CHALMERS

Pinch analysis of Nynas refinery

-an energy efficiency study

Master's Thesis within the Sustainable Energy Systems programme

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Division of Heat and Power Technology
CHALMERS UNIVERSITY OF TECHNOLOGY
Göteborg, Sweden 2011

MASTER'S THESIS

Master's Thesis within the *Sustainable Energy Systems* programme

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Cover:
Picture of Nynas refinery

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ABSTRACT

Energy efficiency is one of many ways to reduce the problems of releasing more greenhouse gases into the atmosphere. The potential for energy savings due to energy efficiency measures can be large in industrial processes, especially in energy intensive businesses such as the refinery industry. In this project the oil refinery Nynas AB, situated near the harbour in Gothenburg, is studied. The plant is specialized in producing bitumen and naphthenic oils, mainly for the Nordic market.

The objective of this thesis is to investigate opportunities to increase process energy efficiency at the refinery Nynas AB. The main aim of the project is to locate energy savings by use of pinch analysis and to evaluate these.

The primary focus of this study is the distillation section of the plant due to its large energy usage, 60 % of the total energy demand at the plant. Within the distillation area two distillation columns and a furnace are located. The present energy demand in this area is 6.5 MW of heating and 5.0 MW of cooling. The results of the pinch analysis indicate that the theoretical minimum energy demand for the plant is 3.7 MW heating and 3.1 MW cooling. This implies that there is a potential for savings of around 42 % for heating and 37 % for cooling. However a more advanced analysis of the plant indicates that it is complicated to retrofit, thus reaching these energy targets is not reasonable from an economic perspective. The focus of the study was therefore on identifying measures that save energy without attempting to reach the maximum energy savings target.

Two retrofit suggestions was performed which both added a new heat exchanger to the heat exchanger train preheating crude oil with bitumen at the plant. The placement of the additional heat exchanger is different in the retrofit suggestions and gives different advantages concerning material, wear and size. The suggestions both decrease the energy demand in the furnace by 24 % and reduce about 55 % of the pinch violations. It would also decrease the CO_2 emissions by 7000 tonnes/yr due to reduced fuel demand.

In both retrofit suggestions the steam generator producing 2 barg steam to preheat the feedwater was reduced. This was solved by utilizing the excess heat below the pinch to preheat the fresh water.

Finally retrofit suggestion 1 with a new heat exchanger close to the furnace, using an appropriate construction material was recommended. This retrofit option was considered to protect of the existing heat exchangers. An economic analysis of retrofit options 1 and 2 showed a payback period corresponding to about 9 months for retrofit 1 and 18 months for retrofit 2. The conclusion is that further heat exchange in the process is something to recommend.

Key words: Pinch analysis, process integration, retrofit, refinery, bitumen

Pinch analys av Nynas raffinaderi, en energieffektiviseringsstudie
Examensarbete inom masterprogrammet *Sustainable Energy Systems*
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SAMMANFATTNING

Energieffektivisering är ett av många sätt att reducera problemen med utsläpp av växthusgaser till atmosfären. Potentialen för energibesparingar är stor i industriella processer, speciellt i energiintensiva företag som raffinaderier. I detta projekt studeras oljeraffinaderiet Nynas AB, beläget i hamnen utanför Göteborg. Anläggningen är specialiserad på att tillverka bitumen och nafteniska oljor, framförallt för den Nordiska marknaden.

Syftet med examensarbetet är att studera möjligheterna för att energieffektivisera processen på raffinaderiet Nynas AB. Huvudmålet med projektet är att lokalisera energibesparingar med hjälp av pinchanalys och sedan utvärdera dessa.

Studien är fokuserad till anläggningens destillationsområde på grund av dess stora energikonsumtion, 60 % av den totala energikonsumtionen. Inom destillationsområdet finns två destillationskolonner och en ugn. Dagens energibehov i den här delen av raffinaderiet är 6.5 MW värme och 5.0 MW kyla. Pinchanalysen resulterar i ett minsta värmebehov på 3.7 MW och ett minsta kylbehov på 2.1 MW. Det innebär en besparingspotential på cirka 42 % av värmebehovet respektive 37 % av kylbehovet. Dock visar mer avancerad pinch analys av processen att den är komplicerad att förändra, vilket betyder att dessa energimål inte är rimliga i ett ekonomiskt perspektiv. Studiens fokus var därför att identifiera åtgärder som sparar energi utan att nå de maximala energimålen.

Två olika ombyggnadsförslag görs och i båda adderas en extra värmeväxlare till den befintliga värmeväxlarkedjan som förvärmer råolja med bitumen på anläggningen. Placeringen av den nya värmeväxlaren sker på varsin ända av värmeväxlarkedjan i de båda förslagen och ger olika fördelar angående material, slitage och storlek. Båda förslagen sänker energibehovet i ugnen med 24 % och reducerare över 55 % av pinchbrotten. De skulle också minska CO_2 utsläppen med 7000 ton/år.

Båda ombyggnadsförslagen resulterar i en minskad ånggenerator på bitumenströmmen som producerar 2 barg ånga för att förvärma ångpannornas matarvatten. Detta löstes genom att använda överskottsvärmen under pinchen för att förvärma matarvattnet med ytterligare två nya värmeväxlare.

Slutligen var ombyggnadsförslag 1 med en ny värmeväxlare nära ugnen med lämpligt material rekommenderat. Förslaget skyddar de befintliga värmeväxlarna i kedjan. En ekonomisk analys av förslag 1 och 2 visar en återbetalningstid motsvarande 9 månader för förslag 1 och 18 månader för förslag 2. Slutsatsen är att ytterligare värmeväxling i processen är något att rekommendera.

Nyckelord: Pinchanalys, processintegration, raffinaderi, bitumen

Contents

Preface

In this master's thesis project, the distillation area at Nynas AB was investigated using pinch analysis. The study was performed in cooperation with Nynas AB, COWI AB and the Division of Heat and Power Technology at Chalmers University of Technology.

We would like to thank our examiner Simon Harvey, supervisor Rickard Fornell at Chalmers and our supervisor Sofia Ohlsson at COWI AB for always taking time to answer our questions and for given support. A special thanks to Stefan Nyh who has helped us a lot during our work at Nynas AB. We would also like to thank Anna Keereweer at Nynas AB for giving us the opportunity to perform this master's thesis.

Finally, we would like to express our gratitude to the rest of the staff at Nynas AB and COWI AB for their useful help and for taking care of us.

Göteborg June 2011

Anna Gunnarsson, Carin Magnusson

Notations

ΔT_{min}	Overall Minimum temperature difference, is used to determine the total utility demand
AD	Atmospheric Distillation
ACLC	Actual Cooling Load Curve
AHLC	Actual Heat Load Curve
CC	Composite Curve
Cp	Specific heat capacity [J/(kg K)]
CUC	Cold Utility Curve
CW	Cooling water
ECLC	Extreme Cooling Load Curve
EHLC	Extreme Heat Load Curve
F	Mass flow [kg/s]
FCp	F*Cp [W/K]
GCC	Grand Composite Curve
HEN	Heat Exchanger Network
HUC	Hot Utility Curve
MER	Maximum Energy Recovery
P	Pressure [barg]
T	Temperature [˚C]
TCLC	Theoretical Cooling Load Curve
THLC	Theoretical Heat Load Curve
VD	Vacuum Distillation

1 Introduction

Global warming caused by emissions of greenhouse gases such as carbon dioxide is an environmental issue frequently discussed today. Resulting policies provide clear incentive for industry to reduce fossil fuel consumption and consequently avoid emissions leading to climate change.

Energy efficiency is one of many ways to reduce the problems of releasing more greenhouse gases into the atmosphere. The potential for energy savings due to energy efficiency measures can be large in industrial processes, especially in energy intensive businesses such as the refinery industry (Sundlöf, 2002). Increases in energy efficiency will decrease the need for firing fossil fuel to provide heat for industrial plants and consequently the carbon dioxide emissions will also decrease. Hence, the climate change issue together with a concern about depletion of fossil fuel reserves provides incentive to investigate energy efficiency.

Increased energy efficiency also provides economic benefits since reduced utility usage implies reduced fuel costs. The price of fossil fuels is rising and is predicted to continue to rise, which further motivates a reduced fuel usage. Taxes and other costs associated with carbon dioxide emissions are also expected to increase as policies to curb global warming are implemented.

Finally, the EU has put in place an Emissions Trading Scheme for fossil CO_2 emissions constructed according to the cap and trade model. This means that all countries within Europe have a common capped level for total emissions and the companies can trade their assigned emission rights within this market. The total cap will be progressively tightened in the future which provides a clear incentive for industrial plants to conduct long term efforts to reduce emissions.

1.1 Background

This project investigates the oil refinery Nynas AB, situated near the harbour area in Göteborg. The plant is specialized in producing bitumen and naphthenic oils mainly for the Nordic market. Bitumen, the main product at the refinery, is the heaviest fraction of the crude oil and is primarily used as binder in asphalt for the paving industry, but also for coating and insulation (Nynas AB, 2011).

The refinery has a capacity to process 438 000 tonnes of crude oil per year (Nynas AB, 2009). The only feedstock used in the process, Heavy Venezuelan crude oil (ÅF, 2007), contains about 71 % of bitumen, 17 % of naphthenic specialty oils and 12 % of fuel. This composition can be compared to fuel oriented oil refineries where the crude oil consists of approximately 96 % fuel and 4 % bitumen, naphthenic speciality oils and lubricant oils.

An overview of the refinery plant is shown in Figure 1. The figure shows the crude oil cistern, the furnace with two combustion chambers, the atmospheric distillation column, the vacuum distillation column, the oxidation system and the storage cisterns for the products.

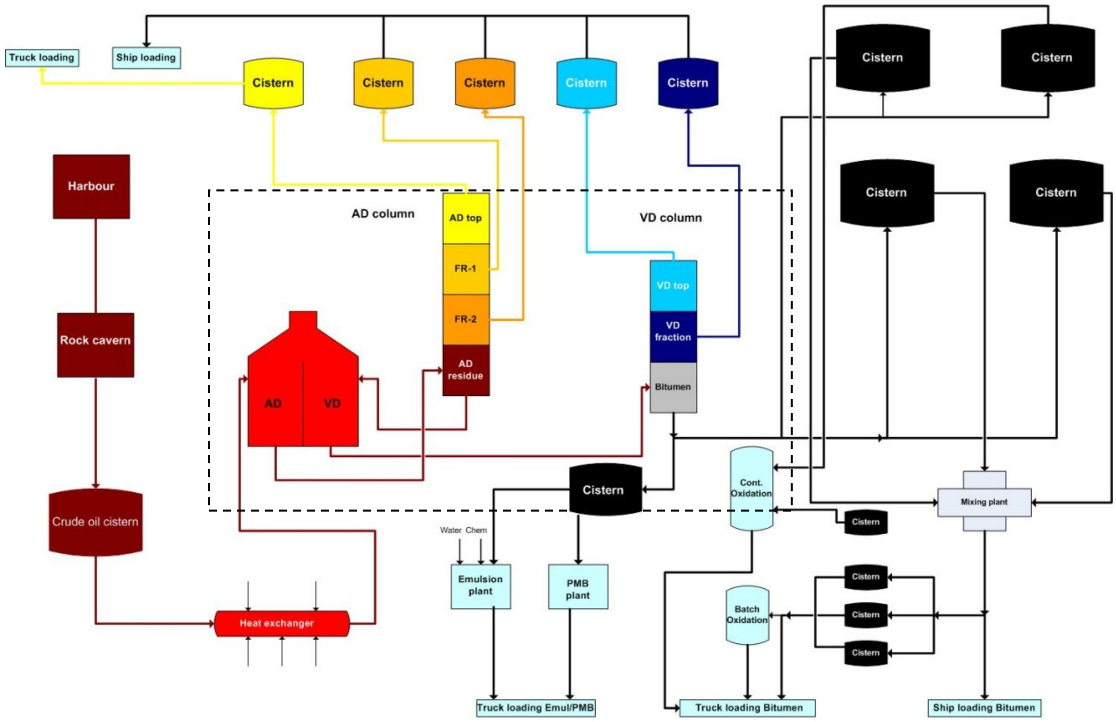

Figure 1. Process plant overview of Nynas refinery operations in Göteborg.

Three products are fractioned from the first distillation column (the Atmospheric Distillation column, AD) and transferred to storage cisterns. The most volatile part is straight run naphtha (AD top), the second part is kerosene (FR1) and the third part is gas oil (FR2). In the second distillation (the Vacuum Distillation column, VD) three more fractions are formed. The main fraction is bitumen which is taken out at the bottom of the VD column, and the other two fractions are vacuum gas oils with different volatility. The bitumen can either be sent directly to the cisterns or be further processed to oxidized bitumen, polymer modified bitumen or emulsions. The fraction of the bitumen sent directly to the storage cistern is mostly utilized for paving, while the oxidized bitumen is used in construction material. Polymer modified bitumen is also used for paving but it is improved to get better resistance to wear than ordinary bitumen. Finally, the bitumen emulsions are used for cold paving.

Bitumen is a product that causes severe equipment fouling which causes difficulties at the plant. One example of this is the decreasing capacity of the heat exchangers over time. The refinery has a yearly stop to perform maintenance work during the winter, which is a period of low demand for asphalt in the Nordic market. The production stop does not concern all parts of the plant; the oxidation of bitumen continues all year around, and the cisterns and some pipelines require heating even when the plant's production units are not running.

The total yearly energy consumption of the refinery was 114.2 GWh/year in 2007 (ÅF, 2007). The main consumers are situated within the distillation section including utility services, steam boilers and a hot oil boiler. Examples of other energy consumers at the plant are cisterns, pipelines, cold water towers and pumps. Both steam and electrical tracing is used for heating of storage cisterns and pipelines. The reason for heating the cisterns is that the bitumen and crude oil are stored at specific temperatures to avoid solidification.

Table 1. Energy consumption at Nynas 2007, (ÅF, 2007). The total energy consumption here is a sum of all energy consumption at the plant and not only the large consumers listed in the table.

	Total plant [MWh/yr]	Furnace AD/VD [MWh/yr]	AD column [MWh/yr]	VD column [MWh/yr]	Steam boiler 1 [MWh/yr]	Steam boiler 3 [MWh/yr]	Elect. Steam boiler [MWh/yr]	Hot oil boiler [MWh/yr]
Fuel oil	97000	66700			27040	3400		-
Electricity	17200	820	751	526	936		914	4200
Steam			7100	1440				
Total	**114200**	**67520**	**7851**	**1966**	**27976**	**3400**	**914**	**4200**

The information in Table 1 is taken from a report written in 2007. The energy consumption is a few percent higher today due to an increase in production. However, Table 1 gives a good estimation of the energy distribution at the plant.

The energy demand at the plant is primarily covered by using fuel oil and electricity. The part of the process with the highest total energy demand is the distillation area, where the two distillation columns and the furnace are located. This section of the plant uses 66.7 GWh/year of fuel oil and 2.1 GWh/year of electricity which is equal to about 60 % of the total energy consumption at the plant, see Table 1. Furthermore, large amounts of steam are used in the distillation columns. Therefore this is the section of the plant which was the main focus of this study.

The emissions from the plant result mainly from fuel combustion and are shown in Table 2. The emissions are the levels specified by the plant's environmental permit or just below.

Table 2. Emissions from Nynas AB, (Nynas AB, 2009).

Emission	CO_2	SO_2	NO_X	Particles	Unburned hydrocarbons	Volatile organic compounds
Amount [tonnes/year]	29 779	62	43	3.8	0.38	70

Nynas AB has conducted a number of previous projects related to energy mapping and energy efficiency. Previous work includes an energy study of a possible future process layout. The main difference between the future process layout compared to this present study is a doubled production, replacement of the furnace and one of the distillation columns (Anitha Jacobsson, 2008). Results from the previous study are included in this project as background information. The previous study was performed using the same method as that adopted in this master's thesis, namely pinch analysis. The future process plant scenario is based on the company's plans to increase their production. However, this will not occur within the coming five year period.

