperature.

Table 4.3 Summary of the streams in the VCM plant chosen for retrofit studies

				Heat demand [kW]		*Heat surplus [kW]*		
		At:	*Description:*	*Below pinch:*	*Above pinch:*	*Below pinch:*	*Above pinch:*	*Pinch violations [kW]*
Cold streams	H105	EDC	Reboiler Azeo-column		3900			
	H108	EDC	Reboiler EDC-column		5800			
	H205	VCM	Reboiler HCl-column	660	900			660
	H212	VCM	Preheater before VCM-column	460				460
	H213	VCM	EDC-preheater	1800	160			1800
Hot streams	H302	OXI	Precooler after Oxi-reactors			740	1900	1900
	H202	VCM	Condenser after cooling column			5300	1700	1700

The locations of the streams are found in the map over the VCM plant in Figure 4.7.

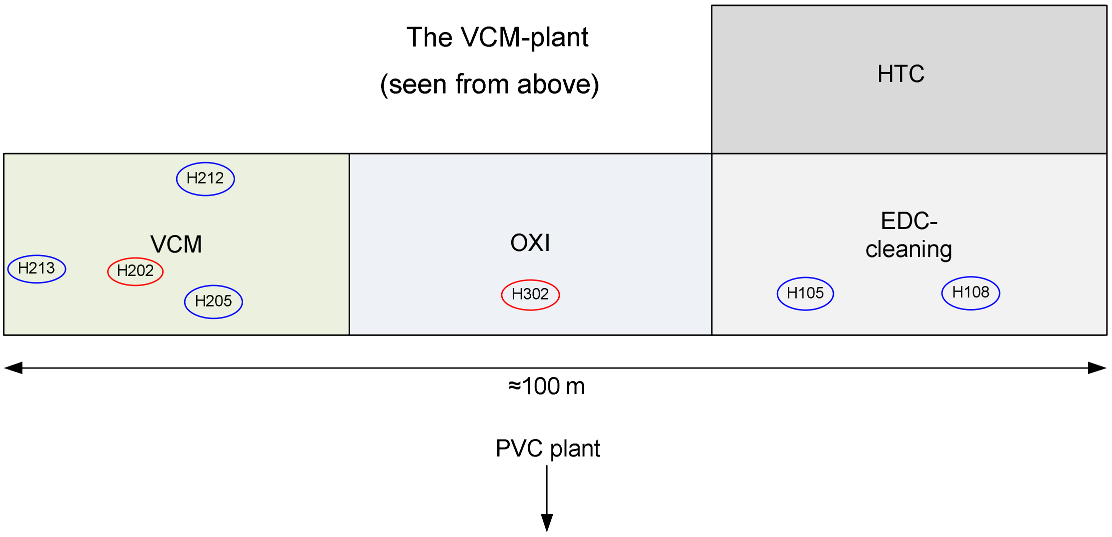

Figure 4.7 A map over the VCM-plant with locations of the discussed heat exchangers.

Different retrofits have been investigated, of which three will be presented. The different heat exchangers in the retrofit will be economically evaluated, as well as the whole retrofit. The following retrofits are made assuming direct heat transfer. Whether that is reasonable or not will be discussed later in Chapter 4.5.

In the following figures, the new heat exchangers are assumed to be placed on the upstream side of the existing heat exchangers, reducing their load. In the calculations, no losses to surroundings are assumed.

As seen in Table 4.2, the steam reduction will be 6.5 MW if all of the chosen pinch violations are eliminated.

4.4.1 Retrofit 1

Retrofit 1 consists of two subsystems, which can be implemented independently of each other. In retrofit 1, large consideration has been taken to the location of the streams.

The streams in system 1 are all located in the VCM unit.

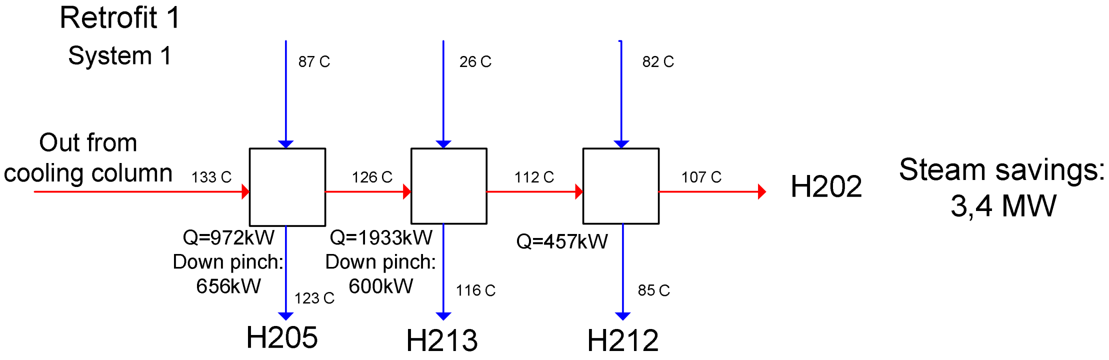

Figure 4.8 Retrofit 1, subsystem 1

The steam savings from system 1 amount to 3.4 MW. Since the location of the streams was of large importance in this retrofit, the stream out from the cooling column is heat exchanged with other streams in the VCM unit. The hot stream has a large amount of heat above the pinch, and since the cold streams are mainly located below the pinch, there will be transfer of heat trough the pinch.

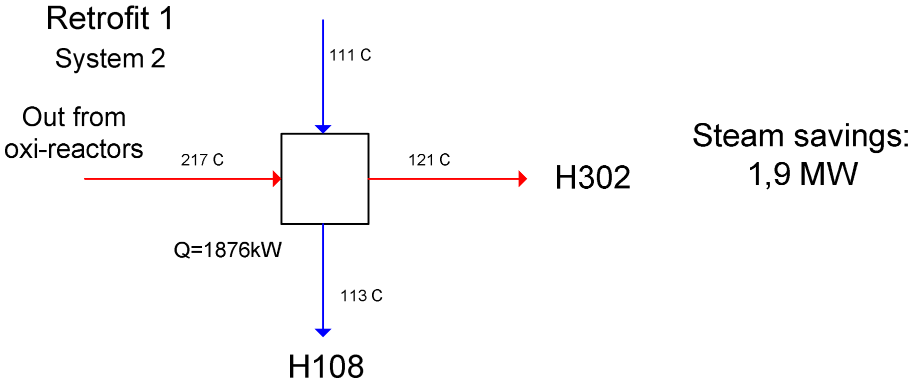

Figure 4.9 Retrofit 1, subsystem 2

The stream leaving the Oxi-reactors is here heat exchanged with reboiler H108. It could as well be heat exchanged with H105, which is located closer, but has a higher temperature and therefore less energy can be transferred. Both streams are above the pinch, and system 2 will not lead to any pinch violations.

Together, the subsystems form retrofit 1.

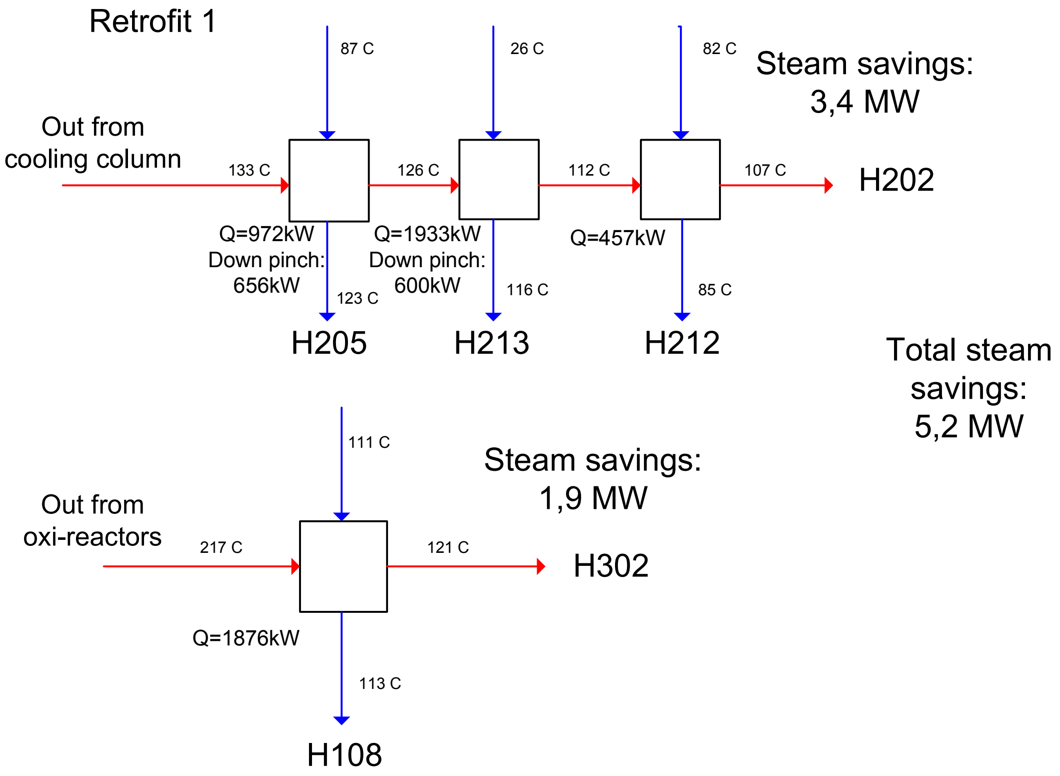

Figure 4.10 Retrofit 1

The reason for the steam savings being only 5.2 MW out of the possible 6.5 MW is the heat transferred through the pinch in subsystem 1. In Figure 4.11, an illustration of the retrofit is shown.

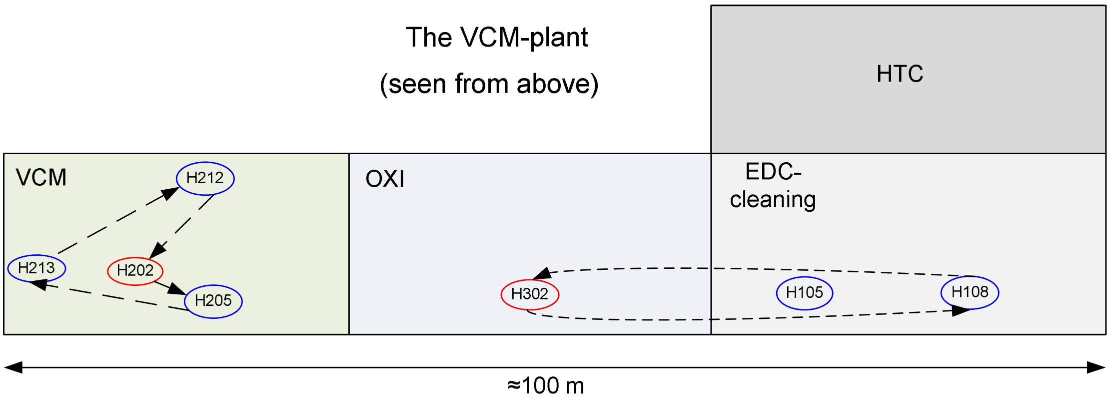

Figure 4.11 Illustration of retrofit 1, with the two subsystems

As seen in Figure 4.11, the streams in the VCM unit are located relatively close to each other. The distance between the stream in the Oxi unit and the EDC-cleaning unit is larger. In this retrofit, the locations of the streams have been of large importance, which resulted in pinch violations. In order to save more energy and to eliminate the pinch violations in retrofit 1, another retrofit is shown, named retrofit 2.

4.4.2 Retrofit 2

Retrofit 2 is an improvement of retrofit 1 regarding steam savings. To avoid the transfer of heat across the pinch that occurs in retrofit 1, the stream out from the cooling could first be heat exchanged with another cold stream.

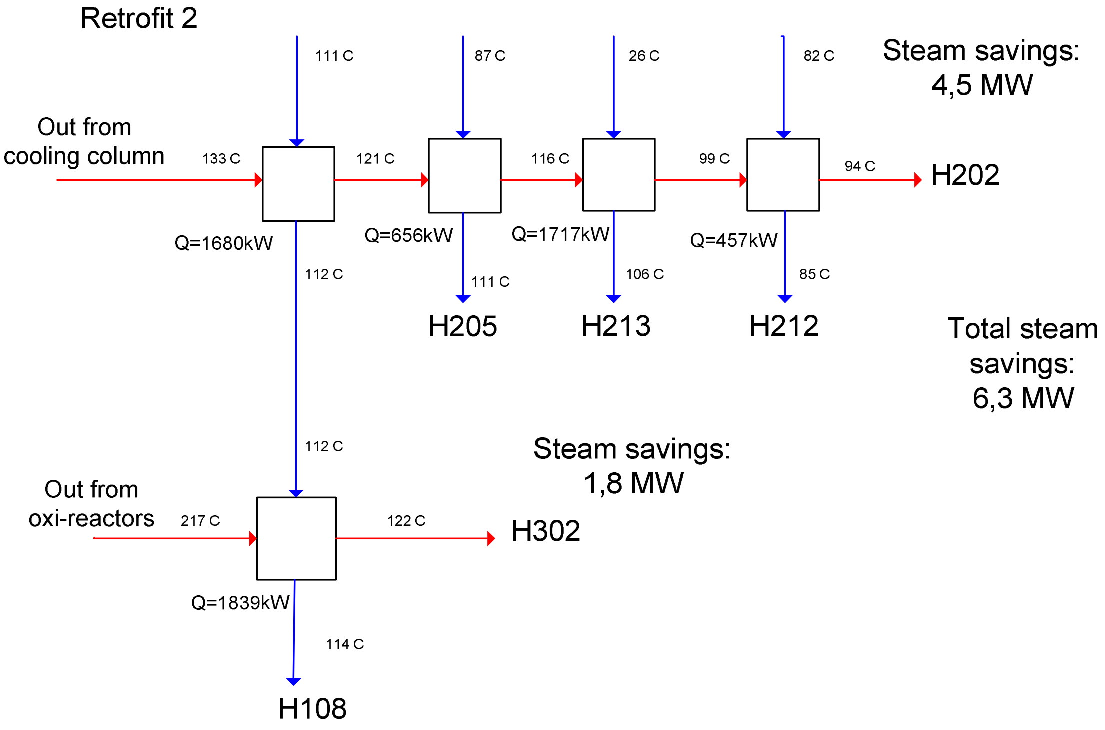

Figure 4.12 Retrofit 2

In retrofit 2, the heat from the cooling column in the VCM unit is first exchanged with the reboiler H108, before returned to the VCM unit and heat exchanged with the same streams as in retrofit 1. This gives no pinch violations, and hence the steam saving is larger. However, since the stream to H108 is first preheated with the stream from the cooling column before heated with the stream from the Oxi-reactors, the stream from the Oxi-reactors can only almost be cooled down to the pinch temperature, and the steam savings do not reach the maximum possible.

From an energy efficiency point of view, this is a better alternative than retrofit 1, saving additional 1.2 MW. But this is at the cost of a more complex network and a large amount of extra piping. An illustration of the network is shown in Figure 4.13.

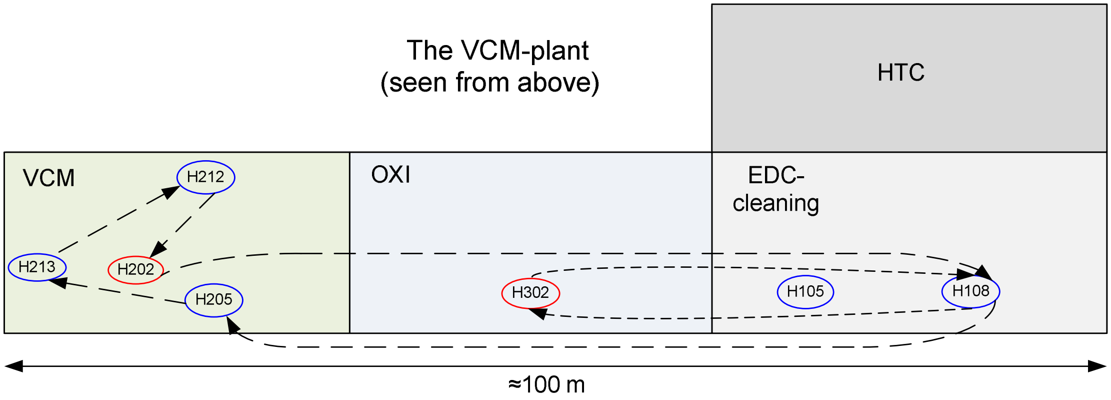

Figure 4.13 Illustration of retrofit 2

As seen, the hot stream has to be transported to the other end of the site and back. The reasonability of this should be compared with the additional steam savings.

4.4.3 Retrofit 3

None of the two earlier proposed retrofits reached the possible target of saving 6.5 MW when eliminating all pinch violations. Therefore, a third retrofit was investigated.

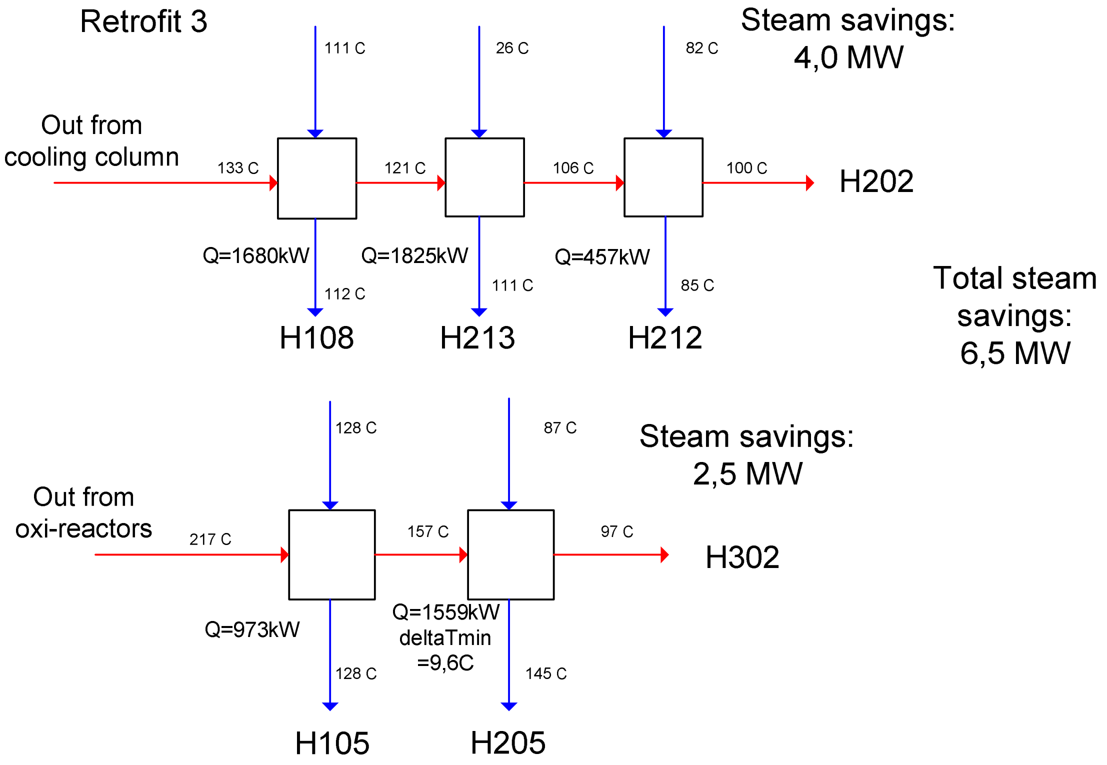

Figure 4.14 Retrofit 3

Retrofit 3 is based on retrofit 2, with two changes. The stream out from the Oxi-reactors is first heat exchanged with the stream in H105. Thereafter, it is also heating the stream in H205, which is no longer heated with the stream from the cooling column. Retrofit 3 is the only one of the three retrofits that reaches the maximum steam savings of 6.5 MW when eliminating all the pinch violations. This is reached with only five heat exchangers at the expense of a temperature difference slightly less than the specified minimum allowable temperature difference on the cold end of one of the heat exchangers. The temperature difference on the cold end of the heat exchanger H302-H205 is only 9.6°C, and the minimum allowable temperature difference was set to 10°C.

As seen in Figure 4.14, retrofit 3 also consists of two subsystems.

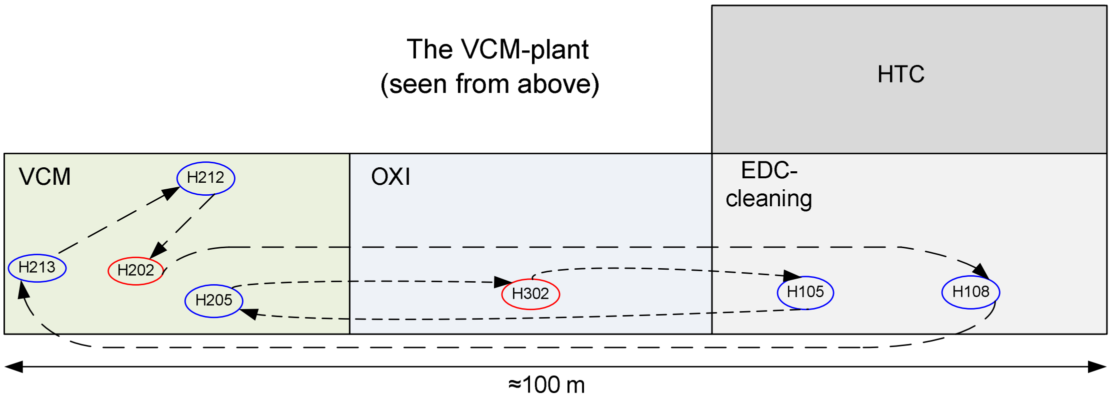

Figure 4.15 Illustration of retrofit 3

It can be seen that even more piping than in retrofit 2 is needed. Retrofit 3 is perhaps not a realistic option, but it is presented to show that it is possible to eliminate all the chosen pinch violations. As with retrofit 2, the increased complexity and the extra piping needed must be seen in proportion to the extra steam savings made.

4.4.4 Evaluation of the retrofits

Retrofit 1 is considered to be the easiest to implement and is therefore the most plausible of the three alternatives. However, retrofit 1includes pinch violations, and hence maximum energy recovery cannot be reached. To avoid pinch violations, it should be built as retrofit 2.

No heat exchangers located in the HTC unit are included in the retrofits. The excess heat from the HTC unit was not needed to cover the selected heat demands. It is also at a lower temperature level than the two chosen hot streams, reducing the possibilities for heat transfer. Another reason for not using the heat in the HTC units in the retrofits is to enable the possibility of instead utilizing it for district heating production or for heating in other parts of the site.

The study is made for future condition, but since none of the retrofits includes units affected by the expansion, the same retrofits can be proposed for the current situation.

The retrofits were all evaluated with respect to economic performance. The calculation procedure is described in Chapter 3.5. For detailed result, please see Appendix 13.

Table 4.4 Cost evaluation of the retrofits

	Q[kW]	*Cost of equipment, Nov 2010 [$]*	*Total fixed capital cost [$]*	*Total fixed capital cost [MSEK]*	*Annual savings (utility cost and reduced CO_2) [MSEK]*	*Payback period [years]*
Retrofit 1						
Subsystem 1						
H202-H205	972	58000[1]	280000	1.9	3.3	
H202-H213	1933	66000[1]	320000	2.2	6.6	
H202-H212	457	49000[1]	240000	1.6	1.6	
		Sum subsystem 1		5.8	11.5	0.50
Subsystem 2						
H302-H108	1876	60000[2]	290000	2.0	6.4	
		Sum subsystem 2		2.0	6.4	0.31
		Sum retrofit 1:		7.8	17.8	0.43
Retrofit 2						
H202-H108	1680	87000[1]	420000	2.9	5.7	
H202-H205	656	55000[1]	260000	1.8	2.2	
H202-H213	1717	66000[1]	320000	2.2	5.9	
H202-H212	457	57000[1]	270000	1.9	1.6	
H302-H108	1839	60000[2]	290000	2.0	6.3	
		Sum retrofit 2:		11	21.6	0.50
Retrofit 3						
H202-H108	1680	87000[1]	420000	2.9	5.7	
H202-H213	1825	66000[1]	320000	2.2	6.2	
H202-H212	457	52000[1]	250000	1.7	1.6	
H302-H105	973	50000[2]	240000	1.6	3.3	
H302-H205	1559	103000[2]	490000	3.4	5.3	
		Sum retrofit 3:		12	22.1	0.54

Material used in heat exchanger: [1]=Monel, [2]=Inconel

From Table 4.4, it can be seen that all of the alternatives are profitable, having a payback time of less than one year. The annual savings is about 20 MSEK. This makes it very interesting from an economic point of view to continue research about the investments.

In the retrofits, one of the most steam saving proposed new heat exchangers is with the stream in H213. In the current process, heater H213 is heated with 1-barg steam. In Chapter 5.1, there is a discussion about how to value the low-pressure steam.

The investment cost might be underestimated. The pipelines that have to be built are probably longer than the standard assumption in the factorial method of cost estimation, especially when heat exchanging between different parts of the plant. The cost of installation could also be higher, due to difficulties of inserting new heat exchangers into the existing units. It is also possible that the material correction factor is too low, since the fluids are corrosive and demand very corrosion-resistant materials. The market price for heat exchanger construction materials has rather high variations, and it is possible that the material price has increased since the correction factors in the model were developed. In reality, there will also be losses of heat, and not as much heat as calculated will be transferred, giving lower savings. Other equipment such as pumps must be installed, increasing the equipment costs and running costs.

On the plus side, there will also be less need for cooling water, which should be accounted for when considering an investment (see Chapter 5.1). The cost for the carbon dioxide emission certificates can be expected to increase in the future, also making the investments more beneficial, even though the profit from the certificates is small compared to the reduced fuel costs.

Table 4.5 CO_2-evaluation and comparison with MER-network

	Reduced heating/cooling demand [MW]	*Steam savings compared to maximum recovery network*	*Annually avoided CO_2 emissions [kton]*
Retrofit 1	5.2	67%	8.1
Retrofit 2	6.3	82%	9.8
Retrofit 3	6.5	83%	10.1

For comparison, the annual carbon dioxide emissions from the whole site in Stenungsund are currently about 60 kton/yr (INEOS ChlorVinyls, 2009).

In Table 4.5, the energy savings from the retrofits compared to the maximum possible savings from internal heat exchange are shown.

4.5 Hot water system and steam production instead of direct heat exchange in the future VCM plant

In the previous chapter, retrofits using direct heat exchanging are discussed. However, that is not always possible. The fluids are often corrosive and expensive materials in the pipelines would be needed. Fouling in the equipment is another problem. The matter of safety is important when considering transporting streams across the site and the case of leakage must be considered. There could be leakage in the heat exchanger as well, and if the fluids to be heat exchanged must not be mixed a heat transfer media should be used. Another advantage with a heat transfer media is that the system could be easier to control.

In the case of direct heat transfer between the streams in the plant being unrealistic, a circulating hot water system might be used. When using a hot water system, the minimum temperature difference between the hot and the cold stream must be doubled, since the minimum temperature should occur both between the hot stream and the water and the water and the cold stream. This reduces the potential for heat exchanging. An additional heat exchanger has to be used, which affects the investment cost.

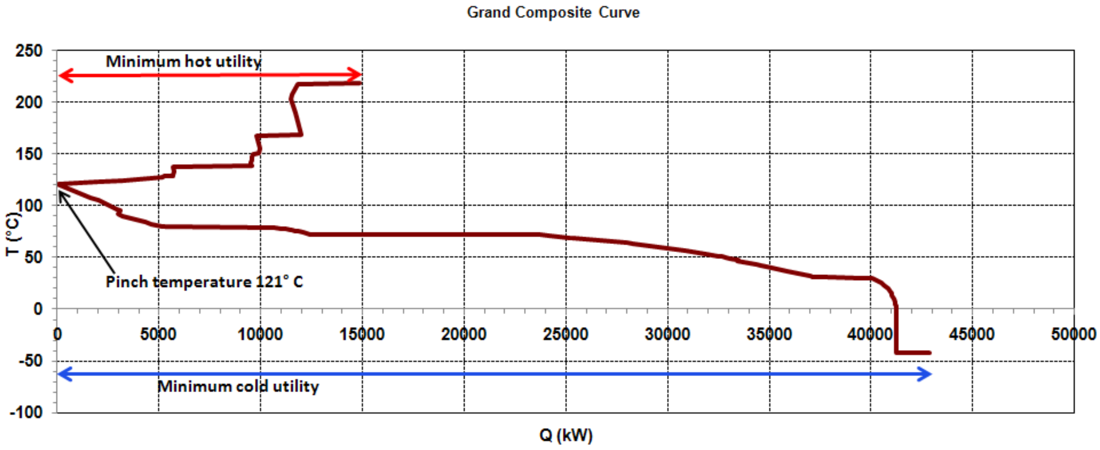

Figure 4.16 Grand composite curve future plant with ΔT_{min}=20°C, for investigation of a hot water system

If the grand composite curve in Figure 4.16 is compared with Figure 4.2, it can be seen that minimum hot and cold utility increases when the minimum temperature difference increases from 10°C to 20°C. The pinch temperature is changed from 116°C to121°C. The change for the utility need is found in Table 4.6.

Table 4.6 Hot and cold utility demand, ΔTmin=10 °C and ΔT_{min}=20 °C

	Hot utility [MW]	*Cold utility [MW]*
Demand future plant	20.8	48.9
Theoretical minimum demand future plant, ΔT_{min}=10 °C	13.1	41.1
Theoretical minimum demand future plant, ΔT_{min}=20 °C	14.9	42.9
Potential for savings in future plant, ΔT_{min}=20 °C	5.9	5.9

From Table 4.6 it is found, that if ΔT_{min} is increased from 10°C to 20°C, the theoretical minimum demand for hot utility increases from 13.1 MW to 14.9 MW. This means that if a circulating hot water system is going to be used instead if direct heat exchanging, the potential for energy saving is almost 2 MW less.

4.5.1 Proposed retrofits, using hot water systems

The first two retrofit proposals are evaluated for a minimum temperature difference of 20 °C. For a higher minimum temperature approach, not as much heat is transferred through the pinch in retrofit 1.

Table 4.7 Reduced heating and cooling demand for retrofit 1 and 2 when having a minimum temperature difference of 20 °C

	Reduced heating/cooling demand [MW]
Retrofit 1	4.5
Retrofit 2	4.7

The reduced utility demand for the different retrofit options is shown in Table 4.7. They should be compared to the total potential for savings of 5.9 MW with a minimum temperature difference of 20°C.

Because of less transfer through the pinch in retrofit 1, retrofit 1 saves almost as much steam as retrofit 2. Retrofit 2 is more complex, with heat from the VCM unit first to the EDC-cleaning to the other side of the plant and then again to the VCM unit. Therefore, it was discarded as a realistic option for a system using circulating hot water for heat transfer.

An investigation of retrofit 1 for a circulating hot water system was conducted. As before, retrofit 1 consists of two subsystems, and two different hot water systems are required. Different flow rates of water in the systems could be used. However, in this preliminary investigation, only one flow rate of water in each system was considered. The flow of water is set to have as large heat recovery as possible. To be able to transfer as much heat as possible to the cold streams, the water flow should be high. However, the capacity rate (FC_p) for the water cannot be higher than for the hot stream, if the temperature difference between the streams is to be kept above the minimum allowable temperature difference at all locations within the heat exchanger. Based on this constraint, the flow rate of water was set to equal the capacity rate of the water and the capacity rate of the hot stream.

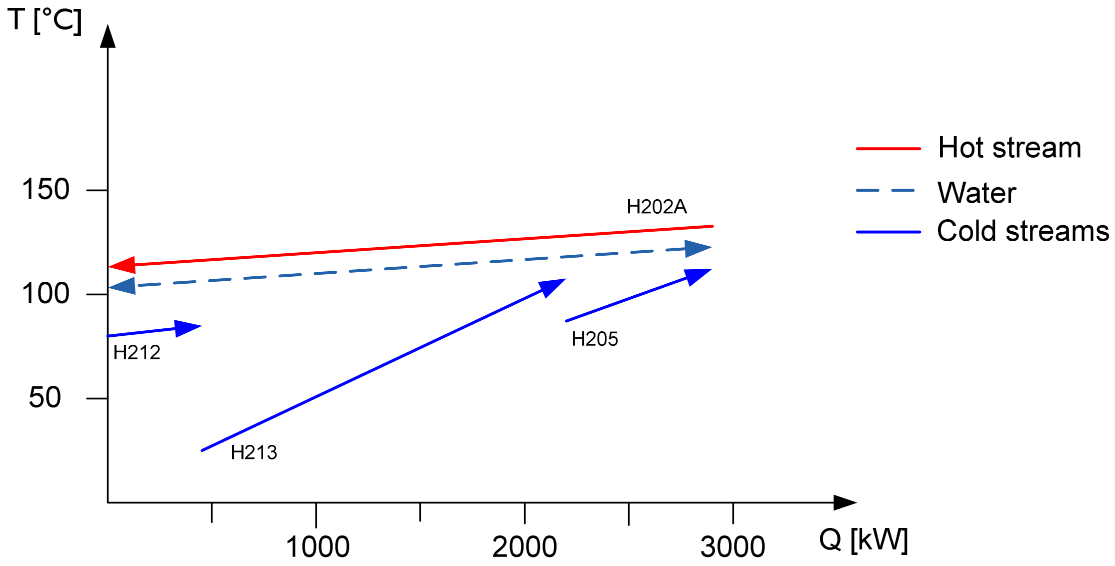

Figure 4.17 Subsystem 1 with high water flow rate, Retrofit 1 using circulating hot water system

In the above figure subsystem 1 is shown. The flow rate of water is set as high as possible, giving the same capacity rate for the hot stream and the water, and the lines are parallel in the diagram. If the flow rate of water was lower, less heat could be transferred to H213, since the minimum temperature difference occurs at the hot end of the heat exchanger. On the other hand, the investment cost will be expensive, since the minimum temperature difference occurs throughout the heat exchanger where the water is heated. It is a trade-off between high energy recovery and large investment costs. The high flow rate of water is chosen in the calculations to recover as much heat as possible, but other flow rates should be tried as well, since the used flow rate could be unreasonably high, resulting in high investment costs. Please see Appendix 14 for a similar discussion about the water flow rate in subsystem 2.

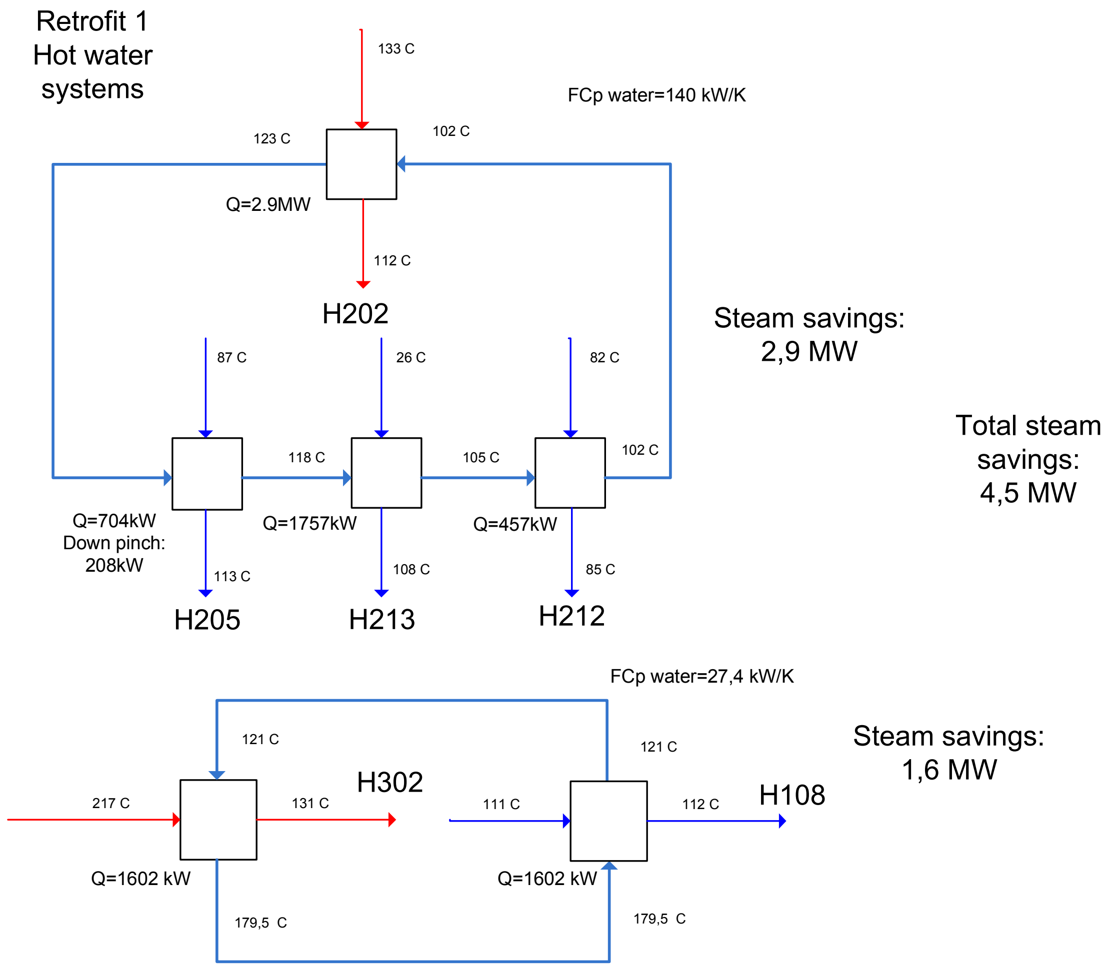

Figure 4.18 Retrofit 1 using circulating hot water systems

The steam savings for retrofit 1 with a circulating hot water system is 4.5 MW, compared to 5.2 MW if direct heat transfer is used.

In Table 4.8, an economic evaluation of the system shown in Figure 4.18 is made. As seen, the investment cost for the heat exchanger between the stream in H202 and the water is very expensive. This is because of the approach of having the same capacity rate to reach as high energy recovery as possible. This leads to a small temperature difference throughout the heat exchanger, and hence a large heat transfer area is needed. Different flow rates of water should be investigated, and the size of the heat exchangers and the investment cost on the one hand and energy recovery and savings on the other hand should be considered.

Table 4.8 Economic evaluation of circulating hot water systems

	Q[MW]	*Cost of equipment, Carbon Steel, Nov 2010 [$]*	*Total fixed capital cost [MSEK]*	*Annual savings [MSEK]:*	*Payback period [years]*
Retrofit 1					
Hot water system 1					
H202-Hot water	2.9	200000[1]	6.5		
Hot water-H205	0.70	57000[1]	1.9	2.4	
Hot water-H213	1.8	68000[1]	2.3	6.0	
Hot water-H212	0.46	52000[1]	1.7	1.6	
		Sum system 1	12.4	10.0	1.25
Hot water system 2					
H302 - Hot water	1.6	121000[2]	4.0		
Hot water -H108	1.6	65000[1]	2.1	5.5	
		Sum system 2	6.1	5.5	1.1
		Sum retrofit 1:	19	15	1.2

Material used in heat exchanger: [1]=Monel, [2]=Inconel

4.5.2 Possibilities for low-pressure steam production

Instead of using a hot water system, there are possibilities to produce low-pressure steam at 1 barg.

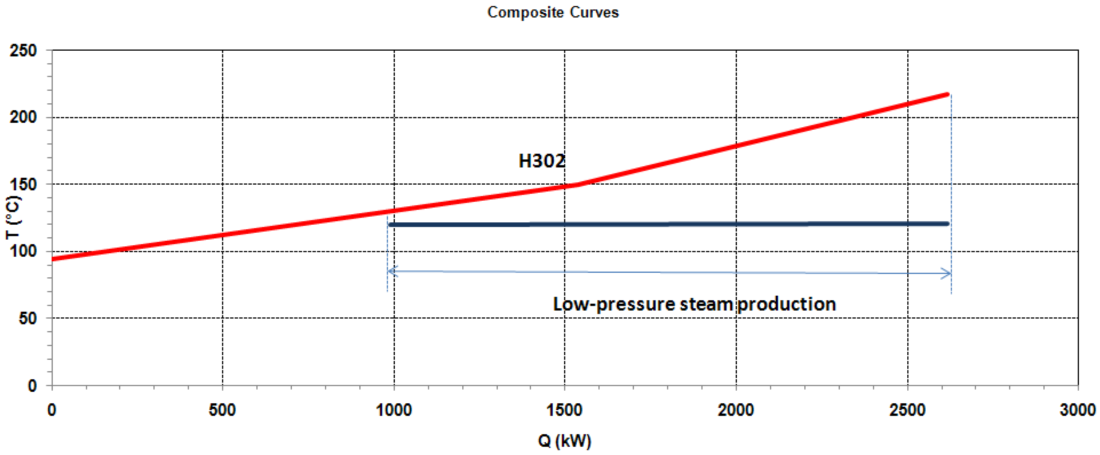

Figure 4.19 Low-pressure steam production from H302

There is a possibility of producing 1.6 MW of low-pressure steam by heat recovery from stream H302. This means that all the heat utilized in the hot water system also could be utilized for steam production.

Investigation of the steam production possibilities from stream H202 indicates that only 0.4 MW of steam could be produced compared to the 2.9 MW that is possible to utilize in a circulating hot water system. The results show that from H302 there could be steam production, but from H202 a hot water system has a much higher potential.

4.6 Use of excess heat from the VCM plant in the PVC plant

In the previous chapters, it was found that there is a large surplus of heat in the VCM plant below the pinch temperature. Options for using this heat in the PVC plant were also investigated.

In the PVC plant, heat can be used both for preheating inlet air for combustion in the drying units, and for heating of low temperature streams. A very rough approximation of the possible heat sinks was conducted, based on an energy mapping of the PVC process (INEOS ChlorVinyls, 2008). In the approximation, it was assumed that the heating fluid has a temperature of 100°C. Assuming a ΔT_{min} of 10°C implies that it is possible to heat the cold stream up to 90°C. For calculations, please see Appendix 16.

In the following preliminary assessment, the same price is assumed for saved steam and saved fuel gas.

Table 4.9 Approximated heat demand in the PVC plant that could be covered with 100°C water, ΔT_{min}=10°C

	Load that can be covered: [MW]	*Annual steam/fuel savings: [GJ/yr]*	*Annual savings, assuming steam price from Chapter 3.5.2 [MSEK/yr]*
Preheating of inlet air for drying units:	5 MW	150000	17
Heating of streams:	2 MW	60000	7
Total:	7 MW	230000	24

From Table 4.9, it can be seen that there is a demand for roughly 7 MW of a hot fluid of 100°C in the PVC plant.

The potential for producing hot utility at 100°C from the VCM plant can be estimated based on the GCC in Figure 4.20, assuming a return temperature of the heating fluid to the VCM plant of 50°C.

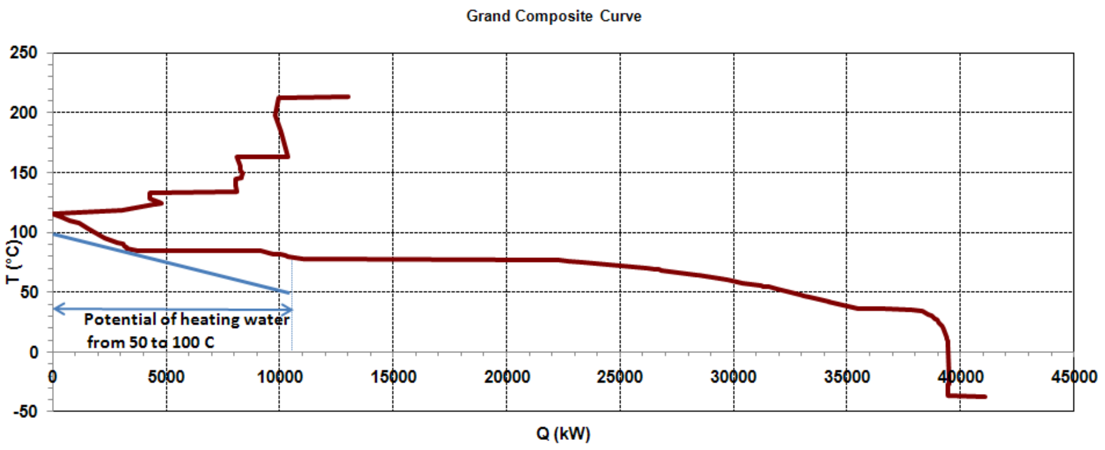

Figure 4.20 Grand composite curve future plant, with potential for hot water production

The grand composite curve for the future plant shows that there is a heat surplus in the VCM plant of more than 10 MW between 50°C and 100°C. Much of this is located in the HTC unit. The condenser at the top of the cooling column, and the three coolers H164, H165 and the new cooler have a combined total heat surplus of almost 19 MW. In Figure 4.21, the hot composite curves for these four coolers are shown, together with the possibilities for producing hot water.

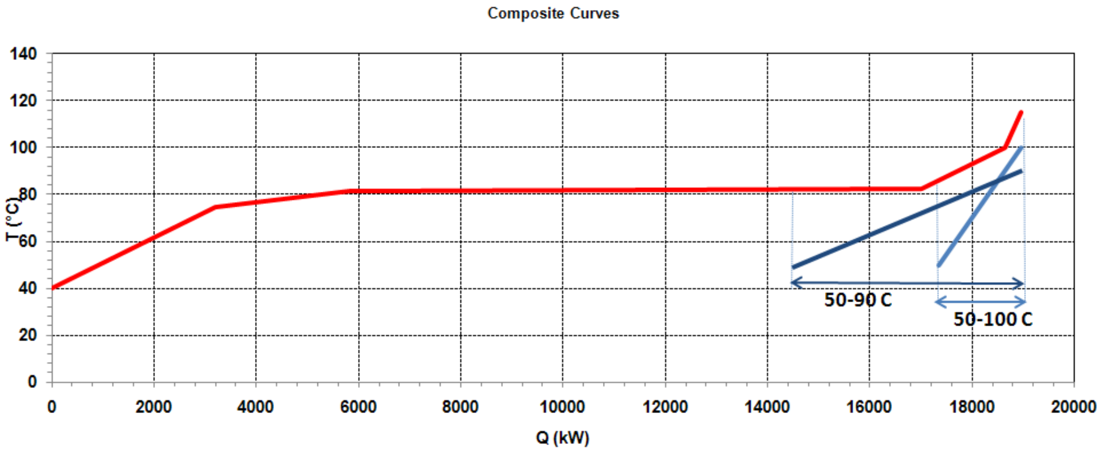

Figure 4.21 Composite curves for H153, H164, H165 and Hnew and the possibilities for hot water production

It can be seen that the potential for heating water up to 90°C is much larger than for heating up to 100°C (4.5 MW compared to 1.6 MW). If the inlet temperature is lowered, the possibilities for transferring more heat are of course increased. As seen in the composite curves, the heat from the condensation in H153 is barely utilised. It is possible to heat almost the same amount without using stream H153, which would be good, since it is a top condenser that is hard to access.

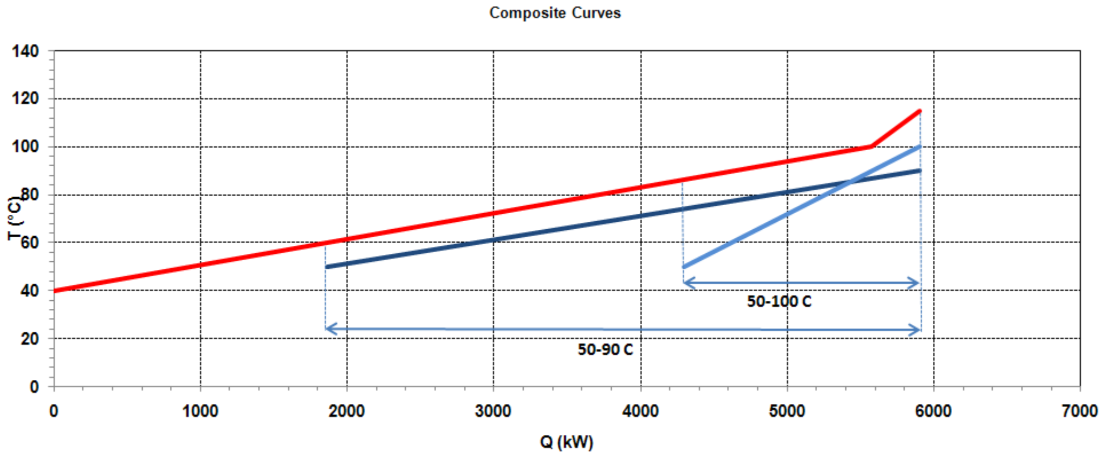

Figure 4.22 Composite curves for H164, H165 and Hnew and the possibilities for hot water production

A table with the possibilities of transferring heat is presented below. For composite curves for the different cases, please see Appendix 17.

Table 4.10 Potential for hot water production from chosen heat exchangers in the HTC unit

Production units:	*Hot water temp in [°C]*	*Hot water temp out [°C]*	*Potential for hot water prod: [MW]*
H153,H164,H165,Hnew	50	100	1.6
	50	90	4.5
	40	90	5.6
H164, H165, Hnew	50	100	1.6
	50	90	4.0
	40	90	5.0

In Table 4.10, it is shown that the potential for heating water up to 100°C is independent of whether H153 is included or not. Also the potential for heating water up to 90°C is about the same. However, as seen when comparing Figure 4.21 and Figure 4.22, the ΔT for heat transfer is lower if H153 is not included, resulting in a larger heat transfer area, although one less heat exchanger is needed.

As mentioned earlier, there is a heat surplus of 10 MW between 50 and 100°C in the VCM plant according to the GCC in Figure 4.21. This means that it is possible to reach maximum energy recovery in the VCM plant even if the heat in the HTC unit is used for producing hot water for the PVC plant. The units in the HTC unit are not included in any of the retrofits proposed in Chapter 4.4. In Appendix 12, a maximum energy recovery network that does not use any heat from the HTC unit is shown. However, it is possible that reaching maximum energy recovery in the VCM plant without using the heat from the HTC unit will be more complicated.

In the assumptions for the future HTC unit, it is assumed that the additional cooler installed will have an inlet temperature of 100°C. If the cooler was placed directly at the reactor outlet, the inlet temperature would instead be 115°C, increasing the possibilities for heating water to higher temperatures.

4.7 Heat from flue gases and Oxi-reactors

In this chapter, an analysis of the VCM plant including heat from flue gases and Oxi-reactors is made. Stream data is found in Appendix 10.

When the flue gases leave the cracking oven, they first pass through a preheater, which heats a stream of EDC. They are further cooled down in a flue gas steam boiler, before being released to the atmosphere. There is excess heat in the flue gases not utilized today. There is a temperature drop of 40°C between the preheater and the flue gas boiler, and thereby losses of about 300 kW. The flue gases leaving the flue gas boiler are still at a high temperature, above 200°C. If they were cooled down to 120°C, either by extending the existing flue gas boiler or by utilizing the heat in some other way, another 800 kW could be recovered. This means that a more efficient usage of the heat in the flue gases could produce more than 1 MW of additional steam. If the excess heat in the flue gases leaving the flue gas boiler is included in the analysis of the VCM plant, the need for hot utility will decrease.

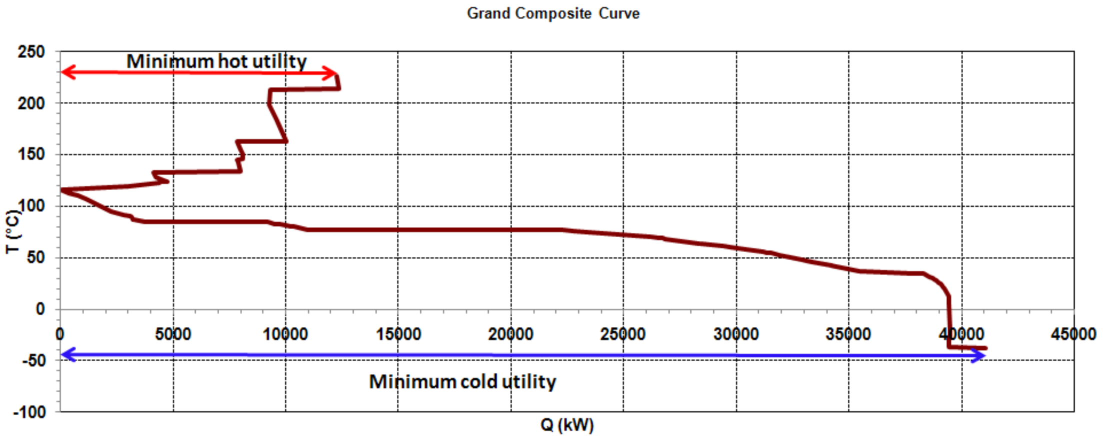

Figure 4.23 Grand composite curve for the future VCM plant including excess heat in the flue gases leaving the flue gas boiler, $\Delta T_{min}=10°C$

From the grand composite curve in Figure 4.23, it can be seen that the minimum need for hot utility is 12.3 MW when the excess heat in the flue gases is included, compared to the minimum need of 13 MW obtained in Chapter 4.2.

1.5 MW of medium pressure steam is produced in the flue gas boiler.

From the Oxi-reactors, steam is produced, utilizing the heat from the exothermic reactions. The steam produced is used in the VCM plant. 7.9 MW of high-pressure steam and 0.9 MW of low-pressure steam are produced. If the steam produced from the exothermic reaction in the Oxi-reactors and the steam produced in the flue gas boiler, as well as the excess heat and losses in the flue gases are included in the analysis, the grand composite curve in Figure 4.24 shows that almost no external heating is needed. Also the preheater is included.