1.2 Objective

The objective of this study is to identify opportunities to increase energy efficiency in the process at the refinery Nynas AB. Pinch analysis is to be used to determine energy targets for the process, as well as to identify pinch rule violations in the existing process energy system which lead to unnecessarily high utility usage. Finally, the study aims to propose measures to improve energy efficiency at the plant and to conduct a simplified economic evaluation of the proposed measures.

1.3 Problem analysis

The study will focus on the most energy consuming part of the plant, namely the distillation area. The first step of the study is to determine how much energy could theoretically be recovered and reused internally in the process. Establishing such energy use targets can be performed using pinch analysis. Pinch analysis can also be used to identify whereabouts in the plant there is surplus or deficit of heat.

The pinch analysis highlight where in the system the main energy flows are located. Further analysis is needed to identify where in the process improvements are needed in order to reduce energy usage. Investigation of the utility system may lead to suggestions of different energy measures. Examples of measures may be more preheating of the fuel oil before entering the furnace or preheating of the boiler feedwater. The heat exchangers can be modified as well as how steam is produced. Further, the expected surplus of process heat could be managed by delivering heat to the district heating network. Another way of dealing with the waste heat is to use it for heating the storage cisterns. Some of these different arrangements and utility systems will be further investigated and some will not, depending on how large energy gains they have.

1.4 Limitations

The primary focus of this study is the distillation section due to its large energy consumption. The steam production will be affected by changes in the distillation section. However, only a brief investigation of the steam system will be done. A flue gas boiler producing 16 barg steam from flue gases in the furnace is situated close to this part of the process. This flue gas boiler is selected to be outside of the system boundaries since its production of steam could be replaced by other steam producers. There are also plans to heat the pipelines and storage cisterns with hot oil in the future. This would heavily reduce the steam consumption at the plant.

The data from the process is collected from a limited time period and thus seasonal variations are not accounted for in this study. The time period selected is from May

2010 to August 2010. Average process data for this period is used in this study. This period was selected since the plant has steady production during the summer in contrast to the winter. The brief time period selected does not allow variation in heat transfer capacity in the heat exchangers over the year to be analysed. These variations mostly occur due to heavy fouling. The heat transfer capacity is larger at the process start up phase in March, directly after the maintenance stop, than at the end of the production year in December.

There are some different production modes at the plant of which almost exclusively 2 modes are used. The study will focus only on the most frequent mode (160/220) which is applied 65 % of the time. The difference between the modes is small in an energy perspective but this study is only representative for mode 160/220. The mode 160/220 denotes that the penetration of the produced bitumen is between 16 and 22 mm, according to the standard SS-EN 1426.

Another limitation to keep in mind is that the bitumen cannot be heated above a certain level to avoid cracking which leads to problems in the furnace. The temperature of the crude oil in the furnace is around the bitumen's maximum temperature which is approximately 378 ^{0}C.

The furnace currently used at Nynas was built in 1956 and is going to be replaced, in about 5 years, to improve heating efficiency. This change will not be taken into consideration in this study since the furnace project is still at a planning stage and consequently there is a lack of data. The maximum allowable temperature of bitumen is already reached in the present situation which implies that a new furnace will affect the fuel consumption more than the effluent stream temperatures.

Finally, the pinch analysis will focus on the existing plant and will not consider reconstruction or future investments. This master's thesis focuses on how the process is today and how process integration can be applied to the plant at Nynas.

2 Process description

This study will focus on the distillation section of the process at Nynas AB, which consists of one furnace with two combustion chambers and two distillation columns. The process flow sheet for this section of the process is shown in Figure 2.

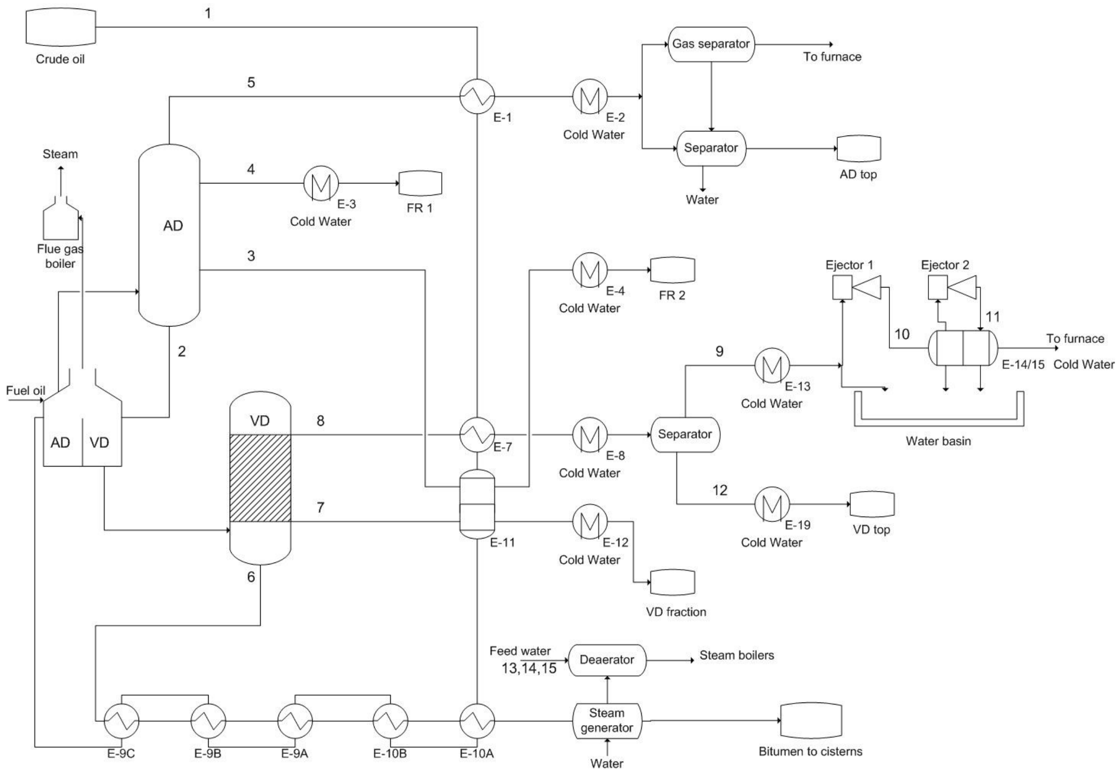

Figure 2. Process flow diagram.

Crude oil is continuously pumped from a rock cavern to a crude oil cistern at the production site. The crude oil is heated by a number of heat exchangers before entering the first furnace, see Figure 2. This results in a preheated crude oil stream which decreases the energy demand of the furnace. The stream continues to the atmospheric distillation column where four different fractions are extracted. Two of the fractions extracted from the AD column (stream 3 and 4) are led into pressure vessels where the gas and liquid are separated and the gaseous part is recycled to the column. The return flows are not of interest in this study since the distillation column is treated as a black box and should not be changed. Therefore these gaseous parts of the streams and the pressure vessels are not shown in the process flow diagram.

The most volatile fraction from the AD column is stream 5 which is a gaseous flow. The gas consists of volatile hydrocarbons and steam from the distillation column. To be able to extract only liquid hydrocarbons later in the process, stream 5 must be condensed. Cooling of stream 5 is achieved in two steps, first by heat recovery to preheat the crude oil and then final cooling in a water-cooler to reach the target temperature. Thereafter, the stream is separated and gas and liquid is led in different directions before further processing. The gaseous part goes to a separator where all the remaining liquid is separated (Nynas AB, 2011). The liquid part is sent to another separator where the water is separated from hydrocarbons in order to deliver a product which is led to a storage cistern.

The bottom fraction from the AD column (stream 2) is reheated in the furnace and led to the vacuum distillation column. The VD column has low pressure to maintain good separation without having to use a too high temperature which would imply a risk of cracking for the bitumen stream.

The VD fraction (stream 7) first goes to a stripper where gas is led back to the distillation column in the same way as in the AD column. The VD top flow (stream 8) contains both steam and hydrocarbons and should be condensed and separated. The method differs from the condensation in the AD column due to higher temperatures. Consequently the water is not separated but follow as steam with stream 9. The gas part of VD top (stream 9) is partly condensed with cooling water and after the heat exchanger the liquid is separated down to a water vacuum basin. The gas is led into the first ejector where high pressure steam is injected and expanded in the ejector to create low pressure in the system. After the first ejector the flow is heat exchanged to condensate out water again. After the last ejector a small gas flow is separated and led to the furnace where it is used as fuel.

From the bottom of the VD column bitumen (stream 6) is extracted. To utilize the heat in the stream it is heat exchanged with crude oil (stream 1) and with a steam generator, which produces steam at 2 barg before it is led to cistern.

2.1 Utility systems

The utility systems used at the plant are the steam system and the cooling system. There is also heating in the furnace with fuel oil used as hot utility. The steam system has been investigated before and showed potential for improvements. Hence the steam system will be investigated further.

2.1.1 Steam system

The steam system at the plant consist of three different levels of steam; 16, 8 and 2 barg. An overview of the total production and distribution is shown in Figure 3. The 16 barg steam is produced in steam boiler 1 and in the flue gas boiler, and the total production is 4.7 MW. When steam boiler 1 is taken out of operation for maintenance it is replaced by an electric steam boiler. Steam boiler 3 is smaller than steam boiler 1 and can be used for peak load production when the demand for steam is high. The electric steam boiler produces steam with higher moisture content than the other steam boilers, this can be a problem in the distillation columns. The reason for the moisture in the steam is that it is not led through a superheater in the boiler. When the steam is only saturated it can condense in the pipelines from the boiler and contribute to higher moisture content compared to superheated steam.

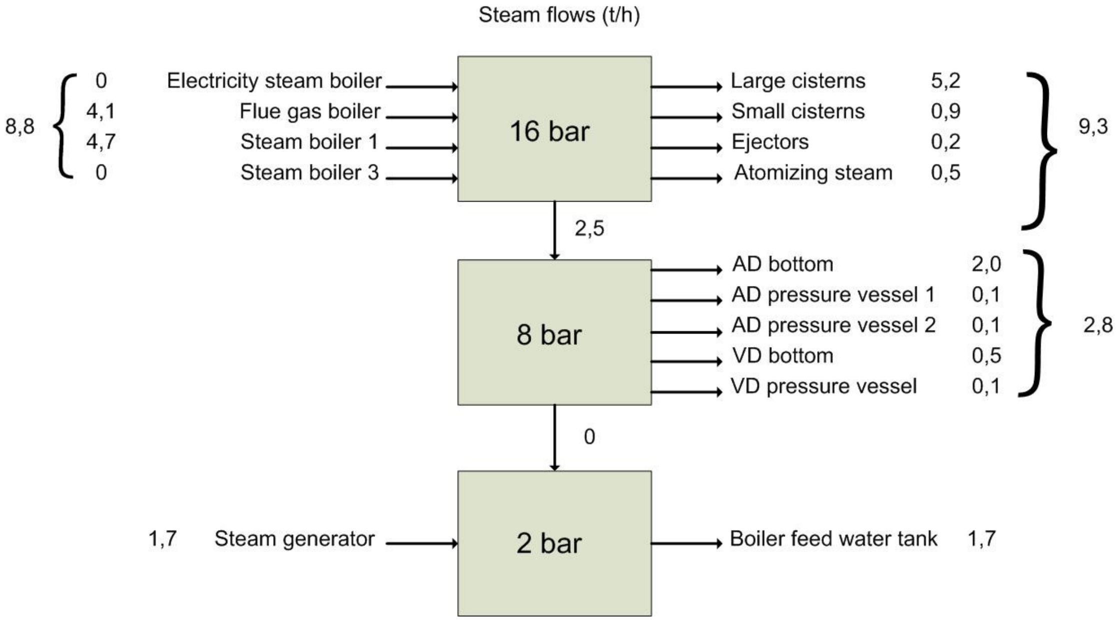

Figure 3. Steam production and distribution. The difference in the production and consumption of steam is due to error in measurement of the steam flows.

Totally 9.3 tonne/hour steam of 16 barg is consumed at the plant as shown in Figure 3. This high pressure steam is used in the ejectors to create vacuum, as atomizing steam and for heating of cisterns, (Jonsson, 2007). Most of the 16 barg steam is used to heat the large cistern park. The reason that this high pressure steam is used for heating the cisterns is that the storage temperature of bitumen is 160-170 ºC and the saturation pressure of high pressure steam lies just above this temperature interval whereas the other steam levels are colder (Nyh, 2011). From the high pressure steam a fraction is let down to 8 bar steam. This medium pressure steam is used in the distillation columns as stripping steam. The last steam level is low pressure steam at 2 barg and is mainly used for preheating boiler feedwater.

The total production of steam at the plant is 10.5 tonne/hour. 2.7 tonne are consumed in the distillation area. The losses in the steam system are large and correspond to 2.3 tonne/hour or 23 % (Jonsson, 2007). One of the reasons for the large losses is that the steam system was built at the same time as the rest of the plant when the importance of energy efficiency was not in focus hence it reveals many areas in need of improvements (Nyh, 2011). This statement implies that there are a lot of leaks in the steam system.

A further possible explanation for the losses is that the flow meters sometimes differ with 2 tonne/h at the largest flows and the flow meters have an error range of 10 %. The temperature measurements are more reliable and the energy balances has been calculated to match the temperatures in larger extent than the flows. The uncertainties of the flow meters are shown for example in the steam balance of 16 barg steam where the production and consumption do not match.

2.1.1.1 Steam generator producing 2 barg steam

The steam generator is situated at the end of the bitumen stream (stream 6). It utilizes the heat available in the bitumen stream after the heat exchanger train and produces 2 barg steam. It also acts as a cooler since the bitumen should not be stored at a too high temperature.

The 2 barg steam is used to heat the boilers feedwater. The feedwater that has to be heated consists of three different streams, the condensate from the large cistern park, the condensate from the small cistern park and fresh water. These flows can be seen as streams 13, 14, 15 in Table 5. In Figure 4 it is shown how the steam produced in the steam generator are fed to the deaerator. In this deaerator the cold feedwater is heat exchanged with steam from 10 or 50 °C to 133°C.

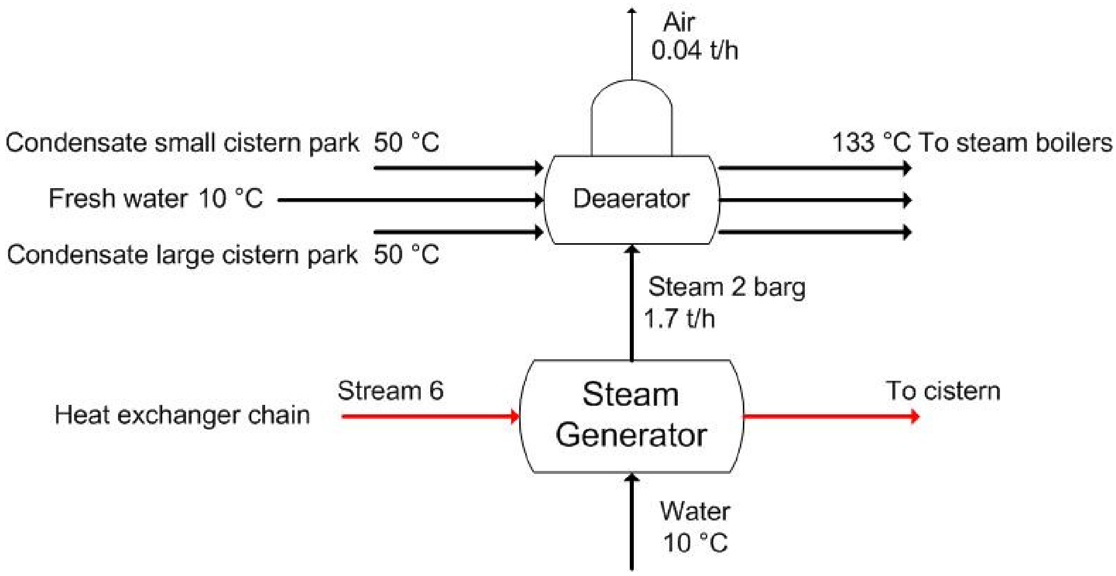

Figure 4. Steam generator and heating of feedwater system.