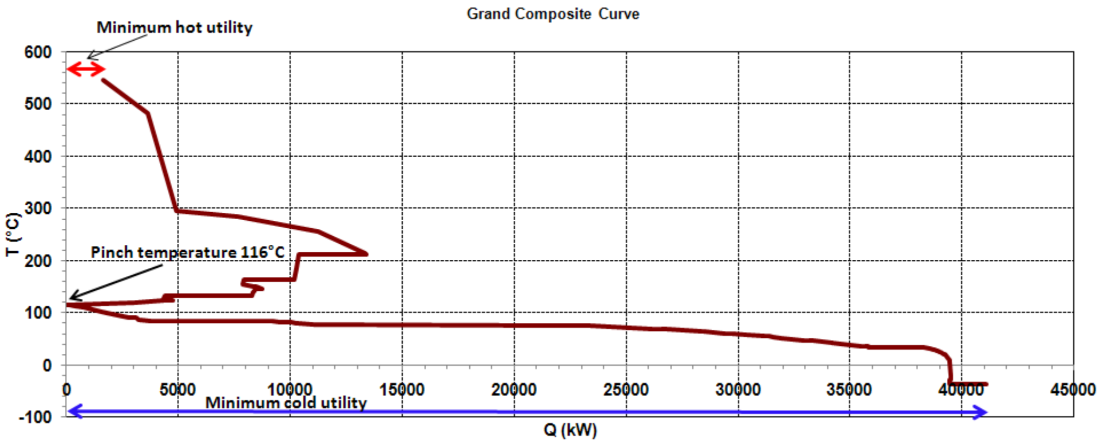

Figure 4.24 Grand composite curve for the future VCM plant including all heat in the flue gases and exothermic reactions in the Oxi-reactors, $\Delta T_{min}=10°C$

From the grand composite curve it is seen that if all heat from the flue gases and the Oxi-reactors were utilized, and maximum heat recovery was implemented, the whole VCM plant should only have a need for 1.6 MW of external hot utility (excluded the heat needed in the cracking oven). The GCC also shows that the additional heat demand could be covered mostly by low-pressure steam.

A new maximum energy recovery (MER) network was designed, including the heat in the flue gases and the heat from the Oxi-reactors, and is shown in Appendix12. In the MER network, utilizing heat from the flue gases and the Oxi-reactors, 40 heat exchangers, 23 coolers and 6 heaters are required. This is quite unrealistic, but shows that it is possible to create a network making the VCM plant to need only 1.6 MW of external heat. As a reminder, the future system without improvements will consist of 2 heat exchangers, 18 heaters and 25 coolers.

The cracking oven is fired with methane fuel. This means that the flue gases are probably not too corrosive, and flue gas condensation could be considered. According to the pinch analysis, there is an excess of heat below the pinch temperature, and therefore, theoretically, there is no point in cooling the flue gases further. But in reality, it could be easier to use heat from the flue gases than from some process streams. There is also the possibility to use the heat in the flue gases in other plants in the site.

4.8 Uncertainties about the results

There are many uncertainties about the results, due to lack of measurement points and lack of detailed validation of the simulation model. In this section of the report, the uncertainties that are mostly likely to influence the results are discussed. As mentioned earlier, a summary of assumptions for all the different streams is found in Appendix 4.

<u>H205</u>

The temperature out from H205, the reboiler at the HCl-column, is unknown. Also the composition is unknown. The composition is set to consist of only two substances, and the relation between them is set to obtain saturated liquid in to the reboiler. The most uncertain parameter is the temperature out. In the simulations, it is assumed that the vapour fraction of the stream leaving the reboiler is equal to 1.0, resulting in a quite high temperature difference between the inlet and outlet temperature for this mixture. This is probably not true. The vapour fraction could be lower, and the vapour could entrain non-evaporated liquid. The only thing that is known is that the outlet temperature is somewhere between the inlet temperature in and the temperature used. A temperature out of the reboiler of 120 C was tried. When performing a pinch analysis, it does not affect the pinch temperature. However, it affects the potential for internal heat transfer, which is increased by 0.5 MW to 8.7 MW. It also affects the amounts of heat transferred in the proposed retrofits. If the temperature out is lower, the capacity rate for the stream is higher since the heat load is known, and more heat can be transferred to the stream in H205 from the stream in H202. This will lower the possibilities for heating H213, but the total potential for heat exchanging will increase. Stream sheets and composite curves are found in Appendix 18.

<u>H302</u>

The flow rate and composition in H302, the cooler after the Oxi-reactors, are not known. They are calculated from the reaction of the ingoing substances (see Appendix 3). The inlet and outlet temperatures are known, but the condensation level and temperatures are dependent on the property method used in Aspen. The uncertainty about the load of the heat exchanger and of the distribution of the load is large, since it is all based on calculations and Aspen simulations.

<u>H202</u>

In the simulations, the inlet temperature and pressure to H202 are varied to obtain a vapour fraction value of 1.0. This means that the measured inlet temperature is lower than the one used in the analysis. If the temperature is lower, the amount of heat that is possible to transfer will be slightly lower. The assumption about all vapour in the inlet might also be incorrect, overestimating the heat content of the stream. The composition in H202 is unknown, and in the simulations it is taken from an old Aspen simulation. If the actual composition is different, this will affect the condensation energy and the temperature levels for which condensation takes place.

5 Discussion

In the previous chapter, the main results were presented and discussed. In this chapter, some additional matters are discussed.

5.1 Cooling water and 1-barg steam

Cooling water is usually seen as almost free. But in this case, the cooling demand will increase when the site is retrofitted. As shown, there will be an increase in the VCM plant, but also in the new chlorine plant, which will have a higher cooling demand than the existing one. As a result, the cooling water system needs to be extended, and investment is needed in new coolers to be able to produce the desired amount of cold water. In this case, cooling water is not for free. If the cooling demand can be reduced due to increased internal heat exchange, the expansion needed of the cold water production can be reduced, and the investment cost decreases.

Another utility today estimated as free in the plant is the low-pressure steam. There is an abundance of it, and the low-pressure steam not used for heating will simply be emitted to the surroundings. Therefore, it can be argued that there is no need to replace heat exchangers heated with low pressure steam with heat exchanging. However, there are always ways in which the low-pressure steam could be used. There are possibilities for expanding the low-pressure steam network to include the PVC plant as well, where there is a heat demand. The steam could also be used for increasing the temperature of the proposed hot water production in the HTC unit for use in the PVC plant. Other ways of utilizing the heat include preheating in the boiler house, or heating of buildings at the site. Finally, some of the heat exchangers currently heated with medium pressure steam could instead be heated with low pressure steam.

5.2 Carbon dioxide emissions price

In the economic evaluation, a carbon dioxide price of 15€/ ton was used. This price is likely to increase in the future. However, since the savings because of reduced carbon dioxide emissions only stand for about 5 % of the total savings, this increase will only have a marginal effect on the total savings.

5.3 Other possibilities

As seen in the grand composite curve in Figure 4.20, there is a potential of more than 10 MW for heating water from 50°C to 100°C. This could be used for producing district heating. Today, there is almost no need for additional district heating in Stenungsund. But if the system were connected to the district heating net in Göteborg, there would be a demand (Hackl, R et al., 2011). If the excess heat in the VCM plant could cover both the heat demand in other parts at the site and produce district heating, and which one to be preferred, needs to be examined.

As mentioned several times, there is a heat surplus below the pinch temperature of 116°C. This surplus could be used to cover the heat demands in the PVC plant, but parts of it could also be used for heating the buildings at the site. There are several buildings, and heating them with waste heat from the plant could be investigated.

In the proposed retrofits, the heat in the condenser after the cooling column is utilized. If instead it was possible to install a heat exchanger or a steam boiler before the cooling column, after the cracker, heat could be recovered at a higher temperature.

Finally, there is a possibility in several units to use low-pressure steam instead of the medium-pressure steam used today. This would lower the needs of producing steam, since there is an abundance of low-pressure steam today.

5.4 Advantages/disadvantages with hot water production and steam production instead of direct heat transfer

A hot water system has both advantages and disadvantages compared to direct heat transfer. There is much less risk that the fluids get accidently mixed, and the often corrosive fluid does not have to be transported across the plant in pipes, reducing both the cost for the pipes and the risks for accidents. However, using a hot water system instead of direct heat transfer reduces the amount of heat possible to transfer. This follows as a result of the larger temperature difference needed when hot water first is to be produced from a hot stream, and later to be heating a cold stream. There could be other disadvantages. A condensing stream, for example, stays at the same temperature during the cooling process. This is not the case for a water stream, which could lead to that a very high flow rate of water is needed to transfer all the heat.

There are different possibilities for designing a hot water system. There could be one or more systems, and the heat exchangers could be connected in series or parallel. As mentioned there could also be different flow-rates of water. Those and other possibilities need to be investigated when considering a hot water system.

Another way of utilizing the heat is for production of low-pressure steam. The industry is used to handle steam, and it could probably be used without modifications of the equipment. It also has a high heat content, reducing the amounts needed. But when producing steam, the heat cannot be utilized down to the same temperature levels as if using heat transfer with cold streams. This limits very much how much heat that it is possible to utilize.

5.5 Other comments

In this report, options for using excess heat in the new chlorine plant are not discussed, since very little is known about the planned process. However, there will be a need for steam for evaporating water to concentrate the diluted NaOH produced. It is possible that excess energy from the VCM plant could be used to decrease the steam consumption.

Another comment has to be made regarding the accuracy of the data. The calculated demand for the VCM plant is less than the demand given in the energy mapping earlier made at the company (INEOS ChlorVinyls, 2008). The reasons for this could be that no start-up heaters were included in the analysis, or that the present production level is lower, but it could also be because of poor agreement of the data used.

The same steam price is assumed for all steam levels, independent on pressure, and is based on the energy content in the steam. No limitations of the steam boilers have been considered, or how the efficiency of the boiler will be affected when the production decreases.

An interesting possibility in a future scenario could be that if the development of fuel cells is continued, and the market expands, it is possible that there will be an increased demand for hydrogen. In that case, it is possible that it will be more profitable to sell the hydrogen produced instead of using it on-site as boiler fuel.

6 Conclusions

This thesis investigated the current and future energy demand of the VCM plant, as well as options for increased energy efficiency by internal heat exchange. The study was conducted using pinch analysis methods and tools.

The results show that today's plant has a heat demand of 21 MW and a cooling demand of 39 MW. When the production is extended in the future, the heat demand will not be affected, but the cooling demand is expected to increase to 49 MW. During the analysis, it was found that there is a large potential for energy savings. Theoretically, the steam use in the future VCM plant could be reduced by 37% by increased internal heat exchange, corresponding to 7.8 MW. Different retrofits have been proposed; of them retrofit 1 is seen as the most reasonable. When considering retrofit 1, it should be remembered that this option does not remove all pinch violations, which means that maximum energy recovery is not reached. To avoid pinch violations, retrofit 2 should be used. Implementing retrofit 1 would lead to a decrease in steam use of 5.2 MW, eliminating 67% of the future pinch violations.

The retrofits were developed assuming direct heat transfer. That may not be possible, due to the composition of the streams used. In that case, implementation of a circulating hot water system may be a possible solution. When using a hot water system, the minimum allowable temperature difference (ΔT_{min}) between the hot stream and the cold stream must be doubled, in order to respect the minimum allowable temperature difference both when heating and cooling the hot water. This implies that the theoretical demand for steam savings is reduced to 5.9 MW. Implementing retrofit 1 using a hot water system with properties as proposed would in this case lead to a steam reduction of 4.5 MW, corresponding to 76% of the possible steam savings when using hot water systems.

A cost evaluation of the proposed retrofit solutions was also conducted. The investment cost estimations are very rough and should thus be considered with caution. They do however indicate that there is money to be saved with a short payback time. Saving 1 MW of steam leads to annual savings of approximately 3 MSEK, and the total savings could reach 20 MSEK/yr.

It is interesting to conclude that none of the streams proposed for heat exchange is affected by the expansion of the plant. The proposed retrofits are therefore possible to implement whether the production is increased or not. It was also noted that there is a lot of excess heat in the flue gases leaving the cracking oven. If the heat in the flue gases were to be used more efficiently with heat recovery down to 120 °C, another 1 MW of steam could be produced.

The future VCM plant will have a pinch temperature of 116°C (interval temperature scale, ΔT_{min}=10°C), corresponding to a pinch temperature of 111°C for cold streams and 121°C for hot streams. This excess heat could be used for different purposes, for example for low temperature needs in the PVC plant. It is interesting to note that maximum energy recovery can be obtained in the VCM plant without the excess heat from the HTC unit. This heat could instead be used for producing hot water, either for use in the PVC plant or for district heating. Approximately 5 MW could be produced in the HTC unit for use in the PVC plant, saving as much steam as the proposed retrofits for the VCM plant.

In conclusion, there is a large potential for increased energy efficiency measures which could reduce both costs and environmental impact.

7 Future work

The calculated heat loads on some of the heat exchangers needs to be confirmed with measurements. This especially is true for H302, and also for H202. The temperature at the outlet of H205 needs to be measured and compared with the assumed value.

The possibilities for direct heat transfer should be evaluated. If it is considered as not possible, a hot water system should be designed and studied more in detail.

The need for low-temperature heat in the PVC plant should be further investigated, and utilizing excess heat from the VCM plant should be evaluated. But before heat is transferred from the VCM plant to the PVC plant, a similar study should be performed for the PVC plant to see if there is excess heat also in the PVC plant, which should be primarily used. It could also be interesting to perform a study of the whole site to find possibilities for utilizing the excess heat in the VCM plant.

Also, a more detailed economic evaluation should be performed.

8 References

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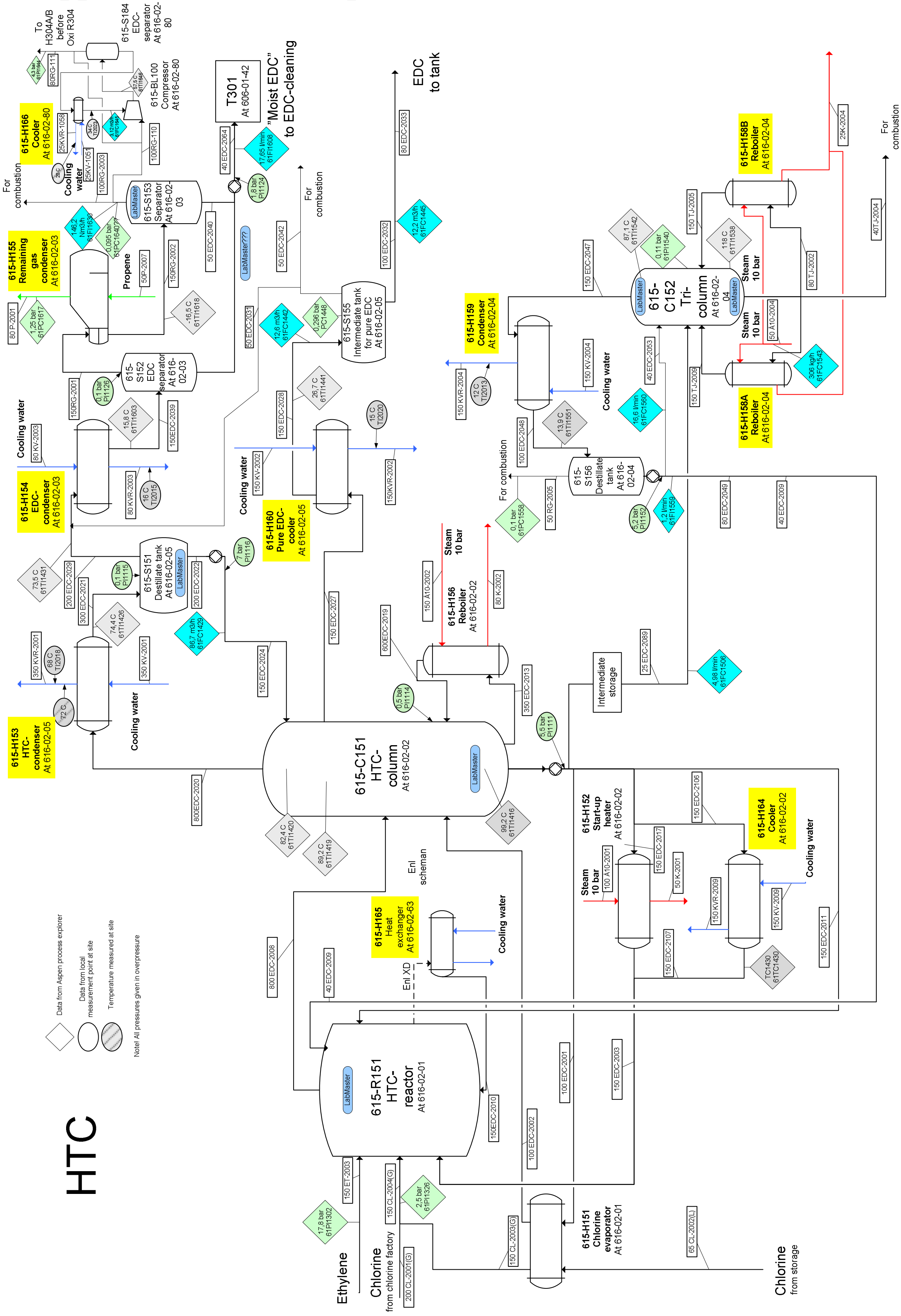
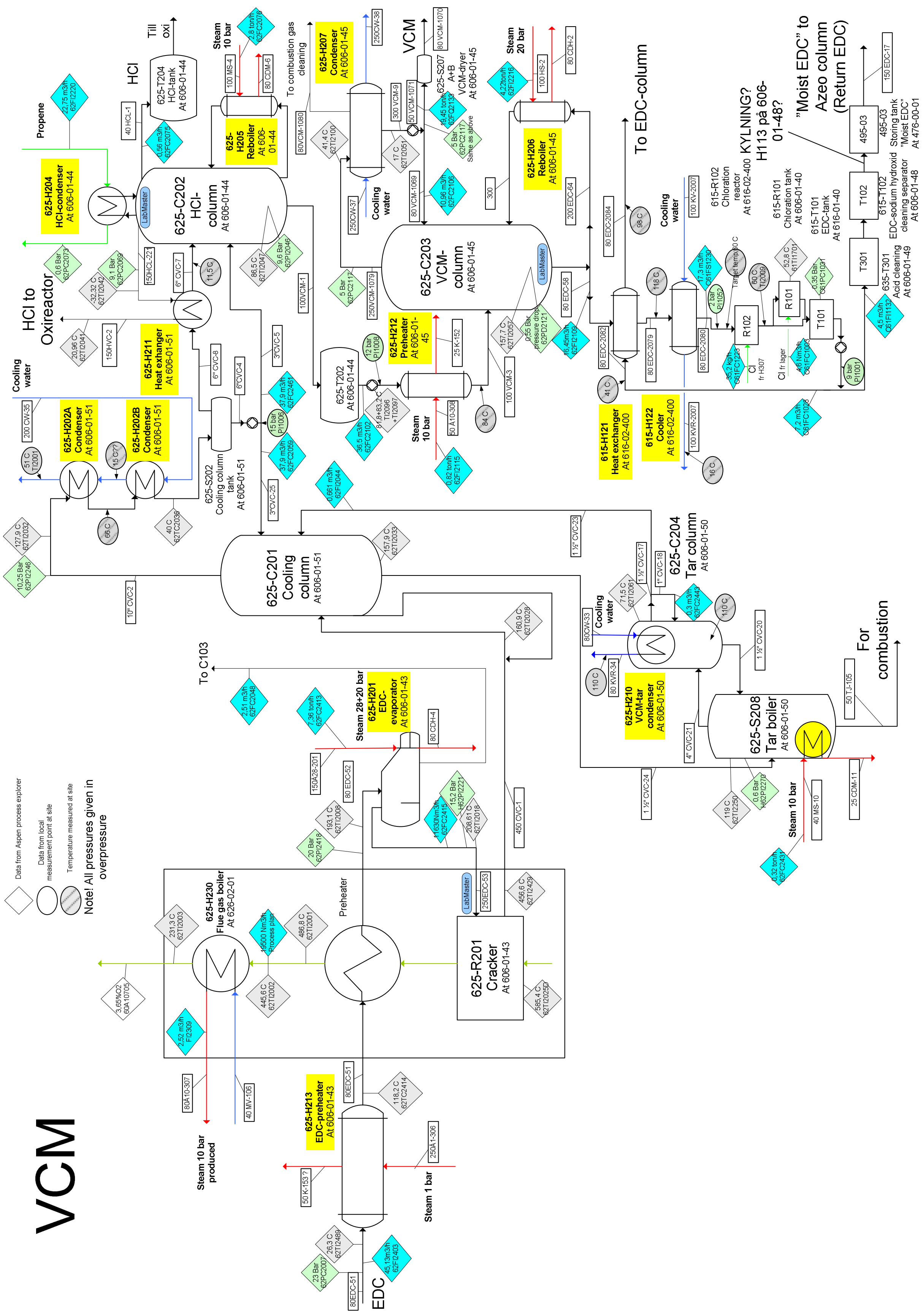
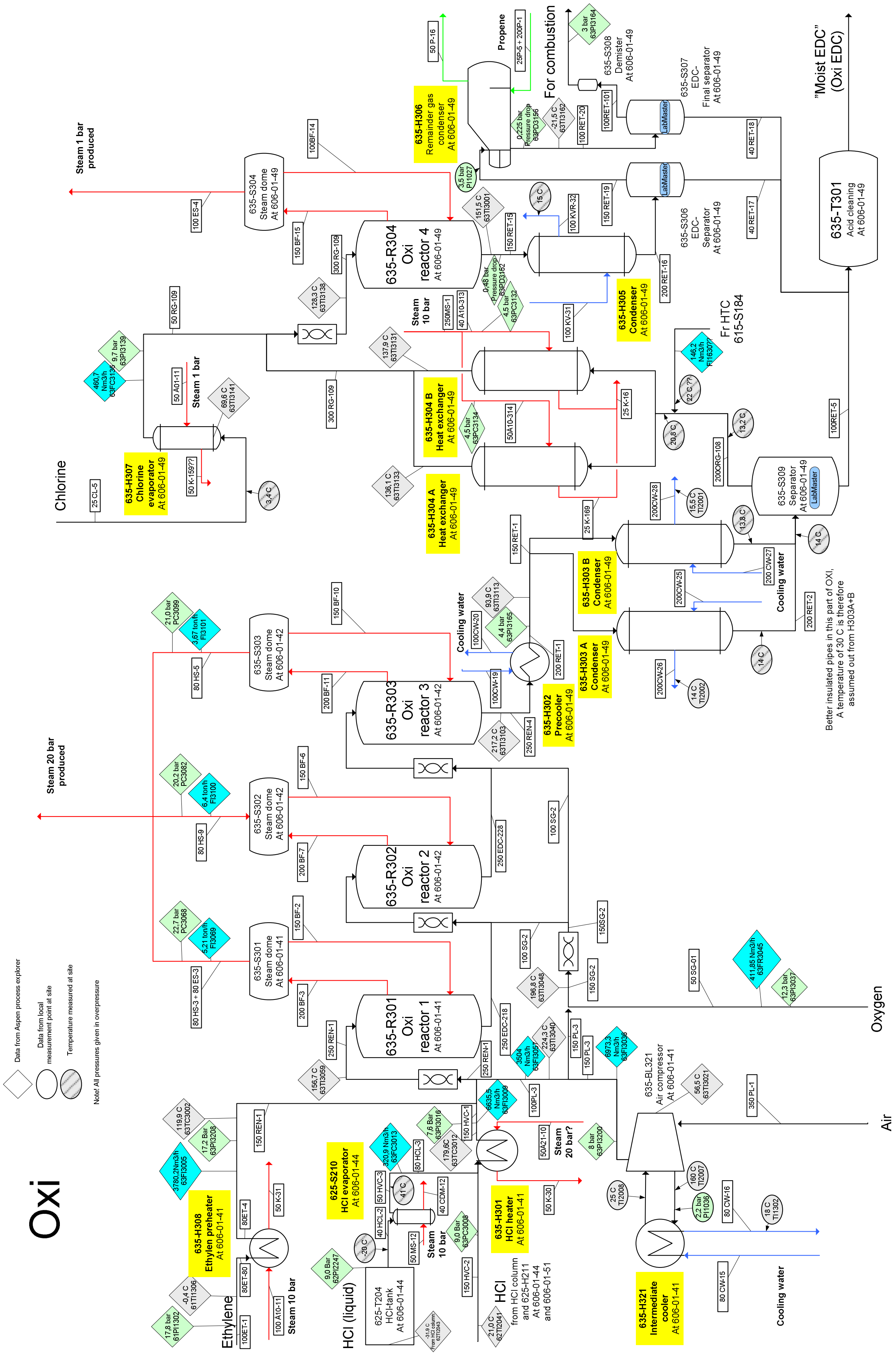
9 Appendix

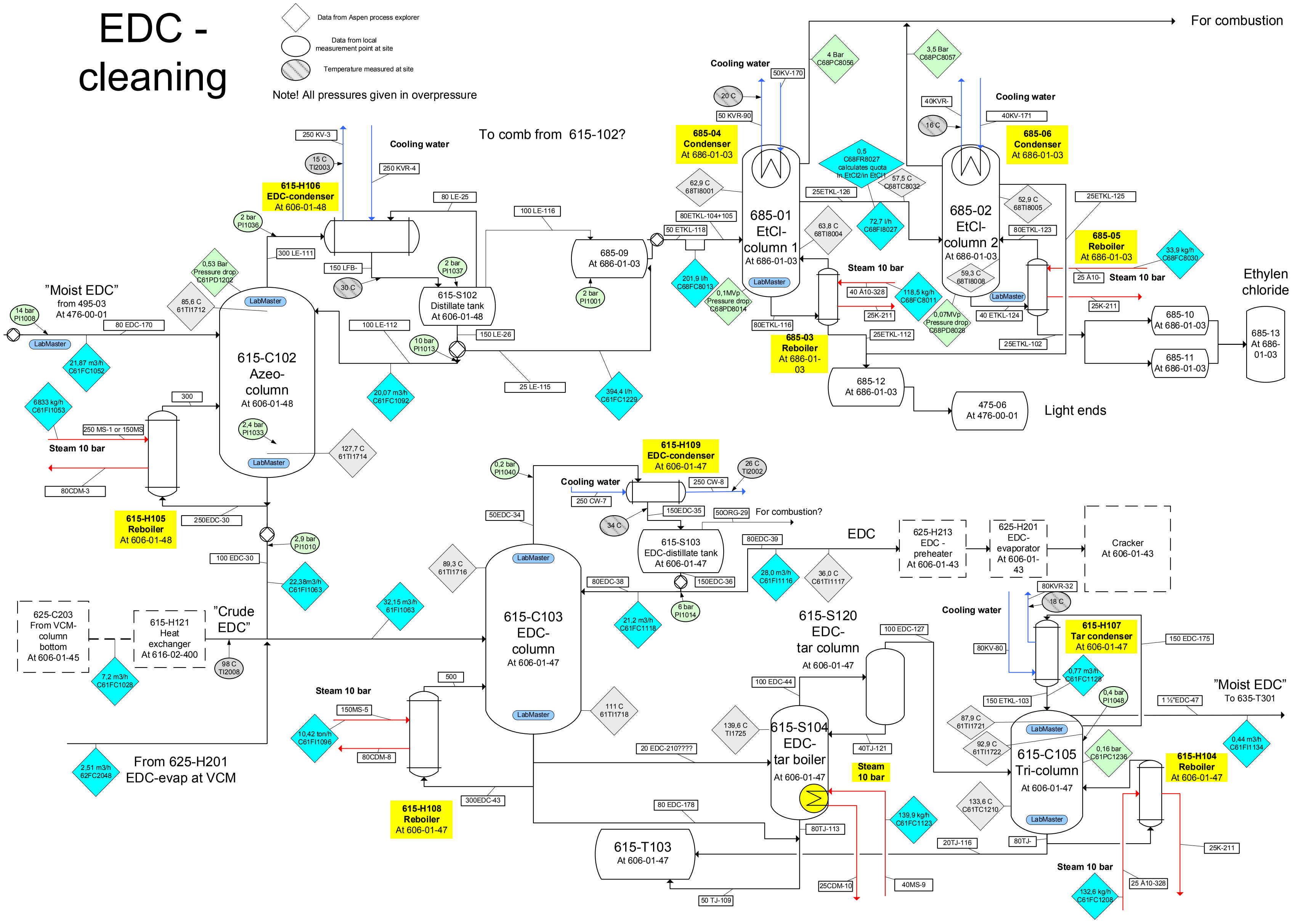
Appendix 1 – Detailed process flowsheets




VCM
EDC
Data from Aspen process explorer
Data from local measurement point at site
Temperature measured at site
Note! All pressures given in overpressure
625-H213 EDC-preheater At 606-01-43
Steam 1 bar
Steam 10 bar produced
625-H230 Flue gas boiler At 626-02-01
Preheater
625-R201 Cracker At 606-01-43
625-H201 EDC-evaporator At 606-01-43
Steam 28+20 bar
To C103
625-C201 Cooling column At 606-01-51
625-S202 Cooling column tank At 606-01-51
625-H202A Condenser At 606-01-51
625-H202B Condenser At 606-01-51
Cooling water
625-H211 Heat exchanger At 606-01-51
HCl to Oxireactor
625-H204 HCl-condenser At 606-01-44
Propene
625-C202 HCl-column At 606-01-44
625-T204 HCl-tank At 606-01-44
HCl
Till oxi
625-H205 Reboiler At 606-01-44
Steam 10 bar
To combustion gas cleaning
625-H207 Condenser At 606-01-45
625-T202 At 606-01-44
625-H212 Preheater At 606-01-45
625-C203 VCM-column At 606-01-45
625-S207 A+B VCM-dryer At 606-01-45
VCM
625-H206 Reboiler At 606-01-45
Steam 20 bar
To EDC-column
615-H121 Heat exchanger At 616-02-400
615-H122 Cooler At 616-02-400
615-R102 Chloration reactor At 616-02-400
KYLNING? H113 på 606-01-48?
615-R101 Chloration tank At 606-01-40
615-T101 EDC-tank At 616-01-40
635-T301 Acid cleaning At 606-01-49
615-T102 EDC-sodium hydroxid cleaning separator At 606-01-48
495-03 Storing tank "Moist EDC" At 476-00-01
"Moist EDC" to Azeo column (Return EDC)
625-C204 Tar column At 606-01-50
625-H210 VCM-tar condenser At 606-01-50
625-S208 Tar boiler At 606-01-50
For combustion
Steam 10 bar
LabMaster


Oxi
Data from Aspen process explorer
Data from local measurement point at site
Temperature measured at site
Note! All pressures given in overpressure
Ethylene
100ET-1
17,8 bar 61PI1302
-0,4 C 61TI1306
80ET-80
80ET-4
635-H308 Ethylen preheater At 606-01-41
3780,2Nm3/h 63FI3005
17,2 Bar 63PI3208
119,9 C 63TC3002
100 A10-11
Steam 10 bar
50 K-31
150 REN-1
HCl (liquid)
625-T204 HCl-tank At 606-01-44
9,0 Bar 62PI2247
-20 C
-31,9 C From HCl column 61TI3043
40 HCL-2
50 HVC-3
41 C
625-S210 HCl evaporator At 606-01-44
320,9 Nm3/h 63FC3013
50 MS-12
Steam 10 bar
9,0 Bar 63PC3008
40 CDM-12
80 HCL-3
179,6C 63TC3012
7,6 Bar 63PI3016
150 HVC-1
150 HVC-2
HCl from HCl column and 625-H211 At 606-01-44 and 606-01-51
21,0 C 62TI2041
635-H301 HCl heater At 606-01-41
50 K-30
50A21-10
Steam 20 bar?
6535,5 Nm3/h 63FI3009
100PL-3
8 bar 63PI3200
635-BL321 Air compressor At 606-01-41
56,5 C 63TI3021
25 C TI2008
180 C TI2007
2,2 bar PI1036
80 CW-16
18 C TI1302
80 CW-15
635-H321 Intermediate cooler At 606-01-41
Cooling water
350 PL-1
Air
6973,3 Nm3/h 63FI3036
150 PL-3
150 PL-3
224,3 C 63TI3040
3504 Nm3/h 63FI3051
250 EDC-218
250 REN-1
156,7 C 63TI3059
250 REN-1
635-R301 Oxi reactor 1 At 606-01-41
200 BF-3
635-S301 Steam dome At 606-01-41
80 HS-3 + 80 ES-3
150 BF-2
22,7 bar PC3068
5,21 ton/h FI3069
196,8 C 63TI3048
150 SG-2
100 SG-2
150SG-2
50 SG-01
411,85 Nm3/h 63FR3045
12,3 bar 63PI3037
Oxygen
635-R302 Oxi reactor 2 At 606-01-42
200 BF-7
635-S302 Steam dome At 606-01-42
80 HS-9
150 BF-6
20,2 bar PC3082
6,4 ton/h FI3100
250 EDC-228
100 SG-2
Steam 20 bar produced
635-R303 Oxi reactor 3 At 606-01-42
200 BF-11
635-S303 Steam dome At 606-01-42
80 HS-5
150 BF-10
21,0 bar PC3099
3,67 ton/h FI3101
217,2 C 63TI3103
250 REN-4
635-H302 Precooler At 606-01-49
100CW-19
100CW-20
Cooling water
4,4 bar 63PI3165
93,9 C 63TI3113
200 RET-1
150 RET-1
635-H303 A Condenser At 606-01-49
635-H303 B Condenser At 606-01-49
200CW-26
14 C TI2002
200CW-25
200 CW-27
Cooling water
200 RET-2
14 C
14 C
25 K-169
200CW-28
15,5 C TI2001
13,8 C
635-S309 Separator At 606-01-49
LabMaster
Better insulated pipes in this part of OXI, A temperature of 30 C is therefore assumed out from H303A+B
100RET-5
200ORG-108
13,2 C
20,8 C
22 C ??
146,2 Nm3/h FI163072
Fr HTC 615-S184
Chlorine
25 CL-5
3,4 C
635-H307 Chlorine evaporator At 606-01-49
50 K-159??
50 A01-11
Steam 1 bar
69,6 C 63TI3141
460,7 Nm3/h 63FC3135
9,7 bar 63PI3139
50 RG-109
300 RG-109
136,1 C 63TI3133
635-H304 A Heat exchanger At 606-01-49
635-H304 B Heat exchanger At 606-01-49
4,5 bar 63PC3134
50A10-314
25 K-16
137,9 C 63TI3131
Steam 10 bar
250MS-1
40 A10-313
4,5 bar 63PC3132
0,48 bar Pressure drop 63PD3152
128,3 C 63TI3138
300 RG-109
635-R304 Oxi reactor 4 At 606-01-49
150 BF-15
635-S304 Steam dome At 606-01-49
100 ES-4
Steam 1 bar produced
100BF-14
151,5 C 63TI3001
3,5 bar PI1027
150 RET-15
15 C
100 KVR-32
100 KV-31
635-H305 Condenser At 606-01-49
200 RET-16
150 RET-19
635-S306 EDC-Separator At 606-01-49
LabMaster
635-H306 Remainder gas condenser At 606-01-49
0,225 bar Pressure drop 63PD3156
-21,5 C 63TI3162
100 RET-20
100RET-101
50 P-16
Propene
25P-5 + 200P-1
For combustion
3 bar 63PI3164
635-S308 Demister At 606-01-49
635-S307 EDC-Final separator At 606-01-49
LabMaster
40 RET-18
40 RET-17
635-T301 Acid cleaning At 606-01-49
"Moist EDC" (Oxi EDC)

EDC - cleaning


Note! All pressures given in overpressure

Appendix 2 – Gathered data

In the process descriptions found in Appendix 1, measurement points are showed. Information about the measurement points, and the time period used, is found in this appendix.

Note! All pressures data is stated as gauge pressure

From Aspen process explorer:

Indicator	Value	Unit	Time period*	At:	Comment
60A10705	3.648	%	365 days, at 2011-04-14	VCM	Oxygen in flue gases
61FC1401	1	ton/h	2010-11-18 to 2011-02-17	HTC	The present value is 0. Only used for start-up
61FC1411:me	0	ton/h	2010-11-18 to 2011-02-17	HTC	Not in use
61FC1429:me	86.65	m3/h	2011-01-03 to 2011-02-17	HTC	NOTE! According to Peter, it must not exceed 120 m3/h
61FC1442:me	12.64	m3/h	2011-01-03 to 2011-02-17	HTC	
61FC1445	12.2	m3/h	40 days, at 2011-02-23	HTC	
61FC1506:me	4.975	l/min	2 days, at 2011-03-01	HTC	Tricolumn not used all the time. Used for some week and then turned off.
61FC1543:me	306	kg/h	2 days, at 2011-03-01	HTC	Tricolumn not used all the time. Used for some week and then turned off. The value increases with time
61FC1560:me	16.6	l/min	2 days, at 2011-03-01	HTC	Tricolumn not used all the time. Used for some week and then turned off.
61FC1645:me	3.12	m3/h	2011-01-03 to 2011-02-17	HTC	
61FI1063	32.15	m3/h	2011-01-09 to 2001-02-21	EDC	Description: "Feed EDC-column C103" Compare with C61FI1063.
61FI1559	1.2	l/min	1 day, at 2011-03-01	HTC	Tricolumn not used all the time. Used for some week and then turned off. The value has quick variations
61FI1608:av	17.65	l/min	2010-11-18 to 2011-02-17	HTC	
61FI1630:av	187.49	Nm3/h	2010-11-18 to 2011-02-17	HTC	
61FI1630:av	146.2	Nm3/h	30 days, at 2011-02-23	OXI	
61PC1448:me	0.296	Bar	30 days, at 2011-03-16	HTC	
61PC1558:me	0.102	bar	2 days, at 2011-03-01	HTC	Tricolumn not used all the time. Used for some week and then turned off.
61PC1617:me	1.25	Bar	30 days, at 2011-02-23	HTC	
61PC1640:me	0.095	Bar	30 days, at 2011-02-23	HTC	
61PD1625.1:av	0.03	MVp	30 days, at 2011-02-23	HTC	Pressure drop 0,0029 bar
61PI1302:av	17.8	Bar	30 days, at 2011-02-23	HTC	
61PI1326:av	2.5	Bar	30 days, at 2011-02-23	HTC	
61PI1540:av	0.114	bar	2 days, at 2011-03-01	HTC	Tricolumn not used all the time. Used for some week and then turned off.
61PI1644:av	4.265	bar	30 days, at 2011-03-01	HTC	
61PS1316.1:av	0.037	Bar	30 days, at 2011-02-23	HTC	Not in use
61TI1306:av	-0.395	C	30 days, at 2011-03-16	OXI	Seems to be variations because of the outdoor temp. Varies between -10 and 25
61TI1416	99.21	C	2010-11-18 to 2011-02-17	HTC	
61TI1419	89.2	C	2010-11-18 to 2011-02-17	HTC	
61TI1420	82.4	C	2010-11-18 to 2011-02-17	HTC	
61TI1426	74.37	C	2010-11-18 to 2011-02-17	HTC	
61TI1431	73.45	C	2010-11-18 to 2011-02-17	HTC	
61TI1441	26.73	C	2010-11-18 to 2011-02-17	HTC	
61TI1538:me	118	C	2 days, at 2011-03-01	HTC	Tricolumn not used all the time. Used for some week and then turned off.
61TI1542	87.1	C	2 days, at 2011-03-01	HTC	Tricolumn not used all the time. Used for some week and then turned off.
61TI1551	13.9	C	2 days, at 2011-03-01	HTC	Tricolumn not used all the time. Used for some week and then turned off.
61TI1603	15.84	C	2010-11-18 to 2011-02-17	HTC	

61TI1618:me	-16.5	C	2010-11-18 to 2011-02-17	HTC	
61TI1646:av	57.49	C	2011-01-03 to 2011-02-17	HTC	High?
61TI1701	52.8	C	30 days, at 2011-02-25	VCM	
61TI1712	85.58	C	2011-01-09 to 2011-02-21	EDC	
61TI1714	127.73	C	2011-01-09 to 2011-02-21	EDC	
61TI1716	89.32	C	2011-01-09 to 2011-02-21	EDC	
61TI1718	111.05	C	2011-01-09 to 2011-02-21	EDC	
61TI1721	87.86	C	2011-01-09 to 2011-02-21	EDC	irregular data has been removed
61TI1722	92.91	C	2011-01-09 to 2011-02-21	EDC	irregular data has been removed
61TI1725	139.6	C	30 days, at 2011-03-16	EDC	
62FC2048:me	2.51	m3/h	30 days, at 2011-03-01	VCM	
62FC2059:me	37.91	m3/h	2011-01-03 to 2011-02-17	VCM	
62FC2075:me	0.56	m3/h	2010-11-18 to 2011-02-17	VCM	
62FC2076:me	2801.62	kg/h	2011-01-03 to 2011-02-17	VCM	
62FC2102:me	36.48	m3/h	2010-11-18 to 2011-02-17	VCM	
62FC2106:me	10.96	m3/h	2010-11-18 to 2011-02-17	VCM	
62FC2413:me	7.36	ton/h	2011-01-03 to 2011-02-17	VCM	
62FC2415:me	11630.36	Nm3/h	2011-01-03 to 2011-02-17	VCM	
62FC2431:me	0.32	ton/h	2011-01-03 to 2011-02-17	VCM	Step functions.
62FC2443:me	0.3	m3/h	2010-11-18 to 2011-02-17	VCM	Goes up to 0,33 stays, and then goes down quickly again
62FC2461:me	37.94	m3/h	2011-01-03 to 2011-02-17	VCM	
62FI2044	0.661	m3/h	30 days at 2011-03-25	VCM	
62FI2109	16.45	m3/h	2010-11-18 to 2011-02-17	VCM	
62FI2115	0.823	ton/h	2010-11-18 to 2011-02-17	VCM	
62FI2216	4215.3	kg/h	2011-01-03 to 2011-02-17	VCM	
62FI2220	22.75	m3/h	2011-01-03 to 2011-02-17	VCM	
62FI2403	45.13	m3/h	2011-01-03 to 2011-02-17	VCM	
62FQ2133	19.45	ton/h	2011-01-03 to 2011-02-17	VCM	
62PC2007:me	23	Bar	30 days, at 2011-02-23	VCM	
62PC2069:me	9.1	Bar	30 days, at 2011-02-23	VCM	
62PC2073:me	0.6	Bar	30 days, at 2011-02-23	VCM	
62PC2117:me	5.0	Bar	30 days, at 2011-02-23	VCM	
62PD2121	0.55	Bar	30 days, at 2011-02-23	VCM	
62PI2046	9.6	Bar	30 days, at 2011-02-23	VCM	
62PI2246	10.25	Bar	15 days, at 2011-02-23	VCM	
62PI2247	9.0	Bar	30 days, at 2011-02-23	OXI	
62PI2418	20	Bar	30 days, at 2011-02-23	VCM	
62TC2036:me	40	C	2010-11-18 to 2011-02-17	VCM	
62TC2414:me	118.12	C	2011-01-03 to 2011-02-17	VCM	
62TI2001	486.8	C	2011-01-03 to 2011-02-17	VCM	
62TI2002	445.59	C	2011-01-03 to 2011-02-17	VCM	
62TI2003	231.33	C	2011-01-03 to 2011-02-17	VCM	
62TI2008	193.14	C	2011-01-03 to 2011-02-17	VCM	
62TI2018	208.61	C	2011-01-03 to 2011-02-17	VCM	
62TI2025D	585.4	C	365 days, at 2011-04-14	VCM	
62TI2028	160.91	C	2010-11-18 to 2011-02-17	VCM	
62TI2032	127.93	C	2010-11-18 to 2011-02-17	VCM	
62TI2033	157.9	C	30 days at 2011-03-25	VCM	
62TI2041	20.96	C	2010-11-18 to 2011-02-17	VCM	
62TI2042	-32.32	C	2010-11-18 to 2011-02-17	VCM	
62TI2043	-31.9	C	30 days, at 2011-03-01	VCM	
62TI2047	86.48	C	2011-01-03 to 2011-02-17	VCM	

62TI2051	16.96	C	2010-11-18 to 2011-02-17	VCM	
62TI2057	157.65	C	2010-11-18 to 2011-02-17	VCM	
62TI2061	71.46	C	2011-01-03 to 2011-02-17	VCM	
62TI2096	81.81	C	2010-11-18 to 2011-02-17	VCM	
62TI2097	83.21	C	2010-11-18 to 2011-02-17	VCM	
62TI2100	41.4	C	2010-11-18 to 2011-02-17	VCM	
62TI2250	119.09	C	2010-11-18 to 2011-02-17	VCM	
62TI2429:av	456.6	C	30 days, at 2011-03-16	VCM	
62TI2489	26.3	C	2010-11-18 to 2011-02-17	VCM	
63FC3013:me	320.9	Nm3/h	2010-11-22 to 2011-02-21	OXI	High variations
63FC3136:me	460.7	Nm3/h	2011-01-09 to 2011-02-21	OXI	Bad data have been removed. Quite high variations
63FI3005:av	3780.2	Nm3/h	2011-01-09 to 2011-02-21	OXI	
63FI3009:av	6635.5	Nm3/h	2011-01-09 to 2011-02-21	OXI	
63FI3036:av	6973.3	Nm3/h	2011-01-09 to 2011-02-21	OXI	Chosen because of the similar values to XD. Slightly irregular in the end of the interval
63FI3057:av	3504	Nm3/h	2011-01-09 to 2011-02-21	OXI	Bad data have been removed. Quite high variations
63FI3069:av	5.21	ton/h	30 days, at 2011-03-16	OXI	
63FI3100:av	6.4	ton/h	30 days, at 2011-03-16	OXI	
63FI3101:av	3.67	ton/h	30 days, at 2011-03-16	OXI	
63FR3045:me	411.854	Nm3/h	30 days, at 2011-03-16	OXI	
63PC3008:me	9.0	Bar	30 days, at 2011-02-23	OXI	
63PC3068:me	22.7	Bar	30 days, at 2011-03-16	OXI	
63PC3082:me	20.2	Bar	30 days, at 2011-03-16	OXI	
63PC3099:me	21	Bar	30 days, at 2011-03-16	OXI	
63PC3132:me	4.5	Bar	30 days, at 2011-02-23	OXI	
63PC3134:me	4.5	Bar	30 days, at 2011-02-23	OXI	
63PC3164:me	3	Bar	30 days, at 2011-03-16	OXI	
63PD3152:av	2297	mmVp	30 days, at 2011-02-23	OXI	0,225 bar
63PD3152:av	4.9	MVp	30 days, at 2011-03-01	OXI	0,48 bar
63PD3156:av	4.9	MVp	30 days, at 2011-02-23	OXI	0,48 bar
63PD3156:av	2297	mmVp	30 days, at 2011-03-01	OXI	0,225 bar
63PI3016:av	7.6	Bar	30 days, at 2011-02-23	OXI	
63PI3037:av	12.3	Bar	30 days, at 2011-03-16	OXI	
63PI3139:av	9.7	Bar	30 days, at 2011-02-23	OXI	
63PI3165:av	4.4	Bar	30 days, at 2011-02-23	OXI	
63PI3200:av	8	Bar	30 days, at 2011-02-23	OXI	
63PI3208:av	17.2	Bar	30 days, at 2011-02-23	OXI	
63TC3002:me	119.9	C	2011-01-09 to 2011-02-21	OXI	
63TC3012:me	179.6	C	2011-01-09 to 2011-02-21	OXI	
63TI3001:av	151.5	C	2011-01-09 to 2011-02-21	OXI	
63TI3021:av	56.5	C	2011-01-09 to 2011-02-21	OXI	
63TI3040:av	224.25	C	2011-01-09 to 2011-02-21	OXI	
63TI3048	196.8	C	30 days, at 2011-03-16	OXI	
63TI3059:av	156.7	C	2011-01-09 to 2011-02-21	OXI	
63TI3103:av	217.2	C	2011-01-09 to 2011-02-21	OXI	
63TI3113:av	93.9	C	2011-01-09 to 2011-02-21	OXI	
63TI3131:av	137.9	C	2011-01-09 to 2011-02-21	OXI	
63TI3133:av	136.1	C	2011-01-09 to 2011-02-21	OXI	
63TI3138:av	128.3	C	2011-01-09 to 2011-02-21	OXI	
63TI3141:av	69.6	C	2011-01-09 to 2011-02-21	OXI	
63TI3162:av	-21.5	C	2011-01-09 to 2011-02-21	OXI	
68TI8001	62.88	C	2010-11-28 to 2011-02-21	EDC	
68TI8004	63.78	C	2010-11-28 to 2011-02-21	EDC	

68TI8005	52.94	C	2010-11-28 to 2011-02-21	EDC	
68TI8008	59.27	C	2010-11-28 to 2011-02-21	EDC	
C61FC1003	4.6	Nm3/h	30 days, at 2011-02-25	VCM	
C61FC1028	7.19	m3/h	2011-01-03 to 2011-02-17	VCM	
C61FC1052	21.87	m3/h	2011-01-09 to 2011-02-21	EDC	
C61FC1092	20.07	m3/h	2011-01-09 to 2011-02-21	EDC	
C61FC1118	21.22	m3/h	2011-01-09 to 2011-02-21	EDC	
C61FC1123	139.91	kg/h	2011-01-09 to 2011-02-21	EDC	Irregular data has been removed
C61FC1128	0.77	m3/h	30 days at 2011-03-25	EDC	
C61FC1208	132.63	kg/h	2011-01-09 to 2011-02-21	EDC	Irregular data has been removed. Quite unstable, varies between 120 and 170
C61FC1223	35.24	kg/h	2011-01-03 to 2011-02-17	VCM	
C61FC1229	394.4	l/h	2010-11-28 to 2011-02-21	EDC	Very high variations. Differs between 0 and 1000.Should probably not be used.
C61FI1053	6833	kg/h	2011-01-09 to 2011-02-21	EDC	
C61FI1063	22.38	m3/h	2011-01-09 to 2011-02-21	EDC	Description: "EDC from Azeocolumn to EDC column" Compare with 61FI1063.
C61FI1096	10.42	ton/h	2011-01-09 to 2011-02-21	EDC	
C61FI1116	28.01	m3/h	2011-01-09 to 2011-02-21	EDC	Bad data has been removed
C61FI1132	4.5	m3/h	30 days, at 2011-02-23	VCM	
C61FI1134	0.44	m3/h	2011-01-09 to 2011-02-21	EDC	Very quick variations
C61FS1230	17.29	m3/h	2011-01-03 to 2011-02-17	VCM	
C61PC1031	0.35	Bar	30 days, at 2011-02-23	VCM	
C61PC1236	0.16	bar	30 days at 2011-03-25	EDC	
C61PD1202	0.53	Bar	30 days, at 2011-02-23	EDC	
C61TC1210	133.6	C	2011-01-09 to 2011-02-21	EDC	Irregular data has been removed
C61TI1117	36.02	C	2011-01-09 to 2011-02-21	EDC	
C68FC8011	118.51	kg/h	2010-11-28 to 2011-02-21	EDC	
C68FC8013	201.94	l/h	2011-01-09 to 2001-02-21	EDC	Looks like some kind of step function. Differs quite much.
C68FC8030	33.87	kg/h	2010-11-28 to 2011-02-21	EDC	
C68FI8027	72.72	l/h	2010-11-28 to 2011-02-21	EDC	Quite high variations, between 35 to 110
C68FR8027	0.5		2010-11-28 to 2011-02-21	EDC	"Ratio In etylkol.2/In etylkol.1" Varies between 0,3 and 0,7
C68PC8056	4	Bar	30 days, at 2011-02-23	EDC	
C68PC8057	3.5	Bar	30 days, at 2011-02-23	EDC	
C68PD8014	0.1	MVp	30 days, at 2011-02-23	EDC	
C68PD8028	0.07	MVp	30 days, at 2011-02-23	EDC	
C68TC8032	57.5	C	30 days, at 2011-02-23	EDC	
FI2309	2.545	ton/h	30 days, at 2011-04-14	VCM	
H62PI2221	15.2	Bar	30 days, at 2011-02-23	VCM	
H62PI2270	0.606	Bar	30 days, at 2011-02-25	VCM	

**If date to date: Data gathered with the following properties in Aspen: Sampling: Average, 1 hour Method: Integral, Stepped. Average calculated in excel*
If days at date: Data gathered with the following properties in Aspen: Sampling: Average, Method: Integral, Stepped. For time, see time period

Data from local measurement points:

Name of unit:	At:	Number of point	Value	Unit	Description	Date	Outdoor temp:
615- S103	EDC	PI1014/1015	6	bar	Pressure after pump after S103	28-feb	ca 0 C
615-C102	EDC	PI1008+1009	14	bar	Pressure after pump, inlet stream Azeocolumn	15-mar	ca 5 C
615-C102	EDC	PI1010+1011	2.9	bar	Pressure after pump, bottom outlet Azeocolumn	15-mar	ca 5 C
615-C102	EDC	PI1033	2.4	bar	Pressure bottom Azeo column	28-feb	ca 0 C
615-C105	EDC	PI1048	0.4	bar	Pressure Tri-column	28-feb	ca 0 C
615-H106	EDC	PI1036	2	bar	Pressure top Azeo column/ in EDC condenser	28-feb	ca 0 C
615-H106	EDC	TI2003	15	C	Temp. outgoing cooling water EDC condenser H106	28-feb	ca 0 C
615-H109	EDC	PI1040	0.2	bar	Pressure in EDC-condenser and out from top EDC column	28-feb	ca 0 C
615-H109	EDC	TI2002	26	C	Temp. outgoing cooling water EDC condenser H109	28-feb	ca 0 C
615-S102	EDC	PI1013	10	bar	Pressure after pump after S102	28-feb	ca 0 C
615-S102	EDC	PI1037	2	bar	Pressure distillate tank S102	28-feb	ca 0 C
685-09	EDC	PI1001	2	bar	Pressure tank 685-09	28-feb	ca 0 C
615-C151	HTC	PI1111+1112	5.5	bar	Pressure after pump, bottom HTC-column	28-feb	ca 0 C
615-C151	HTC	PI1114	0.5	bar	Pressure HTC column	15-mar	ca 5 C
615-H153	HTC	TI2018	68	C	Temperature outgoing cooling water condenser H153	28-feb	ca 0 C
615-H154	HTC	TI2015	16	C	Temperature outgoing cooling water condenser H154	28-feb	ca 0 C
615-H159	HTC	TI2013	12	C	Temperature outgoing cooling water from cooler H159	28-feb	ca 0 C
615-H160	HTC	TI2020	15	C	Temperature outgoing cooling water from cooler H160	28-feb	ca 0 C
615-H166	HTC	TI2023	34	C	Temperature outgoing stream from H166	28-feb	ca 0 C
615-S151	HTC	PI1115	0.1	bar	Pressure S151	15-mar	ca 5 C
615-S151	HTC	PI1116+1117	7	bar	Pressure after pump after S151	15-mar	ca 5 C
615-S152	HTC	PI1126	0.1	bar	Pressure EDC-separator S152	28-feb	ca 0 C
615-S156	HTC	PI1152+1153	5.2	bar	Pressure after pump after S156	28-feb	ca 0 C
T301	HTC	PI1124+1125	1.8	bar	Pressure after pump after 615-S152+615-S153	28-feb	ca 0 C
635-H303A	OXI	TI2002	14	C	Temperature outgoing cooling water condenser H203A	28-feb	ca 0 C
635-H303B	OXI	TI2001	15.5	C	Temperature outgoing cooling water condenser H203B	28-feb	ca 0 C
635-H306	OXI	PI1027	3.5	bar	Pressure in remainder gas condenser H306	28-feb	ca 0 C
635-H321	OXI	PI1036	2.2	bar	Pressure of air going in to intermediate cooler H321	28-feb	ca 0 C
635-H321	OXI	TI1302	18	C	Temperature outgoing cooling water intermediate cooler H321	28-feb	ca 0 C
635-H321	OXI	TI2007	160	C	Temperature of air going in to intermediate cooler H321	28-feb	ca 0 C
635-H321	OXI	TI2008	25	C	Temperature of air going out of intermediate cooler H321	28-feb	ca 0 C

615-H122	VCM	PI1052	2	bar	Pressure out H122	28-feb	ca 0 C
615-H122	VCM	TI2009	60	C	Temperature out R102 after 615-H122 = temp. Out of H122	28-feb	ca 0 C
615-T101	VCM	PI1001+1002	9	bar	Pressure after pump after T101	15-mar	ca 5 C
625-H202A	VCM	TI2001	50	C	Temperature outgoing cooling water condenser H202A	28-feb	ca 0 C
625-H210	VCM	TI2002	13	C	Temp. outgoing cooling water VCM-tar condenser H210	28-feb	ca 0 C
625-H212	VCM	PI1008	11.5-12	bar	Pressure after pump after 625-T202/ in preheater H212	28-feb	ca 0 C
625-S202	VCM	PI1006	15	bar	Pressure after pump after 625-S202	28-feb	ca 0 C
615-H121	VCM + EDC	TI2008	98	C	Temperature out cold stream H121 to EDC-column	28-feb	ca 0 C

Temperatures measured at pipes:

Name of unit:	At:	Value:	Unit	Description	Comment	Date	Outdoor temp
615-H106	EDC	30	C	Temperature out H106		15-mar	ca 5 C
615-H107	EDC	19	C	Temperature outgoing cooling water tar condenser 615-H107		28-feb	ca 0 C
615-H109	EDC	34	C	Temperature out H109		15-mar	ca 5 C
685-04	EDC	20	C	Temperature outgoing cooling water condenser 685-04		28-feb	ca 0 C
685-06	EDC	16	C	Temperature outgoing cooling water condenser 685-06		28-feb	ca 0 C
615-H153	HTC	72	C	Temperature outgoing cooling water condenser H153 measured very close to the heat exchanger		28-feb	ca 0 C
615-H166	HTC	20	C	Temperature outgoing cooling water cooler H166		28-feb	ca 0 C
625-S210	OXI	-20	C	Temperature in S210		16-mar	ca 3 C
625-S210	OXI	41	C	Temperature out S210		16-mar	ca 3 C
635-H303	OXI	14	C	Temperature out H303, measured before S309	Better insulated pipes. Cannot be used	16-mar	ca 3 C
635-H303A	OXI	14	C	Temperature out H303A	Better insulated pipes. Cannot be used	16-mar	ca 3 C
635-H303B	OXI	13.8	C	Temperature out H303B	Better insulated pipes. Cannot be used	16-mar	ca 3 C
635-H304	OXI	20.8	C	Temperature in H304	Better insulated pipes. Cannot be used	16-mar	ca 3 C
635-H304	OXI	22	C	Temp of stream from HTC to H304	Not sure! Measured very close to the other pipe.	16-mar	ca 3 C
635-H305	OXI	15	C	Temperature outgoing cooling water condenser 635-H305		28-feb	ca 0 C
635-H307	OXI	3.4	C	Temperature in H307		15-mar	ca 5 C
635-S309	OXI	13.2	C	Temperature from top S309		16-mar	ca 3 C
615-H121	VCM	41	C	Temperature in H121 cold stream		15-mar	ca 5 C
615-H122	VCM	118	C	Temperature out from H121/ in H122 at 80EDC-2079		28-feb	ca 0 C
615-H122	VCM	60	C	Temperature out from H122	Note! Target temp is 60 C.	28-feb	ca 0 C
615-H122	VCM	16	C	Temperature outgoing cooling water cooler H122		28-feb	ca 0 C
625-C204	VCM	110	C	Temperature bottom Tar column		28-feb	ca 0 C
625-H202B	VCM	66	C	Temperature media between 625-H202A and 625-H202B		28-feb	ca 0 C
625-H202B	VCM	15	C	Temperature cooling water between 625-H202A and 625-H202B	Not sure about the value	28-feb	ca 0 C
625-H211	VCM	11.5	C	Temperature out Hot stream (6"CVC-7) from HX H211		28-feb	ca 0 C
625-H212	VCM	84	C	Temperature out H212		15-mar	ca 5 C

LabMaster data

Name of unit	Description	Substans	Mean Value	Unit	Substans	Mean Value	Unit	Substans	Mean Value	Unit	Substans	Mean Value	Unit	Substans	Mean Value	Unit	Substans	Mean Value	Unit	Substans	Mean Value	Unit
615-C102	**Azeo-column bottom**	EDC	99,2	Mass%	1.1.2Trikloretan	3531	ViPPM															
615-C102	**Azeo-column top**	Kloroform	38,1	Mass%	Ethylene chloride	29,1	Mass%	Tr1.2dikloreten	10,7	Mass%	1.1dikloretan	5,7	mass%	cis1.2dikloreten	4,9	Mass%	VCM	2,8	Mass%	Koltetraklorid	4,4	Mass%
615-C102	**Feed into EDC-cleaning**	EDC	97,9	Mass%	Etylklorid	4052	ViPPM	Kloroform	4550	ViPPM	1.1.2Trikloretan	3302	ViPPM									
615-C103	**EDC-column bottom**	EDC	65,7	Mass%	1.1.2Trikloretan	15,7	Mass%	Tetrakloretan	1,9	Mass%	Pentakloretan	2,7	Mass%	Bromtrikloretan	1,3	Mass%						
615-C103	**EDC-column top**	EDC	99,45	Mass%	Koltetraklorid	2575	ViPPM	Trikloreten	1366	ViPPM	Bensen	1000	ViPPM									
615-C105	**EDC tricolumn**	1.1.2Trikloretan	50,7	Mass%	EDC	6,5	Mass%	Tetrakloretan	6	Mass%	Pentakloretan	8,5	mass%	Bromtrikloretan	3,4	Mass%	Hexakloretan	1,9	Mass%			
615-C105	**EDC tricolumn**	EDC	99,1	Mass%	1.1.2Trikloretan	1	Mass%															
615-C151	**HTC-column bottom**	EDC	83,2	Mass%	1.1.2Trikloretan	14,6	Mass%	1Brom2Kloretan	1,7	Mass%												
615-C152	**HTC tricolumn**	EDC	98,3	Mass%	1Brom2Kloretan	0,5	Mass%	1.1.2Trikloretan	1,1	Mass%												
615-C152	**HTC tricolumn**	1.1.2Trikloretan	80,5	Mass%	EDC	9,2	Mass%	1Brom2Kloretan	6,7	Mass%	Tetrakloretan	5,1	mass%									
615-R151	**HTC-reactor**	EDC	59,3	Mass%	1.1.2Trikloretan	18,5	Mass%	1Brom2Kloretan	2,7	Mass%	Tetrakloretan	1,8	mass%									
615-S151	**HTC destillate tank**	EDC	99,9	Mass%																		
615-S152 + 615-S153	**HTC EDC-separator**	EDC	99,85	Mass%	Eten	471	ViPPM															
615-S153	**HTC vent. Gas**	Kväve	68,6	Vol%	Eten	24,31	Vol%	Etan	1	Vol%	EDC	1	Vol%	Syre	3,7	Vol%						
625-C202	**HCl column top**	HCl	99,7	Vol%																		
625-C203	**VCM column bottom**	EDC	98,5	Mass%	Koltetraklorid	5386	ViPPM															
625-H213	**Feed to R201**	EDC	99,6	Mass%	Koltetraklorid	2030	ViPPM															
635-S306	**OXI EDC-separator**	EDC	98,5	Mass%	Etylklorid	3830	ViPPM	Kloral	3469	ViPPM												
635-S307	**OXI EDC final separator**	EDC	97,1	Mass%	Etylklorid	18553	ViPPM	Tr1.2dikloreten	2980	ViPPM												
635-S309	**OXI EDC-separator**	EDC	98,2	Mass%	Etylklorid	4440	ViPPM	1.1.2Trikloretan	3766	ViPPM												
685-01	**EtCl column 1 bottom**	Kloroform	54,8	Mass%	tr1.2Dikloreten	14,8	Mass%	1.1Dikloretan	7,6	Mass%	cis1.2Dikloreten	6,9	Mass%	Koltetraklorid	6,4	Mass%	Etylklorid	5,5	Mass%			
685-02	**EtCl column 2 bottom**	Etylklorid	98,4	Mass%	Etenoxid	15631	ViPPM															

Appendix 3 - Calculations of composition and flow rate leaving Oxi-reactors

The flow rate and composition out from the Oxi-reactors are not measured. In this appendix, calculations are done based on the ingoing substances to the reactors.

R301-R303

Inlet flow of substances to reactor 1-3:

Component	**Volume flow into reactor R301-R303 [Nm3/h] (from gathered data)**	**Mole flow[mol/s] (from Aspen plus)**	Measurement point
Ethylene	3780.2	**42.35994**	63FI3005
HCl	6635.5	**74.35481**	63FI3009
Air	6973.3	78.14007	63FI3036
Air-nitrogen		**61.7306553**	
Air-oxygen		16.4094147	
Oxygen	411.85	4.61503	63FR3045
Total oxygen flow		**21.0244447**	

Reaction: $C_2H_4+2HCl+½O_2 \rightarrow C_2H_4Cl_2+H_2O$

All HCl is assumed to react. However, some will dissolve into water, and hydrochloric acid with a concentration of 8% is leaving the Oxi-reactors.

Reactions:

For 1 mole of HCl reacted;		
Needed:	Ethylene	½ mole
	Oxygen	1/4 mole
Produced:	EDC	½ mole
	Water	½ mole
Dissolved*	HCl	0.0438

*8% HCl-acid out

In the following table, the outlet concentrations are calculated

Outlet concentrations and flow rate from R303:

Compound:	**In to R301-R303[mol/s]**	**Change [mol/s]**	Dissolved [mol/s]	**Out from R303[mol/s]**	Molefraction out
HCl	74.355	-71.257	-3.098	0.000	0,000
Ethylene	42.360	-35.628		6.732	0,046
Nitrogen	61.731	0.000		61.731	0,423
Oxygen	21.024	-17.814		3.210	0,022
EDC	0.000	35.628		35.628	0,244
Water	0.000	35.628		35.628	0,244
Dissolved HCl			3.098	3.098	0,021
		Total mole flow:		146.027	

R304

The flow consists of two parts. Chlorine from a chlorine evaporator, where the flow is measured, and a flow from a separator after the first three reactors, which had to be simulated in Aspen plus.

Reaction: $C_2H_4+Cl_2 \rightarrow C_2H_4Cl_2$

Outlet concentrations and flow rate from R304:

From 63FC3135: Into R304 from H307:				
Compound:	In to R304 [mol/s]	Change [mol/s]	Out from R304[mol/s]	
Chlorine	5.162	-5.162	0.000	0.000
From Aspen simulations: Into reactor R304 from H304:				
Compound:	In to R304 [mol/s]	Change [mol/s]	Out from R304 [mol/s]	Molefraction
Ethylene	5.931	-5.162	0.769	0.011
EDC	0.864	5.162	6.026	0.085
Water	0.122	0.000	0.122	0.002
Oxygen	3.264	0.000	3.264	0.046
Nitrogen	62.880	0.000	62.880	0.890
Ethane	0.016	0.000	0.016	0.000
		Total mole flow:	73.0775811	

Appendix 4 - Assumptions and input data for the heat exchanger calculations

Heat exchanger calculations, and simulations in Aspen plus

Aspen plus was used to calculate the loads in the heat exchangers. Input needed was inlet and outlet temperature, pressure, flow rate and composition. In the case where not everything could be measured, assumptions had to be done. In this appendix, a summary for assumptions about each heat exchanger, used for the Aspen simulations, can be found.

HTC-unit, existing

H153

Process data

P = 0.1 bar(g)	(PI1115)
T_{in}= 82.4 °C	(61TI1420)
T_{out} = 74.4 °C	(61TI1426)

Assumptions

●Flow rate and composition was calculated by adding:

-*Reflux*, flow rate 86.7m^3/h (61FC1429), pressure 7 bar(g) (PI1116), temperature 74.4 C (61TI1426) and composition given in LabMaster.

-*Flow to T301¸* flow rate 17.65 l/min (61FC1608), pressure 1.8 bar(g) (PI1124) and composition given in LabMaster.

-*Flow from top S153*, flow rate 146.2 Nm^3/h (61FI1630), pressure 0.1 bar(g) (61PC1640), temperature 74.4 C (61TI1426) and composition given in LabMaster.

●To get the vapour fraction value 1.0 out of the top of the cooling column and in to the heat exchanger, the pressure was lowered to 0.98 bar. The lowered pressure is kept through the heat exchanger, but in the flash afterwards, the pressure is increased to the measured value, to simulate how much that leaves as vapour and liquid. This means there will be a small load on the flash, which is neglected. If it is vapour fraction 1.0 into the heat exchanger, or if the gas is bringing some small amount of liquid, is not known.

Results from Aspen plus:

Q = 9482kW

H154

Process data

P = 0.1 bar(g)	(PI1115)
T_{in}= 73.5 °C	(61TI1431)
T_{out} = 15.8 °C	(61TI1603)

Assumptions

●Flow rate and composition was calculated by adding:

-*Flow to T301¸* flow rate 17.65 l/min (61FC1608), pressure 1.8 bar(g) (PI1124) and composition given in LabMaster.

-*Flow from top S153*, flow rate 146.2 Nm^3/h (61FI1630), pressure 0.1 bar(g) (61PC1640), temperature 74.4 C (61TI1426) and composition given in LabMaster.

●To keep it all vapour before the heat exchanger, the pressure had to be lowered to 1.05 bar.

Results from Aspen plus:

Q =150kW

H155

Process data

P = 0.1 bar(g)	(PI1126)
T_{in}= 15.8 °C	(61TI1603)
T_{out} = -16.5 °C	(61TI1618)

Assumptions

● Flow rate and composition was obtained by simulation of H154 and a flash afterwards.

Results from Aspen plus:

Q = 5kW

H158AB
Process data
P= 0.1 bar(g) in middle of column (61PI1540)
T_{in}= 118 °C (61TI1538)
Composition from LabMaster
$\dot{m}_{steam}$=306 kg/h (61FC1543)
-Only run 1/3 of the time
Assumptions
- Pressure set to 1.3 bar to obtain saturated liquid in.
- Assuming saturated vapour out.
- 1-bromo-2-chloro-ethane does not exist in the Aspen database. It was approximated with 50% $C_2H_4Cl_2$ and 50% $C_2H_4Br_2$

Results from Aspen plus:
Q=170 kW
T_{out} = 123.5 °C

H159
Process data
P= 0.1 bar(g) in middle of column (61PI1540)
T_{in}= 87.1 °C (61TI1542)
T_{out} = 13.9 °C (61TI1551)
Composition from LabMaster
-Only run 1/3 of the time
Assumptions
- Flow rate was calculated by adding:

-*Reflux*, flow rate 16,6 l/min (61FC1560) and pressure 5.2 bar(g) (PI1152),
- *Return HTC reactor*, flow rate 1.2 l/min (61FI1559) and pressure 5.2 bar(g) (PI1152)

Results from Aspen plus:
Q=157 kW

H160
Process data
P= 0.3 bar(g) after cooler (61PC1448)
T_{in}= 89.2 °C (61TI1419)
T_{out} = 26.7 °C (61TI1441)
Flow rate=12.6 m^3/h (61FC1442)
Assumptions
- Composition: Pure EDC

Results from Aspen plus:
Q = 364 kW

H166
Process data
P= 4.3 bar(g) after S184 (61PI1644)
T_{in}= 57.5 °C (61TI1646)
T_{out} = 34 °C (TI2023)
Flow rate=3.12 m^3/h (61FC1645)
Composition from LabMaster
Results from Aspen plus:
Q = 33kW

HTC-unit, future

No temperatures or pressures are assumed to change in the future, but will be the same as in the previous calculations for the present HTC unit.

As mentioned, the reflux is limited, and cannot increase by more than 38%. However, the production of light products is assumed to be doubled, like the production of all other products. This means that a new steady-state is assumed to be reached, with a slightly higher content of volatile compounds in the stream leaving the top of the HTC-column. Since the HTC-condenser is substantially larger than the following coolers, it is possible for the reflux to increase by only 38%, and the mass flow through the following coolers to increase with 100%. The increased flow in the other heat exchangers will barely increase the load of the HTC-condenser, which increases with about 38%.

H153
Current load
Q = 9482kW
Assumptions
As mentioned, the reflux is limited, and cannot increase by more than 38%. However, the production of light products is assumed to be doubled, like the production of all other products. This means that a new steady-state is assumed to be reached, with a slightly higher content of volatile compounds in the stream leaving the top of the HTC-column. Since the HTC-condenser is substantially larger than the following coolers, it is possible for the reflux to increase by only 38%, and the mass flow through the following coolers to increase with 100%. The increased flow in the other heat exchangers will barely increase the load of the HTC-condenser, which increases with about 38%.
Future load
Q=13065kW

H154
Current load
Q=150 kW
Assumptions
Load doubled. See assumptions for H153.
Future load
Q=300kW

H155
Current load
Q=5 kW
Assumptions
Load doubled. See assumptions for H153.
Future load
Q=10kW

H158AB
Current load
Q=170 kW
Assumptions
- Today, it is only running 1/3 of the time. In the future, when the production is doubled, it is assumed to be running continuously, but at 2/3 of today’s level.

Future load
Q=114 kW

H159
Current load
Q=157 kW
Assumptions
- Today, it is only running 1/3 of the time. In the future, when the production is doubled, it is assumed to be running continuously, but at 2/3 of today’s level.

Future load
Q=104 kW

H160
Current load
Q=364 kW
Assumptions
Production doubled. However, after discussion with the company, the flow rate is expected to increase to 35 m^3/h in the future
Future load
Q=1011kW

H166
Current load
Q =33 kW

Assumptions
Load doubled. See assumptions for H153
Future load
Q =65 kW

New started heat exchangers not in use today:

H164

- Simulated in Aspen Exchanger design and rating. The simulations were based on an information sheet for the heat exchanger. The results are very insecure, but will not affect the result since the additional load is placed on Hnew
- Assumed inlet temperature: 100°C
- Assumed outlet temperature: 40°C

Future load
1200 kW

H165

- Simulated in Aspen Exchanger design and rating. The simulations were based on an information sheet for the heat exchanger. The results are very insecure, but will not affect the result since the additional load is placed on Hnew
- Assumed inlet temperature: 115 °C
- Assumed outlet temperature: 40 °C

Future load
1600 kW

Hnew

- Calculated by doubling the load of the current H153, and subtracting the load for the future H153, H164 and H165
- Assumed inlet temperature: 100 °C
- Assumed outlet temperature: 40 °C

Future load
3099 kW

VCM-unit

H121

Process data
Hot side:

P= 5.5 bar(g)	(62PC2117+62PD2121)
T_{in}= 157.7 °C	(62TI2057)
T_{out} = 118 °C	(Measured at pipe)
Flow rate=16.45 m^3/h	(62FI2109)

Composition from LabMaster, VCM column bottom
Cold side:

P= 9 bar(g)	(PI1001)
T_{in}= 41 °C	(Measured at pipe)
T_{out} = 98 °C	(TI2008)
Flow rate=7.2 m^3/h	(C61FC1028)

Composition unknown
Assumptions

- Bad agreement between the hot and the cold side of the heat exchanger. The load is determined from the hot side, since the data measured are closer to the heat exchanger, and are believed to be more trustworthy.

Results from Aspen plus:
Q=300 kW

H122

Process data

P= 5.5 bar(g)	(62PC2117+62PD2121)
T_{in}= 118 °C	(Measured at pipe)
T_{out} = 60 °C	(TI2009 after R102)
Flow rate=16.45 m^3/h	(62FI2109)

Composition from LabMaster, VCM column bottom

Assumptions

●Same flow rate as in H121 assumed, even though the flow rate measured after the heat exchanger is larger.

Results from Aspen plus:

Q=389 kW

H201

Process data

P_{in}= 20 bar(g) (62PI2418)
P_{out}= 15.2 bar(g) (H62PI1221)
T_{in}= 193.1 °C (62TI2008)
T_{out} =208.61 °C (62TI2018)
Composition from LabMaster, Feed to R201

Assumptions

●Flow rate calculated by adding 11630 Nm^3/h (62FI2109) to cracker and 2.51 m^3/h (62FC2048) to C103.

●The flow to C103 is assumed to have same pressure as the outlet stream to the cracker, but is assumed to leave as liquid, and therefore at a temperature of 207.5 °C

Results from Aspen plus:

Q=3442 kW

H202A

Process data

P= 10.25 bar(g) (62PI2246)
T_{in}= 127.9 °C (62TI2032)
T_{out} = 66 °C (Measured at pipe)

Assumptions

●The condensers and the cooling column tank afterwards were modeled. The condensate flow rate was calculated by adding 37.9 m^3/h (62FC2059) to the cooling column and 37.9 m^3/h (62FC2461) to the HCl-column. The flow rate in the condensers was set to obtain the calculated condensate flow.

●The composition is unknown. It is taken from the old Aspen simulation, made in 2001, and is changed slightly because of a mass balance made for ingoing and outgoing HCl to the HCl-column

●Pressure lowered to 10 bar(g), and temperature increased to 132.8 C, to obtain vapour fraction 1.

Results from Aspen plus:

Q=7007 kW

H202B

Process data

P= 10.25 bar(g) (62PI2246)
T_{in}= 66°C (Measured at pipe)
T_{out} = 40°C (62TC2036)

Assumptions

●The same assumptions as for H202A

Results from Aspen plus:

Q=2098 kW

H204

Process data

P= 9.1 bar(g) (62PC2117+62PD2121)
T_{out} =-32.3°C (62TI2042)
Composition from LabMaster, HCl column top
Flow rate propene=22.75 m^3/h (62FI2220)
Pressure propene=0.6 bar(g) (62PC2073)

Assumptions

●The load is calculated by assuming all propene liquid, and vapour out.

●The inlet temperature is set to obtain vapour fraction 1.0 in.

Results from Aspen plus:

T_{out} =-31.7°C
Q =1640kW

H205

Process data

P_{in}= 9.6 bar(g) (62PI2046)
T_{in}= 86.5°C (62TI2047)
Flow rate of 10bar steam: 2.8 ton/h (62FC2076)

Assumptions

- Composition calculated in Aspen plus from the fact that it should be saturated liquid in, and the mixture should contain only EDC and VCM
- Outlet temperature calculated, assuming vapour fraction 1 for the stream leaving the reboiler
- The load calculated from the condensation heat in the steam

Results from Aspen plus:

Q=1559 kW
T_{out}= 145°C

H206

Process data

P= 5.5 bar(g) (62PC2117+62PD2121)
T_{in}= 157.7 °C (62TI2057)
Composition from LabMaster, VCM column bottom
Flow rate of 20bar steam: 4.22 ton/h (62FI2216)

Assumptions

- Outlet temperature is calculated, assuming vapour fraction 1.0 for the stream leaving the reboiler
- The load is calculated from the condensation heat in the steam

Results from Aspen plus:

Q=2210 kW
T_{out} = 158.1°C

H207

Process data

P_{in}= 5 bar(g) (62PC2117)
T_{in}= 41.4°C (62TI2100)
T_{out} =17°C (62TI2051)

Assumptions

- The flow rate is calculated by adding 10.96 m^3/h (62FC2106) to VCM column and 19.45 ton/h (62FQ2133) to S207.
- The composition is unknown, and the composition in the old Aspen plus simulation is used

Results from Aspen plus:

Q=2880 kW

H210

Process data

P_{in}= 0.6 bar(g) in S208 (H62PI2270)

Assumptions

- Flow rate calculated by adding 0.3 m^3/h (62FC2443) back to tar column, and 0.661 m^3/h (62FI2044) to cooling column
- The inlet temperature is calculated assuming vapour fraction 1.0, and the outlet temperature is calculated assuming vapour fraction 0.
- The composition is unknown, and the composition in the old Aspen plus simulation is used. However, this gave an unreasonable low outlet temperature, and the composition had to be approximated with no HCl.

Results from Aspen plus:

Q=113 kW
T_{in}=92.4°C
T_{out}=52.1°C

H211

Process data

Hot side:
T_{in}= 40°C (62TC2036)
T_{out} = 13°C (Measured at pipe)

Cold side:
P= 9.1 bar(g) (62PC2069)
T_{in}= -32.3°C (62TI2042)
T_{out} =20.96°C (62TI2041)

Composition from LabMaster, HCl-column top

Assumptions

●Bad agreement between the hot and the cold side of the heat exchanger. The load is determined from the cold side, since more data is known.

●The flow rate of the hot side is calculated from information in the Oxi-unit. The flow rate is calculated by subtracting 320.9Nm^3/h (63FC3013) ,which comes from the HCl evaporator, from 6635.5 Nm^3/h (63FI3009) which goes into R301.

●The pressure of the cold stream was lowered to 9.8 bar(g) to obtain vapour fraction 1 in.