2.1.2 Cold water system

The cooling system consists of 6 cooling towers and a cold water basin from where the water is used in coolers in the process. The cooling water inlet temperature is about 20 °C during the summer. The number of cooling towers used at the plant depends on the cooling demand, with more towers in use during summer than during winter (Risan, 2011). The cooling towers at the plant have a capacity of 5-6 MW (Jonsson, 2007).

3 Methodology

The main steps in this work included a literature study, a pinch analysis with evaluation of process integration opportunities and an economic analysis.

The literature study was performed to gather knowledge about the plant, the theory of pinch analysis and the process at Nynas. A previous study performed at Nynas involved an energy mapping of the process. With this report as a base a brief energy mapping is performed to get an overview of where in the process there is potential for energy savings. The data gathering following the energy mapping is a part of the pinch analysis, see also Section 3.1. Data was gathered from the data log system PI process book. Additional input to the analysis was gathered by energy and mass balances of the process, simulations and measurements at the site.

When the pinch analysis is completed, different technical measures are suggested in order to achieve possible energy savings. Finally, an economical evaluation of the proposed measures is presented.

3.1 Pinch analysis

Pinch analysis is a useful process integration tool to identify the potential for reduction of the energy consumption of industrial plants. Pinch analysis was developed as a method for process integration during the 1970's at ETH Zurich and Leeds University. Further research was done by professor Linnhoff at Manchester Institute of Science and Technology. The technology has been employed on many chemical plants as well as in a wide range of industries (Kemp, 2007). The method is a way to systematically analyze the heating and cooling demand of a process and to set targets for the theoretical maximum amount of heat that can be recovered initially within the process. A process usually includes both heating and cooling demand. Part of the heating demands can be met by internal heat recovery and the remaining heating demand must be met by hot utility. The hot utility demand can be decreased by an optimization of the heat exchanger network.

In a pinch analysis all streams are defined on the basis of their start and target temperatures (T), heating value (Cp) and mass flow (F) and are divided into either hot or cold streams. A hot stream is defined as a stream which needs to be cooled to reach its target temperature. A cold stream needs to be heated to reach its target temperature. Adding up all cold streams and all hot streams separately gives Composite Curves (CC), see Figure 5. The CC shown in a temperature and heat flow diagram indicates how much heat that is recovered in the process and how large cooling and heating utility that is needed (Harvey, 2009).

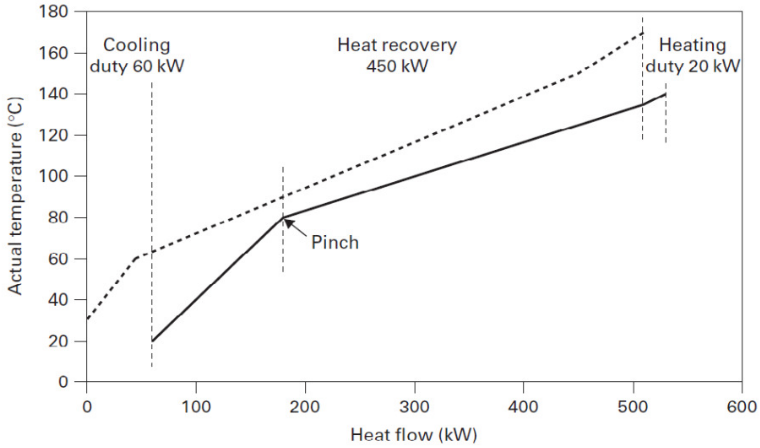

Figure 5. Composite curves, the dashed line depicts the hot streams while the continuous line depicts the cold streams (Kemp, 2006).

When all the streams are defined they can be divided into interval temperatures. All hot stream temperatures are decreased by ΔT_{min} /2 and all cold stream temperatures are increased by ΔT_{min} /2. The ΔT_{min} is the minimum temperature difference allowed between streams that exchange heat. The smaller ΔT_{min} the more heat can be transferred in the heat exchanger, but this will also lead to larger heat exchanger area which is costly. Hence, ΔT_{min} should be selected based on economic considerations and experience (Harvey, 2009).

The reason to shift the temperatures is that the minimum temperature difference between the hot and cold streams will be zero where a pinch point is located. The interval temperatures are used to compose a Grand Composite Curve (GCC) that is shown in a temperature and heat flow diagram and gives an overview of the process, see Figure 6.

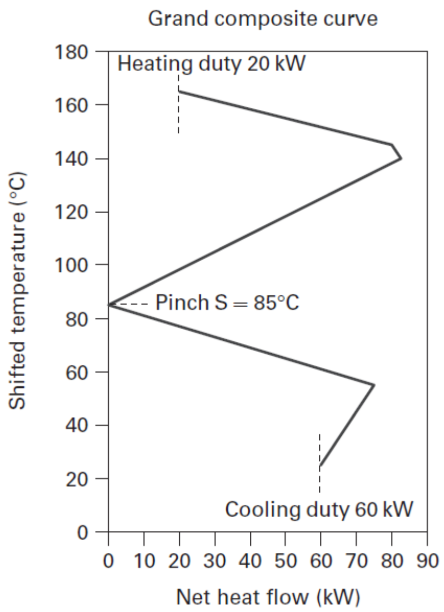

Figure 6. Example of Grand Composite Curve (Kemp, 2006).

The point in the GCC where the net heat flow is zero corresponds to the pinch temperature. At this point in the process the driving forces for heat exchange are at the minimum, see Figure 6. The GCC shows the total heating and cooling demand for the process and is also helpful in a later phase when implementing different measures and utilities such as warm water, different steam pressure levels and gas turbines.

When designing a heat exchanger network there are three rules that are very important to keep in mind. These are:

- Do not cool above the pinch
- Do not heat below the pinch
- Do not transfer heat through the pinch

Violations of any of these rules results in unnecessary heating or cooling demand in the process, see Figure 7. Cooling above the pinch implies that heat is extracted from a system which has a deficit of heat and the same amount of heat must be added by an external heater. Heating below the pinch results in heating in a place that already has excess of heat and the same amount of heat must be cooled with external coolers. Heat transfer through the pinch imply more heat added above the pinch than needed, the heat follows through the whole system and in the end has to be cooled away below the pinch (Harvey, 2009).

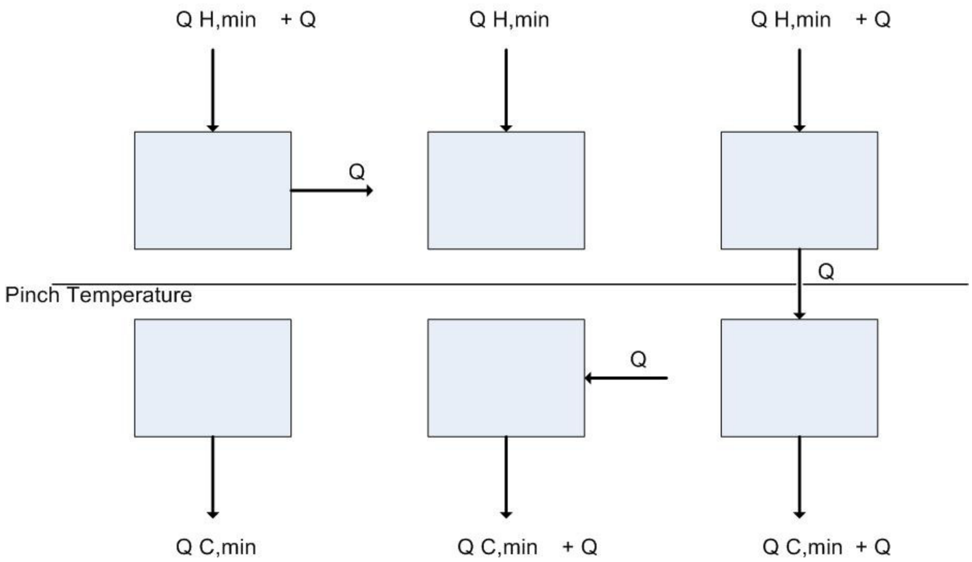

Figure 7. The three golden rules. From the left; cooling above the pinch, heating below the pinch and heat transferred through the pinch.

ProPi, an add-in to Microsoft Excel, is a program to perform the pinch analysis that was developed by Chalmers Industriteknik AB. In this program it is for example possible to analyse process stream heating and cooling requirements so as to establish the overall energy targets for the process as well as location on the process pinch. The programme can be used to obtain CC and GCC curves as well as to design heat exchanger networks layouts. Finally, the tool can be used to identify pinch rule violations in existing heat exchanger networks.

3.1.1 Maximum Energy Recovery

The minimum utility demand of the plant is given from the CC or GCC and can be achieved by constructing a Maximum Energy Recovery (MER) network. In an MER network there are no pinch violations. However, construction of a MER network usually requires a large number of heat exchangers. To achieve an MER network it is sometimes necessary to split large streams. Stream splitting at the pinch is necessary under the following conditions:

- Directly below the pinch if $FCp_{cold} \leq FCp_{hot}$
- Directly above the pinch if $FCp_{cold} \geq FCp_{hot}$.

In practice, it is rarely profitable to construct an MER network, particularly when retrofitting an existing HEN. It is however of interest to establish a reference MER network which can be compared to the present network to discover any similarities or to identify suitable retrofit measures.

3.1.2 Advanced composite curves

When performing a retrofit of a plant it is essential to take into account the existing heat exchanger network (HEN). This can be done by using advanced composite curves. The aim of these curves is to identify heat recovery projects which reduce pinch violations and are economically feasible.

The advanced pinch analysis consists of three curves on each side of the pinch. On the hot side of the pinch it is of interest to plot Actual Heat Load Curve (AHLC), the Extreme Heat Load Curve (EHLC), and the Theoretical Heat Load Curve (THLC). The AHLC represents the process streams going through the heaters and thereby consider the existing heat exchanger system. The THLC shows the theoretical lowest temperature where heat can be introduced to the system. The EHLC shows the maximum temperature for heating in the process and is established from the CC of the original process, see Figure 8. Similar curves can be established below the pinch temperature.

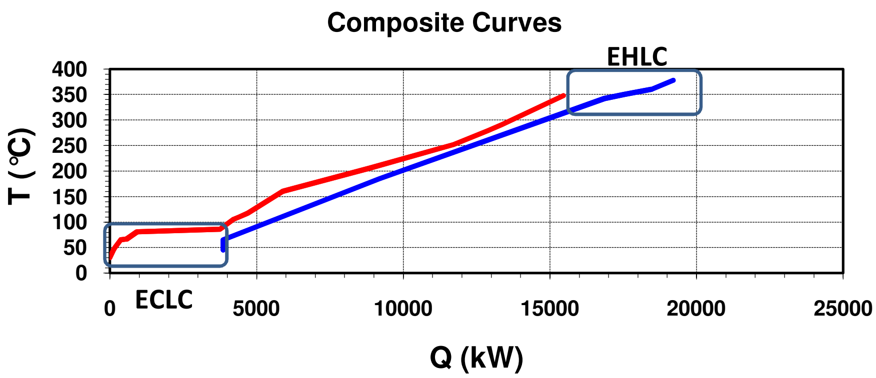

Figure 8. The EHLC and the ECLC are represented by the highlighted curve sections in the CC.

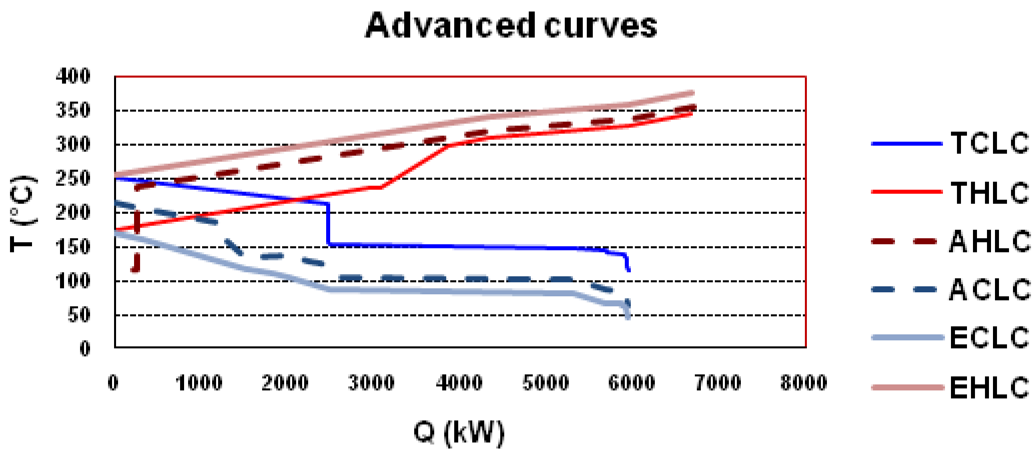

Figure 9. Advanced pinch curves.

The AHLC should lie between the EHLC and the THLC, see Figure 9. If the AHLC is close to the EHLC the heat exchanger network is complicated to retrofit. It indicates that it has coolers placed at low process stream temperature levels and heaters placed at high process stream temperature levels, far away from the pinch, see Figure 10. The system then requires many rearrangements to retrofit and it is necessary to use a small ΔT_{min} and a large area which can lead to high costs.

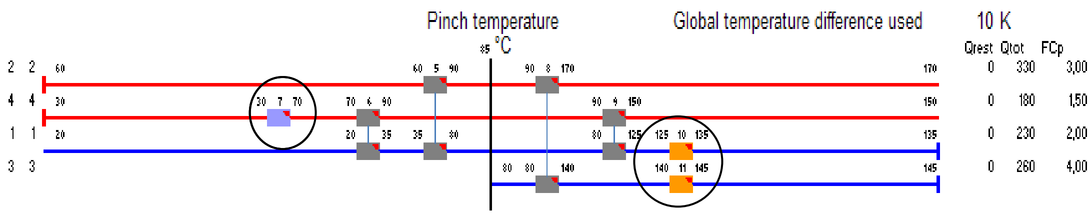

Figure 10. Example of a system which is hard to change with heaters (yellow boxes) placed at high temperature levels and coolers (blue boxes) placed at low temperature levels, far away from the pinch.

If the AHLC is instead close to the THLC it is much cheaper to add extra area. Then heaters are placed at low temperature levels and coolers placed at high temperature levels, close to the pinch, see Figure 11. This situation is most beneficial for retrofit.

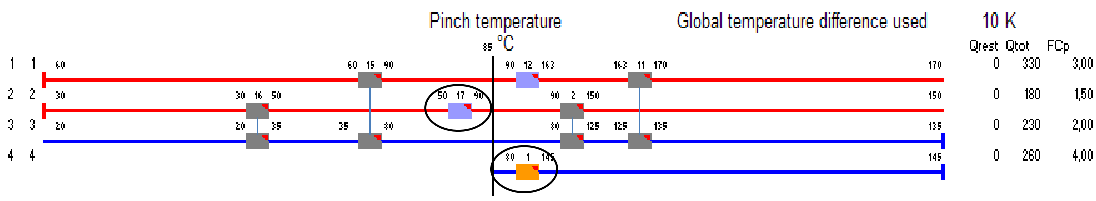

Figure 11. Example of a system which is easy to change with coolers placed high and heaters placed low close to the pinch.

All curves are represented in a temperature versus heat load diagram. The same theory applies both above and below the pinch (Nordman, 2005).

3.2 Assumptions

In this analysis assumed process stream data was used for certain streams in order to perform the calculations. This was done when there was no measured data available or when the span of data was too wide.

3.2.1 Condensing streams

Some of the process streams are condensing. A simulation in Aspen HYSYS of the distillation section at Nynas was available from a previous study. This simulation was used to retrieve data for stream compositions which was then used to establish pinch data for condensing process streams.

Stream 5, see Figure 2, comes from the top of the AD column and contains a lot of steam and some light hydrocarbons that all condense. The stream was simulated in Aspen HYSYS and the condensing curve is shown in Figure 12. The temperature profile contains a clear threshold because the stream mainly consists of steam. The light hydrocarbons condense over a wider temperature range but the amounts are small thus impact on the temperature profile is not significant. The enthalpy distribution in the stream is shown in the temperature profile below.

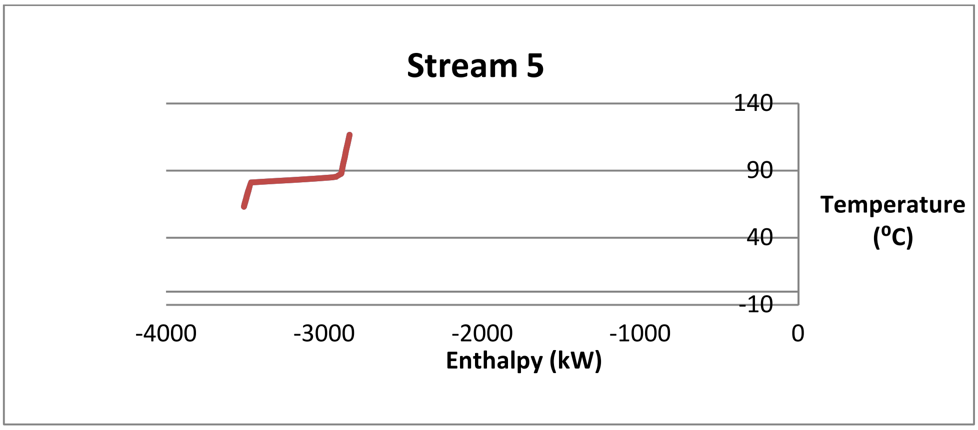

Figure 12. Enthalpy plot for stream 5.

Streams 10 and 11, see Figure 2, are situated by the ejectors and are condensed between the ejectors to extract water. These streams are assumed to behave as pure steam since the composition of the streams only contains negligible amounts of different light hydrocarbons compared to the steam content. The streams also contain a small amount of air which is neglected in the simulation because it is not condensable. Aspen HYSYS is used to generate the condensing curves which show almost the same behaviour as stream 5.

To demonstrate these condensing streams in ProPi they must be divided into different temperature ranges according to the temperature profiles, see Figure 12. The condensing part of the stream is then represented as a single stream and the part before and after condensation are represented as separate streams as well. The separate streams are used because ProPi needs a single value of the specific heat capacity and when dividing it like above the condensing steam can retrieve representative Cp values.

Stream 8, see Figure 2, cannot be handled in the same way as the previously discussed streams. This stream consists of larger amounts of different hydrocarbons which all condense at different temperatures. The enthalpy curve from a simulation is showed in Figure 13 and is an almost linear graph since these various hydrocarbons evaporate at different temperatures. The data in the simulation was uncertain since they were obtained in an already existing HYSYS simulation over the plant and they do not correspond to the measured data. The measured data is taken at the same period of time as the rest of the data in this pinch analysis and are therefore more consistent. The specific heating value of this stream is therefore instead solved by an energy balance over the stream using process data.

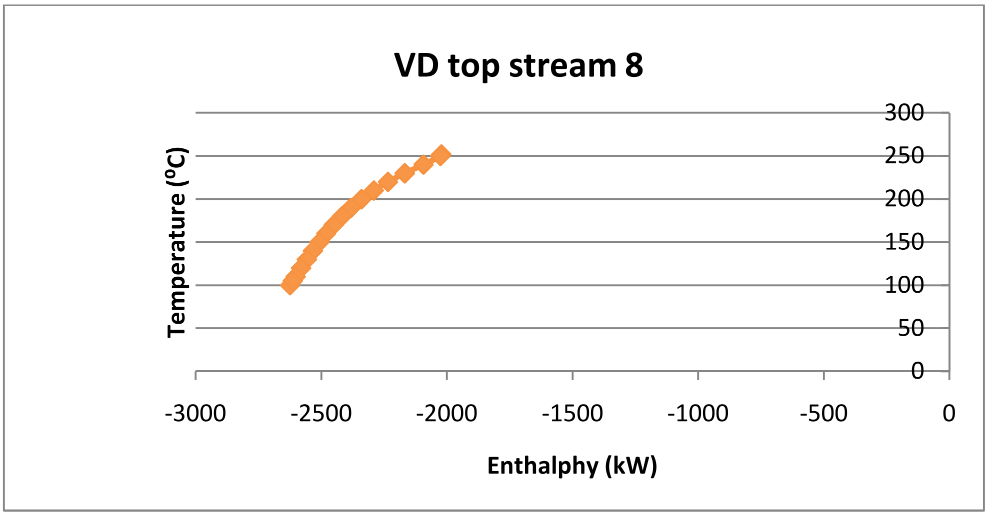

Figure 13. Enthalpy plot for stream 8.

3.2.2 Variations of Cp value as a function of temperature

The crude oil stream (stream 6) is a stream which covers a large span of temperatures, 65-360 °C. Since the existing heat exchanger network includes a number of heat exchangers for heat recovery from this stream there are many temperature measurements. Comparing the calculations in ProPi using a constant Cp value (2.3 kJ/kgK) with the existing measured data of the process they do not correspond. The inaccuracy using only one Cp value started at 185 °C with differing temperatures.

To correct this error the streams was divided in two parts with different Cp values. At 185 °C the Cp value was increased with 2.5 kJ/kgK to match the measured temperatures in the heat exchangers. With these new values of Cp the calculated values in ProPi corresponded to the measured values.

3.2.3 Soft targets

All products that go to the storage cisterns are given “soft” temperature targets in the heat exchanger network analysis. This implies a possibility to recover more heat from these streams than the current level. This is not a process requirement since the products can be discharged to storage within a range of possible temperature levels. If the heat recovery from these streams is increased for internal heat recovery this will result in a lower cold utility demand.

3.3 Economy

The alternatives for retrofitting the heat exchanger network are evaluated with respect to economic performance. The retrofit suggestions imply new equipment. Below follows the calculation methods for estimating the costs of such equipment.

The total investment cost includes the capital cost of heat exchangers as well as installation, engineering and contingency costs. The investment cost is calculated according to the method presented in Sinnot & Towler (2009). The equipment cost C_E is estimated according to Equation (1) where a and b are cost constants, n is the scaling factor specific for the type of equipment, S (m^2) is the capacity of the equipment, see Table 4. The capital costs is calculated from tables valid for previous years and updated to today’s value using Chemical engineering cost index data (Towler, 2009).

$$C_E = a + b \cdot S^n \qquad (1)$$

The capital cost does not concern any costs for installing the equipment and has to be multiplied with an installation factor concerning for example piping and electrical cost, see Table 3. The equation used can be seen as Equation (2).

$$C = \sum_{i=1}^{i=M} C_{e,i,CS}\ [(1 + f_p)f_m + (f_{er} + f_{el} + f_i + f_c + f_s + f_l)] \qquad (2)$$

Table 3. Factors for estimation of project fixed capital cost, (Towler, 2009) .

	Factors for fluids	C_E
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	Structural and buildings	0,2
f_l	Lagging and paint	0,1

Equation 2 does also include a material factor f_m which can be seen in Table 4.

Table 4. Material cost factors relative to plain carbon steel, (Towler, 2009).

Material	f_m
Carbon steel	1
Cast steel	1.1

Additional to this capital costs C_E an engineering cost is calculated as 30 % of the capital cost. The engineering cost is such a large part of the total investment cost due to small amount of equipment in the project compared to the engineering work.

The contingency charge is an extra cost in the project to allow for variations from the initial estimated cost. These variations can be due to changes in project scope, currency fluctuations and price variations. Since heat exchangers are a well known technology the recommended minimum value of 10 % of the capital cost is used in this project (Towler, 2009).

A method of estimating the time it will take for the reduced utility costs (i.e. annual savings) to cover the total investment cost for each retrofit suggestion is the payback period. It is obtained by dividing the total investment cost by the annual savings, see Equation (3).

$$Payback\ period = \frac{Total\ Investment\ cost}{Savings\ per\ year} \qquad (3)$$

Since Nynas has a scheduled annual production stop during which maintenance and reconstruction can be done, they will not lose any revenue because of stop in production due to retrofits, i.e. the shut down period does not need to be accounted for in the cost calculations.

4 Results

This chapter presents the results from the performed pinch analysis. The presented results include composite curves, the GCC as well as advanced curves. The energy targets for the process are established. Thereafter the existing network is analysed and pinch violations are identified. Based on these results, two retrofit proposals are suggested.

4.1 Establishing energy targets for the process

As described above, step one in the analysis is to conduct a basic pinch analysis of the process streams so as to locate the process pinch temperature and establish energy targets for the process (i.e. minimum hot and cold utility requirements). The streams included in pinch analysis of the distillation section in the system are shown in Figure 14. The red streams are hot streams that need to be cooled and the blue streams are cold streams that need to be heated. Most of the streams are hot streams but the flows in these streams are small compared to the largest flow i.e. the crude oil stream that is a cold stream with a major heating demand, see Table 5.

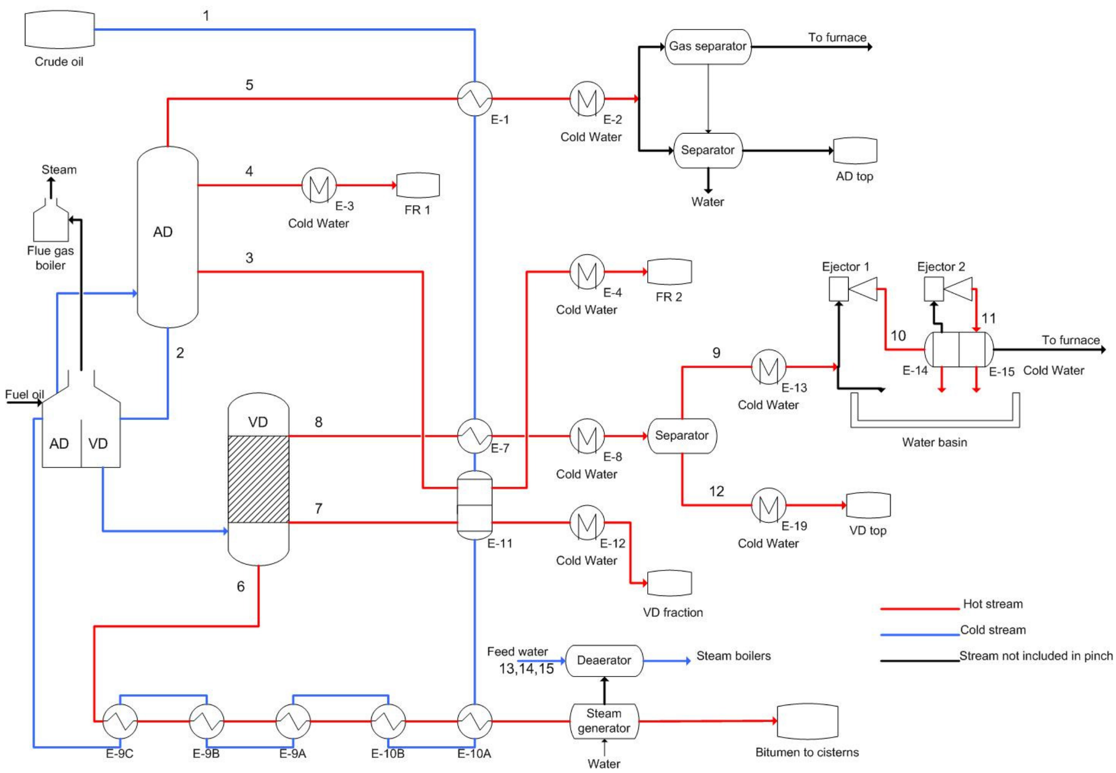

Figure 14. Process streams in the distillation area.

Table 5. Stream data for pinch analysis.

Stream	**Stream content**	**Hot/ Cold**	**T start**	**T target**	**F (kg/s)**	**Cp (kJ/(kg* K))**	**Q (kW)**
1A	Crude oil	Cold	65	185	19,30	2,3	5315
1B	Crude oil	Cold	185	360	19,30	2,5	8543
2	AD residue	Cold	342	378	16,90	2,4	1477
3	FR2-gas oil	Hot	279	30	1,40	2,2	-769
4	FR1-kerosin	Hot	206	30	1,20	2,2	-456
5A	AD-top	Hot	117	86	5,00	2,2	-343
5B	AD-top	Hot	86	81	5,00	109,8	-2769
5C	AD-top	Hot	81	66	5,00	2,4	-180
6	Bitumen	Hot	348	160	14,20	2,6	-6937
7	VDFR	Hot	291	50	1,50	2,2	-792
8	VD-top	Hot	252	105	3,30	5,7	-2764
9	VD-gas	Cold	110	45	0,30	4,2	-83
10A	VD gas+steam	Hot	118	66	0,06	1,9	-6
10B	VD gas+steam	Hot	66	65	0,06	2340,0	-146
10C	VD gas+steam	Hot	65	45	0,06	4,4	-6
11A	VD gas+steam	Hot	161	66	0,01	2,0	-3
11B	VD gas+steam	Hot	66	65	0,01	2340,0	-33
11C	VD gas+steam	Hot	65	45	0,01	4,4	-1
12	VGO	Hot	110	30	1,00	2,1	-166
13	Fresh water	Cold	10	133	1,39	4,2	718
14	Condensate large cistern park	Cold	50	133	0,72	4,2	252
15	Condensate small cistern park	Cold	50	133	0,21	4,2	73

Fuel oil and cold water are used as utilities in the distillation section. The current utility usage for the plant is shown in Table 6. Steam is also used as a utility at the plant but is not covered in this analysis. However, the steam production might be affected by the retrofit suggestions.

Table 6. Present utility demand.

Heating demand [kW]	6456
Cooling demand [kW]	4990

4.1.1 Composite Curves

To establish composite curves ΔT_{min} of the process streams has to be known, this temperature difference can either be set as a global value where all process streams have the same value or as individual ΔT_{min}. In the process streams at Nynas the heat transfer properties differ significantly between the streams, which motivate use of individual ΔT_{min} contributions. Stream 10 and 11 mainly consist of condensing steam and are given a ΔT_{min} of 5 K (Axelsson E. , 2008) The rest of the streams consist of hydrocarbon products and were allocated a ΔT_{min} value of 15 K based on plant experience from Stefan Nyh, process engineer at Nynas (Nyh, 2011).

The pinch is represented in Figure 15 as the location where the vertical distance between the hot and cold composite curves is at its minimum allowable value. With the chosen values of ΔT_{min} the pinch temperature of the system is 237.1 °C for the cold process streams and 252.1 °C for the hot process streams.

It can be seen in the figure that a large part of the system, (i.e. from 3500 to 16 000 kW) show nearly parallel curves and has a temperature difference that is close to the ΔT_{min} which represents the pinch temperature. Since the area around the pinch temperature indicate low temperature driving forces in the heat recovery and this gives a hint that the system will be hard to retrofit.

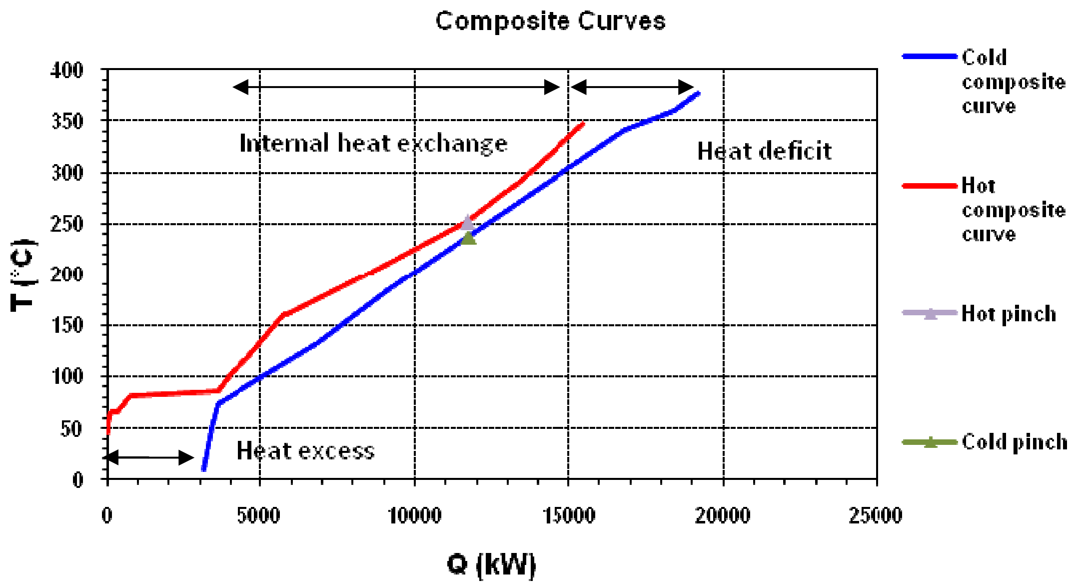

Figure 15. Composite curves for the process base case (based on individual stream contribution for ΔT_{min}).