Results from Aspen plus:

Q=110 kW

H212

Process data

P_{in}=12 bar(g) after pump (PI1008)
T_{in}= 81.8°C before pump (62TI2096)
T_{out}=84°C (measured at pipe)
Flow rate=36.5 m^3/h before pump (62FC2102)
Flow rate of 10bar steam: 0.82 ton/h (62FI2115)

Assumptions

●Composition calculated in Aspen plus from the fact that it should be saturated liquid in the bottom of the HCl-column, and the mixture should only contain EDC and VCM

●The load calculated from the condensation heat in the steam

●Bad agreement between the hot and the cold side. The pressure was lowered to 8bar(g), and the outlet temperature was increased to 85 C to obtain the same heat load.

Results from Aspen plus:

Q=457 kW
T_{out}= 85 C

H213

Process data

P_{in}=23 bar(g) (P62PC2007)
T_{in}= 26.3°C (62TI2489)
T_{out}=118.2°C (62TC2414)
Flow rate=45.13 m^3/h (62TC2414)
Composition from LabMaster, Feed to R201

Assumptions

-

Results from Aspen plus:

Q=1980 kW

S208

Process data

P=0.6 bar(g) (H62PI2270)
T at the top of the tar boiler: 119C (62TI2250)
Flow rate of 10 bar steam: 0.32 ton/h (62FC2431)

Assumptions

●The inlet and outlet temperatures are taken from the energy mapping performed at site in 2006. The inlet temperature is set to 140 C, and the outlet temperature to 141 C.

●The load is calculated from the condensation heat in the steam

Results from Aspen plus:

Q=178 kW

OXI-unit

H301

Process data

P_{in}=9 bar(g) (63PC3008)
T_{out}=179.6°C (63TC3012)
Flow rate=6635.5Nm^3/h (63FI3009)

Assumptions

●The flow rate from the HCl –column is calculated from the flow rate after the H301, subtracting the flow rate from S210 of 320.9Nm^3/h (63FC3013)

●The inlet temperature calculated by adding the two inlet streams. The inlet stream from the HCl-column has a temperature of 21 C (62TI2041) and the stream from S210 has a temperature of 41 C (measured at pipe)
●Assumed to be pure HCl
Results from Aspen plus:
Q=342 kW
T_{in}= 22°C

H302

Process data

P_{out}=4.4 bar(g)	(63PI3165)
T_{in}= 217.2°C	(63TI3103)
T_{out}=93.9 °C	(63TI3113)

Assumptions
●The flow rate and composition calculated, please see Appendix 3.
● Since hydrochloric acid is formed in the reactors, the property method ENRTL-RK is used
Results from Aspen plus:
Q=2618 kW

H303AB

Process data

P_{in}=4.4 bar(g)	(63PI3165)
T_{in}=93.9°C	(63TI3113)

Assumptions
●The outlet temperature is assumed to be 30 C
● The flow rate and composition calculated, please see Appendix 3.
● Since hydrochloric acid is formed in the reactors, the property method ENRTL-RK is used
Results from Aspen plus:
Q=1641 kW

H304AB

Process data

T_{out}=136.1 C+ 137.9°C	(63TI3133+63TI3131)

Assumptions
●The outlet temperature is set to the middle value, 137°C
●The flow rate from S309 is simulated in Aspen plus
●The inlet temperature is calculated by adding the inlet streams, assuming the stream from S309 to have a temperature of 30°C, and the stream from HTC to have a temperature of 57.5°C (61TI1646)
●The inlet pressure is set to the same as the stream from HTC, since it has the lowest pressure of 9.3 bar(g) (61PI1644)
Results from Aspen plus:
Q=245 kW

H305

Process data

T_{in}= 151.5°C	(63TI3001)

Assumptions
●The flow rate and composition calculated, please see Appendix 3
●The outlet temperature is set to 40°C, based on the process description.
● The pressure is set to 4 bar(g), based on the inlet pressure to H306 of 3.5 bar(g) (PI1027) and assuming a pressure drop of 0.5 bar.
Results from Aspen plus:
Q=389 kW

H306

Process data

P_{in}=3.5 bar(g)	(PI1027)
T_{out}=-21.5°C	(63TI3162)

Assumptions
●S 306 is simulated to obtain the flow rate and composition. The flow rate and composition into S309 is calculated, please see Appendix 3.
● The inlet temperature is set to 40°C, based on the process description

Results from Aspen plus:
Q=240 kW

H307

Process data
P_{out}=9.7 bar(g) (63PI3139)
T_{in}= 3.4°C (measured at pipe)
T_{out}=69.6°C (63TI3141)
Flow rate out = 460.7 Nm^3/h (63FC3136)
Assumptions
- Pure chlorine

Results from Aspen plus:
Q=109 kW

H308

Process data
P_{in}=17.8 bar(g) (63PI1302)
T_{in}= -0.4°C (61TI1306)
T_{out}=119.9°C (63TC3002)
Flow rate out = 3780.2 Nm^3/h (63FI3005)
Assumptions
- Pure ethylene

Results from Aspen plus:
Q=237 kW

S210

Process data
P_{in}=9 bar(g) (62PI2247)
T_{in}= -20°C (measured at pipe)
T_{out}=41°C (measured at pipe)
Flow rate out = 320.9 Nm^3/h (63FC3013)
Assumptions
- Pure HCl

Results from Aspen plus:
Q=57 kW

EDC-cleaning unit

H104

Process data
P=0.16 bar(g)at top of column (C61PC1236)
T_{in}= 133.6°C (C61TC1210)
Composition from LabMaster , EDC tricolumn bottom
Flow rate of 10 bar steam: 132.6 kg/h (C61FC1208)
Assumptions
- Pressure increased to 0.8 bar(g) to get saturated liquid in.
- Vapour fraction 1 assumed out from reboiler
- Heat exchanger load calculated from condensation energy in steam

Results from Aspen plus:
Q=74 kW
T_{out}= 145.6°C

H105

Process data
P=2.4 bar(g) (PI1033)
T_{in}= 127.7°C (61TI1714)
Composition from LabMaster , Azeo-column bottom
Flow rate of 10 bar steam: 6833 kg/h (C61FI1053)
Assumptions
- Vapour fraction 1 assumed out from reboiler
- Heat exchanger load calculated from condensation energy in steam

Results from Aspen plus:

Q=3858 kW
T_{out}= 128.5°C

H106

Process data
P_{in}=2 bar(g) (PI1036)
T_{in}= 85.6°C (61TI1712)
T_{out}=30°C (measured at pipe)
Composition from LabMaster, Azeocolumn top

Assumptions
- Flow rate out calculated by adding 20.07 m^3/h (C61FC1092) and 394.4 l/h (C61FC1229)

Results from Aspen plus:
Q=2296 kW

H107

Process data
P=0.16 bar(g) (C61PC1236)
T_{in}= 87.9°C (61TI1721)
Flow rate out: 0.77 m^3/h (C61FC1128)
Composition from LabMaster, EDC tri-column top

Assumptions
- Assuming vapour fraction 0 out.

Results from Aspen plus:
Q=80 kW
T_{out}= 86.5°C

H108

Process data
T_{in}= 111°C (61TI1718)
Composition from LabMaster , EDC-column bottom
Flow rate of 10 bar steam: 10.42 ton/h (C61FI1096)

Assumptions
- Pressure set to obtain saturated liquid in to the reboiler
- Vapour fraction 1 assumed out from reboiler
- Heat exchanger load calculated from condensation energy in steam

Results from Aspen plus:
Q=5803 kW
T_{out}= 119.1°C

H109

Process data
P_{in}=0.2 bar(g) (PI1040)
T_{in}= 89.3°C (61TI1716)
T_{out}=36°C after pump after S103 (C61TI1117)
Composition from LabMaster, Azeocolumn top

Assumptions
- Flow rate out calculated by adding 21.2 m^3/h (C61FC1118) and 28 m^3/h (C61FI1116)

Results from Aspen plus:
Q=6634 kW
T_{out}=35.4°C

S104

Process data
T_{in}= 139.6°C at top of column (TI1725)
Flow rate of 10 bar steam: 139.9 kg/h (C61FC1123)

Assumptions
- Inlet temperature set to 140°C
- An outlet temperature of 141°C is guessed
- Heat exchanger load calculated from condensation energy in steam

Results from Aspen plus:
Q=78 kW

685-03

Process data

P=4 bar(g) at top of column (C68PC8056)

T_{in}= 63.8°C (68TI8004)

Composition from LabMaster , EtCl-column 1 bottom

Flow rate of 10 bar steam: 118.5kg/h (C68FC8011)

Assumptions

- The outlet temperature is set to 65°C
- Heat exchanger load calculated from condensation energy in steam

Results from Aspen plus:

Q=66 kW

685-04

Process data

P= 4 bar(g) (C68PC8056)

T_{in}= 62.9°C (68TI8001)

Assumptions

- Composition approximated with calculations, and the outlet temperature is obtained by assuming vapour fraction 0 out, giving an outlet temperature of 59.1°C
- The heat exchanger load is approximated to be slightly less than the load of the reboiler, since all outgoing streams from the EtCl-column1 have higher temperatures than the ingoing.

Results from Aspen plus:

Q=60 kW

685-05

Process data

P=3.5 bar(g) at top of column (C68PC8057)

T_{in}= 59.3°C (68TI8008)

Composition from LabMaster , EtCl-column 2 bottom

Flow rate of 10 bar steam: 33.9 kg/h (C68FC8030)

Assumptions

- The inlet temperature is lowered to 58.8 °C to get saturated liquid in to the reboiler
- The outlet vapour fraction is assumed to be 1.0
- Heat exchanger load calculated from condensation energy in steam

Results from Aspen plus:

Q=19 kW

T_{out}= 58.9°C

685-04

Process data

P=3.5 bar(g) (C68PC8057)

T_{in}= 52.9°C (68TI8005)

Assumptions

- Composition approximated with calculations, and the outlet temperature is obtained by assuming vapour fraction 0 out, giving an outlet temperature of 32.8°C
- The heat exchanger load is approximated to be the same as the load of the reboiler

Results from Aspen plus:

Q=19 kW

Appendix 5 – Properties for the fuel gas

Composition of the purchased combustion gas

Average value February 2011	Composition [Mol %]	LHV [MJ/kg]	CO2 Emission factor [ton/TJ fuel]
H2	46.60		
Methane	49.17		
Ethane	1.64		
Ethylene	0.57		
Propane	1.82		
Propene	0.01		
C4	0.07		
Pentanes	0.01		
C6	0		
N2	0.06		
CO2	0		
CO	0.05		
Mixture		56.441	44.275

Appendix 6 – Heat transfer coefficient

The overall heat transfer coefficient U can be estimated with:

$$\frac{1}{U} = \frac{1}{h_1} + \frac{1}{h_2}$$

	h value: [W/m^2°C] *
Condensating organic vapours	800-1700
Boiling organics	1000-1400
Heavy organics	550-870
River or sea water	900-1200

* (Sinnot & Towler, 2009) pp 821

The h-values used for each stream are found in the following table:

		h value: [W/m2C]	*Approximated with:*
Stream in:	H202	1250	Condensation organic vapours,
	H205	1200	Boiling organics
	H213	850	Heavy organics. Chosen in the upper part, not heavy
	H212	850	Heavy organics. Chosen in the upper part, not heavy
	H302	1250	Condensation organic vapours
	H108	1200	Boiling organics
	H105	1200	Boiling organics
	Water	1000	Water

Appendix 7 – Current load of the heat exchangers

In this appendix, the heat loads for the heat exchangers in the current VCM plant are found. The load is divided into two or more temperature intervals in the case of evaporation or condensation, where the difference in capacity rate could be high.

Present situation. Results from Aspen simulations, put into pinch analysis

Unit	At:	Description	Tstart [C]	Ttarget [C]	Q [kW]	Utility
685-03	EDC	Reboiler EtCl-column 1	63.8	65	66	10 bar steam
685-04	EDC	Condenser EtCl column1	62.9	59.1	60	Cooling water
685-05	EDC	Reboiler EtCl-column 2	58.8	59	19	10 bar steam
685-06	EDC	Condenser EtCl-column 2	52.9	32.8	19	Cooling water
H104	EDC	Reboiler Tri-column EDC	133.6	145.6	74	10 bar steam
H105	EDC	Reboiler Azeo column	127.7	128.5	3858	10 bar steam
H106	EDC	Condenser Azeo column	85.6	30	2296	Cooling water
H106a	*EDC*	*Condenser Azeo column, cooling*	*85.6*	*80.7*	*26*	
H106b	*EDC*	*Condenser Azeo column, condensation*	*80.7*	*60.6*	*1994*	
H106c	*EDC*	*Condenser Azeo column, cooling*	*60.6*	*30*	*276*	
H107	EDC	Tar condenser (Condenser Tri-column EDC)	87.9	86.5	80	Cooling water
H108	EDC	Reboiler EDC-column	111	119.1	5803	10 bar steam
H108a	*EDC*	*Reboiler EDC column , cond1*	*111*	*113.75*	*3397*	
H108b	*EDC*	*Reboiler EDC column , cond2*	*113.75*	*119.1*	*2406*	
H109	EDC	Condenser EDC-column	89.3	35	6634	Cooling water
H109a	*EDC*	*Condenser EDC-column, condensation*	*89.3*	*89.2*	*5413,8*	
H109b	*EDC*	*Condenser EDC-column, cooling*	*89.2*	*35*	*1220*	
S104	EDC	EDC tar boiler	140	141	78	10 bar steam
H153	HTC	HTC-condenser	82.4	74.4	9482	Cooling water
H153a	*HTC*	*HTC-condenser, condensation part 1*	*82.4*	*81.5*	*8404*	
H153b	*HTC*	*HTC-condenser, condensation part 2*	*81.5*	*74.4*	*1078*	
H154	HTC	EDC-condenser	73.5	15.8	150	Cooling water
H154a	*HTC*	*EDC-condenser, condensation part 1*	*73.5*	*57.5*	*98,5*	
H154b	*HTC*	*EDC-condenser, condensation part 2*	*57.5*	*15.8*	*51,5*	
H155	HTC	Remaining gas condenser	15.8	-16.5	5	Propene
H158AB	HTC	Reboiler Tri-column HTC	118	123.5	170	10 bar steam
H159	HTC	Condenser Tri-column HTC	87.1	13.9	157	Cooling water
H159a	*HTC*	*Condenser Tri-column HTC, condensation*	*87.1*	*86.4*	*121*	
H159b	*HTC*	*Condenser Tri-column HTC, cooling*	*86.4*	*13.9*	*36*	
H160	HTC	Pure EDC-cooler	89.2	26.7	364	Cooling water
H166	HTC	Cooler	57.5	34	33	Cooling water
H301	OXI	HCl heater	22	179.6	342	10 bar steam
H302	OXI	Precooler	217.2	93.9	2618	Cooling water
H302a	*OXI*	*Precooler, condensation part 1*	*217.2*	*150*	*1082*	
H302b	*OXI*	*Precooler, condensation part 2*	*150*	*93.9*	*1536*	
H303AB	OXI	Condenser after precooler H302 after R303	93.9	30	1641	Cooling water

H303ABa	*OXI*	*Condenser after precooler, condensation part 1*	*93.9*	*65*	*1005,7*	
H303ABb	*OXI*	*Condenser after precooler, condensation part 2*	*65*	*30*	*635,2*	
H304AB	OXI	Heater before R304	30.7	137	245	10 bar steam
H305	OXI	Condenser after R304	151.5	40	389	Cooling water
H305a	*OXI*	*Condenser after R304, cooling*	*151.5*	*57.5*	*236,3*	
H305b	*OXI*	*Condenser after R304, condensation*	*57.5*	*40*	*152,8*	
H306	OXI	Remainder gas condenser	40	-21.5	240	Propene
H307	OXI	Chlorine evaporator	3.4	69.6	109	1 bar steam
H307a	*OXI*	*Chlorine evaporator, heating 1*	*3.4*	*37.1*	*14,6*	
H307b	*OXI*	*Chlorine evaporator, evaporation*	*37.1*	*37.2*	*88,6*	
H307c	*OXI*	*Chlorine evaporator, heating 2*	*37.2*	*69.6*	*5,8*	
H308	OXI	Ethylene preheater	-0.4	119.9	237	10 bar steam
H321	OXI	Intermediate air cooler	160	25	307	Cooling water
S210	OXI	HCl- evaporator	-31.9	41	57,2	10 bar steam
S210a	*OXI*	*HCl- evaporator, evaporation*	*-31.9*	*-31.7*	*49,6*	
S210b	*OXI*	*HCl- evaporator, heating*	*-31.7*	*41*	*7,6*	
H121-cold	VCM	From VCM-column/To EDC column	41	98	300	
H121-hot	VCM	From VCM-column/To EDC column	157.7	118	300	
H122	VCM	Cooler before R102	118	60	389	Cooling water
H201	VCM	EDC-evaporator	193.1	208.61	3442	20 bar steam
H201a	*VCM*	*EDC-evaporator, heating*	*193.1*	*207.5*	*385*	
H201b	*VCM*	*EDC-evaporator, evaporation*	*207.5*	*207.8*	*3042,7*	
H201c	*VCM*	*EDC-evaporator, heating*	*207.8*	*208.61*	*14,6*	
H202A	VCM	Condenser cooling column	132.8	66	7007,5	Cooling water
H202Aa	*VCM*	*Condenser cooling column, cond. Part1*	*132.8*	*112.4*	*2905,5*	
H202Ab	*VCM*	*Condenser cooling column, cond. Part2*	*112.4*	*66*	*4102*	
H202B	VCM	Condenser 2 cooling column	66	40	2098	Cooling water
H204	VCM	HCl condenser	-31.7	-32.3	1640	Propene
H205	VCM	Reboiler HCl-column	86.5	144.7	1559	10 bar steam
H206	VCM	Reboiler VCM column	157.7	158.1	2210	20 bar steam
H207	VCM	Condenser VCM column	41.4	17	2880	Cooling water
H207a	*VCM*	*Condenser VCM column, cooling*	*41.4*	*39.8*	*12,33*	
H207b	*VCM*	*Condenser VCM column, condensation*	*39.8*	*39.7*	*2501,3*	
H207c	*VCM*	*Condenser VCM column, cooling*	*39.7*	*17*	*365,8*	
H210	VCM	VCM tar condenser	92.4	52.1	113	Cooling water
H210a	*VCM*	*VCM tar condenser, cond part 1*	*92.4*	*85*	*59,3*	
H210b	*VCM*	*VCM tar condenser, cond part 2*	*85*	*52.1*	*53,9*	
H211-cold	VCM	HCl preheater/cooler in HCl column	-32.3	21	110	
H211-hot	VCM	HCl preheater/cooler in HCl column	40	13	110	
H212	VCM	Preheater ingoing stream to VCM-column	81.8	85	457	10 bar steam
H213	VCM	EDC-preheater	26.3	118.2	1980	1 bar steam
S208	VCM	VCM tar boiler	140	141	178	10 bar steam

Appendix 8 – Pinch analysis of the existing VCM plant

The existing plant has currently a heat demand of 20.9 MW and a cooling demand of 38.6 MW. 10.3 MW of steam is produced in the process. For the detailed result of the heat exchangers on which the pinch analysis is based, please see Appendix 7.

For the pinch analysis, a minimum temperature difference in the heat exchangers (ΔT_{min}) of 10 °C was chosen.

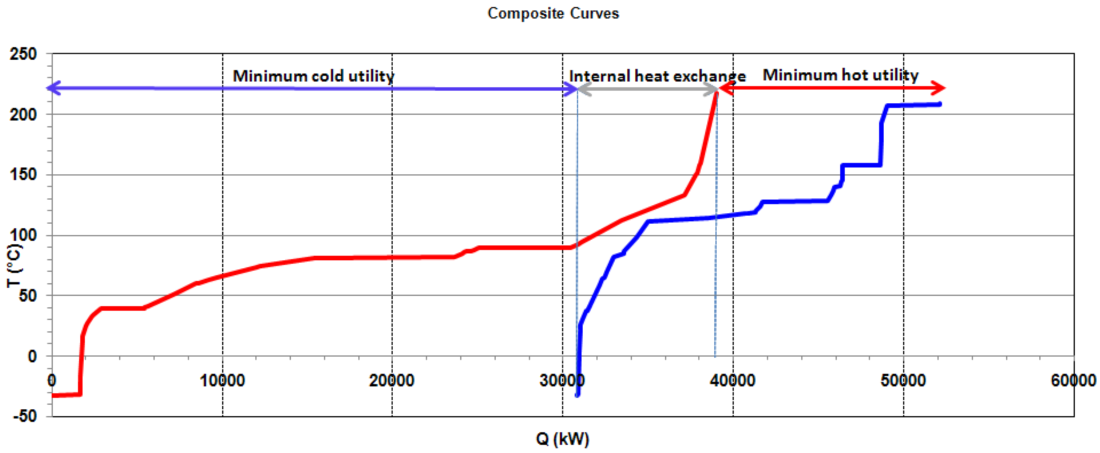


Figure 9.1 Composite curves existing plant

From the composite curves in Figure 9.1, the potential for internal heat exchange of the plant can be determined to 8.2 MW, of which 0.4 MW is utilized today. The pinch temperature for the system is 116°C (interval temperature). This gives that below 111°C, there should be no need for external heating of cold streams, and hot streams above 121°C should not be cooled with cooling utility.

Table 9.1 Current demand and theoretical minimum demand existing process

	Demand existing process [MW]	*Theoretical minimum demand [MW]*	*Potential for savings [MW]*
Hot utility	20.9	13.1	7.8
Cold utility	38.6	30.8	7.8

The demands for hot and cold utility are shown in Table 9.1. The theoretical potential for steam saving is 37% of the current steam demand.

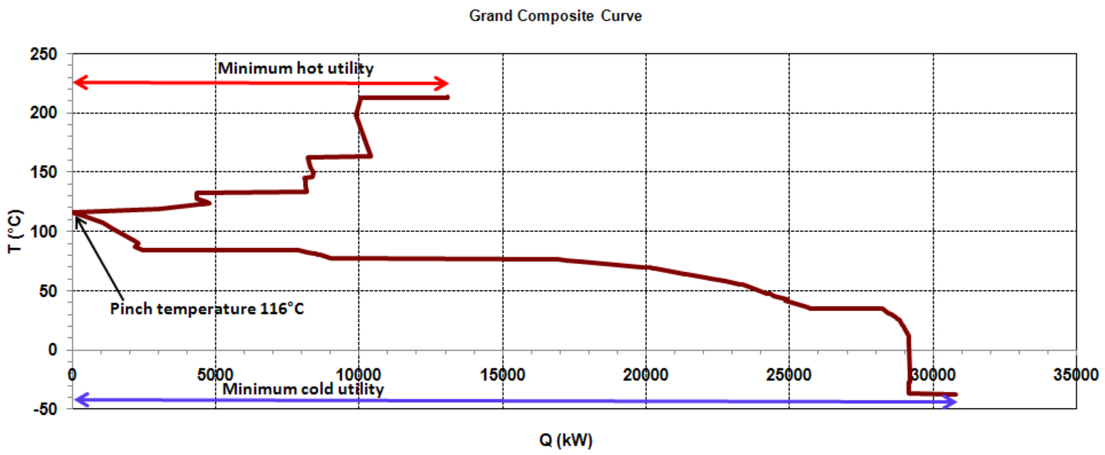


Figure 9.2 Grand composite curve existing plant

As seen in Figure 9.1 there is a large surplus of heat below the pinch temperature.

Appendix 9 – Future load of the heat exchangers

In this appendix, the heat loads for the heat exchangers in the future VCM plant are found. The load is divided into two or more temperature intervals in the case of evaporation or condensation, where the difference in capacity rate could be high.