The CC show that the process has a minimum heat demand of 3.7 MW and a cooling demand of 3.1 MW, see Figure 15. By comparing the minimum (target) utility demands with the utility demands of the existing system it is possible to establish the potential for savings, see Table 7.

Table 7. Potential heating and cooling savings.

	Present Demand [kW]	Minimum Demand [kW]	Potential for savings [kW]	Potential for savings [%]
Heating	6456	3718	**2738**	**42**
Cooling	4990	3145	**1845**	**37**

4.1.2 Demand curves

The demand curves are established to investigate whether any specific global ΔT_{min} has an impact on the pinch analysis. A threshold in the demand curves indicate a drastic change in the minimum utility demand for the process and should be investigated further.

The ΔT_{min} of this analysis is set as individual values. Since only two streams differ in individual ΔT_{min} compared to the rest of the streams, an investigation of the demand curves (which only use global ΔT_{min}) is performed. It is shown from the demand curves in the analysis that an increase or decrease of ΔT_{min} does not have a significant impact of the minimum utility load of the pinch analysis. This is due to the parallel demand curves shown in Figure 16.

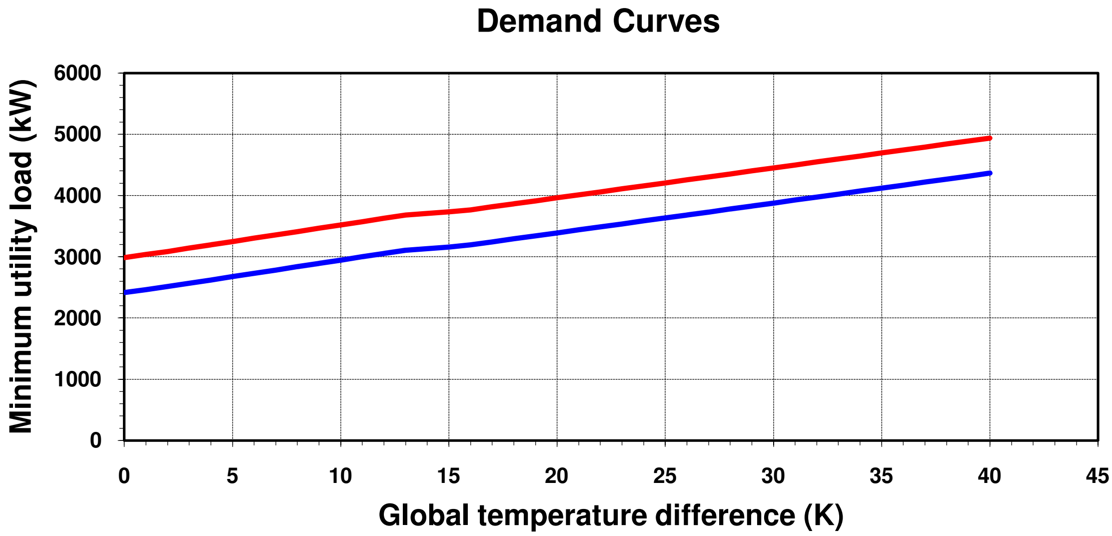

Figure 16. Demand curves for the process (with a global ΔT_{min}).

4.1.3 Grand Composite Curve

The GCC provides further insights about energy flows within the process if it is design for maximum energy recovery, as well allowing a good overview of how to select suitable utility levels for heating and cooling the process, see Figure 17. The pinch temperature of the system is here represented at 246 °C. The figure shows where the process theoretically could be integrated, the so called pockets. The current process has a large heating demand at around 350 °C. It also has a large cooling demand at about 60 °C. In the GCC the pinch of the system is indicated as a peak which coincide with zero heat transfer (i.e. Q=0). However, in this system there is a second peak which is close to be a pinch as well. This implies that there are two areas where it is complicated to integrate the process. An indication of this was already seen in the CC which showed parallel composite curves around the ΔT_{min}.

Having the second peak in the system also makes the system sensitive to changes since the second peak can become the main process pinch if a small change of process pinch stream data is made. It can almost be said that the system has double pinch temperature.

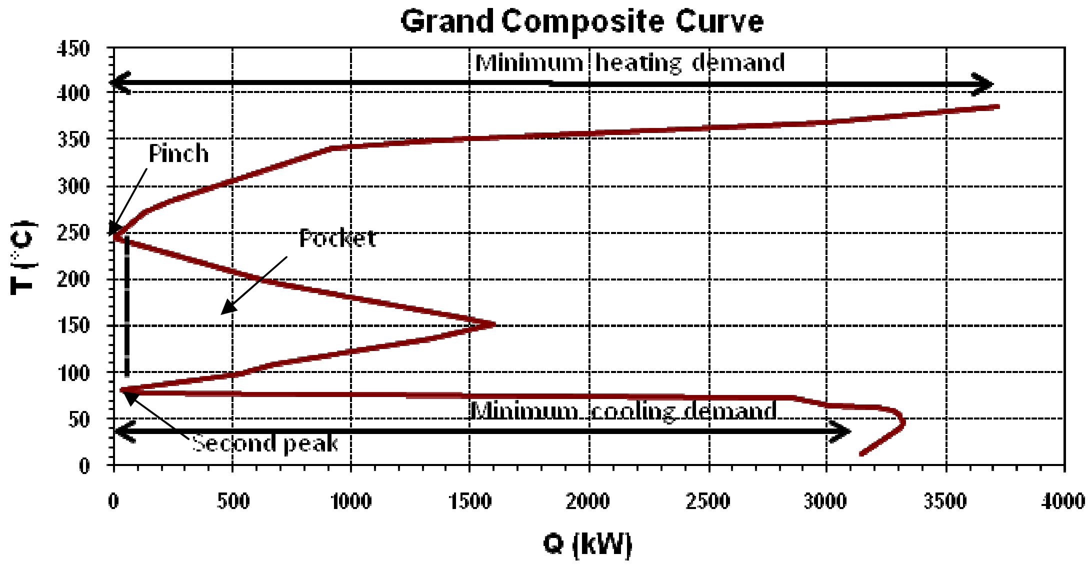

Figure 17. Grand composite curve of the process.

4.1.4 Pinch violations

Pinch violations occur when one of the three golden rules of pinch analysis is violated. In this study the only violated rule is heat transferred through the pinch. The sum of the pinch violations corresponds to 2.7 MW, see Table 8.

Table 8. Pinch violations.

Heat exchanger	**Pinch violation**	**Heat load [kW]**
E-11	Heat through pinch	91
E-11	Heat through pinch	133
E-9B	Heat through pinch	414
E-9C	Heat through pinch	2050
Total		**2687**

The largest violation is in heat exchanger E-9C. The second largest pinch violation is in E-9B, situated next to E-9C in the crude oil heating train.

There is also a small pinch violation in heat exchanger E-11, which transfer heat from both stream 3 and stream 7 through the pinch. The pinch violation is so small that it is probably not economically feasible to correct it. Another reason not to remove this pinch violation is the design of the heat exchanger which is complex using two different hot streams and two tube bundles in one shell, see Figure 18. Because of these difficulties this pinch violation is left unchanged.

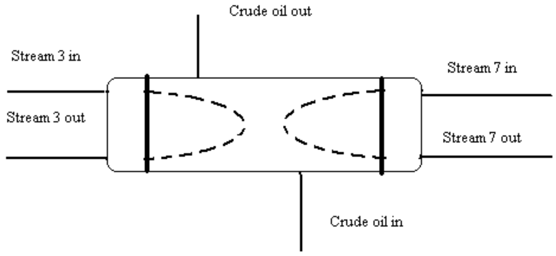

Figure 18. Design of heat exchanger E-11.

The pinch violations give an indication of where in the process one could start to look for retrofit possibilities and are shown in Figure 19.

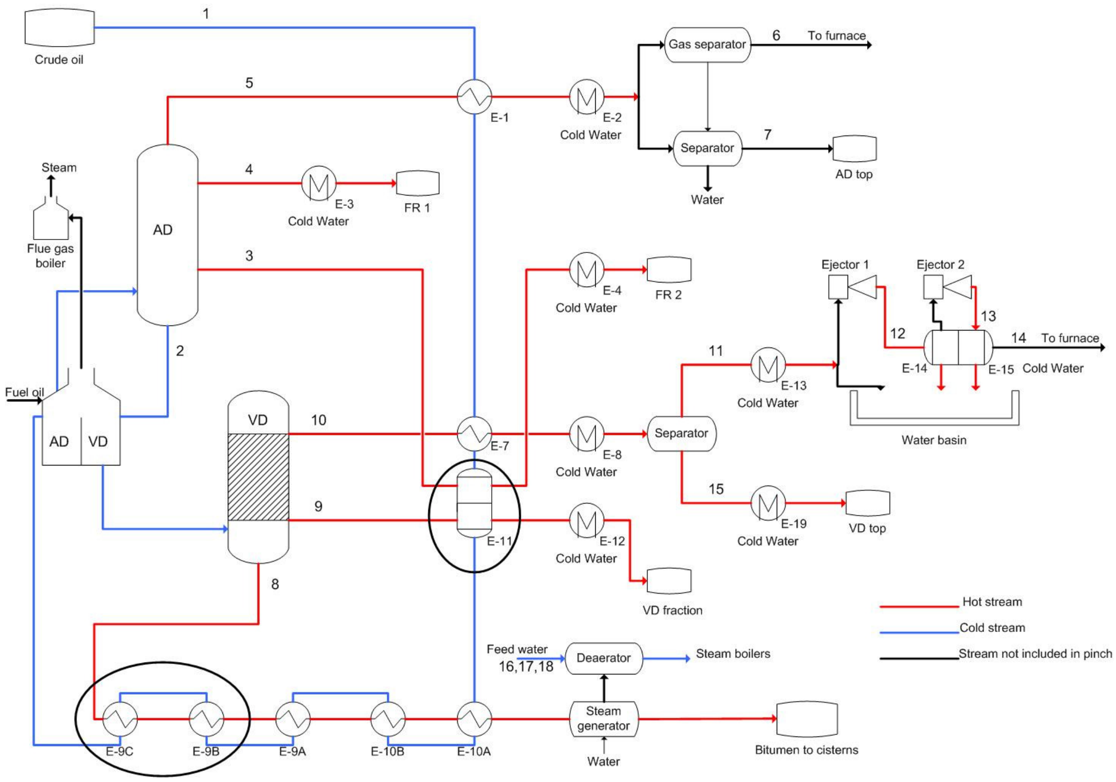

Figure 19. Pinch violations are present within the circled areas.

The heat exchanger in E-11 correspond to 8 % of the total pinch violations and the exchangers E-9C and E-9B in the heat exchanger train correspond to 92 % of the total pinch violations. In the heat exchanger network shown in Figure 20 the pinch violations in E-9B and E-9C are circled.

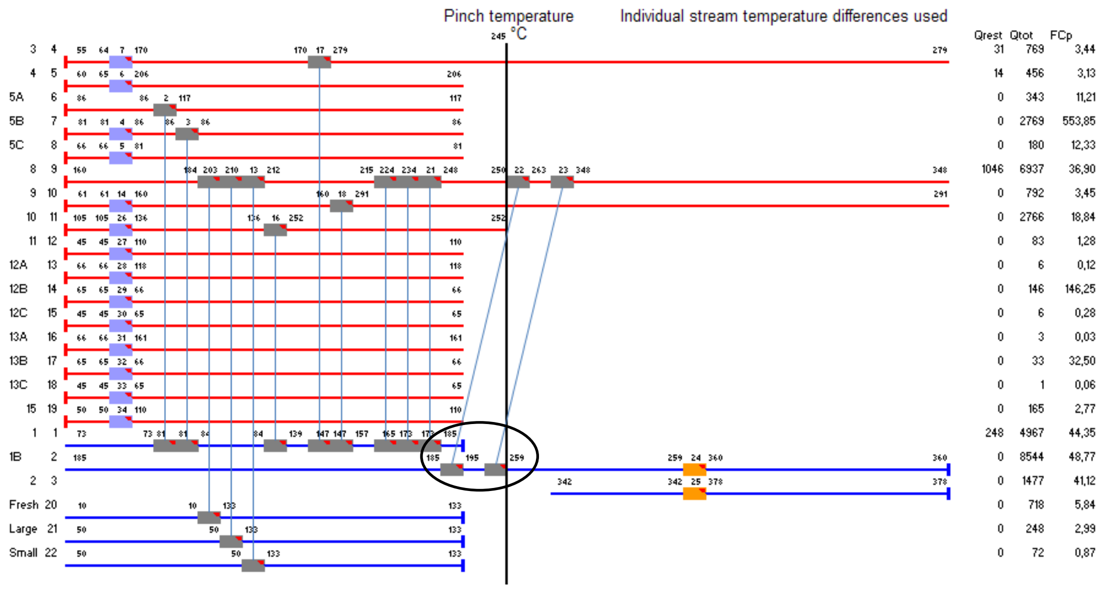

Figure 20. Heat exchanger network. Pinch violation for the heat exchanger train.

It can be seen in Figure 20 that the crude oil stream (stream 1 and 1B) needs additional heating to remove the pinch violations shown in Figure 19. The deficit of heat in the crude oil stream implies a fraction of the heat exchangers heat load is transferred heat through the pinch.

Shown in Figure 21 are the current heat exchangers in the heat exchanger train between bitumen and crude oil. The hot and cold pinch temperatures are acknowledged in the figure to represent were the pinch violations occurs.

When the plant was built 1956 there was only one heat exchanger between the crude oil and bitumen streams. The existing heat exchanger train was built in many steps and the other four heat exchangers have been added during the years.

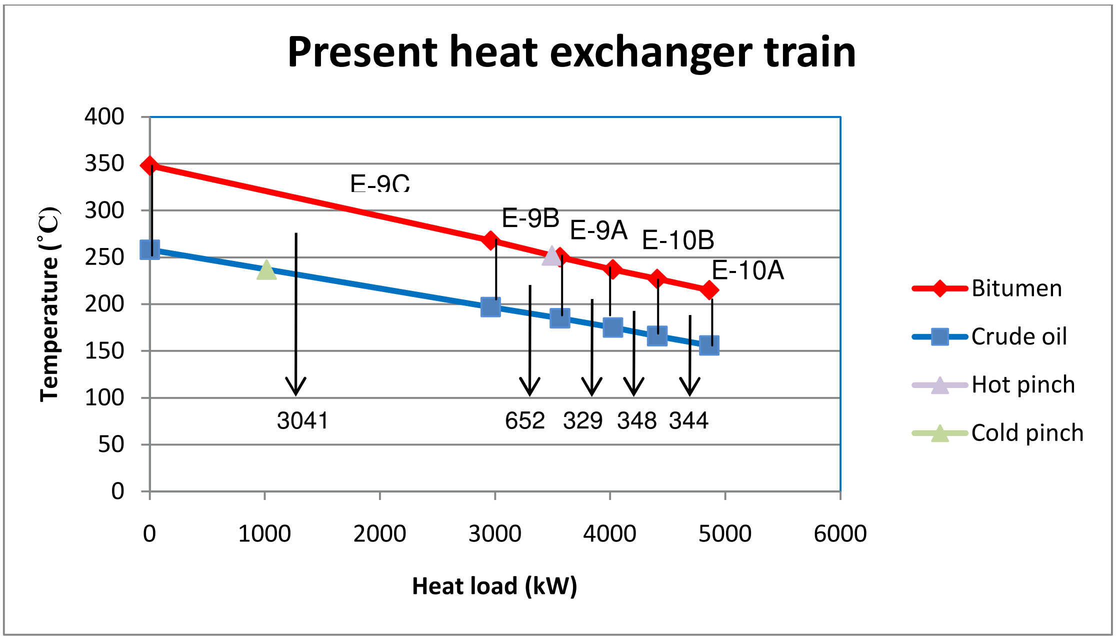

Figure 21. Existing crude oil and bitumen heat exchanger train.

4.1.5 Advanced pinch curves

Advanced pinch curves are useful to provide insight as to show how easily the current heat exchanger network can be retrofitted. The curves for this process are shown in Figure 22.

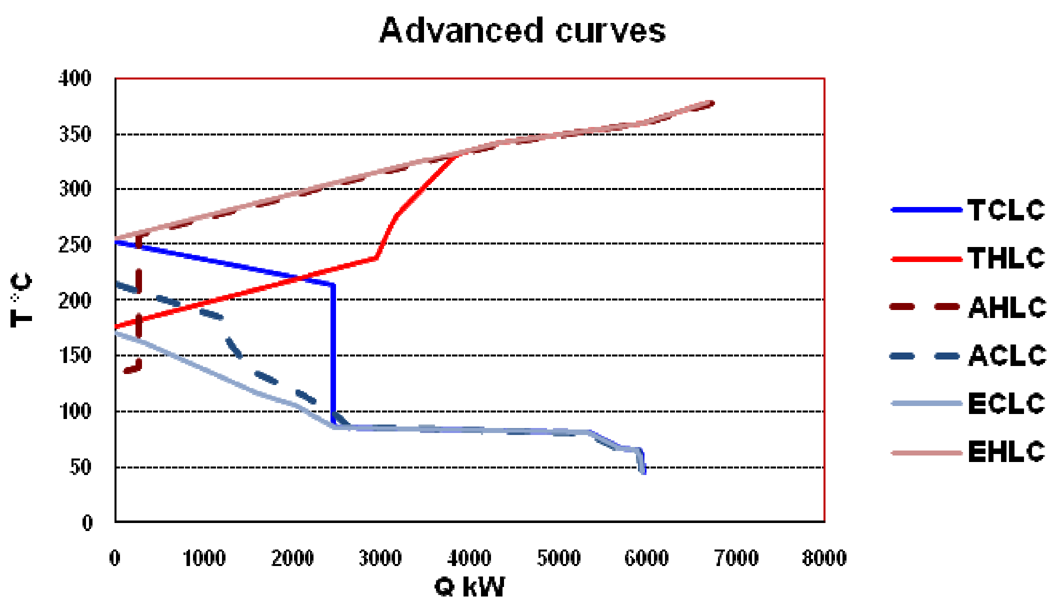

Figure 22. Advanced curves.

In the figure the hot curves are represented with red colours and the cold curves are represented with blue colours. The streams can be hard to distinguish from each other because they more or less coincide at higher heat loads. However, all the hot curves and all the cold curves respectively have the same total heat load.

The advanced pinch curves show that the Actual Heat Load (AHLC) lies close to the Extreme Heat Load Curve (EHLC), see Figure 22. This implies that the existing network design is difficult to change without incurring large costs. The Actual Cooling Load Curve (ACLC) is on the other hand located in between the Extreme Cooling Load Curve (ECLC) and Theoretical Cooling Load Curve (TCLC), which indicate that retrofits could be made with less difficulty. This implies that the limiting factor to achieve good retrofits is within the hot streams. The results from the advanced curves corresponds well to the results from the pinch analysis where it has been seen that the pinch violations are located close to the pinch temperature and thus make it complicated to perform a retrofit. In the first pinch analysis there can also be seen that the coolers are placed at low temperature levels and the heaters are placed at high temperature levels which additionally confirm the results from the advanced curves, i.e. that the process is complicated to retrofit.

4.1.6 MER network

In order to provide guidelines for retrofitting the process heat exchanger network, it is of interest to design a Maximum Energy Recovery (MER) network of the process.

At Nynas there is only one cold stream in the network that can be used for heat exchanging in a large extent, namely the crude oil stream. The other cold streams are at too high temperatures or have small flows and therefore do not contribute so much, see Table 5. To match the whole crude oil stream with a number of hot streams it requires a large amount of stream splitting to be done.

Below the pinch the cold stream has to be split four times and above the pinch three times, see Figure 23. This is done to keep the driving forces and to match the hot and cold process streams to utilize as much heat in each stream as possible.

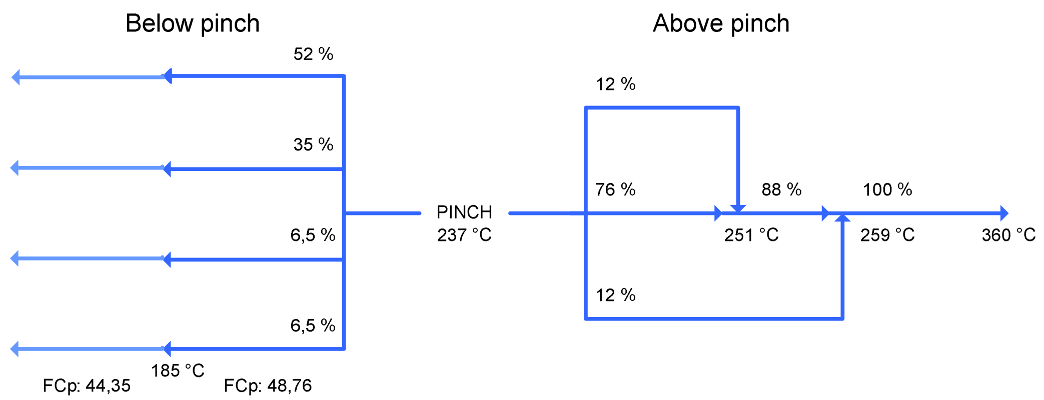

Figure 23. Stream splitting illustration.

The negative aspects with stream splitting are that it makes the process difficult to control. Constant flows are desirable if implementing stream splitting which is not the case in this plant. Hence a completely new network would be very complex and not economically feasible to build, which is the reason why the MER network is not a proposal for Nynas. The MER network shows further that to retrofit in this process is complicated.

4.2 Retrofit suggestions

Below two retrofit suggestions are presented. They are similar since they are both dealing with pinch violations situated in the same place in the process, but there are some different ways of solving these pinch violations. The primary target for the two retrofits is the largest pinch violations that occur in the heat exchange between crude oil and bitumen (stream 1 and 6). The two streams have an existing train of heat exchangers and it is in the end of this train that the largest violation occurs. The existing heat exchangers are at the moment in the same material and are all tube and shell heat exchangers.

The increased heat exchange will result in a warmer crude oil stream going into the furnace. This increase in the ingoing temperature to the furnace results in reduced fuel consumption which also decreases the CO_2 emissions by about 7000 tonnes/year. However, the reduced heat load in the furnace will reduce the amount of steam produced in the flue gas boiler. The increased heat exchange will also influence the steam generator situated on one side of the heat exchanger train as it will decrease the amount of steam produced.

The reason for heating the crude oil with bitumen instead of any other hot streams is that this stream has the largest FCp and a heat load at a reasonable temperature. The bitumen stream can however not be utilized without making large changes in the existing heat exchanger train. The existing heat exchangers have different designs and it would be complicated and unprofitable to add the extra area to the existing heat exchangers. Instead the existing areas of the current heat exchanger train are kept constant and an addition of new heat exchangers is instead investigated. To be able to keep the areas of the already existing heat exchangers constant while calculating the additional areas a simulation in HYSYS was performed.

The simulations in HYSYS are done by modeling the heat exchanger train and include the new exchangers. Information about the existing exchangers is put into the program where they are modeled by steady state rating, see Appendix 10.2. Then the areas are held constant and with the given information about the existing exchangers for example the Ft factor (temperature correction factor for heat exchanger design) is calculated in HYSYS. The additional heat exchangers are modeled with data input only on temperatures and pressure drops. The whole heat exchanger train is connected and the ingoing and outgoing temperatures of bitumen, at each end of the train, and the ingoing temperature of the crude oil are set. The outgoing temperature of the crude oil is calculated.

To calculate the heat exchanger areas an overall heat transfer coefficient (U-value) has to be used. These U values were obtained by performing a simulation over just the existing heat exchanger train. To this simulation all the temperatures between the exchangers was adjusted to correspond to the temperatures between the heat exchangers obtained in the pinch analysis. Then with the information of the heat exchanger design, the fluids and the temperatures a U value for each exchanger was obtained as shown in Table 9.

Table 9. U values for the existing heat exchangers.

Heat exchanger	E-9C	E-9B	E-9A	E-10B	E-10A
U value [kW/m^{2}K]	0,22	0,13	0,09	0,04	0,03

There can be seen in Table 9 that the U values differs even for heat exchangers with the same design i.e. E-9B and E-9A, see Appendix 2. This implies that the overall heat transfer coefficient changes with the temperature of the fluids. The U values for E-9B with higher temperatures are larger than for E-9A with lower temperatures.

The U values are therefore plotted in a temperature vs. U values diagram to show these variations, see Figure 24. This figure is the basis for the U values in further simulations and for extracting U values for the additional heat exchanger to obtain their area.

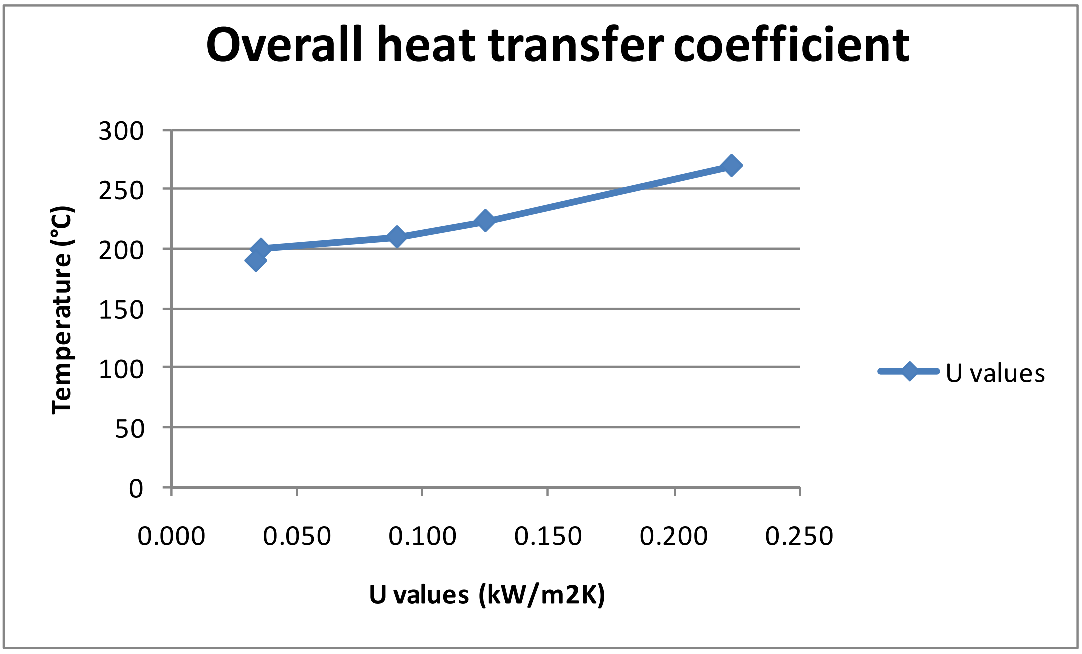

Figure 24. Variation of U values with temperature.

This simulation as a whole results in an estimate on the new heat exchanger areas that is the basis for further economical calculations.

The pressure drops for the heat exchanger train was extracted from piping, instrumental and design diagrams of the plant. However, the information was unspecific and only pressure drops in each ends of the streams in the heat exchanger train could be obtained. Therefore the pressure drop within each heat exchanger in the simulation was assumed equal to 1.5 bar.

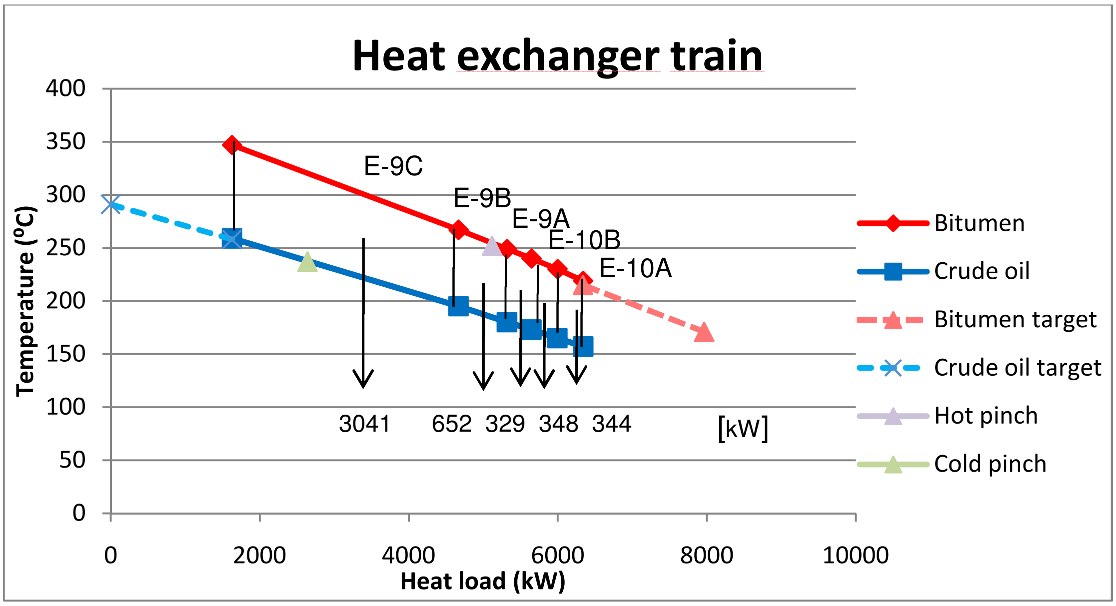

Figure 25. Existing heat exchanger train with targets of increased heat exchanging.

In Figure 25 the existing heat exchangers before any retrofit proposals are shown with dashed lines representing the streams capacities of increased heat recovery. The existing heat exchangers together have a duty of 4.8 MW and there is possible to increase this with 1.6 MW to a new level of 6.4 MW if the bitumen stream is used to its target temperatures to heat the crude oil. The possibility to increase the heat load more than 1.6 MW is restricted due to the temperature difference on the cold side of the streams that then is at ΔT_{min}. The aim for the retrofit suggestions is to reach this level of 6.4 MW of heat recovery.

The available space at the plant is an important factor when analyzing a retrofit suggestion. The heat exchangers at Nynas are placed on a four level construction. The space here is very limited so placing any new equipment on the existing platform cannot be done. The new heat exchangers both have to be placed about 100 meters away on the ground.

4.2.1 Feedwater

To be able to add new heat exchangers at the heat exchanger train the existing steam generator would have to be heavily reduced. This results in a too low production of 2 barg steam which currently is used for preheating the feedwater to the steam boilers.

Heating the feedwater with steam is not the most efficient way in an energy perspective and here is room for improvements. A suggestion is preheating the feedwater using the excess heat below the pinch. This would result in a release of heat in the bitumen stream which could be used for further heat exchanging instead.

There are several hot streams in the system that have a heat load that is currently cooled away and instead could be used for heat exchanging. To select streams the first criterion is that they should heat the feedwater enough to compensate for the decrease of steam heating. When this criterion is fulfilled other aspects can be considered, among them distance, temperature interval and heating the feedwater more than in the base case.

4.3 Retrofit suggestion 1

The first way of introducing an additional heat exchanger to the heat exchanger train is to place a new exchanger at the hot end of the crude oil stream before the AD furnace, see Figure 26. This is the first alternative when studying the results from the pinch analysis. There is also shown two new heat exchangers on streams 3 and 4 for fresh water heating.