Future situation. Results from Aspen simulations, put into pinch analysis

Unit	At:	Description	Tstart [C]	Ttarget [C]	Q [kW]	Utility
685-03	EDC	Reboiler EtCl-column 1	63.8	65	66	10 bar steam
685-04	EDC	Condenser EtCl column1	62.9	59.1	60	Cooling water
685-05	EDC	Reboiler EtCl-column 2	58.8	59	19	10 bar steam
685-06	EDC	Condenser EtCl-column 2	52.9	32.8	19	Cooling water
H104	EDC	Reboiler Tri-column EDC	133.6	145.6	74	10 bar steam
H105	EDC	Reboiler Azeo column	127.7	128.5	3858	10 bar steam
H106	EDC	Condenser Azeo column	85.6	30	2296	Cooling water
H106a	*EDC*	*Condenser Azeo column, cooling*	*85.6*	*80.7*	*26*	
H106b	*EDC*	*Condenser Azeo column, condensation*	*80.7*	*60.6*	*1994*	
H106c	*EDC*	*Condenser Azeo column, cooling*	*60.6*	*30*	*276*	
H107	EDC	Tar condenser (Condenser Tri-column)	87.9	86.5	80	Cooling water
H108	EDC	Reboiler EDC-column	111	119.1	5803	10 bar steam
H108a	*EDC*	*Reboiler EDC column , cond1*	*111*	*113.75*	*3397*	
H108b	*EDC*	*Reboiler EDC column , cond2*	*113.75*	*119.1*	*2406*	
H109	EDC	Condenser EDC-column	89.3	35	6634	Cooling water
H109a	*EDC*	*Condenser EDC-column, condensation*	*89.3*	*89.2*	*5413,8*	
H109b	*EDC*	*Condenser EDC-column, cooling*	*89.2*	*35*	*1220*	
S104	EDC	EDC tar boiler	140	141	78	10 bar steam
H153	HTC	HTC-condenser	82.4	74.4	13065	Cooling water
H153a	*HTC*	*HTC-condenser, condensation part 1*	*82.4*	*81.5*	*11105*	
H153b	*HTC*	*HTC-condenser, condensation part 2*	*81.5*	*74.4*	*1960*	
H154	HTC	EDC-condenser	73.5	15.8	300	Cooling water
H154a	*HTC*	*EDC-condenser, condensation part 1*	*73.5*	*57.5*	*197*	
H154b	*HTC*	*EDC-condenser, condensation part 2*	*57.5*	*15.8*	*103*	
H155	HTC	Remaining gas condenser	15.8	-16.5	10	Propene
H166	HTC	Cooler	57.5	34	65	Cooling water
H160	HTC	Pure EDC-cooler	89.2	26.7	1011	Cooling water
H159	HTC	Condenser Tri-column HTC	87.1	13.9	104	Cooling water
H159a	*HTC*	*Condenser Tri-column HTC, cond.*	*87.1*	*86.4*	*81*	
H159b	*HTC*	*Condenser Tri-column HTC, cooling*	*86.4*	*13.9*	*23*	
H158AB	HTC	Reboiler Tri-column HTC	118	123.5	114	10 bar steam
H165	HTC	Cooler R151	115	40	1600	Cooling water
H164	HTC	Cooler HTC-column	100	40	1200	Cooling water
Hnew	HTC	New cooler HTC	100	40	3099	Cooling water
H301	OXI	HCl heater	22	179.6	342	10 bar steam
H302	OXI	Precooler	217.2	93.9	2618	Cooling water
H302a	*OXI*	*Precooler, condensation part 1*	*217.2*	*150*	*1082*	

H302b	*OXI*	*Precooler, condensation part 2*	*150*	*93.9*	*1536*	
H303AB	OXI	Condenser after precooler H302 after R303	93.9	30	1641	Cooling water
H303ABa	*OXI*	*Condenser after precooler, cond. part 1*	*93.9*	*65*	*1005,7*	
H303ABb	*OXI*	*Condenser after precooler, cond. part 2*	*65*	*30*	*635,2*	
H304AB	OXI	Heater before R304	30.7	137	245	10 bar steam
H305	OXI	Condenser after R304	151.5	40	389	Cooling water
H305a	*OXI*	*Condenser after R304, cooling*	*151.5*	*57.5*	*236,3*	
H305b	*OXI*	*Condenser after R304, condensation*	*57.5*	*40*	*152,8*	
H306	OXI	Remainder gas condenser	40	-21.5	240	Propene
H307	OXI	Chlorine evaporator	3.4	69.6	109	1 bar steam
H307a	*OXI*	*Chlorine evaporator, heating 1*	*3.4*	*37.1*	*14,6*	
H307b	*OXI*	*Chlorine evaporator, evaporation*	*37.1*	*37.2*	*88,6*	
H307c	*OXI*	*Chlorine evaporator, heating 2*	*37.2*	*69.6*	*5,8*	
H308	OXI	Ethylene preheater	-0.4	119.9	237	10 bar steam
H321	OXI	Intermediate air cooler	160	25	307	Cooling water
S210	OXI	HCl- evaporator	-31.9	41	57,2	10 bar steam
S210a	*OXI*	*HCl- evaporator, evaporation*	*-31.9*	*-31.7*	*49,6*	
S210b	*OXI*	*HCl- evaporator, heating*	*-31.7*	*41*	*7,6*	
H121-cold	VCM	From VCM-column/To EDC column	41	98	300	
H121-hot	VCM	From VCM-column/To EDC column	157.7	118	300	
H122	VCM	Cooler before R102	118	60	389	Cooling water
H201	VCM	EDC-evaporator	193.1	208.61	3442	20 bar steam
H201a	*VCM*	*EDC-evaporator, heating*	*193.1*	*207.5*	*385*	
H201b	*VCM*	*EDC-evaporator, evaporation*	*207.5*	*207.8*	*3042,7*	
H201c	*VCM*	*EDC-evaporator, heating*	*207.8*	*208.61*	*14,6*	
H202A	VCM	Condenser cooling column	132.8	66	7007,5	Cooling water
H202Aa	*VCM*	*Condenser cooling column, cond. Part1*	*132.8*	*112.4*	*2905,5*	
H202Ab	*VCM*	*Condenser cooling column, cond. Part2*	*112.4*	*66*	*4102*	
H202B	VCM	Condenser 2 cooling column	66	40	2098	Cooling water
H204	VCM	HCl condenser	-31.7	-32.3	1640	Propene
H205	VCM	Reboiler HCl-column	86.5	144.7	1559	10 bar steam
H206	VCM	Reboiler VCM column	157.7	158.1	2210	20 bar steam
H207	VCM	Condenser VCM column	41.4	17	2880	Cooling water
H207a	*VCM*	*Condenser VCM column, cooling*	*41.4*	*39.8*	*12,33*	
H207b	*VCM*	*Condenser VCM column, condensation*	*39.8*	*39.7*	*2501,3*	
H207c	*VCM*	*Condenser VCM column, cooling*	*39.7*	*17*	*365,8*	
H210	VCM	VCM tar condenser	92.4	52.1	113	Cooling water
H210a	*VCM*	*VCM tar condenser, cond part 1*	*92.4*	*85*	*59,3*	
H210b	*VCM*	*VCM tar condenser, cond part 2*	*85*	*52.1*	*53,9*	
H211-cold	VCM	HCl preheater/cooler in HCl column	-32.3	21	110	
H211-hot	VCM	HCl preheater/cooler in HCl column	40	13	110	
H212	VCM	Preheater ingoing stream to VCM-column	81.8	85	457	10 bar steam
H213	VCM	EDC-preheater	26.3	118.2	1980	1 bar steam
S208	VCM	VCM tar boiler	140	141	178	10 bar steam

Appendix 10 – Heat in the Oxi-reactors and flue gases

Generated in Oxi-reactors and flue gases

	At:	Unit:	Tstart [C]	Ttarget [C]	Q [kW]	Utility produced
R301	OXI	Reactor R301 Steam prod	300	290	2679	20 bar steam
R302	OXI	Reactor R302 Steam prod	290	260	3336	20 bar steam
R303	OXI	Reactor R303 Steam prod	260	217,2	1905	20 bar steam
R304	OXI	Reactor R304 Steam prod	160	151,5	921	1 bar steam

	At:	Unit:	Tstart [C]	Ttarget [C]	Q [kW]	Utility produced
Preheater-cold	VCM	Preheater, stream side	118.2	193.1	1991	
Preheater-hot	VCM	Preheater, flue gas side	550	486.8	1991	

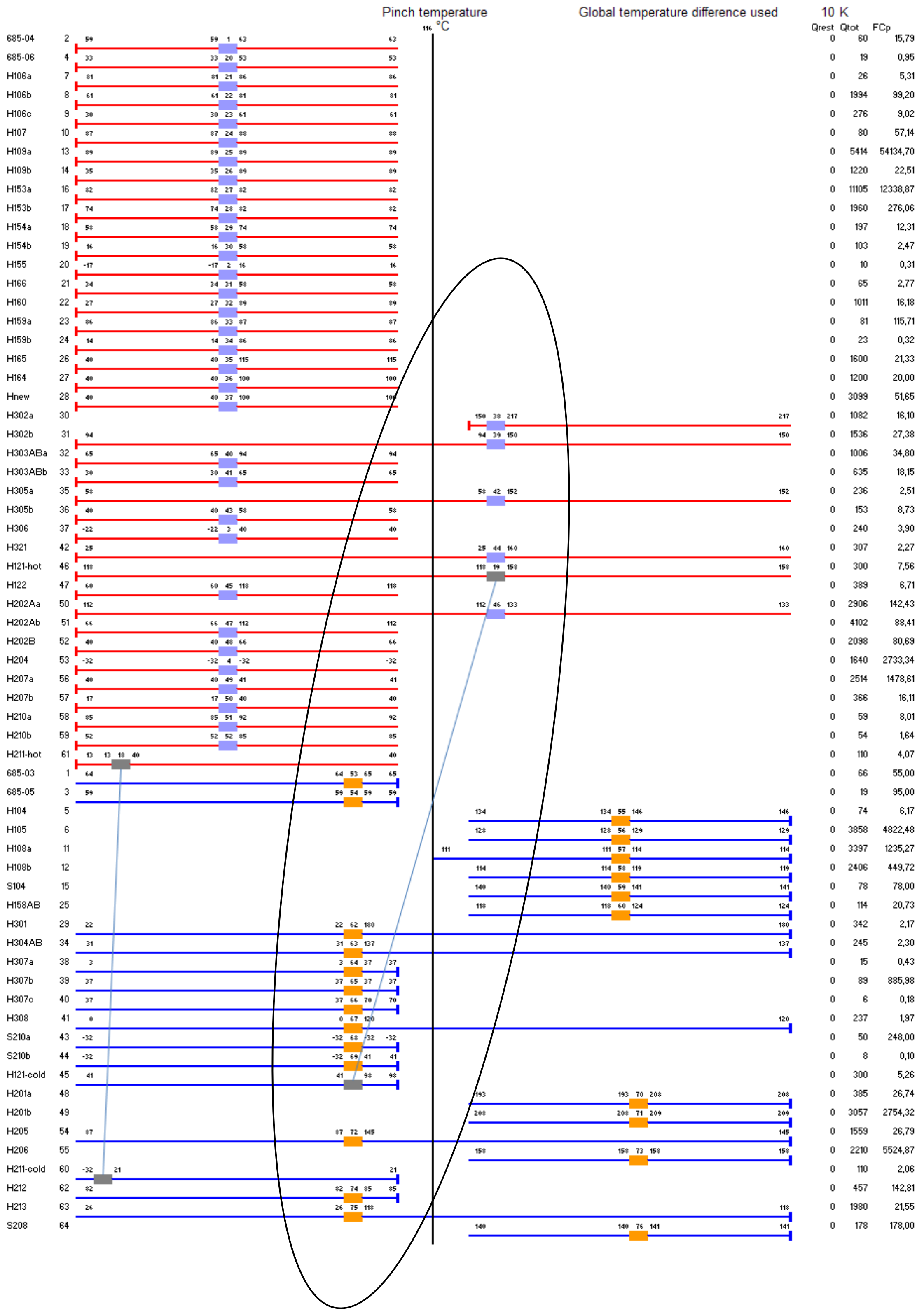
	At:	Unit:	Tstart [C]	Ttarget [C]	Q [kW]	Utility produced
Losses flue gases	VCM	Losses before flue gas boiler	486.8	445.6	303	
Flue gas boiler	VCM	Steam production in flue gas boiler	445.6	231.3	1493	10 bar steam
Excess energy flue gases	VCM	Excess energy in flue gases leaving the flue gas boiler	231.3	121	763	
Energy flue gases	VCM	Total energy in flue gases leaving preheater, used for pinch analysis	486.8	121	2594	

Appendix 11 – Pinch violations in future VCM plant

Pinch violations in future plant

	At:	Heat through pinch [kW]	Cooling above pinch [kW]	Heating below pinch [kW]
H121	VCM	277		
H302	OXI		1876	
H305	OXI		77	
H321	OXI		89	
H202A	VCM		1680	
685-03	EDC-cleaning			66
685-05	EDC-cleaning			19
H301	OXI			193
H304AB	OXI			185
H307	OXI			109
H308	OXI			219
S210	OXI			57
H205	VCM			656
H212	VCM			457
H213	VCM			1825
Total:		277	3722	3787

A stream representation showing the pinch violations is found on next page:



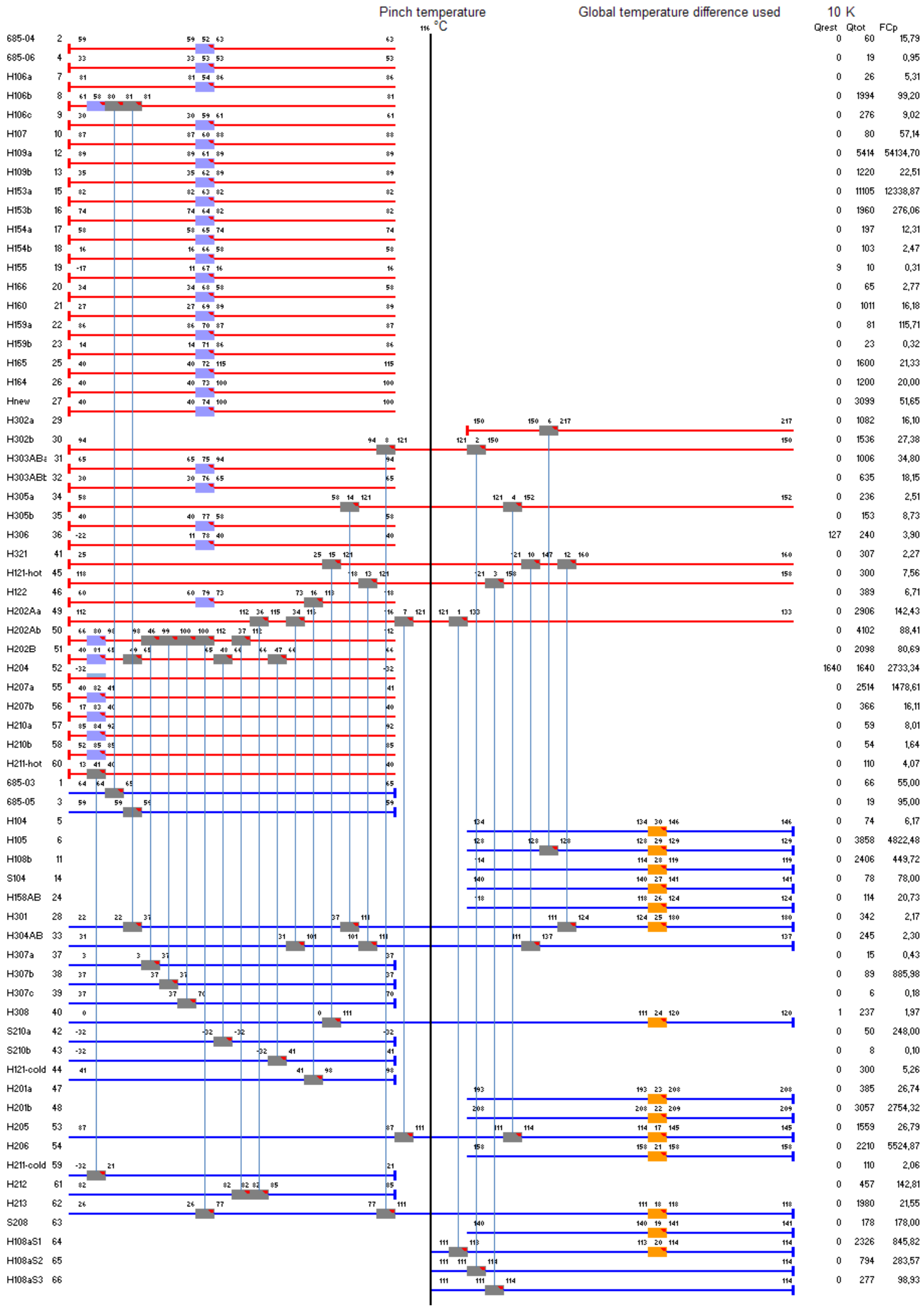
		Qrest	Qtot	FCp
685-04	2	0	60	15,79
685-06	4	0	19	0,95
H106a	7	0	26	5,31
H106b	8	0	1994	99,20
H106c	9	0	276	9,02
H107	10	0	80	57,14
H109a	13	0	5414	54134,70
H109b	14	0	1220	22,51
H153a	16	0	11105	12338,87
H153b	17	0	1960	276,06
H154a	18	0	197	12,31
H154b	19	0	103	2,47
H155	20	0	10	0,31
H166	21	0	65	2,77
H160	22	0	1011	16,18
H159a	23	0	81	115,71
H159b	24	0	23	0,32
H165	26	0	1600	21,33
H164	27	0	1200	20,00
Hnew	28	0	3099	51,65
H302a	30	0	1082	16,10
H302b	31	0	1536	27,38
H303ABa	32	0	1006	34,80
H303ABb	33	0	635	18,15
H305a	35	0	236	2,51
H305b	36	0	153	8,73
H306	37	0	240	3,90
H321	42	0	307	2,27
H121-hot	46	0	300	7,56
H122	47	0	389	6,71
H202Aa	50	0	2906	142,43
H202Ab	51	0	4102	88,41
H202B	52	0	2098	80,69
H204	53	0	1640	2733,34
H207a	56	0	2514	1478,61
H207b	57	0	366	16,11
H210a	58	0	59	8,01
H210b	59	0	54	1,64
H211-hot	61	0	110	4,07
685-03	1	0	66	55,00
685-05	3	0	19	95,00
H104	5	0	74	6,17
H105	6	0	3858	4822,48
H108a	11	0	3397	1235,27
H108b	12	0	2406	449,72
S104	15	0	78	78,00
H158AB	25	0	114	20,73
H301	29	0	342	2,17
H304AB	34	0	245	2,30
H307a	38	0	15	0,43
H307b	39	0	89	885,98
H307c	40	0	6	0,18
H308	41	0	237	1,97
S210a	43	0	50	248,00
S210b	44	0	8	0,10
H121-cold	45	0	300	5,26
H201a	48	0	385	26,74
H201b	49	0	3057	2754,32
H205	54	0	1559	26,79
H206	55	0	2210	5524,87
H211-cold	60	0	110	2,06
H212	62	0	457	142,81
H213	63	0	1980	21,55
S208	64	0	178	178,00

Pinch violations

Future production with existing net and additional coolers

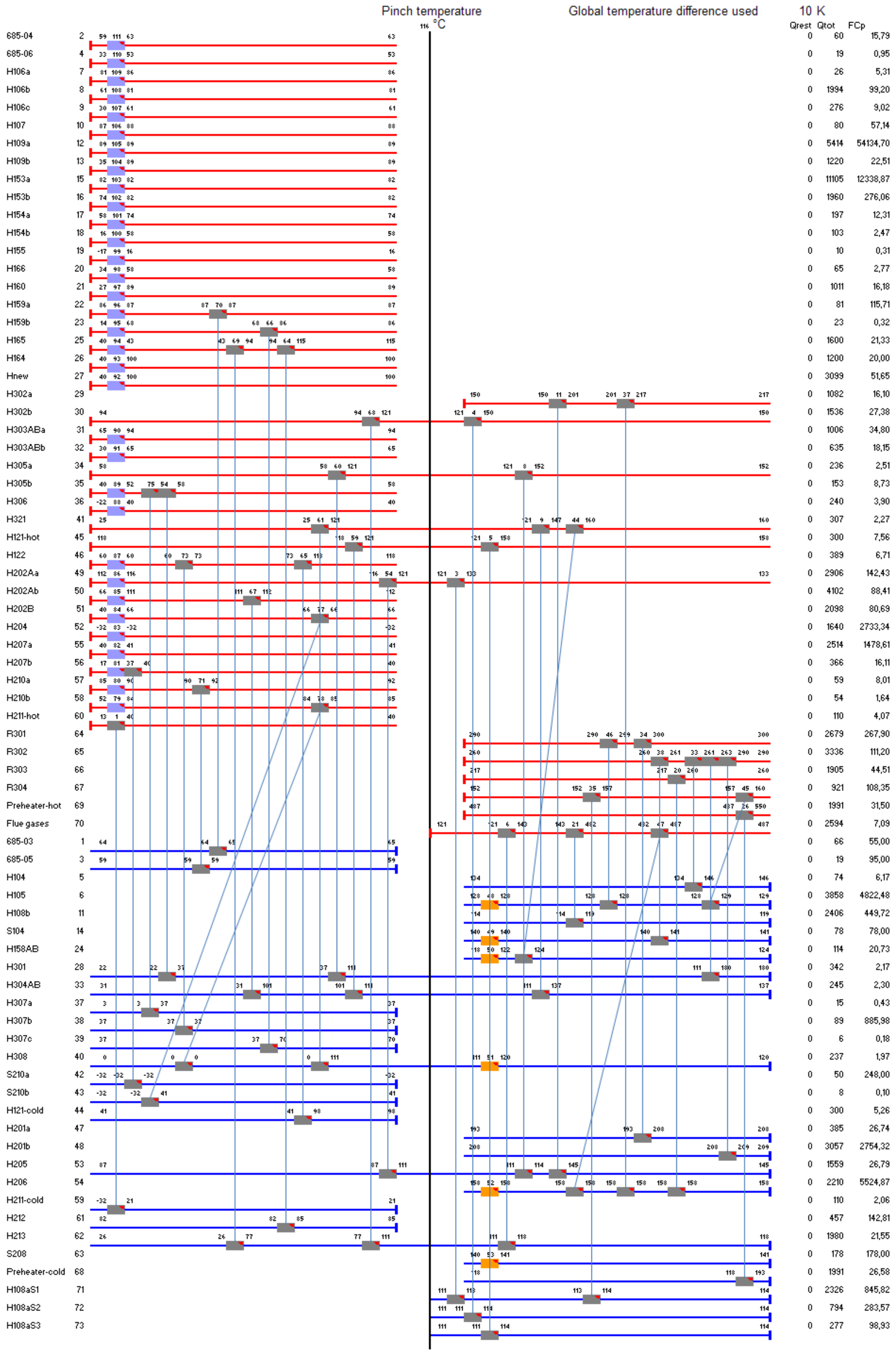
Appendix 12 – MER network future VCM plant

A stream representation of a Maximum Energy Recovery (MER) network for the future VCM plant is found below. Note that one stream, H108, is splitted into three.



MER network future VCM plant

A stream representation of a Maximum Energy Recovery (MER) network for the future VCM plant, including heat from flue gases, and steam produced in the Oxi-reactors is found below. Note that also here is one stream, H108, splitted into three.



		Qrest	Qtot	FCp
685-04	2	0	60	15,79
685-06	4	0	19	0,95
H106a	7	0	26	5,31
H106b	8	0	1994	99,20
H106c	9	0	276	9,02
H107	10	0	80	57,14
H109a	12	0	5414	54134,70
H109b	13	0	1220	22,51
H153a	15	0	11105	12338,87
H153b	16	0	1960	276,06
H154a	17	0	197	12,31
H154b	18	0	103	2,47
H155	19	0	10	0,31
H166	20	0	65	2,77
H160	21	0	1011	16,18
H159a	22	0	81	115,71
H159b	23	0	23	0,32
H165	25	0	1600	21,33
H164	26	0	1200	20,00
Hnew	27	0	3099	51,65
H302a	29	0	1082	16,10
H302b	30	0	1536	27,38
H303ABa	31	0	1006	34,80
H303ABb	32	0	635	18,15
H305a	34	0	236	2,51
H305b	35	0	153	8,73
H306	36	0	240	3,90
H321	41	0	307	2,27
H121-hot	45	0	300	7,56
H122	46	0	389	6,71
H202Aa	49	0	2906	142,43
H202Ab	50	0	4102	88,41
H202B	51	0	2098	80,69
H204	52	0	1640	2733,34
H207a	55	0	2514	1478,61
H207b	56	0	366	16,11
H210a	57	0	59	8,01
H210b	58	0	54	1,64
H211-hot	60	0	110	4,07
R301	64	0	2679	267,90
R302	65	0	3336	111,20
R303	66	0	1905	44,51
R304	67	0	921	108,35
Preheater-hot	69	0	1991	31,50
Flue gases	70	0	2594	7,09
685-03	1	0	66	55,00
685-05	3	0	19	95,00
H104	5	0	74	6,17
H105	6	0	3858	4822,48
H108b	11	0	2406	449,72
S104	14	0	78	78,00
H158AB	24	0	114	20,73
H301	28	0	342	2,17
H304AB	33	0	245	2,30
H307a	37	0	15	0,43
H307b	38	0	89	885,98
H307c	39	0	6	0,18
H308	40	0	237	1,97
S210a	42	0	50	248,00
S210b	43	0	8	0,10
H121-cold	44	0	300	5,26
H201a	47	0	385	26,74
H201b	48	0	3057	2754,32
H205	53	0	1559	26,79
H206	54	0	2210	5524,87
H211-cold	59	0	110	2,06
H212	61	0	457	142,81
H213	62	0	1980	21,55
S208	63	0	178	178,00
Preheater-cold	68	0	1991	26,58
H108aS1	71	0	2326	845,82
H108aS2	72	0	794	283,57
H108aS3	73	0	277	98,93

MER network future VCM plant, including heat from flue gases and Oxi-reactors

Appendix 13 – Economic evaluation of the retrofits

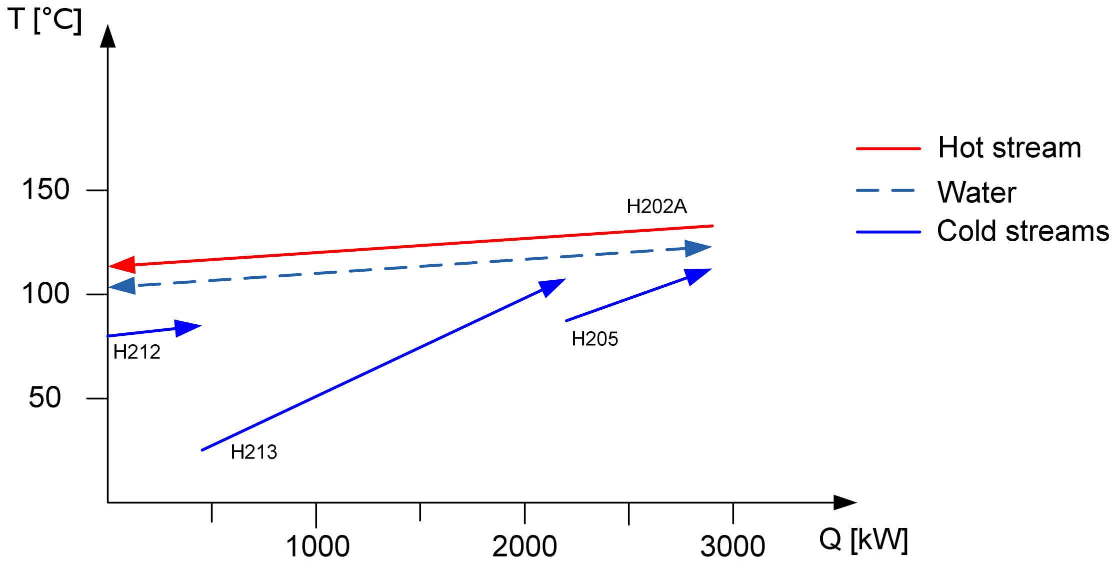
The calculations are made following the calculation procedure described in Chapter 3.5

	Q [kW]	dTlm [C]	U [W/ m2K]	A [m2]	Cost of equipment (Ce) Carbon Steel [$] Jan 2007	Ce, Carbon Steel [$] Nov 2010	Material cost factor fm	ISBL cost C [$]	Total fixed capital cost CFC [$]	Total fixed capital cost CFC [MSEK]	Annual steam savings [GJ]	Approx. annual savings [MSEK]:	Payback period [years]
Retrofit 1													
H202 - H205	972	21	612	75	32117	35078	1.65	153293	278993	2	29477	3.31	0.58
H202-H213	1933	35	506	108	36696	40080	1.65	175151	318774	2	58621	6.58	0,33
H202-H212	457	26	506	35	27251	29764	1.65	130068	236724	2	13859	1.56	1,05
								Sum subsystem 1		6	101957	11.45	0.50
H302-H108	1876	40	612	76	32357	35340	1.7	157618	286865	2	56892	6.39	0,31
								Sum subsystem 2		2	56892	6.39	0.31
								Sum retrofit 1:		8	158850	17.84	0.43
Retrofit 2													
H202-H108	1680	15	612	185	48187	52631	1.65	229996	418593	3	50948	5.72	0,51
H202 - H205	656	18	612	60	30265	33056	1.65	144453	262904	2	19894	2.23	0,81
H202-H213	1717	32	506	107	36543	39913	1.65	174419	317443	2	52070	5.85	0,38
H202-H212	457	13	506	70	31482	34385	1.65	150263	273479	2	13859	1.56	1,22
H302-H108	1839	40	612	75	32224	35195	1.7	156971	285686	2	55770	6.26	0,32
								Sum retrofit 2:		11	192542	21.63	0.50
Retrofit 3													
H202-H108	1680	15	612	185	48187	52631	1.65	229996	418593	3	50948	5.72	0,51
H202-H213	1825	34	506	107	36553	39924	1.65	174468	317531	2	55346	6.22	0,35
H202-H212	457	19	506	46	28600	31237	1.65	136505	248438	2	13859	1.56	1,10
H302-H105	973	54	612	30	26692	29153	1.7	130024	236644	2	29508	3.31	0,49
H302-H205	1559	11	612	232	55742	60882	1.7	271533	494191	3	47279	5.31	0,64
								Sum retrofit 3:		12	196940	22,12	0.54

Appendix 14 – Hot water calculations

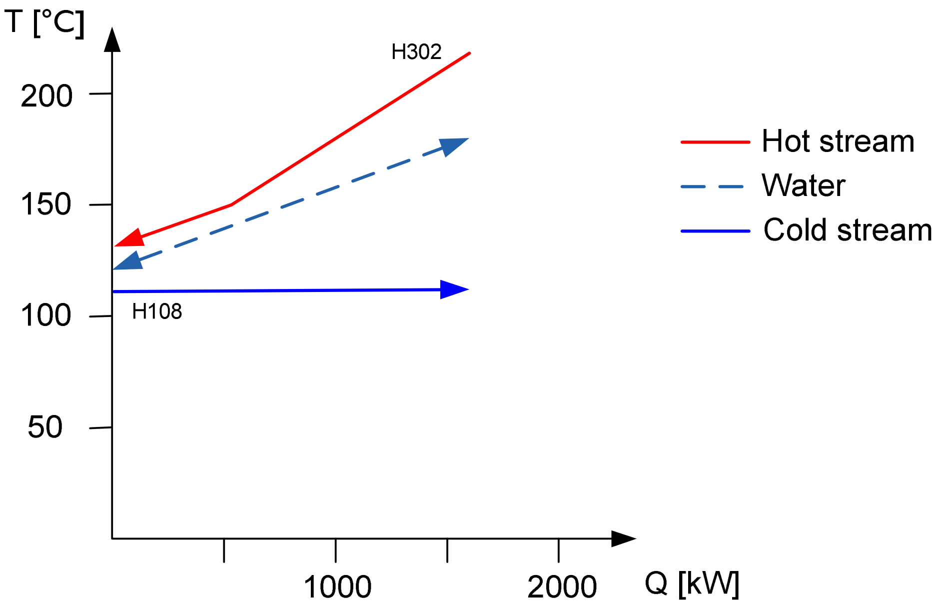
Both hot streams are condensing streams and the capacity rate changes. The flow of water is set to keep minimum temperature difference. How the flow rate of water was set for subsystem 1 was explained in Chapter 4.5.1, but is explained here as well.