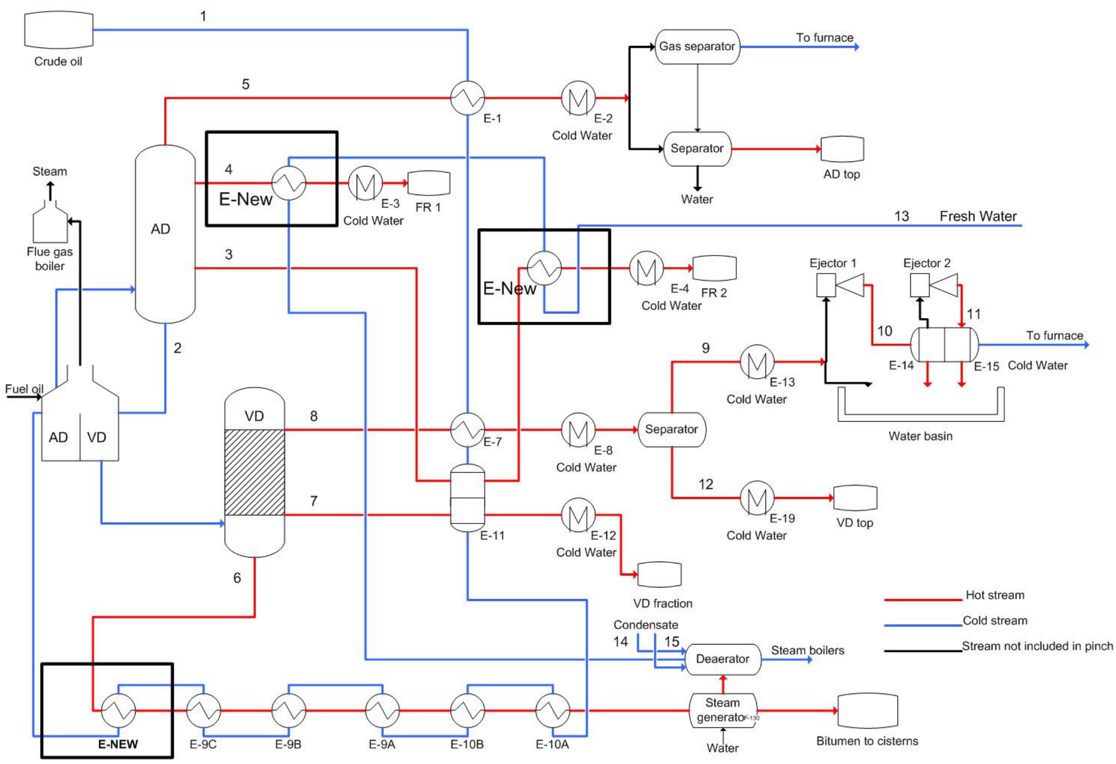

Figure 26. Retrofit 1 with new heat exchangers.

The targets of increased heat exchanging are reached in this retrofit alternative. The addition of the new heat exchanger to the heat exchanger train is shown in Figure 27.

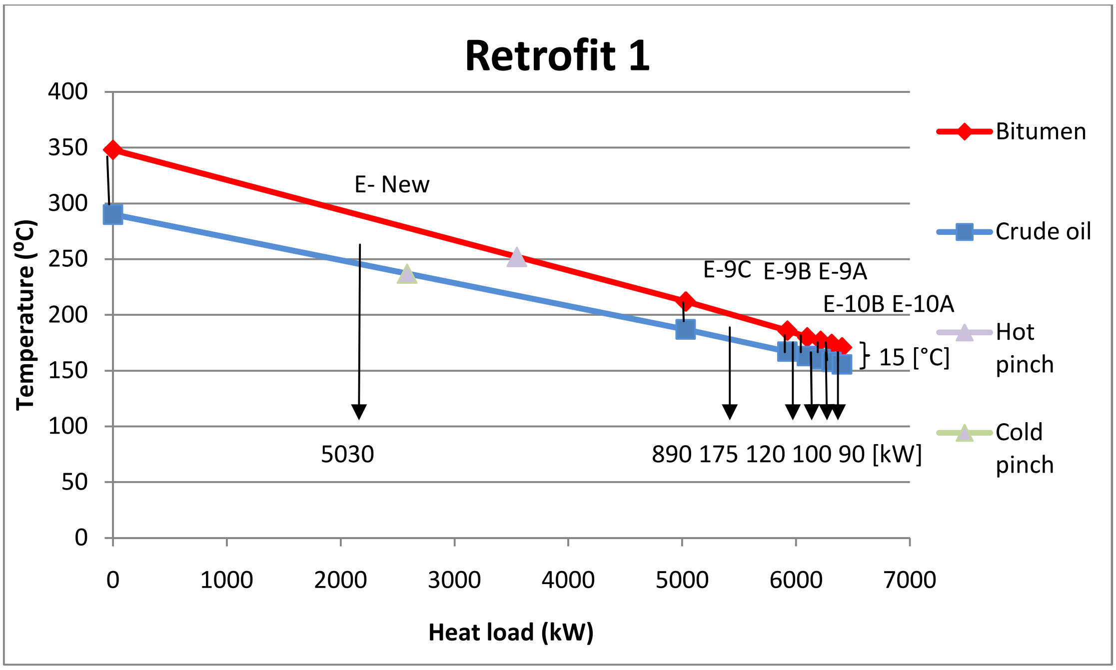

Figure 27. Heat exchanger train after addition of new heat exchanger.

The total heat load of the train has increased with 1.6 MW which implies that maximum amount of heat is utilized in the bitumen stream as shown in Figure 27. This is seen as the ΔT_{min} is used in the cold end of the streams. In Figure 27 the additional heat exchanger is situated to the left at the higher temperature interval. The additional heat exchanger has a large heat load compared to the existing heat exchanger train. The existing heat exchangers have reduced heat loads compared to the base case because the existing heat exchangers are being shifted into operating points with lower temperature, resulting in reduced driving forces for the heat transfer.

All the additional areas of the new heat exchangers are accounted for in Table 10.

Table 10. New heat exchanger areas for retrofit 1.

New heat exchangers	Stream 3	Stream 4	Bitumen/Crude oil	Total
Area [m^2]	10	80	700	790

Retrofit 1 increase the temperature of the crude oil into the furnace to 290.3 ºC (compared to 259 ºC) and thereby reduced the demand of fuel in the furnace by 24.3 %, see Table 11. This corresponds to a 58 % reduction of heat transfer through the pinch.

Table 11. Results for retrofit 1.

	New heating demand [kW]	Decreased heating demand [kW]	Decreased heating demand [%]	New cooling demand [kW]	Decreased cooling demand [kW]	Decreased cooling demand [%]	Pinch violation decrease [%]
Retrofit 1	4 486	1 570	24.3	4 490	449	10	58

Furthermore, in this retrofit the new heat exchanger in the heat exchanger train will be installed in durable material due to the high temperatures in the installation area. This implies a release of the high temperature load on the existing heat exchangers at the high temperatures but also an increased material cost.

4.3.1 Reduced steam generator

The steam generator is reduced in this retrofit suggestion due to further use of the bitumen stream 6 for heat exchanging. The steam generators reduced heat load is only enough to heat the condensate from both cistern parks and a small part of the fresh water stream. The steam generator also has capacity to heat the fresh water in the higher temperature interval from 117 °C to 133 °C. In need of preheating is the fresh water stream in the temperature range 10 °C to 117 °C.

The suggestion in this study is to use stream 3 in combination with stream 4 to heat the fresh water. These streams have large enough heat load at the right temperature range. Furthermore after the heat exchange the streams are discharged to the cistern storage area and therefore have soft targets.

Figure 26 shows that the heat exchange first occurs with stream 3 and then with stream 4. This order is preferable since stream 4 then can be ticked off entirely with this design and does not need cooling at all. But even though the stream can be heat exchanged all the way the existing cooler will still be left in its place to be able to regulate the process. This is done as a safety measure in case of changed process conditions or a shift of operating mode. Stream 4 still needs to be cooled when the new heat exchanger is installed due to its large heat load. The heat load of the new heat exchangers will be 0.5 MW for stream 4 and 0.2 MW for stream 3.

This design will save cooling water corresponding to the amount of cooling that is reduced in stream 3 and 4. This is about 0.5 MW and is a little less than the additional heat exchanging here. This is because the soft target is used on stream 4 which does not reduce the cooling demand.

4.4 Retrofit suggestion 2

This retrofit suggestion is similar to retrofit 1 since it is dealing with the same pinch violations. The difference is that the heat exchanger is added at the cold-end of the heat exchanger train, closest to the steam generator, see Figure 28. The reason to place the new heat exchanger closest to the steam generator is that it can be built in ordinary carbon steel since the temperatures here are the lowest in the train.

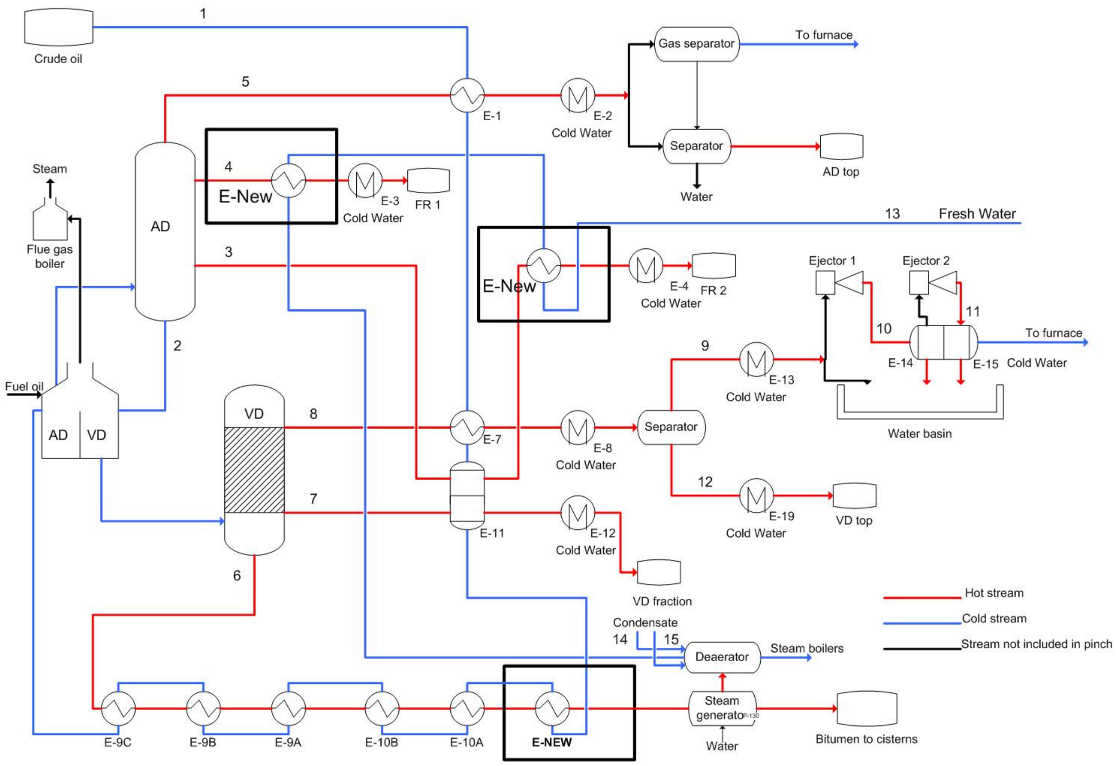

Figure 28. Retrofit 2 with one new heat exchanger closest to the steam generator.

This suggestion also changes the heat loads and temperatures of the existing heat exchangers in the train as shown in Figure 29.

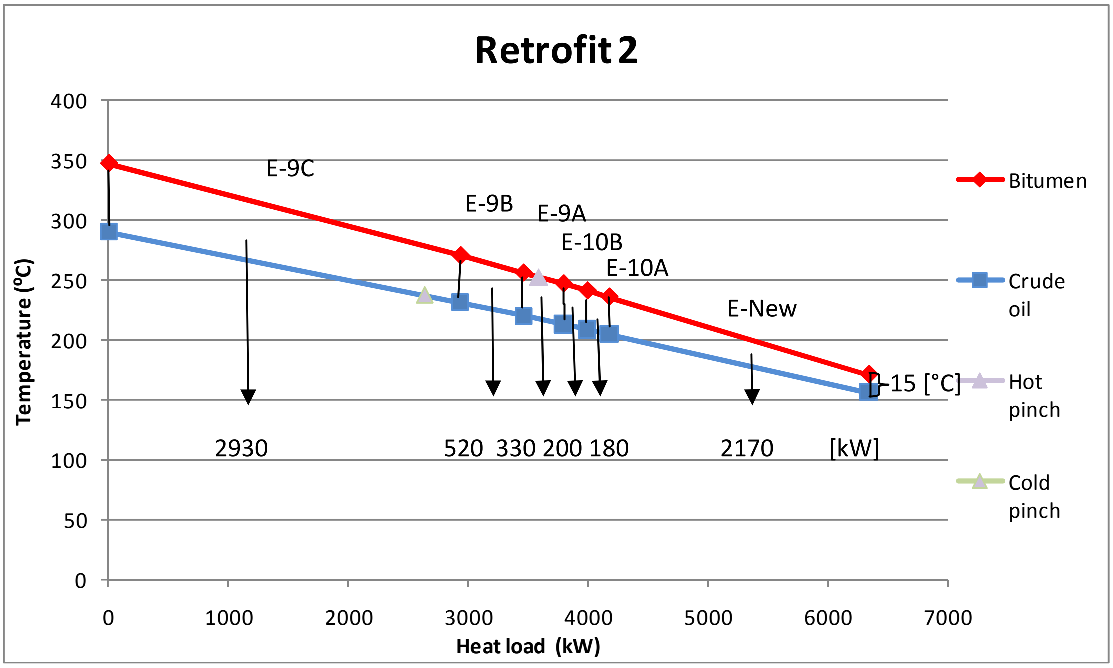

Figure 29. Heat exchanger train after addition of new heat exchanger according to retrofit 2.

The heat loads of the existing heat exchanger train in this retrofit are close to what they currently are. This is due to that a smaller heat load is achieved in the new heat exchanger due to the small driving forces at this end of the train because of low temperatures that affect the heat transfer of the fluid.

The new heat exchanger between the crude oil and the bitumen stream would have an area of 1770 m^2 as shown in Table 11. The area of the new heat exchanger in the train is larger than the corresponding area in retrofit suggestion 1. This is due to that the existing heat exchangers are operating at a higher temperature level which means that the bitumen and the crude oil have enhanced heat transfer. However the temperatures of the streams are lower close to the steam generator compared to the base case and it will demand more area per extra heat load due to lower driving forces.

This suggestion would increase the temperature in the existing heat exchangers close to the furnace. When adding a new heat exchanger the crude oil temperature in the existing exchangers increase which can lead to problems with example wearing on the heat exchanger material. The design temperature of that heat exchanger (E-9C) is 300 °C, see Appendix 10.2, which indicate that the heat exchangers should be sufficient since the target temperatures are just below this temperature. But at Nynas they have seen some damage to the exchangers and it is still a concern regarding this suggestion. If there is a change in production (i.e. due to a change in production mode) the temperatures may shift and be just above the design temperature instead.

The part of this retrofit that concerns the steam generator and the preheating of the fresh water are using the same streams and areas as in retrofit suggestion 1.

The areas of the new heat exchangers are shown in Table 11.

Table 12. New heat exchanger areas for retrofit 2.

New heat exchangers	Stream 3	Stream 4	Bitumen/Crude oil	Total
Area [m^2]	10	80	1770	1860

This retrofit would result in a decreased heating demand of 1.6 MW in the furnace, see Table 12. This is the same as in retrofit 1 with a pinch violation decrease at 58 % and a reduced fuel demand of 24.3 % in the furnace.

Table 13. Results for retrofit 2.

	New heating demand [kW]	Decreased heating demand [kW]	Decreased heating demand [%]	New cooling demand [kW]	Decreased cooling demand [kW]	Decreased cooling demand [%]	Pinch violation decrease [%]
Retrofit 2	4 486	1 570	24.3	4 490	449	10	58

5 Economic evaluation

To identify if the retrofit suggestions are economically feasible a comparison of the profit due to reduced fuel consumption in the furnace and the cost of implementing new heat exchangers is made. The first step is to calculate the fuel savings achieved within the retrofit suggestions.

5.1 Utility savings

When increasing the temperature of the crude oil stream the heat load in the furnace is reduced. This means a reduced amount of fuel oil is needed to heat the crude oil in the furnace. For calculations see Appendix 10.1.

Table 14. Parameters for the fuel and the furnace.

Price EO5 [SEK/m^3]	8 882
Density EO5 [kg/m^3]	930
Heating value EO5 LS [MJ/kg]	41
Load factor [Days/year]	270
Efficiency in furnace	0,765

A comparison of the fuel demand is made between the present situation and the two retrofit suggestions see Table 14. The calculations concern all parameters in Table 13 where the price for the fuel oil includes taxes. These taxes are the sales taxes and Swedish CO_2 tax. Nynas is not included in the EU CO_2 emissions trading system. It can be seen that the fuel cost savings are the same for the two retrofit suggestions.

The cooling utility reduction is neglected both due to the low reduction in cooling demand and due to the relatively low cost of cold utility compared to the cost of fuel oil in the furnace.

Table 15. Fuel cost savings in the furnace.

	Present situation	Retrofit 1,2
Heat load [kW]	6 456	4 886
Fuel cost [SEK/yr]	45 900 000	34 700 000
Fuel cost savings [SEK/yr]		**11 200 000**

5.2 Investment cost

To estimate the cost of the new investments a capital cost of the new heat exchangers must be calculated. The two retrofit suggestions include both a new heat exchanger between the crude oil and bitumen stream and two new heat exchangers for preheating of feedwater. Table 16 shows that the first retrofit suggestion is more expensive than the second retrofit shown in Table 15. This is mostly due to a larger area of the heat exchangers but also because of the more expensive material in the heat exchanger closest to the furnace.

The capital cost is calculated with an over design of 15 % extra area compared to the results from the retrofit suggestions. This is done due to the fact that there are variations in the heat load demanded in the heat exchangers due to fouling.

The heat exchangers are approximated with the material cost of carbon steel. However, the heat exchanger closest to the furnace exchanging bitumen and crude oil in retrofit 1 is specified to cast steel, a more rigid but also more costly material compared to carbon steel. This is due to the high temperatures closest to the furnace. This imply a slightly changed total installation factor in the different heat exchangers depending on the material factor, see Table 4, which is 3.2 for heat exchangers made in carbon steel and 3.38 for heat exchangers made in cast steel.

The heat exchange between the fresh water and stream 3 and stream 4 is considered to have some safety risks. With direct heat exchange there is a risk of organic compounds contaminating the water. This is why the heat exchange maybe has to be done with an intermediate heat exchanger with a carrier fluid. This implies a larger cost than the costs shown in Table 15 and Table 16.

Table 16. Capital cost for heat exchangers in retrofit 1.

Retrofit 1		
Heat exchanger	Crude oil/bitumen	Preheating
Area [m2]	700	90
C_E 2007 [$]	140 000	60 000
C_{Etot} 2011 [$]	170 000	70 000
Cost incl. over design [$]	190 000	80 000
Total HEX cost [SEK]	1 200 000	480 000
Installation factor fm=1 [SEK]		1 550 000
Installation factor fm=1.1 [SEK]	4 050 000	
Total investment cost [SEK]	**5 600 000**	

Table 17. Capital cost for heat exchangers in retrofit 2.

Retrofit 2		
Heat exchanger	Crude oil/bitumen	Preheating
Area [m2]	1 770	90
C_E 2007 [$]	380 000	60 000
C_{Etot} 2011 [$]	450 000	70 000
Cost incl. over design [$]	520 000	80 000
Total HEX cost [SEK]	3 300 000	480 000
Installation factor fm=1 [SEK]	10 430 000	1 550 000
Total investment cost [SEK]	**12 000 000**	

The engineering cost and contingency charge further increase the total investment cost, see Table 17. These costs concern the pre study and planning of the project and are calculated as a percentage of the total capital cost. The total investment cost of retrofit 2 is larger than of retrofit 1 due to larger area needed.

Table 18. Total investment cost for retrofit 1 and retrofit 2.

	Retrofit 1	*Retrofit 2*
Engineering cost [SEK]	1 700 000	3 600 000
Contingency cost [SEK]	560 000	1 200 000
Total investment costs [SEK]	**7 800 000**	**16 800 000**

5.3 Payback period

The payback period is calculated by comparing the total investment cost and the fuel savings per year and is in this case about 9 months for retrofit 1 and 18 months for retrofit 2. Retrofit suggestion 1 is to be considered as a short time period in this context, see Table 18. The calculations are made for a running period at the plant of 270 days per year. This is at the lower limit of time and it implies that the payback period could be even shorter. It can also be noted that the total investment cost does not include cost for extra piping due to large distance from current heat exchangers, new platform or any pumps needed which would increase the payback period.

Table 19. Payback period.

	Retrofit 1	*Retrofit 2*
Payback period	**0.7**	**1.5**

6 Discussion

The discussion mainly concerns the two retrofit suggestions but also different alternatives for using the excess heat at the plant as well as a short analysis of different technical and economical uncertainties.

6.1 Retrofit suggestions

It was revealed that Nynas has a system that is complicated to retrofit and suggestions for further heat exchange have therefore been limited to one place in the process. This is because the process within the system boundaries only has one useful cold stream which reduces the possibilities for heat recovery. However, two retrofit suggestions were proposed for the crude feed preheater train and preheating of feedwater.

In the retrofit suggestions the new heat exchangers, between the bitumen and crude oil streams, would be placed far away from the rest of the heat exchanger train. This implies that a more thorough study of pressure drops and piping is necessary. In this report an assumed value of pressure drop inside the new heat exchangers has been used for calculations.

In retrofit 1 the new heat exchanger on the heat exchanger train would be built in a hard-wearing material compared to the other heat exchangers. In this way the new heat exchanger would sustain the higher temperatures in the bitumen and crude oil and the older heat exchangers would be spared the stress caused by the high temperatures. It is also a cheaper suggestion to place the heat exchanger closest to the furnace due to less area needed. The amount of pinch violations removed in this retrofit suggestion is 58 % of the base case violations.

Retrofit 2 implies a larger heat exchanger as an addition to the heat exchanger train, built in carbon steel material. A negative aspect is that it is no room for protecting the existing heat exchangers with a new exchanger of better material. Further, this proposal is more expensive than suggestion 1. The pinch violations are decreased with 58 % which is the same as for retrofit 1.

The retrofit suggestions are accompanied with a decrease of heat available from the steam generator. In this study an alternative to heat exchange this feedwater with stream 3 and 4 is suggested. This is however not the only option available. If one considers all hot streams that have an inflexible temperature target for this type of heat exchange there are supplementary suggestions to be made.

An aspect from Nynas is that improved insulation of the pipelines of the condensate from the cistern parks could be enough to reduce the need of heating to these streams. Thereby the reduced 2 barg steam would be sufficient to only preheat the fresh water and the new heat exchangers for preheating suggested would not be necessary.

The 16 barg steam production has to be considered since it changes if any of the retrofit proposals will be implemented. All the retrofits that are suggested result in a decreased load in the furnace. This will affect the flue gas boiler which generates steam to the plant since it will decrease parallel to the furnace load. The amount of steam from the flue gas boiler will be reduced with 1 tonne/h compared to the current production of 4 tonne/h. The flue gas boiler is one of the main suppliers of 16 barg steam and a decrease of this unit could affect the process. The thought here is thus that this steam could be replaced with superheated 2 bars steam that could be produced from excess heat in the system, see Chapter 6.2 Excess heat. The demand of steam could also be reduced because Nynas already has plans to use a hot oil boiler system to heat some of the storage cisterns due to safety measures instead of using steam. There have also been discussions of heating the cisterns with electric tracing (Keereweer, 2011).

In the future the emission restrictions will probably be harder and Nynas AB may have to join ETS which would increase the cost of using fossil fuel further. This promotes a retrofit of the system. The fuel Nynas uses emits about 3.5 tonnes CO_2/tonnes fuel which would result in an additional cost of 2.1 MSEK/year if no changes are made. All retrofit suggestions would reduce this cost to about 24 %. For calculations, see Appendix 10.1.

6.2 Excess heat

After the retrofit there is a lot of excess heat that is still cooled away in the process. The excess heat in the plant has been discussed in the terms of district heating. One advantage concerning district heating is that Nynas refinery is situated close to Rya-verket which produces district heating and could probably easy be connected to the network. On the other hand there is large problem with this because the largest need for district heating coincides with the time period during which the plant has its yearly stop. Therefore there has been no further investigation in this direction even though the temperatures at the plant are sufficient for this type of heat use (109 ^{o}C). In the future Nynas has plans for increased production which implies a shorter stop period. If this becomes reality a new investigation concerning district heating possibilities could be suitable.

Another way of solving the excess heat issue is to produce more 2 barg steam. Independently of the retrofit alternatives it has been observed that this type of steam could be produced in additional places at the plant. This could make a base for increased production of 2 barg steam to replace the medium pressure steam (8 barg) in the distillation. It would lead to reduced production of 16 barg steam that currently is let down to 8 barg steam. The problem today is that the 2 barg steam has too high moisture content since it is not superheated. But if the moisture content of the 2 barg steam could be superheated the steam could be used in the distillation columns. This could also be a solution to part of the problem with less steam produced in the flue gas boiler with a reduced heating demand as in the retrofit alternatives.

Nynas has been considering heating the cisterns with hot oil instead of steam. This is both because of safety and energy efficiency reasons. If the proposal of cisterns heated with hot oil will be realized the amount of steam needed will be heavily reduced. Then the problem with decreased steam production in the flue gas boiler could be avoided. The plan is to heat oil in a boiler and lead it to the cisterns and later return it to the boiler for reheating. This idea would lead to a large decrease of 16 barg steam since the main part of this steam level is used to heat the cisterns (65 %). A result of this could be that steam boiler 1 is unnecessary and the production of steam could be focused on the flue gas boiler alone in cooperation with steam boiler 3. This would lead to lower fuel costs of the steam boiler but has to be investigated further.

6.3 Uncertainties

The uncertainties in the analysis are mostly concerning the measured data at the plant. The data used for the calculations is average values of the summer month for one of the plants operating modes.

A lot of parameters have been given specific values even if they in reality should vary in time or with the streams. This includes the specific heating value and the fouling. The heat transfer coefficients used are representative for a range of products approximated to all organic compounds which are a simplification. Neither have there been any additional losses represented in the HYSYS simulations except the fouling in the heat exchangers.

The proposal from Nynas of changing fuel from fuel oil to natural gas is an important parameter that has to be considered. Since there is no timeframe of when to change the fuel the economic evaluation does not include this. Another parameter to consider economically is the planned process capacity increase which implies that the payback time for this studies retrofit proposals has to be low.

7 Conclusions

This master thesis has shown that the theoretical minimum heating demand is 3.7 MW and the minimum cooling demand is 3.1 MW at the distillation area at Nynas refinery. This can be compared to the current utility demand of 6.5 MW (hot utility) and 6.0 MW (cold utility) which means that 42 % of the heating demand and 37 % of the cooling demand could be saved. However to achieve this savings a lot of stream splitting has to be done and many heat exchangers would have to be bought which would not be profitable. It would also make the process hard to control and sensitive to changes.

Instead two retrofit suggestions which have different benefits concerning areas, material and economic performances have been proposed. Both the suggestions reduce the heating demand in the furnace by about 24 %. The proposal that is considered most beneficial within the limits of this study is retrofit suggestion 1. In retrofit suggestion 1 there is increased heat exchange in the crude oil preheater train due to added area at the hot end of the train. The steam generator at the bitumen stream is partly replaced by heat exchange with waste heat recovery at the plant. The new heat exchanger in the heat exchanger train is also constructed in stronger material reliving the stress on the existing heat exchangers. This suggestion has the shortest payback period of 9 months compared to 15 months for retrofit 2.

Further the CO_2 emissions will be decreased by 24 % due to the decreased fuel demand with both suggestions. This corresponds to a reduction of CO_2 emissions of 7000 tonnes/year. The short payback time, the reduction of fuel in the furnace and reduced carbon dioxide emissions motivate a change of the system.

8 Further work

There is naturally work left to do in the field of energy efficiency improvements at the refinery. Firstly there is further work to do concerning the retrofit suggestions given in the report. There could be more detailed calculations of new process design and costs, this is to actually be able to choose one of the retrofit alternatives combined with more preheating of the feed boiler water. The possibilities to implement the suggestions at the plant can thus be further investigated.

The pinch analysis made in this study is concentrated to the process streams and the distillation area. A larger pinch analysis including utility streams and the cistern parks would also be of interest. The result of such analysis could confirm the results here or add suggestions of different energy efficiency measures. This would be especially interesting concerning the utility systems. There is a lot of heat excess in the return of the cooling water streams and there is also room for improvements of the steam system. The question raised in this study about preheating feedwater is something that might have even larger potential with a more thorough investigation of the utility systems.

Further work in making the plant more energy efficient could also include reviewing the pipelines at the plant and their insulation.

Nynas is planning to increase the production during the next ten years. The measures of this will be implemented stepwise and as they are planned one might incorporate the retrofit suggestions into these plans. Time perspective of doing a retrofit now compared to later in the increased production process concerning the value is not considered in this report. Hence, the results of this study should either be implemented soon or be the basis of a completely new heat exchanger network.

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10 Appendix

Appendix 1 - Economical evaluation

Appendix 2 - Heat exchanger designs

10.1 Appendix 1 - Economic calculations

Fuel cost savings

Fuel cost savings in Furnace	
Price EO5 [SEK/m3]	8 882
Density EO5 [kg/m3]	930
Number of days per year (min)	270
Heat load in present case [kW]	6 456
Heat load in retrofit 1 [kW]	4 886
Heat load in retrofit 2 [kW]	4 886
Heating value EO5 LS [MJ/kg]	41
Efficiency in furnace	0,765
Fuel Cost in present case [SEK/yr]	45 850 000
Fuel Cost in retrofit 1 [SEK/yr]	34 710 000
Fuel Cost in retrofit 2 [SEK/yr]	34 710 000
Fuel cost savings retrofit 1 [SEK/yr]	11 200 000
Fuel cost savings retrofit 2 [SEK/yr]	11 200 000

Parameters for calculating capital cost of heat exchangers. (Towler, 2009)

Investment cost parameters for equipment	
a	24 000
b	46
n	1,2

Additional CO_2 cost if Nynas will be included in ETS

CO_2 cost	
CO_2 cost [euro/ton CO_2]	18
Exchange SEK/Euro	9
Emission EO5 [tonne CO_2/tonne fuel]	3,52
CO_2 emission in base case [ton/day]	48
CO_2 emission in retrofit 2 [ton/day]	36
CO_2 emission in retrofit 3 [ton/day]	36
CO_2 cost base case [SEK/yr]	2 100 000
CO_2 cost retrofit 1 [SEK/yr]	1 570 000
CO_2 cost retrofit 2 [SEK/yr]	1 570 000
CO_2 cost savings retrofit 1 [SEK/yr]	520 000
CO_2 cost savings retrofit 2 [SEK/yr]	520 000

10.2 Appendix 2 - Heat exchanger designs

	E-9C	**E-9B**	**E-9A**	**E-10B**	**E-10A**
Surface Area	247,5	104,7	104,7	159,5	159,5
Shell Side	Crude oil	Bitumen	Bitumen	Bitumen	Bitumen
Material	CS	CS	CS	CS	CS
Shell ID (mm)	1022	830	854	838	838
Material thickness (mm)	14	12	12	12	12
No of passes	1	1	1	1	1
Design pressure (barg)	16	16	16	12	12
Design temperature (˚C)	300	370	370	300	300
Tube Side	Bitumen	Crude oil	Crude oil	Crude oil	Crude oil
Tube number	848	136	136	610	610
Tube OD (mm)	19,05	38	38	19,05	19,05
Wall Thickness (mm)	2,1	2,6	2,6	2,6	2,6
Effective length (mm)	4877	4880	4880	4703	4871
No of passes	8	6	6	2	2
Design pressure (barg)	16	16	16	16	16
Design temperature (˚C)	385	240	240	300	300

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