Retrofit 1 – Subsystem 1

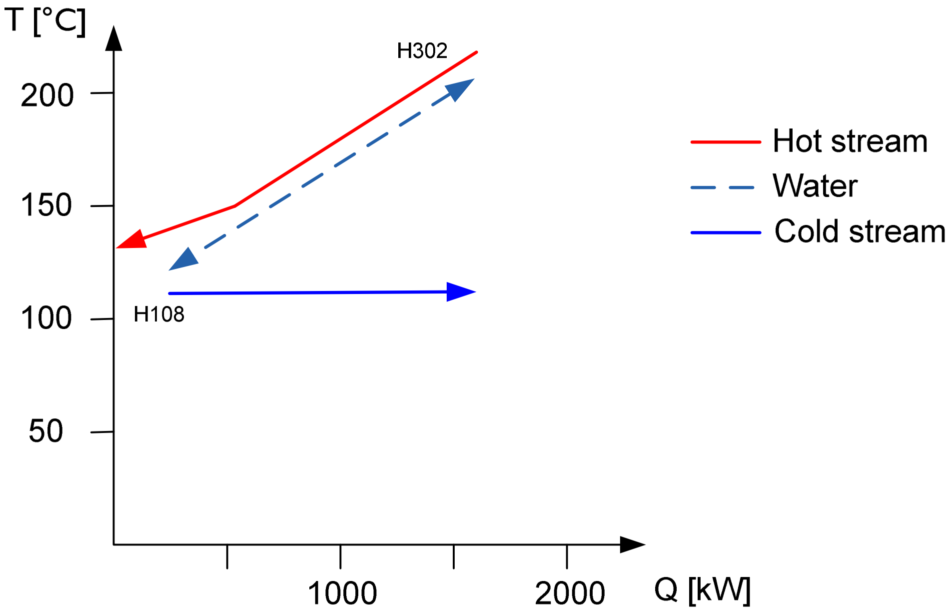


In the above figure, the flow rate of water is set as high as possible, giving the same capacity rate for the hot stream and the water, and the lines are parallel in the diagram. If the flow rate of water was lower, less heat could be transferred to H213, since the minimum temperature difference occurs at the hot end of the heat exchanger. On the other hand, the investment cost will be expensive, since the minimum temperature difference occurs throughout the hot heat exchanger. It is a trade-off between high energy recovery and large investment costs. The high flow rate of water is chosen in the calculations to recover as much heat as possible, but other flow rates should be tried as well, since the used flow rate could be unreasonably high, resulting in high investment costs.

Retrofit 1 – Subsystem 2



Here, the minimum temperature difference occurs on the cold end. The capacity rate of the water is set equal to the capacity rate at the lower part of the hot stream. If the water flow is decreased, the line will be steeper, and less heat can be transferred to the cold stream.



The above figure shows that if the flow rate of water is lowered, less heat can be transferred. The flow rate used for the calculations is the flow rate showed in the first picture.

The reason for the bend on the curve for H302 is that condensation takes place at different levels.

Appendix 15 – Economic evaluation hot water system

The calculations are made following the calculation procedure described in Chapter 3.5

	Q [kW]	dTlm [C]	U [W/ m2K]	A [m2]	Cost of equipment (Ce) Carbon Steel [$] Jan 2007	Ce, Carbon Steel [$] Nov 2010	Material cost factor fm	ISBL cost C [$]	Total fixed capital cost CFC [$]	Total fixed capital cost CFC [MSEK]	Annual steam savings [GJ]	Approx. annual savings [MSEK]	Payback period [years]
Hot water system, Retrofit 1, system 1													
H202-cirk water	2926	10,0	556	527	108843	118879	1.65	519502	945493	6.5			
Cirk water-H205	702	18,6	545	69	31446	34346	1.65	150091	273166	1.9	21289	2.39	0.790
Cirk water-H213	1768	33,5	459	115	37648	41119	1.65	179690	327036	2.3	53617	6.02	0.376
Cirk water - H212	457	20,2	459	49	28923	31590	1.65	138050	251251	1.7	13859	1.56	1.116
								Sum hot water, system 1:		12,4	88765	9.97	1.247
Hot water system, Retrofit 1, system 2													
H302 - cirk water	1602	10,0	556	288	65179	71190	1.7	317506	577861	4.0			
Cirk water - H108	1602	30,1	545	98	35214	38461	1.65	168075	305896	2.1	48583	5.46	0.388
								Sum hot water, system 2:		6,1	48583	5.46	1.120
								Sum hot water retrofit 1		18,5	137348	15.43	1.202

Appendix 16 – Estimation of the low-temperature energy demand in the PVC plant

The following values are based on an energy mapping made at the site, where also an explanation of the different units can be found (INEOS ChlorVinyls, 2008).

		From energy mapping				Load that can be covered with 100 °C water:		
In:	Descr.	Q [GJ/yr]	Q [kW]	Temp in [C]	Temp out [C]	[kW]	[GJ/yr]	
PM7	Hot water tank	7000	230.8	50	60	230.8	7000	
PM7	Before stripping	10000	329.7	75	88	329.7	10000	
PM7	Stream dryer	55814	1840.4	8	110	1479.6	44870	Air preheater
PM7	Fluid dryer	4596	151.6	8	50	151.6	4596	
PM8	Hot water tank	5000	164.9	?	50	164.9	5000	
PM8	Before stripping	3000	98.9	?	?	98.9	3000	
PM8	Spray dryer preheating	13680	451.1	8	20	3082.5	93480	Air preheater, extrapolated
PM8	After treatment	5000	164.9	8	22	164.9	5000	
PM82	Fluid dryer	8000	263.8	8	70	263.8	8000	
PM9	Stream dryer	59097.6	1948.7	8	170	986.4	29914	Air preheater
PM9	Fluid dryer	12000	395.7	?	55	395.7	12000	
PM9	Hot water tank	3500	115.4	40	52	115.4	3500	
PM9	Before stripping	6000	197.8	75	88	197.8	6000	

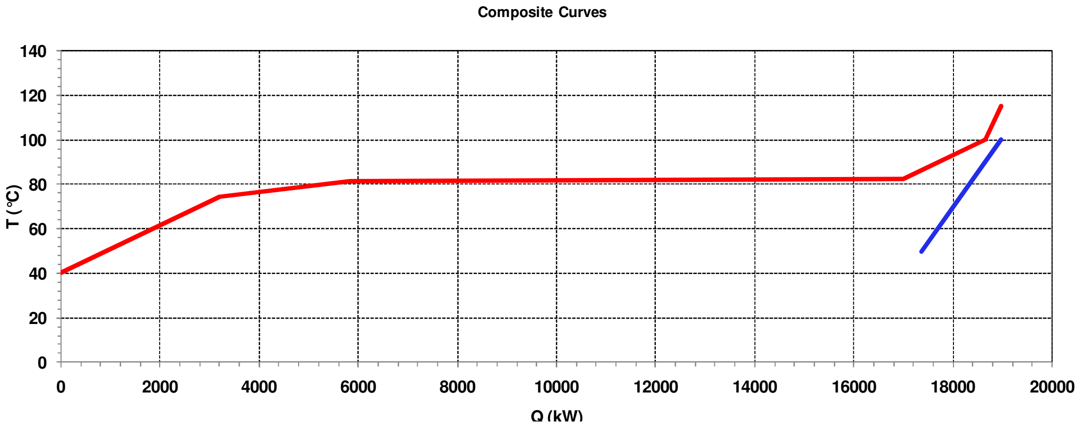
A summary of the loads that can be covered with 100 °C water is found below:

	[kW]	[GJ/yr]
Preheating of inlet air to combustion:	5548.4	168263.7
Steam saving for heating of fluids:	2113.5	64096.0
Total:	7662.0	232359.7

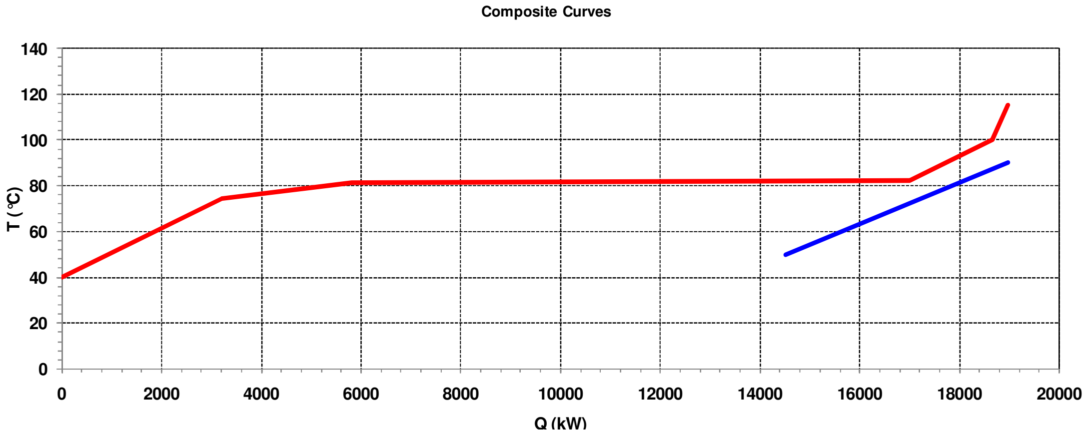
Appendix 17 – Composite curves using heat from the HTC unit for producing hot water

In this Appendix, composite curves for producing hot water from the HTC-unit are found. Two different combinations of hot streams are shown, and the possibility for producing hot water of different temperatures can be seen.

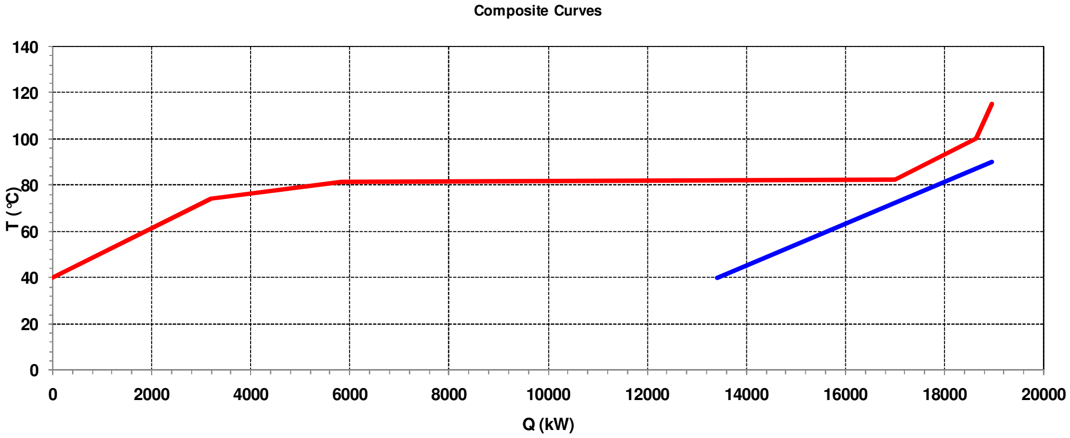
From H153, H164, H165 and Hnew



Composite curves heating water 50-100 C, from H153, H164, H165 and Hnew

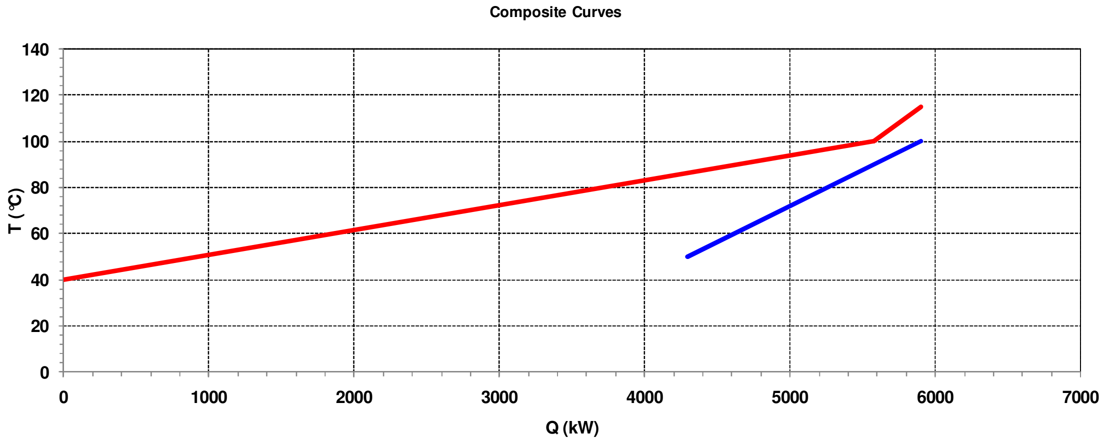


Composite curves heating water 50-90 C, from H153, H164, H165 and Hnew

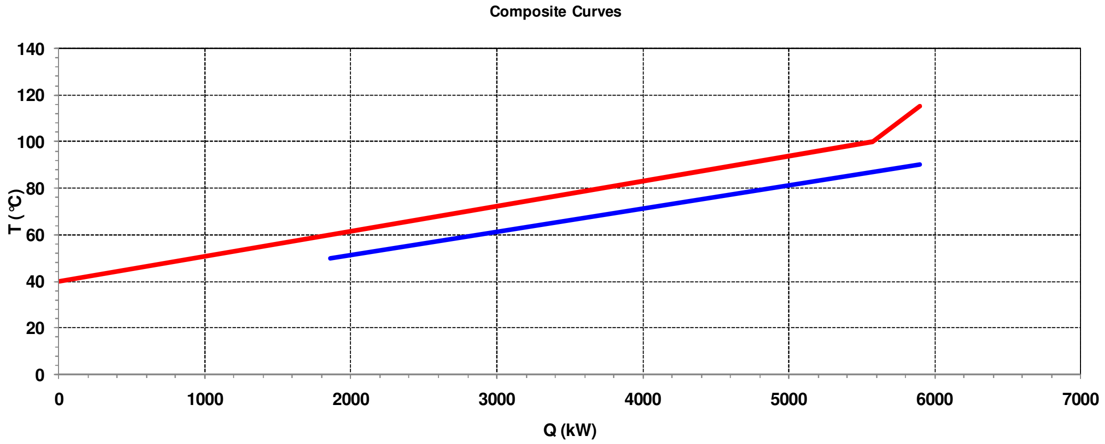


Composite curves heating water 40-90 C, from H153, H164, H165 and Hnew

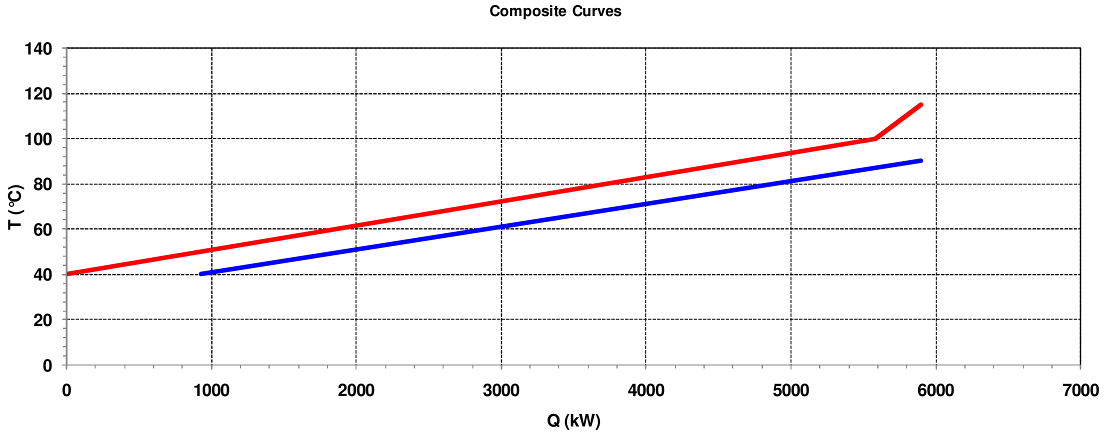
From H164, H165 and Hnew



Composite curves heating water 50-100 C, from H164, H165 and Hnew



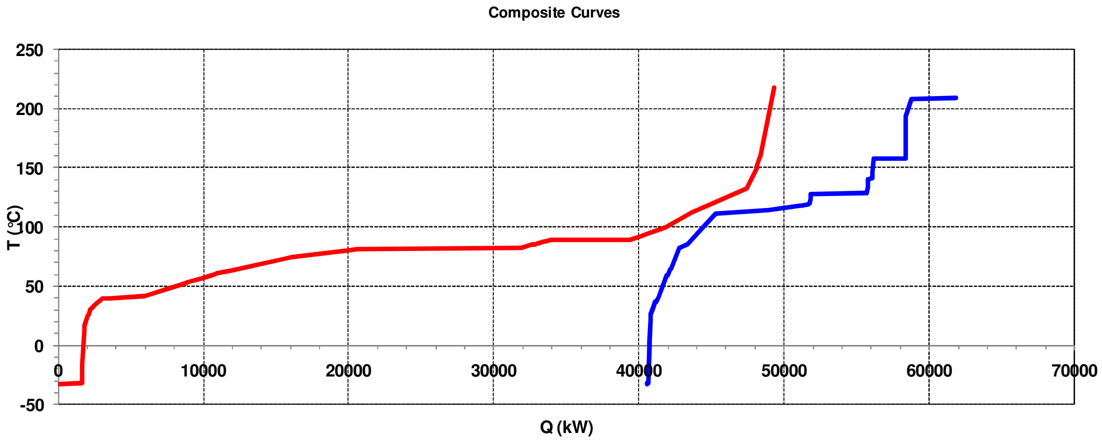
Composite curves heating water 40-90 C, from H164, H165 and Hnew



Composite curves heating water 40-90 C, from H164, H165 and Hnew

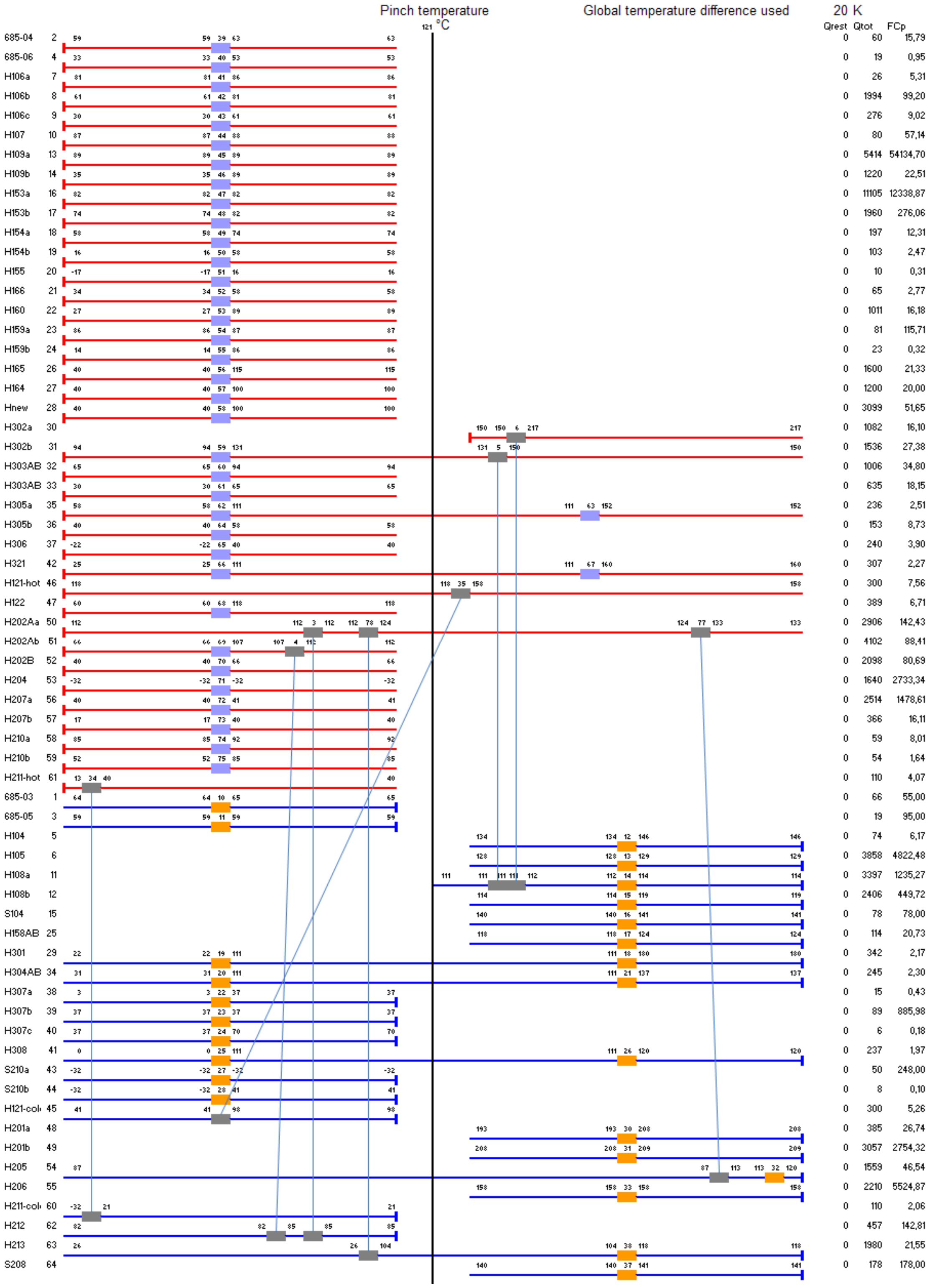
Appendix 18 – H205 with an outlet temperature 120°C

The outlet temperature from H205 is unknown. In the calculations, it is set to the temperature where the vapour fraction is 1.0, according to Aspen plus. This temperature is 145°C. If the temperature is lowered to 120°C, the pinch temperature will not be changed. But the potential for internal heat transfer will increase to 8.7 MW, of which 0.4 MW is utilised today



Composite curves future VCM plant, ΔT_{min}= 10°C, and an outlet temperature from H205 of 120°C

If the outlet temperature of H205 is only 120°C, the network for retrofit 1 with hot water (temperature difference of 20°C) is shown in the following figure. In the system 5.0 MW is transferred, instead of the 4.5 MW when the outlet temperature was higher.



		Qrest	Qtot	FCp
685-04	2	0	60	15,79
685-06	4	0	19	0,95
H106a	7	0	26	5,31
H106b	8	0	1994	99,20
H106c	9	0	276	9,02
H107	10	0	80	57,14
H109a	13	0	5414	54134,70
H109b	14	0	1220	22,51
H153a	16	0	11105	12338,87
H153b	17	0	1960	276,06
H154a	18	0	197	12,31
H154b	19	0	103	2,47
H155	20	0	10	0,31
H166	21	0	65	2,77
H160	22	0	1011	16,18
H159a	23	0	81	115,71
H159b	24	0	23	0,32
H165	26	0	1600	21,33
H164	27	0	1200	20,00
Hnew	28	0	3099	51,65
H302a	30	0	1082	16,10
H302b	31	0	1536	27,38
H303AB	32	0	1006	34,80
H303AB	33	0	635	18,15
H305a	35	0	236	2,51
H305b	36	0	153	8,73
H306	37	0	240	3,90
H321	42	0	307	2,27
H121-hot	46	0	300	7,56
H122	47	0	389	6,71
H202Aa	50	0	2906	142,43
H202Ab	51	0	4102	88,41
H202B	52	0	2098	80,69
H204	53	0	1640	2733,34
H207a	56	0	2514	1478,61
H207b	57	0	366	16,11
H210a	58	0	59	8,01
H210b	59	0	54	1,64
H211-hot	61	0	110	4,07
685-03	1	0	66	55,00
685-05	3	0	19	95,00
H104	5	0	74	6,17
H105	6	0	3858	4822,48
H108a	11	0	3397	1235,27
H108b	12	0	2406	449,72
S104	15	0	78	78,00
H158AB	25	0	114	20,73
H301	29	0	342	2,17
H304AB	34	0	245	2,30
H307a	38	0	15	0,43
H307b	39	0	89	885,98
H307c	40	0	6	0,18
H308	41	0	237	1,97
S210a	43	0	50	248,00
S210b	44	0	8	0,10
H121-col	45	0	300	5,26
H201a	48	0	385	26,74
H201b	49	0	3057	2754,32
H205	54	0	1559	46,54
H206	55	0	2210	5524,87
H211-col	60	0	110	2,06
H212	62	0	457	142,81
H213	63	0	1980	21,55
S208	64	0	178	178,00

Stream representation for retrofit 1 with hot water system, ΔT_{min}= 20°C, and an outlet temperature from H205 of 120°C